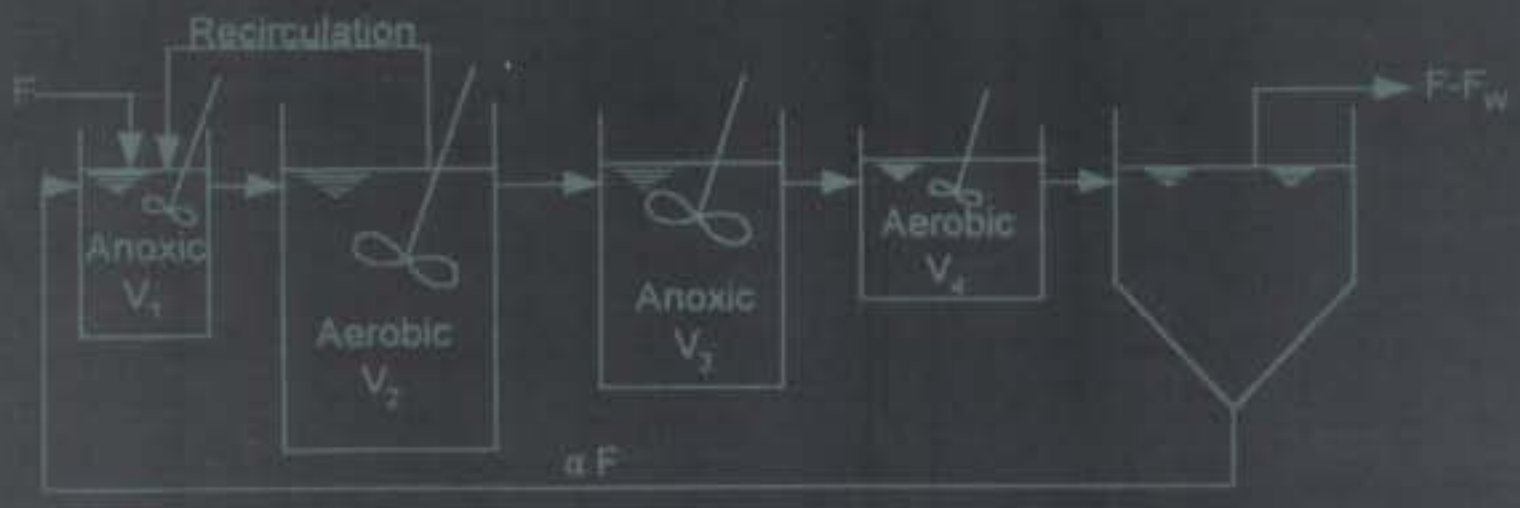
Biological Wastewater Treatment

Second Edition, Revised and Expanded



C. P. Leslie Grady, Jr.
Glen T. Daigger
Henry C. Lim

H2O

Biological Wastewater Treatment

Environmental Science and Pollution Control Series

1. Toxic Metal Chemistry in Marine Environments, *Muhammad Sadiq*
2. Handbook of Polymer Degradation, *edited by S. Halim Hamid, Mohamed B. Amin, and Ali G. Maadhah*
3. Unit Processes in Drinking Water Treatment, *Willy J. Masschelein*
4. Groundwater Contamination and Analysis at Hazardous Waste Sites, *edited by Suzanne Lesage and Richard E. Jackson*
5. Plastics Waste Management: Disposal, Recycling, and Reuse, *edited by Nabil Mustafa*
6. Hazardous Waste Site Soil Remediation: Theory and Application of Innovative Technologies, *edited by David J. Wilson and Ann N. Clarke*
7. Process Engineering for Pollution Control and Waste Minimization, *edited by Donald L. Wise and Debra J. Trantolo*
8. Remediation of Hazardous Waste Contaminated Soils, *edited by Donald L. Wise and Debra J. Trantolo*
9. Water Contamination and Health: Integration of Exposure Assessment, Toxicology, and Risk Assessment, *edited by Rhoda G. M. Wang*
10. Pollution Control in Fertilizer Production, *edited by Charles A. Hodge and Neculai N. Popovici*
11. Groundwater Contamination and Control, *edited by Uri Zoller*
12. Toxic Properties of Pesticides, *Nicholas P. Cheremisinoff and John A. King*
13. Combustion and Incineration Processes: Applications in Environmental Engineering, Second Edition, Revised and Expanded, *Walter R. Niessen*
14. Hazardous Chemicals in the Polymer Industry, *Nicholas P. Cheremisinoff*
15. Handbook of Highly Toxic Materials Handling and Management, *edited by Stanley S. Grossel and Daniel A. Crow*
16. Separation Processes in Wast Minimization, *Robert B. Long*
17. Handbook of Pollution and Hazardous Materials Compliance: A Sourcebook for Environmental Managers, *Nicholas P. Cheremisinoff and Nadelyn Graffia*
18. Biosolids Treatment and Management, *Mark J. Girovich*
19. Biological Wastewater Treatment: Second Edition, Revised and Expanded, *C. P. Leslie Grady, Jr., Glen T. Daigger, and Henry C. Lim*

Additional Volumes in Preparation

To the student in each of us.
It is our sincere hope that by learning together,
we can further the cause of
water quality and environmental protection.

This page intentionally left blank

BIOLOGICAL WASTEWATER TREATMENT

SECOND EDITION, REVISED AND EXPANDED

C. P. LESLIE GRADY, JR.
Clemson University
Clemson, South Carolina

GLEN T. DAIGGER
CH2M Hill
Englewood, Colorado

HENRY C. LIM
University of California, Irvine
Irvine, California

MARCEL DEKKER, INC. NEW YORK • BASEL

This book has been prepared based on information presented in the technical and professional literature and the knowledge and experience of the authors. The authors' intention is to present, to the best of their ability, their profession's current understanding of the design and operation of biological wastewater treatment processes. The reader must recognize, however, that both the authors' understanding of the current state of the art and the profession's understanding of the principles on which the processes operate are unavoidably incomplete. This book was prepared primarily for instructional purposes, and it is the knowledge and experience of the designer and operator that determine its success, not the use of any particular design or operational procedure. Thus, while the information presented in this book may serve to supplement the expertise of a competent practitioner, it is not a replacement. It is the user's responsibility to independently verify and interpret information found in this book prior to its application. Consequently, use of the information presented in this book does hereby release the authors, the publisher, and the authors' employers from liability for any loss or injuries of any nature that may result from use of the information presented.

ISBN: 0-8247-8919-9

The first edition was published as *Biological Wastewater Treatment: Theory and Applications*, by C. P. Leslie Grady, Jr., and Henry C. Lim (Marcel Dekker, 1980).

This book is printed on acid-free paper.

Headquarters
Marcel Dekker, Inc.
270 Madison Avenue, New York, NY 10016
tel: 212-696-9000; fax: 212-685-4540

Eastern Hemisphere Distribution
Marcel Dekker AG
Hutgasse 4, Postfach 812, CH-4001 Basel, Switzerland
tel: 44-61-8482; fax: 44-61-261-8896

World Wide Web
http://www.dekker.com

The publisher offers discounts on this book when ordered in bulk quantities. For more information, write to Special Sales/Professional Marketing at the address above.

Current printing (last digit):
10 9 8 7 6 5 4 3 2

PRINTED IN THE UNITED STATES OF AMERICA

Preface

The components in wastewater treatment processes may be conveniently categorized as physical, chemical and biochemical unit operations. A thorough understanding of the principles governing their behavior is a prerequisite for successful process design. This "unit operations approach" to the study of process engineering has been widely accepted in the field of environmental engineering, just as in chemical engineering where it was developed, and a number of books are now available presenting the principles of physical and chemical operations and illustrating their applications. The purpose of this book is to provide similar coverage for the biochemical unit operations, replacing *Biological Wastewater Treatment: Theory and Applications*, which was published in 1980.

Much has happened in the field of biological wastewater treatment since 1980. A particularly significant occurrence was recognition of the many events that can happen simultaneously in biological processes and the role that the designer has in determining which predominate. The impact on process design of this recognition demanded that new environmental engineers be trained from the outset to think in terms of multiple events, rather than compartmentalizing them as they had in the past. In addition, because of the efforts of task groups within the International Association on Water Quality (IAWQ), process simulation has been widely adopted as a tool by design engineers, allowing them to visualize how those events will interact in a proposed design. There have been other significant changes in practice as well, with the development or maturation of several new processes. Consequently, to incorporate all of these changes, it was necessary to completely rewrite—not just revise, as is customary in second editions.

Another change that has occurred since the appearance of the first edition is the background and preparation of environmental engineering students. Most have much better preparation in microbiology and reactor engineering than their predecessors. Consequently, the two sections dealing with these topics have been eliminated. However, background information essential to an understanding of the text is presented in the three chapters of Part I. Elimination of the two background sections provided space for expansion of later sections to include many of the new developments in biological wastewater treatment that have evolved since 1980.

The book is organized into six parts: Part I, Introduction and Background; Part II, Theory: Modeling of Ideal Suspended Growth Reactors; Part III, Applications: Suspended Growth Reactors; Part IV, Theory: Modeling of Ideal Attached Growth Reactors; Part V, Applications: Attached Growth Reactors; and Part VI, Future Challenges.

Part I seeks to do three things. First, it describes the various "named biochemical operations" in terms of their treatment objectives, biochemical environment, and reactor configuration. This helps to remove some of the confusion caused by the somewhat peculiar names given to some biochemical operations early in their history. Second, it introduces the format and notation that will be used to present the models describing the biochemical operations. Finally, it presents the basic stoichiometry and kinetics of the various microbial reactions that form the key for quantitative description of biochemical operations. In Part II, the stoichiometry and kinetics are used in mass balance equations to investigate the theoretical performance of biological reactors containing microorganisms growing suspended in the wastewater as it moves through the system. Part II is at the heart of the book because it provides the reader with a fundamental understanding of why suspended growth reactors behave as they do. In Part III, the theory is applied to the various named suspended growth biochemical operations introduced in Part I. In that application, however, care is taken to point out when practical constraints must be applied to ensure that the system will function properly in the real world. In this way, the reader obtains a rational basis for the design of biological wastewater treatment operations that incorporates knowledge that has been obtained through practice. In other words, we have sought to make Part III as practical as possible. Parts IV and V parallel Parts II and III in organization, but focus on biochemical operations in which the microorganisms grow attached to solid surfaces. This mode of growth adds complexity to the analysis, even though the operations are sometimes simpler in application. Finally, Part VI looks briefly at the use of biochemical operations to remove xenobiotic organic chemicals from wastewaters. The intention is to introduce this topic so the reader can continue learning with the rest of us as we seek to solve the world's environmental problems.

Our plan in preparing this new edition was to provide a text for use in a graduate-level environmental engineering course of three semester-hours credit. In reality, the amount of information provided is more than can be covered comfortably. This provides some latitude for the instructor, but also makes the book a resource for the student interested in knowing more than the minimum. It is our hope, furthermore, that our professional colleagues will find the book to be worthwhile as a reference and as a resource for self-guided study.

At this point, we would like to add a note of caution to the students using this book. It relies heavily upon modeling to provide a conceptual picture of how biochemical operations function. Although these models are based on our best current ideas, one must always remember that they are simply just someone's way of trying to describe in simple terms very complex phenomena. Their purpose is to help the reader learn to think about the processes described by providing "experience." One should not fall into the trap, however, of substituting the models and their simulated experience for reality. Engineering requires the application of judgment in situations lacking sufficient information. The reader can use the background provided by this book to help gain sound judgment, but should not hesitate to discard concepts when real-world experience indicates that they are incorrect or don't apply. Theories are constantly evolving, so be prepared to change your ideas as our knowledge advances.

As with any book, many people have had a hand in its preparation, either directly or indirectly. First and foremost, we would like to thank Henry C. Lim, coauthor of the first edition. Although his career has followed another path, his

thoughts and ideas continue to permeate the work. For example, much of the material on modeling of attached growth systems, written by him for the first edition, has been retained in this edition. Second, CPLG owes a great deal to M. Henze of the Technical University of Denmark, W. Gujer of the Swiss Federal Institute for Environmental Science and Technology, G. v. R. Marais of the University of Cape Town (now retired), and T. Matsuo of the University of Tokyo for all that he learned through long discussions and arguments about the modeling of suspended growth biological reactors as he studied with them on an IAWQ task group. Third, this book would not have been possible had it not been for the hundreds of researchers who generated the knowledge upon which it is based. In the first edition we asked the forbearance of those we overlooked or neglected. This time, we must ask for that tenfold because the information explosion since 1980 has been phenomenal! It simply is not possible for two individuals to have read every article of merit. Fourth, we thank the thousands of practitioners (both designers and operators) who have had the foresight and faith to use biological processes to treat such a wide variety of wastewaters. Their observations and factual documentation of the performance and operational characteristics of these processes have provided both a sound basis for process design and operation and the development of new process options. It is the combination of thoughtful and creative research and application that has provided the factual basis for this book. Finally, we would like to express our appreciation to the many people directly involved in the preparation of this book. The students in ESE 804 at Clemson University, who suffered through early drafts of the manuscript, did yeoman's service in helping us write more clearly. Professorial colleagues at other universities provided very constructive criticism. In particular we would like to thank Nancy G. Love at Virginia Tech, Lutgarde Raskin at the University of Illinois, Barth F. Smets at the University of Connecticut, Timothy G. Ellis at Iowa State University, Robert M. Cowan at Rutgers University, and Wen K. Shieh at the University of Pennsylvania. And last, but far from least, we would like to recognize the hard work of three wonderful people who provided valued assistance: Patsy A. Phillips for her word-processing skills, Rebecca E. Laura for her artwork, and Joni K. Grady for preparing the index.

C. P. Leslie Grady, Jr.
Glen T. Daigger

This page intentionally left blank

Contents

This page intentionally left blank

Part I
Introduction and Background

As with any subject, the study of the biochemical operations used in wastewater treatment systems requires an understanding of the terminology used. The purpose of Chapter 1 is to provide that understanding by defining the nature of biochemical operations in terms of the biochemical transformation being performed, the environment in which the transformation is occurring, and the reactor configuration employed. Engineering design is greatly facilitated by application of mathematical models to quantitatively describe system performance. Construction of such models for biochemical operations must be based on a fundamental understanding of the microbiological events occurring in them. Chapter 2 provides that understanding, as well as an appreciation of the complex interactions occurring among the microorganisms that form the ecosystems in the operations. That appreciation is crucial to recognition of the simplified nature of the models, thereby encouraging their appropriate usage. Finally, construction of the models requires knowledge of the stoichiometry and kinetics of the major reactions occurring in biochemical operations. Chapter 3 provides that knowledge.

This page intentionally left blank

1

Classification of Biochemical Operations

The purpose of wastewater treatment is to remove pollutants that can harm the aquatic environment if they are discharged into it. Because of the deleterious effects of low dissolved oxygen (DO) concentrations on aquatic life, wastewater treatment engineers historically focused on the removal of pollutants that would deplete the DO in receiving waters. These so-called oxygen-demanding materials exert their effects by serving as a food source for aquatic microorganisms, which use oxygen in their metabolism and are capable of surviving at lower DO levels than higher life forms. Most oxygen-demanding pollutants are organic compounds, but ammonia nitrogen is an important inorganic one. Thus, early wastewater treatment systems were designed to remove organic matter and sometimes to oxidize ammonia nitrogen to nitrate nitrogen, and this is still the goal of many systems being built today. As industrialization and population growth continued, another problem was recognized, eutrophication, which is the accelerated aging of lakes and estuaries, etc., due to excessive plant and algal growth. This is the result of the discharge of nutrients such as nitrogen and phosphorus. Hence, engineers became concerned with the design of wastewater treatment systems that could remove these pollutants in an efficient and cost effective manner, and much research during the past two decades has focused on processes for doing that. Most recently, we have become concerned about the discharge of toxic organic chemicals to the environment. Many of them are organic, and thus the processes used to remove oxygen-demanding materials are effective against them as well. Consequently, much current research is directed toward a better understanding of the fate and effects of toxic organic chemicals in those processes.

In addition to the categories listed above, pollutants in wastewaters may be characterized in a number of ways. For example, they may be classified by their physical characteristics (e.g., soluble or insoluble), by their chemical characteristics (e.g., organic or inorganic), by their susceptibility to alteration by microorganisms (e.g., biodegradable or nonbiodegradable), by their origin (e.g., biogenic or anthropogenic), by their effects (e.g., toxic or nontoxic), etc. Obviously, these are not exclusive classifications, but overlap. Thus, we may have soluble, biodegradable organic material; insoluble, biodegradable organic material; and so on. The job of the wastewater treatment engineer is to design a process train that will remove all of them in an efficient and economical manner. This requires a sound understanding of process engineering, which must be built on a thorough knowledge of unit operations. Unit operations, which are the components that are linked together to form a process train, are commonly divided on the basis of the fundamental mechanisms

acting within them, i.e., physical, chemical, and biochemical. Physical operations are those, such as sedimentation, that are governed by the laws of physics. Chemical operations, as the name suggests, are those in which strictly chemical reactions occur, such as precipitation. Biochemical operations, on the other hand, are those that use living microorganisms to destroy or transform pollutants through enzymatically catalyzed chemical reactions. In this book we will examine the role of biochemical operations in wastewater treatment process trains and develop the methods for their design.

1.1 THE ROLE OF BIOCHEMICAL OPERATIONS

The most effective way to define the role of biochemical operations in wastewater treatment systems is to examine a typical process flow diagram, as shown in Figure 1.1. Four categories of pollutants are traced through the process, with the widths of the arrows indicating their mass flow rates. They are soluble organic matter (SOM), insoluble organic matter (IOM), soluble inorganic matter (SIM), and insoluble inorganic matter (IIM). For the most part, the transformation rates of insoluble inorganic matter by microorganisms are too low to be of practical importance. Thus, insoluble inorganic matter is typically removed by preliminary physical unit operations and taken elsewhere for treatment and disposal. Wastewaters occur in large volume, but the pollutants are relatively dilute. Thus, engineers attempt to remove pollutants in the most efficient way, concentrating them where possible to reduce the volumes that must be handled. For insoluble constituents this can be accomplished by the physical operation of sedimentation, which is why it is often one of the first unit operations in a treatment system. The effluent from a sedimentation basin (overflow) contains all of the soluble constituents in the influent, plus those insoluble ones that were too small to be removed. The bulk of the insoluble material, however, exits from the bottom of the vessel (underflow) as a thick suspension called a sludge. Both the overflow and the underflow require further treatment, and that is where biochemical operations come into play.

Most unit operations used for the destruction or transformation of soluble pollutants in the overflow are biochemical ones. This is because biochemical operations function more efficiently than chemical and physical ones when the concentrations of reacting constituents are low. In biochemical operations, the soluble pollutants are converted either into an innocuous form, such as carbon dioxide or nitrogen gas, or into new microbial biomass, which can be removed by a physical operation because it is particulate. In addition, as the microorganisms grow, they entrap insoluble organic matter that escaped removal upstream, thereby allowing it to be removed from the wastewater by the physical operation as well. Consequently, the effluent from the physical operation is relatively clean and can often be discharged with little or no additional treatment. A portion of the insoluble materials removed by the physical operation may be returned to the upstream biochemical operation while the remainder is transferred to another portion of the process train for further treatment.

The other major use of biochemical operations is in the treatment of sludges, as shown in Figure 1.1. Primary sludges are those resulting from sedimentation of the wastewater prior to application of any biochemical operations. Secondary sludges are those produced by biomass growth in the biochemical operations and by entrap-

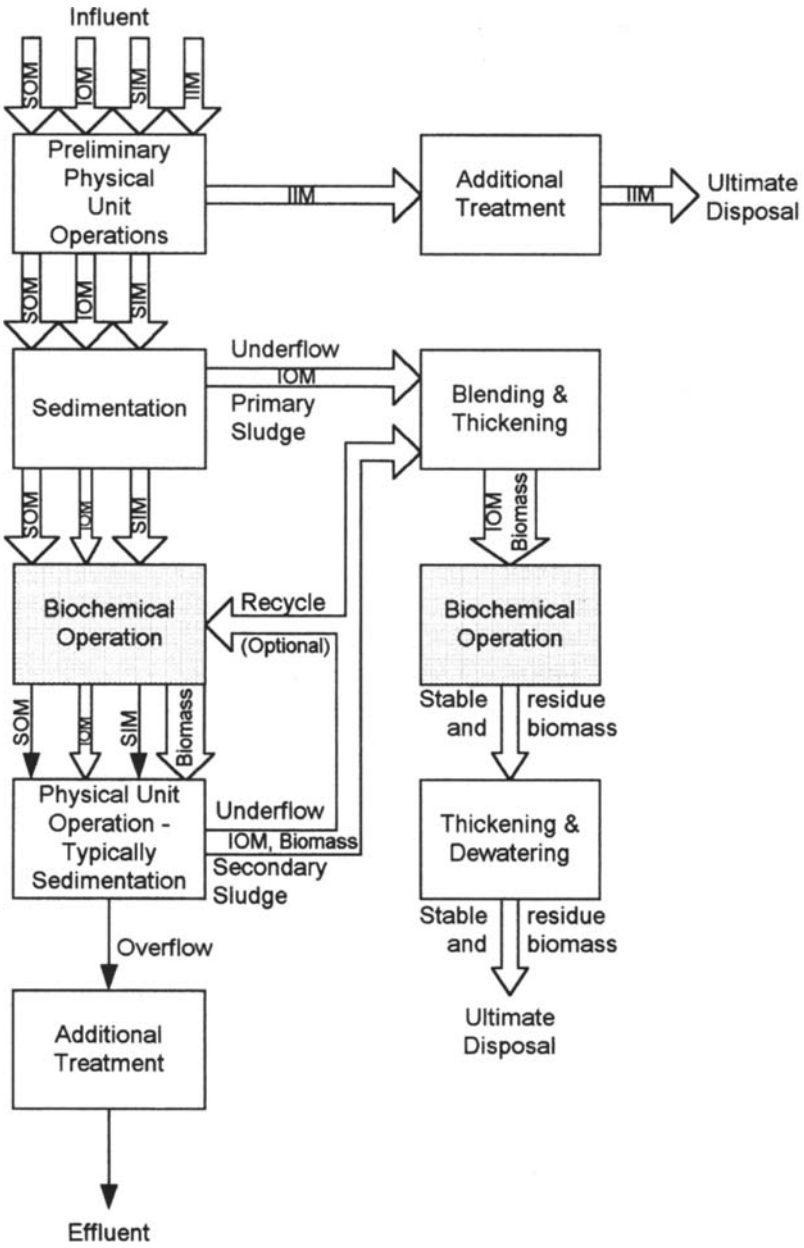


Figure 1.1 Typical process flow diagram for a wastewater treatment system illustrating the role of the biochemical operations. SOM = soluble organic matter; IOM = insoluble organic matter; SIM = soluble inorganic matter; IIM = insoluble inorganic matter.

ment of insoluble organic matter by that biomass. The nature of the materials in primary sludges tend to be very diverse because of the multitude of sources from which they arise, whereas secondary sludges are more uniform, being mainly microbial biomass. Sometimes the two sludges are blended and treated together as shown in the figure, but at other times they are treated separately. This is because the efficacy of a biochemical operation in treating a sludge depends strongly on the nature of the materials in it.

In spite of the major role of biochemical operations in the treatment of wastewaters, if a visitor to a treatment facility were to ask the name of the particular biochemical operation being used, the answer generally would give little indication of its nature. In fact, the most common operation, activated sludge, was named before its biochemical nature was even recognized. Consequently, before starting the study of the various biochemical operations it would be beneficial to establish what they are and what they do.

1.2 CRITERIA FOR CLASSIFICATION

The classification of biochemical operations may be approached from three points of view: (1) the biochemical transformation, (2) the biochemical environment, and (3) the bioreactor configuration. If all are considered together, the result is a detailed classification system that will aid the engineer in choosing the operation most appropriate for a given need.

1.2.1 The Biochemical Transformation

Removal of Soluble Organic Matter. The major application of biochemical operations to the main wastewater stream is the removal of SOM. This occurs as the microorganisms use it as a food source, converting a portion of the carbon in it into new biomass and the remainder into carbon dioxide. The carbon dioxide is evolved as a gas and the biomass is removed by sedimentation, leaving the wastewater free of the original organic matter. Because a large portion of the carbon in the original organic matter is oxidized to carbon dioxide, removal of SOM is also often referred to as carbon oxidation.

Aerobic cultures of microorganisms are particularly suitable for the removal of organic matter in the concentration range between 50 and 4000 mg/L as biodegradable chemical oxygen demand (COD). At lower concentrations, carbon adsorption is often more economical, although biochemical operations are being used for treatment of contaminated groundwaters that contain less than 50 mg/L of COD. Although they must often be followed by aerobic cultures to provide an effluent suitable for discharge, anaerobic cultures are frequently used for high strength wastewaters because they do not require oxygen, give less excess biomass, and produce methane gas as a usable product. If the COD concentration to be removed is above 50,000 mg/L, however, then evaporation and incineration may be more economical. Anaerobic cultures are also used to treat wastewaters of moderate strength (down to ~ 1000 mg/L as COD), and have been proposed for use with dilute wastewaters as well. It should be emphasized that the concentrations given are for soluble organic matter. Suspended or colloidal organic matter is often removed more easily from the main

wastewater stream by physical or chemical means, and then treated in a concentrated form. However, mixtures of soluble, colloidal, and suspended organic matter are often treated by biochemical means.

Stabilization of Insoluble Organic Matter. Many wastewaters contain appreciable quantities of colloidal organic matter which are not removed by sedimentation. When they are treated in a biochemical operation for removal of the SOM, much of the colloidal organic matter is entrapped with the biomass and ultimately converted to stable end products that are resistant to further biological activity. The formation of such stable end products is referred to as stabilization. Some stabilization will occur in the biochemical operation removing the soluble organic matter, but most will occur in operations designed specifically for that purpose.

Insoluble organic matter comes from the wastewater itself and from the growth of microorganisms as they remove soluble organic matter. Because these solids can be removed from the wastewater by settling, they are normally concentrated by sedimentation before being subjected to stabilization by biochemical means. Stabilization is accomplished both aerobically and anaerobically, although anaerobic stabilization is more energy efficient. The end products of stabilization are carbon dioxide, inorganic solids, and insoluble organic residues that are relatively resistant to further biological activity and have characteristics similar to humus. In addition, methane gas is a product from anaerobic operations.

Conversion of Soluble Inorganic Matter. Since the discovery, during the 1960s, of the effects of eutrophication, engineers have been concerned about the removal of inorganic nutrients from wastewater. Two of the prime causes of eutrophication are nitrogen and phosphorus, and a number of biological nutrient removal processes have been developed to remove them. Phosphorus is present in domestic wastewater in inorganic form as orthophosphate, condensed phosphates (e.g., pyrophosphate, tripolyphosphate, and trimetaphosphate), and organic phosphate (e.g., sugar phosphates, phospholipids, and nucleotides). Both condensed phosphates and organic phosphate are converted to orthophosphate through microbial activity. Orthophosphate, in turn, is removed through its uptake by specialized bacteria possessing unique growth characteristics that allow them to store large quantities of it in granules within the cell. Nitrogen is present in domestic wastewater as ammonia and organic nitrogen (e.g., amino acids, protein, and nucleotides), which is converted to ammonia as the organic matter is biodegraded. Two groups of bacteria are required to convert the ammonia into an innocuous form. First, nitrifying bacteria oxidize the ammonia to nitrate in a process called nitrification. Then denitrifying bacteria convert the nitrate to nitrogen gas in a process called denitrification. The nitrogen gas escapes to the atmosphere. Other inorganic transformations occur in nature, but few are exploited on a large scale in biochemical operations.

1.2.2 The Biochemical Environment

The most important characteristic of the environment in which microorganisms grow is the terminal acceptor of the electrons they remove as they oxidize chemicals to obtain energy. There are three major types of electron acceptors: oxygen, inorganic compounds, and organic compounds. If dissolved oxygen is present or supplied in sufficient quantity so as to not be rate limiting, the environment is considered to be

aerobic. Growth is generally most efficient in this environment and the amount of biomass formed per unit of waste destroyed is high. Strictly speaking, any environment that is not aerobic is anaerobic. Within the wastewater treatment field, however, the term anaerobic is normally reserved for the situation in which organic compounds, carbon dioxide, and sulfate serve as the major terminal electron acceptor and in which the electrode potential is very negative. Growth is less efficient under this condition. When nitrate and/or nitrite are present and serve as the primary electron acceptor in the absence of oxygen, the environment is called anoxic. The presence of nitrate and/or nitrite causes the electrode potential to be higher and growth to be more efficient than under anaerobic conditions, although not as high or as efficient as when oxygen is present.

The biochemical environment has a profound effect on the ecology of the microbial community. Aerobic operations tend to support complete food chains from bacteria at the bottom to rotifers at the top. Anoxic environments are more limited, and anaerobic are most limited, being predominantly bacterial. The biochemical environment influences the outcome of the treatment process because the microorganisms growing in the three environments may have very different metabolic pathways. This becomes important during the treatment of industrial wastewaters because some transformations can be carried out aerobically but not anaerobically, and vice versa.

1.2.3 Bioreactor Configuration

The importance of classifying biochemical operations according to bioreactor type follows from the fact that the completeness of a given biochemical transformation will be strongly influenced by the physical configuration of the bioreactor in which it is being carried out. Therefore, it is important to get a clear picture of the many bioreactor types available.

Wastewater treatment bioreactors fall into two major categories, depending on the way in which microorganisms grow in them: suspended in the liquid under treatment or attached to a solid support. When suspended growth cultures are used, mixing is required to keep the biomass in suspension, and some form of physical unit operation, such as sedimentation, is used to remove the biomass from the treated effluent prior to discharge. In contrast, attached growth cultures grow as a biofilm on a solid support and the liquid being treated flows past them. However, because organisms can slough from the support, a physical unit operation is usually required before the treated effluent may be discharged.

Suspended Growth Bioreactors. The simplest possible continuous flow suspended growth bioreactor is the continuous stirred tank reactor (CSTR), which consists of a well mixed vessel with a pollutant-rich influent stream and a treated effluent stream containing microorganisms. The liquid volume is constant and the mixing is sufficient to make the concentrations of all constituents uniform throughout the reactor and equal to the concentrations in the effluent. Consequently, these reactors are also called completely mixed reactors. The uniform conditions maintain the biomass in a constant average physiological state. Considerable operational flexibility may be gained by the addition of a physical unit operation, such as a sedimentation basin, which captures the biomass, as shown in Figure 1.1. As discussed previously, the overflow from the sedimentation basin is relatively free of biomass, while the underflow contains a concentrated slurry. Most of that concentrated slurry is recycled

to the bioreactor, but a portion is wasted. Because the wasted biomass is organic, it must be treated in an appropriate process before release to the environment.

Connecting several CSTRs in series offers additional flexibility as feed may be added to any or all of them. Furthermore, biomass recycle may be employed about the entire chain or any portion of it. The behavior of such systems is complex because the physiological state of the biomass changes as it passes from bioreactor to bioreactor. Nevertheless, many common wastewater treatment systems use bioreactors with split influent and recycle streams. One advantage of multistage systems is that different environments may be imposed upon different stages, thereby allowing multiple objectives to be accomplished. This is very common in biological nutrient removal processes.

A batch reactor is a completely mixed reactor without continuous flow through it. Instead, a "batch" of material is placed into the vessel with the appropriate biomass and allowed to react to completion as the microorganisms grow on the pollutants present. As growth proceeds, reaction conditions change and consequently, so does the growth environment. Batch processes can be very flexible and are particularly well suited for situations with low or highly variable flows. Furthermore, by changing the nature of the electron acceptor temporally, it is also possible to accomplish nutrient removal in a single bioreactor. Because their operation follows a sequence of events, they are commonly called sequencing batch reactors (SBRs).

A perfect plug–flow reactor (PFR) is one in which fluid elements move through in the same order that they enter, without intermixing. Thus, the perfect PFR and the CSTR represent the two extreme ends of the continuum representing all possible degrees of mixing. Because of the lack of intermixing, perfect PFRs may be considered to contain an infinite number of moving batch cultures wherein changes occur spatially as well as temporally. Both, however, cause the biomass to go through cycles of physiological change that can have strong impacts on both community structure and activity. Because perfect PFRs are difficult to achieve in practice, plug–flow conditions are generally approximated with a number of CSTRs in series. In Chapter 4, we will examine ways of characterizing the mixing conditions in suspended growth bioreactors.

Attached Growth Bioreactors. There are three major types of attached growth bioreactors: (1) packed towers, (2) rotating discs, and (3) fluidized beds. The microorganisms in a packed tower grow as a film on an immobile support, such as plastic media. In aerobic bioreactors, the wastewater flows down the media in a thin film. If no recirculation of effluent is practiced, there is considerable change in reaction environment from top to bottom of the tower as the bacteria remove the pollutants. Recirculation of effluent tends to reduce the severity of that change, and the larger the recirculation flow, the more homogeneous the environment becomes. The performance of this bioreactor type is strongly influenced by the manner in which effluent is recirculated. Organisms are continually sloughed from the support surface as a result of fluid shear. If they are removed from the effluent prior to recirculation, pollutant removal is caused primarily by the activity of the attached biomass. On the other hand, if flow is recirculated prior to the removal of the sloughed-off microorganisms, the fluid stream will resemble that of a suspended growth bioreactor and pollutant removal will be by both attached and suspended biomass. In anaerobic packed towers, the media is submerged and flow may be either upward or downward.

Microorganisms in a rotating disc reactor (RDR) grow attached to plastic discs that are rotated in the liquid. In most situations, the horizontal shaft on which the discs are mounted is oriented perpendicularly to the direction of flow and several reactors in series are used to achieve the desired effluent quality. Consequently, environmental conditions are uniform within a given reactor, but change from reactor to reactor down the chain. This means that both the microbial community structure and the physiological state change from reactor to reactor.

In fluidized bed bioreactors (FBBRs), the microorganisms grow attached to small particles, such as sand grains, which are maintained in a fluidized state by the upward velocity of the wastewater undergoing treatment. The effluent from such bioreactors generally contains little suspended biomass, but particles must be continually removed and cleaned to maintain a constant mass of microorganisms in the system. The cleaned particles are continually returned to the bioreactor while the wasted biomass is sent to an appropriate treatment process. Recirculation of effluent around the bioreactor is usually needed to achieve the required fluidization velocity and thus the system tends to behave as if it were completely mixed.

1.3 COMMON NAMES OF BIOCHEMICAL OPERATIONS

In almost all fields, certain operations have gained common names through years of use and development. Although such names are not always logical, they are recognized and accepted because of their historical significance. Such is the case in environmental engineering. In fact, some of the names bear little resemblance to the process objectives and are even applied to more than one reactor configuration. For purposes of discussion, twelve common names have been chosen and are listed in Table 1.1. To relate those names to the classification scheme presented above, Table 1.2 was prepared. It defines each name in terms of the bioreactor configuration, the treatment objective, and the reaction environment. Many other named biochemical operations are used, but they can all be related to those described in Table 1.2.

1.3.1 Suspended Growth Bioreactors

Activated Sludge. Eight different types of activated sludge systems are listed in Table 1.2, suggesting that the name is not very descriptive. The common char-

Table 1.1 Common Biochemical Operations

Suspended growth reactors	Attached growth reactors
Activated sludge	Fluidized bed
Biological nutrient removal	Rotating biological contactor
Aerobic digestion	Trickling filter
Anaerobic contact	Packed bed
Upflow anaerobic sludge blanket	Anaerobic filter
Anaerobic digestion	
Lagoon	

acteristic of all of them, however, is that they use a flocculent suspended growth culture of microorganisms in an aerobic bioreactor and employ some means of biomass recycle. Further examination of the table reveals that the primary treatment objective is the removal of soluble organic matter and oxidation of the carbon contained in it. Under appropriate conditions, nitrification will also occur, and thus it is listed as an objective for those systems in which it is most likely. Extended aeration activated sludge (EAAS) systems are often used on wastewaters that have not been treated in a physical operation to remove suspended organic matter. In that case, the insoluble organic matter becomes trapped in the biofloc and undergoes some oxidation and stabilization. Thus, that objective is marked for it. Most other activated sludge types are used on wastewaters from which settleable solids have been removed. As discussed earlier, however, those wastewaters still contain colloidal organic matter, most of which will be removed along with the soluble organic matter. Even though the colloidal material is insoluble and will be partially stabilized during treatment, the main event governing system performance is removal of the soluble organic matter, which is listed as the main treatment objective.

The first uses of activated sludge were on a batch basis. At the end of each aeration period suspended solids (referred to as sludge) were present and they were left in the bioreactor when the clear wastewater was withdrawn after settling. As this batch procedure was repeated the quantity of suspended solids increased, giving more complete removal of organic matter within the allotted reaction time. Although this increase in suspended solids with the associated improvement in removal activity was due to the growth of a viable microbial culture, the reason was unknown to the early researchers, who characterized the sludge as being "activated," thereby giving the process its name.[3] Use of the batch process waned as larger facilities were required, but during the 1970s there was a resurgence of interest in the use of batch reactors because of the flexibility offered small installations. Now referred to as sequencing batch reactor activated sludge (SBRAS), many are in use treating both municipal and industrial wastewaters.

As the need to treat larger flows increased, the early batch operation was converted to continuous flow through the use of long aeration chambers similar to plug–flow reactors, followed by sedimentation and biomass recycle. Such systems are called conventional activated sludge (CAS). Various modifications of the plug–flow reactor were tried, among them introduction of the wastewater at various points along the tank, in what has been called step feed activated sludge (SFAS). In the mid-50s, various engineers began advocating the CSTR with cell recycle as an alternative to the CAS reactor because of its inherent stability. That stability, plus the advantages regarding the maintenance of the microbial community in a relatively constant physiological state, caused wide adoption of the completely mixed activated sludge (CMAS) process, particularly for the treatment of industrial wastewaters. The process, however, tended to produce sludges which did not settle as well as sludges from systems containing concentration gradients, so that today many bioreactor systems in use employ several small CSTRs in series before a large one, thereby achieving desired environmental conditions. Such systems are referred to as selector activated sludge (SAS) systems. Other innovations that require CSTRs in series, such as the use of high purity oxygen (HPOAS), have also been adopted. The history of the activated sludge process is very interesting and the reader is encouraged to learn

Table 1.2 Classification of "Named" Biochemical Operations

Name	Acronym	Bioreactor configuration	Objective: Removal of soluble organic matter			Objective: Stabilization of insoluble organic matter			Objective: Conversion of soluble inorganic matter		
			Aerob.	Anaerob.	Anoxic	Aerob.	Anaerob.	Anoxic	Aerob.	Anaerob.	Anoxic
Suspended growth reactors											
Activated sludge		All with biomass recycle									
Completely mixed	CMAS	CSTR	X						N[a]		
Contact stabilization	CSAS	CSTRs in series	X								
Conventional	CAS	CSTRs in series or plug–flow with dispersion	X						N		
Extended aeration	EAAS	CSTR, CSTRs in series, or plug–flow with dispersion	X			X			N		
High purity oxygen	HPOAS	CSTRs in series	X						N		
Selector	SAS	CSTRs in series	X						N		
Sequencing batch reactors	SBRAS	Completely mixed batch	X						N		
Step feed	SFAS	CSTRs in series or plug–flow with dispersion, both with multiple feed points	X						N		
Biological nutrient removal	BNR	All with biomass recycle									
Biological phosphorus removal		CSTRs in series	X	X						P[b]	
Separate stage denitrification		CSTR			X						D[c]
Separate stage nitrification		CSTR							N		

Sequencing batch reactors		Completely mixed batch	X	X	X				N	P	D
Single sludge systems		CSTRs in series with internal recirculation streams	X	X	X				N	P	D
Aerobic digestion											
Conventional	CAD	CSTR				X			N		
Anoxic/aerobic	A/AD	CSTRs in series				X		X	N		D
Anaerobic contact	AC	CSTR with biomass recycle		X			X				
Upflow anaerobic sludge blanket	UASB	Upflow sludge blanket reactor		X			X				
Anaerobic digestion	AD	CSTR					X				
Lagoon		All without biomass recycle									
Completely mixed aerated	CMAL	CSTR	X			X			N		
Facultative/aerated	F/AL	Large shallow basins in series	X		X	X	X	X	N		D
Anaerobic	ANL	Large deep basins		X			X				
		Attached growth reactors									
Fluidized bed bioreactors	FBBR										
Aerobic		Fluidized bed with oxygenation cell	X						N		
Anaerobic		Fluidized bed		X							
Anoxic		Fluidized bed			X						D
Rotating biological contactor	RBC	Rotating disc	X						N		
Trickling filter	TF	Packed tower with large media	X						N		
Packed bed		Submerged packed tower with small media	X		X				N		D
Anaerobic filter	AF	Submerged packed tower with large media		X							

[a]Nitrification.
[b]Phosphorus uptake or release. Requires both aerobic and anaerobic zones.
[c]Denitrification.

more about it by referring to Refs. 1 and 3. Activated sludge systems are discussed in detail in Chapter 10, which includes flow diagrams for several of the variations.

Biological Nutrient Removal. Biological nutrient removal (BNR) systems are among the most complicated biochemical operations devised for wastewater treatment, and like the activated sludge systems from which they were derived, they come in a number of configurations, as shown in Table 1.2. A biological phosphorus removal system is essentially an activated sludge system employing CSTRs in series, in which the first bioreactor is anaerobic to encourage the growth of specialized phosphorus-storing bacteria. Separate stage nitrification and denitrification systems usually employ single CSTRs with cell recycle for the purpose of converting ammonia to nitrate, and nitrate to nitrogen gas, respectively. They are usually used as downstream treatment additions to existing systems. Sequencing batch reactors can be made to remove phosphorus and nitrogen while they are achieving carbon oxidation by imposing anaerobic and anoxic periods during their cycles, but otherwise are similar to the SBRAS used exclusively for removal of soluble organic matter. The most complex BNR systems are the single-sludge systems which accomplish carbon oxidation, nitrification, denitrification, and phosphorus removal with a single biomass by recycling it through CSTRs in series in which some are aerobic, some anoxic, and some anaerobic. A key characteristic of single sludge BNR systems is their use of internal recirculation streams from downstream reactors to upstream ones. There is currently strong interest in the United States in the removal of inorganic nutrients and much of the construction for municipal wastewater treatment in the next decade will be to convert existing activated sludge systems to BNR systems. BNR systems are covered in Chapter 11, where several flow diagrams are given.

Aerobic Digestion. Aerobic digestion is the name given to the aerobic destruction of insoluble organic matter in a suspended growth bioreactor. Generally, aerobic digesters employ a CSTR with a long retention time, allowing ample time for the conversion of much of the organic matter to carbon dioxide. Although not the primary objective, nitrification also occurs. Aerobic digestion is often used to destroy part of the excess biomass formed during treatment of soluble industrial wastewater and at small "package plant" installations treating domestic wastewater. Conventional aerobic digestion (CAD) maintains the biomass in an aerobic state at all times. Anoxic/aerobic digestion (A/AD) cycles the biomass between anoxic and aerobic conditions to use the nitrate formed during nitrification as an electron acceptor in place of oxygen, thereby reducing costs of aeration and pH control. Sometimes small treatment plants do not have primary sedimentation and allow aerobic digestion of the insoluble organic matter present in the influent to occur in the same bioreactor as the removal of soluble organic matter and the stabilization of the excess biomass formed in the process. In those cases the system is usually considered to be an extended aeration activated sludge process, as discussed above. Aerobic digestion is discussed in Chapter 12.

Anaerobic Contact. The operation used to remove soluble organic matter under anaerobic conditions in a CSTR with cell recycle is called anaerobic contact (AC). It is also used to treat wastes containing a mixture of soluble and insoluble organic matter, as is the activated sludge process. Two groups of microorganisms are involved. The first group is responsible for the conversion of the influent organic matter into acetic acid, molecular hydrogen, and carbon dioxide. Other short chain

volatile fatty acids may accumulate, as will a stable insoluble residue similar to humus. The second group is responsible for the conversion of the acetic acid, hydrogen, and carbon dioxide to methane gas. Anaerobic contact is well suited as a pretreatment method for wastes containing more than 4000 mg/L of biodegradable COD, but less than 50,000 mg/L, because it is less expensive than either activated sludge or evaporation.[2] Its main advantages over activated sludge systems are lower power requirements, less production of excess solids, and the generation of methane gas. Further treatment is required for the effluent from anaerobic contact, however, because many aerobically biodegradable soluble products remain.

Upflow Anaerobic Sludge Blanket Reactor. Like the anaerobic contact system, the primary purpose of the upflow anaerobic sludge blanket (UASB) reactor is the removal of soluble organic matter under anaerobic conditions, with the production of methane gas. The thing that distinguishes the UASB system from anaerobic contact is the absence of an external sedimentation chamber. Instead, the wastewater is introduced at the bottom of the reactor and flows upward at a velocity that matches the settling velocity of the biomass. In this way a sludge blanket is formed and maintained. A special zone is required to allow the gas formed to escape without carrying sludge particles with it. The biomass in these reactors is in the form of compact granules that contain mixed cultures of methanogenic and acidogenic bacteria.[4] Because the retention of biomass in UASBs is good, they are suitable for treating wastewaters with much lower substrate concentrations than anaerobic contact. In fact, they have been demonstrated to be capable of effective treatment of municipal wastewater.[5]

Anaerobic Digestion. By far the largest use of anaerobic cultures is in the stabilization of insoluble organic matter by anaerobic digestion (AD), which involves microbial communities similar to those found in anaerobic contact. Anaerobic digestion is one of the oldest forms of wastewater treatment, yet because of the complex ecosystem involved it has continued to be the subject of research and new process development. Designers currently favor the use of CSTRs because of their uniform environmental conditions, and some utilize CSTRs with solids recycle because smaller bioreactors can be used. Anaerobic digestion, along with AC and UASB systems, is discussed in Chapter 13.

Lagoons. The term lagoon refers to suspended growth bioreactors that do not include biomass recycle from a downstream sedimentation tank. Their name comes from their construction and appearance. Historically, they have been constructed as large earthen basins, that because of their size, resemble typical "South Sea island lagoons." Originally, lagoons were not lined, but this has proven to be unacceptable because of the potential for leakage of the basin contents into groundwater. Consequently, new design practice requires them to be lined with an impermeable liner. A wide range of environmental conditions can exist in lagoons, depending on the degree of mixing imposed. If the lagoon is well mixed and aerated, it can be aerobic throughout, but with lesser degrees of mixing, solids will settle, leading to anoxic and anaerobic zones. Three types of lagoons are characterized in Table 1.2. Completely mixed aerated lagoons (CMALs) can generally be classified as completely mixed reactors that are used for the removal of soluble organic matter, although stabilization of insoluble organic matter and nitrification can also occur. Facultative/aerated lagoons (F/ALs) are mixed, but not sufficiently to keep all solids in suspen-

sion. As a consequence, the upper regions tend to be aerobic, whereas the bottom contains anaerobic sediments. Anaerobic lagoons (ANLs) are not purposefully mixed. Rather, any mixing that occurs is the result of gas evolution within them. ANLs are discussed in Chapter 13, whereas CMALs and F/ALs are covered in Chapter 14.

1.3.2 Attached Growth Bioreactors

Fluidized Bed Bioreactors. Fluidized bed bioreactors (FBBRs) can be operated with any of the three biochemical environments, and the nature of that environment determines what the bioreactor accomplishes. Fluidized bed systems for denitrification were among the earliest developed because all materials to be reacted were present in a soluble state. However, through use of pure oxygen as a means of providing dissolved oxygen at high concentration, aerobic fluidized beds soon followed. Their chief purpose is removal of soluble organic matter, but they are also used for nitrification. Finally, anaerobic fluidized bed systems were developed for the treatment of soluble wastewaters.[4] The key characteristic of fluidized bed systems is their ability to retain very high biomass concentrations, thereby allowing small bioreactor volumes to be used. This is accomplished by using very small particles, which provide a large surface area per unit volume, as the attachment media for biofilm growth. Maintenance of the particles in a fluidized state by control of the upflow velocity ensures better mass transfer characteristics than can be achieved in other attached growth systems. The major use of FBBRs has been for industrial wastewater treatment. This type of attached growth bioreactor is discussed in Chapter 21.

Rotating Biological Contactor. The rotating biological contactor (RBC) is a modern application of an old idea for the removal of soluble organic matter and the conversion of ammonia to nitrate. Microorganisms growing attached to rotating discs accomplish the desired objectives by the same mechanisms used in suspended growth systems, but in a more energy efficient manner because oxygen transfer is accomplished by the rotation of the discs, which are only half submerged. These bioreactors have been popular for the treatment of both domestic and industrial wastewaters, typically at smaller installations. RBCs are discussed in Chapter 20.

Trickling Filter. As indicated in Table 1.2, trickling filter (TF) is the name given to an aerobic attached growth bioreactor in the shape of a packed tower. Until the mid 1960s, TFs were made of stone, which limited their height to around two meters for structural reasons. Now TFs are made of plastic media much like that used as packing in absorption and cooling towers and are self-supporting to heights of around seven meters because of the greater void space and lighter weight of the media. The primary use of TFs is for removal of soluble organic matter and oxidation of ammonia to nitrate. Traditionally, TFs have been used for municipal wastewater treatment in small to medium size installations desiring minimal operating expense. However, since the introduction of plastic media they have also found use as pretreatment devices preceding other biochemical operations. This is because they have the ability to reduce the waste concentration at relatively low operating cost, a bonus when aerobic treatment is being employed. Trickling filters cause relatively little degradation of insoluble organic matter and should not be used for that purpose. They are covered in Chapter 19.

Packed Bed. Packed bed bioreactors utilize submerged media with a particle size on the order of a few millimeters. They are designed and operated with flow either upward or downward. Because of the small particle size, packed beds act as physical filters, as well as biochemical operations. Their primary use is for conversion of soluble inorganic matter, particularly nitrification and denitrification, depending on the biochemical environment provided. They are also used to remove soluble organic matter, especially at low concentrations. They are discussed in Chapter 21.

Anaerobic Filter. Even though the name anaerobic filter (AF) suggests the use of media like that in a packed bed, this is not the case. Rather, an AF is a packed tower containing plastic media like that in a TF. Unlike a TF, however, an AF is operated under submerged conditions to maintain the microbial community in an anaerobic state. Its primary use is to treat high strength soluble wastewaters by converting the bulk of the organic matter to methane. Biomass grows attached to the solid media in the tower, and flow may be from either direction. If flow is upward, suspended biomass may accumulate and have to be removed periodically. Even though AFs are attached growth systems, their anaerobic character is the determinant of their behavior. Anaerobic filters are discussed in Chapter 13.

1.3.3 Miscellaneous Operations

There are many other biochemical operations in use or development. Among those are the hybrid systems, such as the activated biofilter, which combine the attributes of suspended and attached growth bioreactors. Even though these systems are not listed in Tables 1.1 and 1.2, they are covered at appropriate points. However, many other new biochemical operations are not included in this book due to space constraints. Their exclusion should not be construed as a bias against their use. Rather, it is felt that once the fundamental principles of biochemical operations are learned, the reader will be able to apply them to understand and evaluate any biochemical wastewater treatment system.

1.4 KEY POINTS

1. Biochemical operations may be carried out in aerobic, anoxic, or anaerobic environments, and the choice of environment has a profound effect on both the ecology of the microbial community and the outcome of its activity.
2. Three major biochemical transformations may be performed with biochemical operations: removal of soluble organic matter (SOM), stabilization of insoluble organic matter (IOM), and conversion of soluble inorganic matter (SIM).
3. Bioreactors for biochemical operations may be divided into two major categories, depending on the manner in which the microorganisms grow: suspended in the wastewater undergoing treatment, and attached to a solid support.
4. The major suspended growth bioreactors are: continuous stirred tank reactor (CSTR), either alone or in series; batch reactor; and plug–flow reactor (PFR). The major attached growth bioreactors are: fluidized bed (FBBR), packed tower, and rotating disc reactor (RDR).

5. The major suspended growth biochemical operations are: activated sludge, biological nutrient removal (BNR), aerobic digestion, anaerobic contact (AC), upflow anaerobic sludge blanket reactor (UASB), anaerobic digestion (AD), and lagoons. The major attached growth biochemical operations are: fluidized bed, rotating biological contactor (RBC), trickling filter (TF), packed bed, and anaerobic filter (AF).

1.5 STUDY QUESTIONS

1. List and define the three major biochemical transformations that may be performed with biochemical operations.
2. Describe each of the major bioreactor types that find use in biochemical operations.
3. List the twelve named biochemical operations and tell whether each uses a suspended or an attached growth culture.
4. Describe each of the named biochemical operations in terms of the biochemical transformation involved, the reaction environment used, and the bioreactor configuration employed.

REFERENCES

1. Alleman, J. E. and T. B. S. Prakasam. Reflections on seven decades of activated sludge history. *Journal, Water Pollution Control Federation* **55**, 436–443, 1983.
2. Cillie, G. G., M. R. Hensen, G. J. Stander, and R. D. Baillie. Anaerobic digestion. IV. The application of the process in waste purification. *Water Research* **3**, 623–643, 1969.
3. Sawyer, C. N. Milestones in the development of the activated sludge process. *Journal, Water Pollution Control Federation* **37**, 151–170, 1965.
4. Switzenbaum, M. S. Anaerobic treatment of wastewater: recent developments. *ASM News* **49**, 532–536, 1983.
5. Switzenbaum, M. S. and C. P. L. Grady Jr. Feature—Anaerobic treatment of domestic wastewater. *Journal, Water Pollution Control Federation* **58**, 102–106, 1986.

2

Fundamentals of Biochemical Operations

Before we begin the systematic study of biochemical operations, it is necessary to develop a clear picture of what wastewater treatment engineers hope to accomplish through their use. Furthermore, if we are to develop the capability for their design, it is necessary to understand what is happening within them and to recognize the role of various types of microorganisms in those events.

2.1 OVERVIEW OF BIOCHEMICAL OPERATIONS

Biochemical operations only alter and destroy materials that microorganisms act upon, i.e., those that are subject to biodegradation or biotransformation. If soluble pollutants are resistant to microbial attack, they are discharged from a biochemical operation in the same concentration that they enter it, unless they are acted on by chemical or physical mechanisms such as sorption or volatilization (see Chapter 22). Insoluble pollutants entering a suspended growth biochemical operation become intermixed with the biomass and, for all practical purposes, are inseparable from it. Consequently, engineers consider this mixture of biomass and insoluble pollutants as an entity, calling it mixed liquor suspended solids (MLSS). If insoluble pollutants are biodegradable, their mass is reduced. On the other hand, if they are nonbiodegradable, their only means of escape from the system is through MLSS wastage and their mass discharge rate in the wasted MLSS must equal their mass input rate to the system. Attached growth processes usually have little impact on nonbiodegradable insoluble pollutants, although in some cases those pollutants are flocculated and settled along with the biomass discharged from the operation.

When wastewater treatment engineers design biochemical operations they use natural cycles to accomplish in a short time what nature would require a long time to accomplish, often with environmental damage. For example, if biodegradable organic matter were discharged to a stream, the bacteria in that stream would use it as a source of carbon and energy (electrons) for growth (see Chapter 1). In the process, they would incorporate part of the carbon into new cell material and the rest would be oxidized to carbon dioxide to provide the energy for that synthesis. The electrons removed during the oxidation would be transferred to oxygen in the stream, but if the supply of oxygen were insufficient, the dissolved oxygen (DO) concentration would be depleted, killing fish and causing other adverse effects. On the other hand, in a well designed biochemical operation, microbial growth is allowed to occur in

an environment where the appropriate amount of oxygen can be supplied, thereby destroying the organic matter and allowing the treated wastewater to be discharged without environmental harm.

The two major cycles employed in biochemical operations are the carbon and nitrogen cycles. Actually, most biochemical operations only use half of the carbon cycle, i.e., the oxidation of organic carbon, releasing carbon dioxide. While some biochemical operations use algae and plants to fix carbon dioxide and release oxygen, thereby using the other half of the carbon cycle, they are not as widely applied and will not be covered in this book. However, almost all of the nitrogen cycle is used, as illustrated in Figure 2.1. In domestic wastewaters, most nitrogen is in the form of ammonia (NH_3) and organic nitrogen, whereas industrial wastewaters sometimes contain nitrate (NO_3^-) nitrogen as well. Organic nitrogen is in the form of amino groups (NH_2), which are released as ammonia—in the process called ammonification—as the organic matter containing them undergoes biodegradation. The form in which bacteria incorporate nitrogen during growth is as ammonia. If an industrial wastewater has insufficient ammonia or organic nitrogen to meet the growth needs of the bacteria, but contains nitrate or nitrite (NO_2^-) nitrogen, they will be converted to ammonia through assimilative reduction for use in cell synthesis. On the other hand,

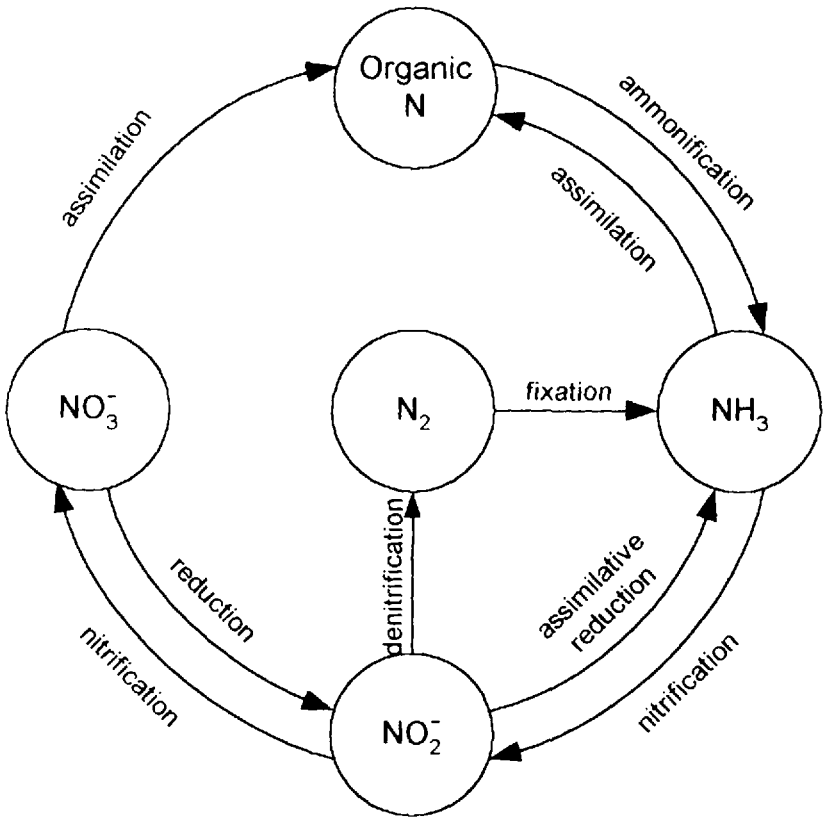


Figure 2.1 The nitrogen cycle.

if a wastewater contains ammonia-N in excess of that needed for cell synthesis, nitrification can occur, where the excess ammonia-N is oxidized to nitrate-N, going through the intermediate, nitrite. Discharge of nitrate to a receiving water is preferable to discharge of ammonia because nitrification in the receiving water can deplete the DO, just as degradation of organic matter can. In some cases, however, discharge of nitrate can have a deleterious effect on the receiving water, and thus some effluent standards limit its concentration. In that case, biochemical operations that use denitrification to convert nitrate and nitrite to nitrogen gas must be used to reduce the amount of nitrogen in the effluent. The only step in the nitrogen cycle not normally found in biochemical operations is nitrogen fixation, in which nitrogen gas is converted to a form that can be used by plants, animals, and microorganisms.

2.2 MAJOR TYPES OF MICROORGANISMS AND THEIR ROLES

Modern molecular biology has allowed scientists to investigate the relatedness among organisms by analysis of the nucleotide sequences within certain segments of their genes. Organization of this information into a phylogenetic tree has revealed that organisms fall into three primary groupings, or domains: Archaea, Bacteria, and Eucarya.[83] Members of the domains Archaea and Bacteria are microscopic and procaryotic,[a] i.e., they lack a nuclear membrane, whereas members of the domain Eucarya are eucaryotic, i.e., they have a nuclear membrane, and vary in size from microscopic (e.g., protozoa) to macroscopic (e.g., animals). The workhorses of biochemical operations belong to the domains Bacteria and Archaea, but protozoa and other microscopic Eucarya have a role as well. Thus it is important to have a clear picture of what various microorganisms do.

2.2.1 Bacteria

Bacteria can be classified in many ways; however, the most important from an engineering perspective is operational. Consequently, we will focus on it.

Like all organisms, members of the domain Bacteria derive energy and reducing power from oxidation reactions, which involve the removal of electrons. Thus, the nature of the electron donor is an important criterion for their classification. The two sources of electrons of most importance in biochemical operations are organic and inorganic compounds that are present in the wastewater or released during treatment. Bacteria that use organic compounds as their electron donor and their source of carbon for cell synthesis are called heterotrophic bacteria, or simply heterotrophs. Since the removal and stabilization of organic matter are the most important uses of biochemical operations, it follows that heterotrophic bacteria predominate in the systems. Bacteria that use inorganic compounds as their electron donor and carbon

[a] Recognition of the distinction between Bacteria and Archaea is relatively recent. Consequently, it is still common for members of both domains to be referred as bacteria, in reference to their procaryotic nature. In this book, the term bacteria (with a lower case "b") will be used to refer to procaryotes in general, without regard to their domain.

dioxide as their source of carbon are chemoautotrophic bacteria, although most wastewater treatment engineers call them autotrophic bacteria, or simply autotrophs. The most important autotrophic bacteria in biochemical operations are those that use ammonia-N and nitrite-N. They are responsible for nitrification, and are referred to as nitrifiers. Other autotrophic bacteria are important in nature and in sewers, but play little role in engineered treatment systems.

Another important characteristic of bacteria is the type of electron acceptor they use. The most important acceptor in biochemical operations is oxygen. Bacteria that use only oxygen are called obligately aerobic bacteria, or simply obligate aerobes. Nitrifying bacteria are the most significant obligately aerobic bacteria commonly found in biochemical operations. At the other end of the spectrum are obligately anaerobic bacteria, which can only function in the absence of molecular oxygen. Between the two obligate extremes are the facultatively anaerobic, or simply facultative, bacteria. They use oxygen as their electron acceptor when it is present in sufficient quantity, but can shift to an alternative acceptor in its absence. Thus they tend to predominate in biochemical operations. Some facultative bacteria are fermentative, meaning that they use organic compounds as their alternative terminal electron acceptor in the absence of oxygen, producing reduced organic end products. Others perform anaerobic respiration, in which an inorganic compound serves as the alternative acceptor. In Chapter 1, mention is made of anoxic environments in which oxygen is absent, but nitrate is present as an electron acceptor. Because of the prevalence of such environments in biochemical operations, the most significant facultative bacteria are those that perform denitrification, i.e., reduce nitrate-N to nitrogen gas. Other facultative and obligately anaerobic bacteria reduce other inorganic compounds, but with the exception of protons (H^+), most are not of general importance in biochemical operations. Proton reduction, which occurs in anaerobic operations, yields hydrogen gas (H_2), which is an important electron donor for methane formation.

Gravity sedimentation is the most common method for removing biomass from the effluent from biochemical operations prior to its discharge. Since single bacteria are so small ($\sim$ 0.5–1.0 μm), it would be impossible to remove them in that way if they grew individually. Fortunately, under the proper growth conditions, bacteria in suspended growth cultures grow in clumps or gelatinous assemblages called biofloc, which range in size from 0.05 to 1.0 mm.[3] Figure 2.2a shows a typical floc particle. The bacteria which are primarily responsible for this are called floc-forming bacteria, and a variety of species fall into this category.

Not all bacteria are beneficial in biochemical operations; some are a nuisance. Two forms of nuisance bacteria can grow in aerobic/anoxic systems. One grows as long strands, or filaments, which become intermeshed with biofloc particles and interfere with sedimentation. They are called filamentous bacteria. Although a small number of filaments can provide strength for the biofloc, preventing its disruption by fluid shear forces, too many can act to hold the biofloc particles apart,[4] as shown in Figure 2.2b. When that occurs, sedimentation is very inefficient and the biomass will not compact into a sufficiently small volume to allow discharge of a clear effluent. The other type of nuisance bacteria forms copious quantities of foam in bioreactors that are being aerated for oxygen transfer. The foam can become so deep as to completely cover both aeration and sedimentation basins, thereby disrupting treatment and posing a danger to plant personnel. The most common nuisance or-

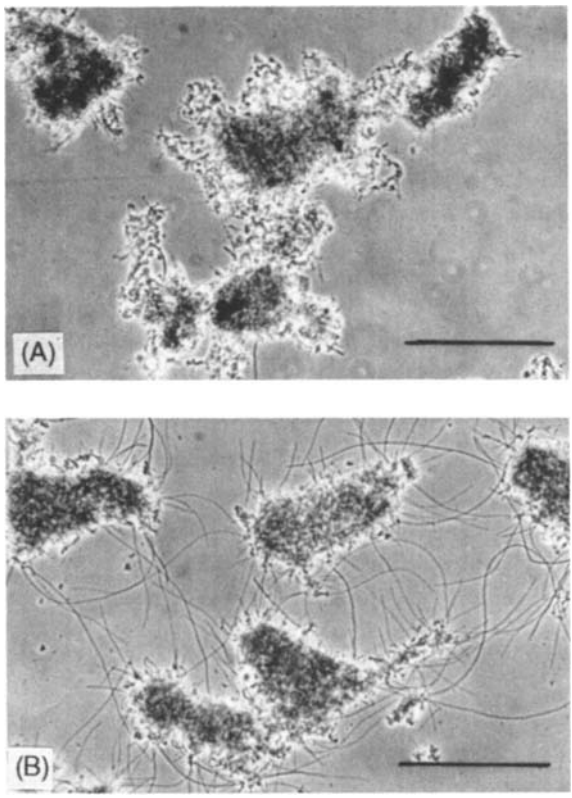


Figure 2.2 Photomicrographs of activated sludge floc: (a) Good settling biomass with optimal filaments; (b) poor settling biomass with excessive filaments. [Courtesy of M. G. Richard (Colorado State Univ.) and David Jenkins (Univ. California, Berkeley.)]

ganisms in anaerobic systems are the sulfate-reducing bacteria. It is generally desirable to design anaerobic operations to produce methane because it is a valuable product. If a wastewater contains high concentrations of sulfate, however, sulfate-reducing bacteria will compete for the electron donor, producing sulfide as a product. This not only reduces the amount of methane produced, but results in a product that is both dangerous and undesirable in most situations. Wastewater treatment engineers need to be aware of the growth characteristics of such nuisance organisms so that systems that discourage or prevent their growth can be designed.

Bacteria can also be classified according to their function in biochemical operations. Many act as primary degraders and attack the organic compounds present in the wastewater, beginning their degradation. If an organic compound is one normally found in nature (biogenic), the primary degraders usually will completely metabolize it in an aerobic environment, converting it to carbon dioxide, water, and new biomass. Such ultimate destruction is called mineralization and is the goal of most wastewater treatment systems. On the other hand, if an organic compound is synthetic and foreign to the biosphere (xenobiotic), it is possible that no single type of bacteria will be able to mineralize it. Instead, a microbial consortium may be

required, with secondary degraders living on the metabolic products excreted by the primary degraders. The more complex the organic compounds found in a wastewater, the more important secondary degraders will be. Secondary degraders are common in anaerobic environments, however, even when biogenic compounds are being degraded, because of the specialized needs of the bacteria involved. Other functions that are important in wastewater treatment systems are the production and elimination of nitrate-N through nitrification and denitrification, respectively. Consequently, it is not surprising that bacteria are classified according to those functions, as nitrifiers and denitrifiers. While the nitrifiers constitute a highly specialized group containing a limited number of species of aerobic, chemoautotrophic bacteria, the denitrifying bacteria constitute a diverse group of facultative heterotrophic bacteria containing many species. Finally, some species of bacteria have the ability to store and release phosphate in response to cyclical environmental conditions. Because they contain quantities of phosphate well in excess of other bacteria, these bacteria are often called phosphate accumulating organisms (PAOs).

As with the classification of pollutants in wastewaters, the classifications listed above are not exclusive, but overlap, with members of the domain Bacteria playing many roles. Nevertheless, these simple classification schemes are very helpful in describing the events occurring in biochemical operations and will be used throughout this book.

2.2.2 Archaea

Many Archaea are capable of growing in extreme environments, such as high temperatures (up to 90°C), high ionic strength, and highly reduced conditions. Consequently, members of this domain were first thought to be restricted to growth in such environments, although that proved to be incorrect. More recent studies have shown that Archaea are abundantly distributed in a wide variety of environments.[21] As our knowledge of the Archaea expands it is likely that wastewater treatment engineers will find more applications for them. Currently, however, their major use in biological wastewater treatment is in anaerobic operations, where they play the important role of producing methane. Methane-producing Archaea, commonly called methanogens, are obligate anaerobes that bring about the removal of organic matter from the liquid phase by producing an energy rich gas of low solubility. This allows capture of the energy in the pollutants in a useful form. Because methanogens are very limited in the substrates they can use, they grow in complex microbial communities with Bacteria, which carry out the initial attack on the pollutants and release the methanogens' substrates as fermentation products.

2.2.3 Eucarya

Although fungi can use soluble organic matter in competition with Bacteria, they seldom compete well in suspended growth cultures under normal conditions, and thus do not usually constitute a significant proportion of the microbial community.[22] On the other hand, when the supplies of oxygen and nitrogen are insufficient, or when the pH is low, fungi can proliferate, causing problems similar to those caused by filamentous bacteria. In contrast to suspended growth cultures, fungi commonly play an important role in attached growth cultures, making up a large part of the

biomass.[76] Under certain conditions, however, they can also become a nuisance in such systems by growing so heavily as to block interstices and impede flow.

Protozoa play an important role in suspended growth cultures by grazing on colloidal organic matter and dispersed bacteria, thereby reducing the turbidity remaining after the biofloc has been removed by sedimentation. Protozoa are also known to contribute to bioflocculation, but their contribution is thought to be less important than that of the floc-forming bacteria.[11] Although some protozoa can utilize soluble organic compounds for growth, it is doubtful that they can compete effectively with bacteria in that role and thus soluble substrate removal is generally considered to be due to bacterial action. Protozoa also play a significant role in attached growth bioreactors where the protozoan community is usually richer than it is in suspended growth cultures. Nevertheless, their role appears to be similar to that in suspended growth cultures.

Other Eucarya in suspended growth cultures are usually limited to rotifers and nematodes, but their presence depends very much on the way in which the culture is grown. Although these organisms feed upon protozoa and biofloc particles, their contribution to biochemical operations using suspended growth cultures is largely unknown because little change in process performance can be attributed to their presence. In contrast, because attached growth bioreactors provide a surface upon which higher organisms can graze, it is not uncommon for such reactors to have highly developed communities of macroinvertebrates in addition to rotifers and nematodes.[22] The nature of those communities depends largely on the physical characteristics of the bioreactor and in some cases the presence of the higher community has no deleterious effect on system performance. In other cases, however, the grazing community can disrupt development of the primary biofilm that is responsible for the removal of the pollutants, leading to a deterioration in system performance.

2.3 MICROBIAL ECOSYSTEMS IN BIOCHEMICAL OPERATIONS

An ecosystem is the sum of interacting elements (both biological and environmental) in a limited universe. Consequently, each biochemical operation will develop a unique ecosystem governed by the physical design of the facility, the chemical nature of the wastewater going to it, and the biochemical changes wrought by the resident organisms. The microbial community which develops in that ecosystem will be unique from the viewpoint of species diversity, being the result of physiological, genetic, and social adaptation. Thus, it is impossible to generalize about the numbers and types of species that will be present. Nevertheless, it would be instructive to consider the general nature of the community structures in biochemical operations and relate them to the environments in which the operations are performed. The objective of such an exercise is not the simple listing of the organisms present, but rather an understanding of the role that each important group plays in the operation.

Because the biochemical processes in aerobic and anoxic environments are based on respiration, whereas those in anaerobic environments are based on fermentation, there are large differences in the microbial communities involved. Thus, the biochemical environment provides a logical way for dividing this discussion.

2.3.1 Aerobic/Anoxic Operations

Suspended Growth Bioreactors. Activated sludge, aerated lagoons, and aerobic digesters have similar microbial ecosystems, although they differ somewhat in the relative importance of various groups. The microorganisms in those operations are all Bacteria and microscopic Eucarya, and generally may be divided into five major classes: (1) floc-forming organisms, (2) saprophytes, (3) nitrifying bacteria, (4) predators, and (5) nuisance organisms.[59] With the exception of nitrifying bacteria, these are not distinct physiological groups and, in fact, any particular organism may fit into more than one category at a time or may change categories as the selective pressures within the community change.

Floc-forming organisms play a very important role in suspended growth biochemical operations because without them the biomass cannot be separated from the treated wastewater nor can colloidal-sized organic pollutants be removed. Figure 2.2a shows typical, well-settling biomass. Originally it was thought that the bacterium *Zooglea ramigera* was primarily responsible for floc formation, but it has now been shown that a variety of bacteria are capable of flocculation,[58] although they constitute only a small percentage of the species found in a floc particle.[51] Classification of organisms into the floc-forming group is complicated by the fact that protozoa and fungi can also cause bacteria to flocculate.[11,58,59] Nevertheless, the predominant floc-forming organisms are generally considered to be bacteria,[11] with *Zooglea ramigera* playing an important role.[65] Flocculation is thought to be caused by aggregative growth and natural polyelectrolytes, although their origin is uncertain.

Saprophytes are organisms responsible for the degradation of organic matter. These are primarily heterotrophic bacteria and include most of those considered to be floc formers. Nonflocculent bacteria are also involved, but are entrapped within the floc particles. The saprophytes can be divided into primary and secondary degraders, as discussed previously, and the larger the number of substrates, the more diverse the community will be. The principal saprophytic genera are gram-negative and include *Achromobacter, Alcaligenes, Bacillus, Flavobacterium, Micrococcus,* and *Pseudomonas.*[57]

Nitrification is the conversion of ammonia-N to nitrate-N and it may be performed by either heterotrophic or autotrophic bacteria.[52] In spite of the fact that over a hundred heterotrophic species have been cited as forming nitrite from ammonia,[78] significant amounts of nitrate are not thought to be generated heterotrophically in natural systems,[14] although studies suggest that this assumption should be investigated further.[52] Nevertheless, nitrification in wastewater treatment systems is generally considered to be due to autotrophic bacteria, primarily of the genera *Nitrosomonas* and *Nitrobacter*, which appear to grow in close physical association.[48] *Nitrosomonas* oxidizes ammonia-N to nitrite-N with hydroxylamine as an intermediate product, whereas *Nitrobacter* oxidizes nitrite-N to nitrate-N in a single step. The fact that nitrifying bacteria are autotrophic does not mean that they cannot incorporate exogenous organic compounds while obtaining their energy from inorganic oxidation, because they can.[32] The amount of such uptake will be small and will vary with the growth conditions, however, so that most equations depicting the stoichiometry of nitrification ignore it and use carbon dioxide as the sole carbon source. Nitrifying bacteria have several unique growth characteristics that are important to their impact on and survival in biochemical operations. The first is that

their maximal growth rate is smaller than that of heterotrophic bacteria. Consequently, if suspended growth bioreactors are operated in a way that requires the bacteria to grow rapidly, the nitrifying bacteria will be lost from the system and nitrification will stop even though organic substrate removal will continue. Second, the amount of biomass formed per unit of nitrogen oxidized is small. As a result, they may make a negligible contribution to the MLSS concentration even when they have a significant effect on process performance.

The main predators in suspended growth bioreactors are the protozoa, which feed on the bacteria. About 230 species have been reported to occur in activated sludge and they may constitute as much as 5% of the biomass in the system.[58] Ciliates are usually the dominant protozoa, both numerically and on a mass basis. Almost all are known to feed on bacteria and the most important are either attached to or crawl over the surface of biomass flocs. On occasion, both amoeba and flagellates may be seen in small numbers, but they are not thought to play a major role in well-settling, stable communities. As discussed earlier, it has been suggested that protozoa play a secondary role in the formation of biomass flocs and contribute to the absence of dispersed bacteria and colloidal organic material in stable communities.[11]

Nuisance organisms are those that interfere with proper operation of a biochemical reactor when present in sufficient numbers. In suspended growth bioreactors, most problems arise with respect to removal of the biomass from the treated wastewater, and are the result of filamentous bacteria and fungi. Although a very small number of filamentous bacteria is desirable to strengthen floc particles, too large a number is undesirable.[70] Even a small percentage by weight in the microbial community can make the effective specific gravity of the biomass flocs so low that the biomass becomes very difficult to remove by gravity settling. This leads to a situation known as bulking. A poor-settling biomass is shown in Figure 2.2b. For many years it was thought that the bacterium *Sphaerotilus natans* was the organism primarily responsible for bulking, but the conditions causing its growth were a puzzle because they appeared to be so contradictory. It was not until the pioneering work of Eikelboom[13] that it was realized that many types of filamentous organisms could be responsible for bulking, and that different organisms were favored by different growth conditions. Today, effective bulking control is based on identification of the causative organism and elimination of the condition favoring its growth.[31] Table 2.1 ranks the most abundant filamentous organisms found in bulking sludges in the United States and Table 2.2 lists the suggested causes for some. In Table 2.2, the term, "low F/M," refers to a low food to microorganism ratio; in other words, the system is being operated with a very low loading of organic matter into it. It should be noted that although *Nocardia* is a commonly found filamentous organism, it does not normally cause bulking because its filaments do not extend beyond the floc particle.[31]

The other major nuisance associated with suspended growth cultures is excessive foaming. This condition is caused primarily by bacteria of the genus *Nocardia* and the species *Microthrix parvicella*.[31] There is still controversy concerning the conditions responsible for excessive foaming in suspended growth cultures. Because *Nocardia* and *M. parvicella* have very hydrophobic cell surfaces, they migrate to air bubble surfaces, where they stay, thereby stabilizing the bubbles and causing foam.[31] Foaming also appears to be related to the concentration of hydrophobic organic compounds at the air–water interface, where they are metabolized by the *Nocardia* and *Nocardia*-like organisms that have collected there.[21]

Table 2.1 Filament Abundance in Bulking and Foaming Activated Sludge in the United States

Rank	Filamentous organism	Percentage of treatment plants with bulking or foaming where filament was observed	
		Dominant	Secondary
1	*Nocardia* spp	31	17
2	type 1701	29	24
3	type 021N	19	15
4	type 0041	16	47
5	*Thiothrix* spp	12	20
6	*Sphaerotilus natans*	12	19
7	*Microthrix parvicella*	10	3
8	type 0092	9	4
9	*Halsicomenobacter hydrossis*	9	45
10	type 0675	7	16
11	type 0803	6	9
12	*Nostocoida limicola* (Types I, II, and III)	6	18
13	type 1851	6	2
14	type 0961	4	6
15	type 0581	3	1
16	*Beggiatoa* spp	1	4
17	fungi	1	2
18	type 0914	1	1
—	all others	1	—

From Ref. 31.

Table 2.2 Conditions Associated with Dominant Filament Types

Suggested causative conditions	Filament types
Low DO	*H. hydrossis*, *M. parvicella*, *S. natans*, type 1701
Low F/M	*M. parvicella*, types 0041, 0092, 0675, 1851
Completely mixed bioreactors	*H. hydrossis*, *Nocardia* spp., *N. limicola*, *S. natans*, *Thiothrix* spp., types 021N, 1701, 1851
Septic wastewater/sulfide	*Beggiatoa*, *Thiothrix* spp., types 021N, 0914
Nutrient deficiency	*S. natans*, *Thiothrix* spp., type 021N; possibly *H. hydrossis*, types 0041, 0675
Low pH	fungi

From Ref. 31.

Although the ecosystems of activated sludge, aerated lagoons, and aerobic digestion are complex, they are not as complicated as those in suspended growth systems accomplishing biological nutrient removal. This is because biological nutrient removal systems also contain anoxic and anaerobic reactors, which provide opportunities for the growth of microorganisms that do not ordinarily grow in totally aerobic systems.

The impact of having appropriately placed anoxic zones in a suspended growth system is to allow the proliferation of denitrifying bacteria. As discussed in Section 2.2.1, they are heterotrophic organisms that use nitrate-N and nitrite-N as electron acceptors in the absence of molecular oxygen. Denitrification can be accomplished by a large number of bacterial genera commonly found in wastewater treatment systems, including *Achromobacter*, *Aerobacter*, *Alcaligenes*, *Bacillus*, *Flavobacterium*, *Micrococcus*, *Proteus*, and *Pseudomonas*,[52] thereby making the establishment of a denitrifying culture relatively easy. However, there is uncertainty concerning the fraction of the heterotrophic bacteria in a biological nutrient removal system that can denitrify,[24] and it may well depend on the nature of the microorganisms entering the system in the wastewater,[25] as well as on the system configuration. Nevertheless, it is evident that the introduction of anoxic zones in suspended growth bioreactors will give a competitive advantage to denitrifying bacteria over heterotrophs that do not denitrify.

As described in Section 2.4.6, the placement of an anaerobic zone at the influent end of an otherwise aerobic suspended growth system establishes the conditions required for proliferation of phosphate accumulating organisms, thereby allowing development of a biomass that is rich in phosphorus. Although bacteria of the genus *Acinetobacter* were originally thought to be the major PAOs,[7] several other bacterial types have also been found to be capable of storing polyphosphate.[34,79] In fact, in one study, *Acinetobacter* was not the predominant PAO present; rather it was an unidentified gram-positive bacteria.[79] Identification of phosphate-accumulating bacteria in wastewater treatment systems is not an easy task because of the complicated growth environment required for the formation of polyphosphate granules. Nevertheless, through the development and application of new techniques, we can expect to learn more in the future about the microbial ecology of these important communities.

The previous discussion has indicated the various types of organisms that can be present in suspended growth bioreactor. However, it is very important to recognize that the types that are present in any given system will depend on the reactor configuration and the biochemical environment imposed. In later chapters we will see how these conditions, which are under engineering control, can be used to select the type of microbial community required to accomplish a specific objective.

Attached Growth Bioreactors. Attached growth bioreactors are those in which the microorganisms grow as a biofilm on a solid support. In a fluidized bed bioreactor (FBBR), the biofilm grows on small particles of sand or activated carbon that are maintained in a fluidized state by the force of water flowing upward. Packed bed bioreactors contain similar support particles, but the water being treated flows over them without displacing them. Thus, in both bioreactor types, the biofilm is surrounded by the fluid containing the substrate being removed. In a trickling filter (TF) or rotating biological contactor (RBC), on the other hand, the biofilm grows on a large surface over which the wastewater flows in a thin film (TF) or moves through

the wastewater (RBC). As a consequence, the fluid shear associated with the latter two is less than that associated with the first two. This has an impact on the type of microbial community involved.

Because FBBRs and packed beds are relatively new, few studies have been done to characterize the microbial communities involved. However, we would expect them to be very similar to those in suspended growth bioreactors, being comprised primarily of bacteria and protozoa. In contrast, TFs and RBC's contain more diverse microbial communities containing many other Eucarya, notably nematodes, rotifers, snails, sludge worms, and larvae of certain insects.[10] This more complex food chain allows more complete oxidation of organic matter, with the net result that less excess biomass is produced. This has the beneficial effect of decreasing the mass of solid material that must be disposed of.

The Bacteria form the base of the food chain by acting on the organic matter in the wastewater being treated. Soluble materials are taken up rapidly, while colloidal-sized particles become entrapped in the gelatinous layer built up by the bacteria to form the biofilm. There they undergo attack by extracellular enzymes, releasing small molecules that can be metabolized. The bacterial community is composed of primary and secondary saprophytes, much like suspended growth bioreactors, including members of the genera *Achromobacterium*, *Alcaligenes*, *Flavobacterium*, *Pseudomonas*, *Sphaerotilus*, and *Zooglea*.[7] Unlike suspended growth cultures, however, the species distribution is likely to change with position in the reactor. Attached growth reactors can also contain nitrifying bacteria, such as members of the genera *Nitrosomonas* and *Nitrobacter*, which tend to be found in regions of the film where the organic substrate concentration is low.

Quite extensive communities of Eucarya are known to exist in trickling filters.[7,10,11] Over 90 species of fungi have been reported, and of these, more than 20 species are considered to be permanent members of the community. Their role is similar to that of the bacteria, i.e., saprophytic. Many protozoa have also been found, with large communities of Sarcodina, Mastigophora, and Ciliata being reported. Their roles are largely those of predators. During warm summer months algae can flourish on the upper surfaces of the biomass. Usually green algae and diatoms predominate. Finally, trickling filters also contain a large metazoan community, consisting of annelid worms, insect larvae, and snails. These feed on the microbial film and in some cases have been responsible for extensive film destruction.

Because of the diverse nature of the microbial community in attached growth bioreactors, the microbial interactions are extremely complex. Unfortunately, even less is known about the impact of these interactions on system performance than is known about them in suspended growth systems.

2.3.2 Anaerobic Operations

The microbial communities in anaerobic operations are primarily procaryotic, with members of both the Bacteria and the Archaea being involved. Although fungi and protozoa have been observed under some circumstances, the importance of eucaryotic organisms is questionable.[5] Thus, the emphasis here will be on the complex and important interactions between the Bacteria and the Archaea that are fundamental to the successful functioning of methanogenic communities. Because those interac-

tions occur in both suspended and attached growth systems, no distinctions will be made between the two.

General Nature of Anaerobic Operations. The multistep nature of anaerobic biochemical operations is depicted in Figure 2.3. Before insoluble organic materials can be consumed, they must be solubilized, just as was necessary in aerobic systems. Furthermore, large soluble organic molecules must be reduced in size to facilitate transport across the cell membrane. The reactions responsible for solubilization and size reduction are usually hydrolytic and are catalyzed by extracellular enzymes produced by bacteria. They are all grouped together as hydrolysis reactions (reaction 1) in Figure 2.3, but in reality many enzymes are involved, such as cellulases, amylases, and proteases. They are produced by the fermentative bacteria that are an important component of the second step, acidogenesis.

Acidogenesis is carried out by members of the domain Bacteria. Amino acids and sugars are degraded by fermentative reactions (reaction 2) in which organic compounds serve as both electron donors and acceptors. The principal products of reaction 2 are intermediary degradative products like propionic and butyric acids and the direct methane precursors, acetic acid and H_2. The H_2 production from fermentative reactions is small and originates from the dehydrogenation of pyruvate by mechanisms that are different from the production of the bulk of the H_2 produced.[18] In contrast, most of the H_2 produced comes from oxidation of volatile and long chain fatty acids to acetic acid (reactions 3 and 4) and arises from the transfer of electrons from reduced carriers directly to hydrogen ions, in a process called anaerobic oxidation.[18] Because of the thermodynamics of this reaction, it is inhibited by high partial pressures of H_2, whereas the production of H_2 from pyruvate is not.

The production of H_2 by anaerobic oxidation is very important to the proper functioning of anaerobic processes. First, H_2 is one of the primary substrates from which methane is formed. Second, if no H_2 were formed, acidogenesis would not result in the oxidized product acetic acid being the major soluble organic product. Rather, the only reactions that could occur would be fermentative, in which electrons released during the oxidation of one organic compound are passed to another organic compound that serves as the electron acceptor, yielding a mixture of oxidized and reduced organic products. Consequently, the energy level of the soluble organic matter would not be changed significantly because all of the electrons originally present would still be in solution in organic form. When H_2 is formed as the reduced product, however, it can escape from the liquid phase because it is a gas, thereby causing a reduction in the energy content of the liquid. In actuality, the H_2 does not escape. It is used as a substrate for methane production, but because methane is removed as a gas, the same thing is accomplished. Finally, if H_2 formation did not occur and reduced organic products were formed, they would accumulate in the liquid because they cannot be used as substrates for methane production. Only acetic acid, H_2, methanol, and methylamines can be used. As shown by reaction 5, some of the H_2 can be combined with carbon dioxide by H_2-oxidizing acetogens to form acetic acid,[86] but since the acetic acid can serve as a substrate for methanogens, the impact of this reaction is thought to be small.

The products of the acidogenic reactions, acetic acid and H_2, are used by methanogens, which are members of the domain Archaea, to produce methane gas. Two groups are involved: (1) aceticlastic methanogens, which split acetic acid into methane and carbon dioxide (reaction 6), and (2) H_2-oxidizing methanogens, which reduce

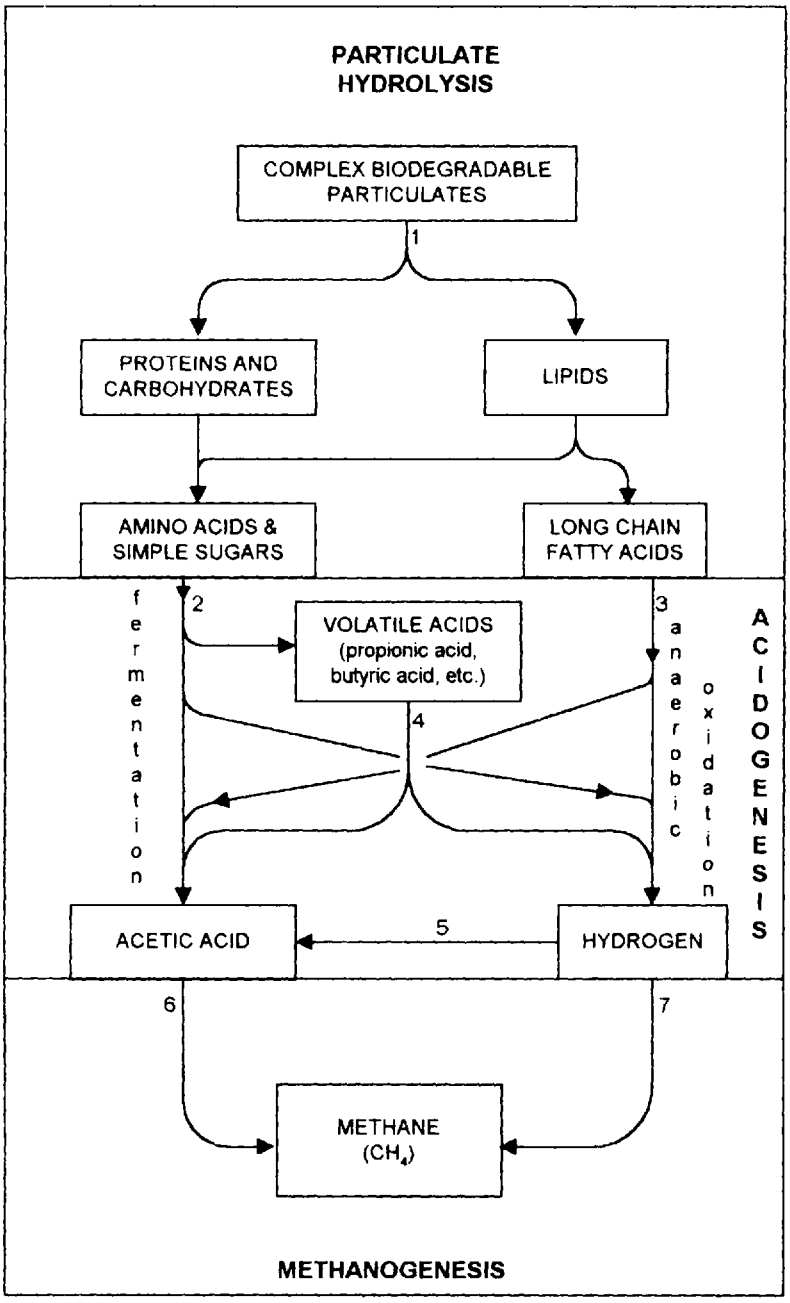


Figure 2.3 Multistep nature of anaerobic operations.

carbon dioxide (reaction 7). It is generally accepted that about two-thirds of the methane produced in anaerobic digestion of primary sludge is derived from acetic acid, with the remainder coming from H_2 and carbon dioxide.[18,86] With the exception of the electrons incorporated into the cell material formed, almost all of the energy removed from the liquid being treated is recovered in the methane. Chemical oxygen demand (COD),[67] a common measure of pollutant strength, is a measure of the electrons available in an organic compound, expressed in terms of the amount of oxygen required to accept them when the compound is completely oxidized to carbon dioxide and water. One mole of methane requires two moles of oxygen to oxidize it to carbon dioxide and water. Consequently, each 16 grams of methane produced and lost to the atmosphere corresponds to the removal of 64 grams of COD from the liquid.[42] At standard temperature and pressure, this corresponds to 0.34 m^3 of methane for each kg of COD stabilized.[43]

Microbial Groups and Their Interactions. The hydrolytic and fermentative bacteria comprise a rather diverse group of facultative and obligately anaerobic Bacteria. Although facultative bacteria were originally thought to be dominant, evidence now indicates that the opposite is true,[33] at least in sewage sludge digesters where the numbers of obligate anaerobes have been found to be over 100 times greater. This does not mean that facultative bacteria are unimportant, because their relative numbers can increase when the influent contains large numbers of them,[29] or when the bioreactor is subjected to shock loads of easily fermentable substrates.[38] Nevertheless, it does appear that most important hydrolytic and fermentative reactions are performed by strict anaerobes, such as *Bacteroides*, *Clostridia*, and *Bifidobacteria*,[66] although the nature of the substrate will determine the species present.

As mentioned previously, the role of H_2 as an electron sink is central to the production of acetic acid as the major end product of acidogenesis. Reactions leading from long chain fatty acids, volatile acids, amino acids, and carbohydrates to acetic acid and H_2 are thermodynamically unfavorable under standard conditions, having positive standard free energies.[86] Thus, when the H_2 partial pressure is high, these reactions will not proceed and instead, fermentations occur, with the results discussed above. Under conditions in which the partial pressure of H_2 is 10^{-4} atmospheres or less, however, the reactions are favorable and can proceed, leading to end products (acetic acid and H_2) that can be converted to methane. This means that the bacteria that produce H_2 are obligately linked to the methanogens that use it. Only when the methanogens continually remove H_2 by forming methane will the H_2 partial pressure be kept low enough to allow production of acetic acid and H_2 as the end products of acidogenesis. Likewise, methanogens are obligately linked to the bacteria performing acidogenesis because the latter produce the substrates required by the former. Such a relationship between two microbial groups is called obligate syntrophy.

While the organisms responsible for the fermentative reactions are reasonably well characterized, less is known about the H_2-producing acetogenic bacteria. This is due in part to the fact that the enzyme system for H_2 production is under very strict control by H_2.[68] As a consequence, early studies which attempted to enumerate the H_2-forming bacteria underestimated them by allowing H_2 to accumulate during testing. However, because H_2 partial pressures are kept low in anaerobic biochemical operations,[71] H_2-forming bacteria play an important role, and thus they have been the subject of more intensive research in recent years. Several species have been

identified and studied, including members of the genus *Syntrophomonas*, which oxidize fatty acids, and the genus *Syntrophobacter*, which oxidize propionate.[56]

As mentioned previously, the major nuisance organisms in anaerobic operations are the sulfate-reducing bacteria, which can be a problem when the wastewater contains significant concentrations of sulfate. Sulfate-reducing bacteria are all obligate anaerobes of the domain Bacteria. They are morphologically diverse, but share the common characteristic of being able to use sulfate as an electron acceptor. Group I sulfate reducers can use a diverse array of organic compounds as their electron donor, oxidizing them to acetate and reducing sulfate to sulfide. A common genus found in anaerobic biochemical operations is *Desulfovibrio*. Group II sulfate reducers specialize in the oxidation of fatty acids, particularly acetate, to carbon dioxide, while reducing sulfate to sulfide. An important genus in this group is *Desulfobacter*.

The H_2-oxidizing methanogens are classified into three orders within the domain Archaea: (1) *Methanobacteriales*, (2) *Methanococcales*, and (3) *Methanomicrobiales*.[4] A wide variety of these microorganisms have been cultured from anaerobic digesters, including the genera *Methanobrevibacter* and *Methanobacterium* from the first order, and the genera *Methanospirillum* and *Methanogenium* from the third.[57] They are all strictly obligate anaerobes which obtain their energy primarily from the oxidation of H_2 and their carbon from carbon dioxide. Because of this autotrophic mode of life, the amount of cell material synthesized per unit of H_2 used is low. During their metabolism they also use carbon dioxide as the terminal electron acceptor,[20] forming methane gas in the process.

$$4H_2 + CO_2 \rightarrow CH_4 + 2H_2O \tag{2.1}$$

Their range of electron donors is very restricted, usually being limited to H_2 and formate.[20] In some cases, short chain alcohols can also be used.[4]

In spite of the importance of the aceticlastic route to methane (reaction 6), fewer aceticlastic methanogens have been cultured and identified. All are of the order *Methanosarcinales*, which contains two families, *Methanosarcinaceae* and *Methanosaetaceae*.[4] *Methanosarcina*, of the first family, can be cultivated from anaerobic operations[57] and is among the most versatile genera of methanogens known, being able to use H_2 and carbon dioxide, methanol, methylamines, and acetic acid as substrates.[20,58] When acetic acid is the substrate, it is cleaved, with all of the methyl carbon ending up as methane and all of the carboxyl carbon as carbon dioxide.

$$^*CH_3COOH \rightarrow {}^*CH_4 + CO_2 \tag{2.2}$$

Methanosarcina grows relatively rapidly at high acetic acid concentrations, although it is very sensitive to changes in that concentration. Furthermore, H_2 exerts a regulatory effect on acetic acid utilization, shutting it down as the H_2 partial pressure increases. The family *Methanosaetaceae* contains a single genus, *Methanosaeta* (formerly *Methanothrix*), the members of which can use only acetic acid as their electron and carbon donor.[4] They grow much more slowly than *Methanosarcina* at high acetic acid concentrations, but are not influenced as strongly by that concentration and can compete effectively when it is low. As a consequence, the manner in which an anaerobic operation is designed and operated will determine the predominant aceticlastic methanogen.

2.3.3 The Complexity of Microbial Communities: Reality Versus Perception

It is apparent from the preceding that the microbial communities in biochemical operations are very complex, involving many trophic levels and many genera and species within a trophic level. Unfortunately, most studies on community structure have been descriptive and the exact roles of many organisms have not even been defined, much less quantified. As a consequence, wastewater treatment engineers have tended to view the communities in biochemical operations as if they were monocultures consisting only of procaryotes of a single species. This is slowly changing, but the models used by engineers still reflect only the procaryotic portion of the community, and its divisions are usually limited to major groups, such as aerobic heterotrophs, floc-formers, denitrifiers, nitrifiers, PAOs, etc. In the chapters to follow, we will be exploring the performance of biochemical operations based on these divisions. While the resulting mathematical descriptions are adequate for establishing a fundamental understanding of system performance, and indeed, even for design, it is important to remember the complex nature of the microbial communities involved, and to temper acceptance of the models accordingly. As engineers and microbiologists continue to work together to understand these fascinating systems, we will eventually be able to consider community structure in a quantitative way, resulting in better system design and performance.

2.4 IMPORTANT PROCESSES IN BIOCHEMICAL OPERATIONS

Regardless of the nature and complexity of the microbial community involved, there are certain fundamental processes that occur universally in biochemical operations. The relative importance of these processes, and hence the outcome from a biochemical operation, depends on the physical configuration of the operation and the manner in which it is operated. Our ability to select and design the appropriate biochemical operation for a specific task depends on our recognition of the importance of the various processes in it and our capability for quantitatively expressing the rates of those processes. In this section we will introduce those processes in qualitative terms; in Chapter 3 we will describe them quantitatively.

2.4.1 Biomass Growth, Substrate Utilization, and Yield

When reduced to their barest essentials, biochemical operations are systems in which microorganisms are allowed to grow by using pollutants as their carbon and/or energy source, thereby removing the pollutants from the wastewater and converting them to new biomass and carbon dioxide, or other innocuous forms. Because of the role of enzymes in microbial metabolism, the carbon and/or energy source for microbial growth is often called the substrate, causing wastewater treatment engineers to commonly refer to the removal of pollutants during biomass growth as substrate utilization. If growth is balanced, which is the case for most (but not all) biochemical operations, biomass growth and substrate utilization are coupled, with the result that the removal of one unit of substrate results in the production of Y units of biomass,

where Y is called the true growth yield, or simply the yield.[b] Because of the coupling between biomass growth and substrate utilization, the rates of the two activities are proportional, with Y as the proportionality factor. Consequently, the selection of one as the primary event (or cause) and the other as the secondary event (or effect) is arbitrary. Both selections are equally correct and benchmark papers have been published using both substrate removal[36] and biomass growth[27] as the primary event. The point of view taken in this book is that biomass growth is the fundamental event, and the rate expressions presented in Chapter 3 are written in terms of it. It should be emphasized, however, that rate expressions for biomass growth and substrate utilization can be interconverted through use of the yield, Y.

Because of the central role that Y plays in the relationship between biomass growth and substrate utilization, it is an intrinsic characteristic. Consequently, a clear understanding of the factors that can influence its magnitude is important. The development of such an understanding requires consideration of the energetics of microbial growth, including energy conservation and energy requirements for synthesis.

Overview of Energetics. Microorganisms require four things for growth: (1) carbon, (2) inorganic nutrients, (3) energy, and (4) reducing power. As mentioned in Section 2.2.1, microorganisms derive energy and reducing power from oxidation reactions, which involve the removal of electrons from the substrate with their ultimate transfer to the terminal electron acceptor. Consequently, the energy available in a substrate depends on its oxidation state, which is indicative of the electrons available for removal as the substrate is oxidized. Highly reduced compounds contain more electrons, and have a higher standard free energy, than do highly oxidized compounds, regardless of whether they are organic or inorganic. As described in Chapter 1, most biochemical operations are used for the removal of soluble organic matter and the stabilization of insoluble organic matter. Consequently, in this discussion we will focus on carbon oxidation by heterotrophic bacteria. Since COD is a measure of available electrons, compounds with a high COD:C ratio are highly reduced, whereas those with a low COD:C ratio are more oxidized. The carbon in methane is in the most highly reduced state possible, with a COD:C ratio of 5.33 mg COD/mg C, whereas the carbon in carbon dioxide is in the most highly oxidized state with a COD:C ratio of zero. Thus, all organic compounds will have a COD:C ratio between these extremes.

As heterotrophic bacteria oxidize the carbon in organic compounds through their catabolic pathways, they convert them to metabolic intermediates of the central amphibolic pathways that are in a higher oxidation state than either the starting compound or the biomass itself. Those metabolic intermediates are used in the anabolic pathways for cell synthesis, but since they are in a higher oxidation state than the cell material being synthesized from them, electrons must be available in an appropriate form for reducing them. Those electrons arise from the original substrate during its catabolism and are transferred to the anabolic pathways through the use of carriers such as nicotinamide adenine dinucleotide (NAD) and nicotinamide adenine dinucleotide phosphate (NADP), which alternate between the oxidized (NAD

[b]Throughout this book, the term "yield" will be considered to be synonymous with "true growth yield."

and NADP) and the reduced (NADH and NADPH) state. Thus NAD and NADP serve as electron acceptors for catabolic reactions, forming NADH and NADPH, which act as electron donors for biosynthetic reactions. The availability of NADH and NADPH is called reducing power.

Biosynthetic reactions also require energy in a form that can be used in coupled reactions to join the amphibolic intermediates into new compounds. That energy is provided primarily by adenosine triphosphate (ATP), and to a lesser degree by other nucleotides. ATP is generated by phosphorylation reactions from adenosine diphosphate (ADP) and when the ATP is used to provide energy in biosynthetic reactions, ADP is released for reuse. ATP can be formed from ADP by two types of phosphorylation reactions: substrate level and electron transport phosphorylation. During substrate level phosphorylation, ATP is formed directly by coupled reactions within a catabolic pathway. Only small amounts of ATP can be generated in this way. Much larger amounts can be generated during electron transport phosphorylation, which occurs as electrons removed during oxidation of the substrate (and carried in NADH) are passed through the electron transport (or terminal respiratory) chain, to the terminal electron acceptor, setting up a proton-motive force.[5] The magnitude of the proton motive force, and consequently, the amount of ATP that can be generated, depends on both the organism and the nature of the terminal electron acceptor.

An important concept to recognize about microbial energetics is that as a compound is degraded, all of the electrons originally in it must end up in the new cell material formed, in the terminal electron acceptor, or in the soluble organic metabolic intermediates excreted during growth. If a compound is mineralized, the amount of metabolic intermediates will be very small, so that essentially all electrons must end up either in the cell material formed or in the terminal acceptor. Because the yield is the amount of cell material formed per unit of substrate destroyed; because the amount of cell material formed depends on the amount of ATP generated; and because the amount of ATP generated depends on the electrons available in the substrate, the organism carrying out the degradation, and the growth environment, it follows that the yield also depends on the nature of the substrate, the organism involved, and the growth environment.

Effects of Growth Environment on ATP Generation. The electron transport chains found in most Bacteria and Eucarya share common features. They are highly organized and are localized within membranes. They contain flavoproteins and cytochromes which accept electrons from a donor like NADH and pass them in discrete steps to a terminal acceptor. All conserve some of the energy released by coupling the electron transfer to the generation of proton motive force, which drives a number of processes, such as the synthesis of ATP from ADP and inorganic phosphate, active transport, and flagellar movement. The electron transport chain in Eucarya is located in the mitochondria and is remarkably uniform from species to species. The electron transport chain in Bacteria is located in the cytoplasmic membrane and exhibits considerable variety among individual species in the identity of the individual components and in the presence or absence of sections of the chain. Nevertheless, the sequential organization of the components of the electron transport chain is determined by their standard oxidation–reduction potentials. Table 2.3 presents the potentials for the array of couples found in mitochondrial electron transport chains.[20] The couples in Bacteria are similar, but not necessarily identical. The transfer is in the direction of increasing redox potential until the final reaction with the terminal

Table 2.3 The Standard Oxidation–Reduction Potentials of a Number of Redox Couples of Interest in Biological Systems

Redox couple	E_0' (mV)
$H_2 = 2H^+ + 2e$	−420
Ferredoxin red./oxid.	−410
NADPH/$NADP^+$	−324
NADH/NAD^+	−320
Flavoproteins red./oxid.	−300 to 0
Cyt. b red./oxid.	+ 30
Ubiquinone red./oxid.	+100
Cyt. c red./oxid.	+254
Cyt. a_3 red./oxid.	+385
$O_2 + 2H^+ = H_2O_2 + 2e$	+820

From Ref. 20.

acceptor is catalyzed by the appropriate enzyme. When the environment is aerobic, oxygen serves as the terminal acceptor and the enzyme is an oxidase.

ATP generation is associated with the transfer of electrons down the electron transport chain through electron transport phosphorylation, although it is not directly coupled to specific biochemical reactions that occur during that transfer.[3,5] Rather, the generation of ATP is driven by the proton motive force through chemiosmosis. The elements of the electron transport chain are spatially organized in the cytoplasmic membrane of Bacteria and the mitochondrial membrane of Eucarya in such a way that protons (hydrogen ions) are translocated across the membrane as the electrons move down the electron transport chain, i.e., toward more positive E_0' values. In Bacteria the transfer is from the cytoplasm (inside the cell) to the periplasmic space (outside the cell); in Eucarya, from inside the mitochondria to outside. The transfer of electrons across the membrane establishes a proton gradient which causes a diffusive counterflow of protons back across the membrane through proton channels established by a membrane-bound ATPase enzyme. This proton counterflow drives the synthesis of ATP from ADP and inorganic phosphate. The number of ATP synthesized per electron transferred to the terminal acceptor depends on the nature and spatial organization of the electron transport chain because they determine the number of protons that are translocated per electron transferred down the chain. In mitochondria, 3 ATP can be synthesized per pair of electrons transferred. However, in Bacteria the number will depend on the organization of the electron transport chain in the particular organism involved. This explains why the amount of ATP synthesized from the oxidation of a given substrate depends on the organism performing the oxidation.

In the absence of molecular oxygen, other terminal acceptors may accept electrons from the electron transport chain, and the redox potentials for them, as well as for various donors, are given in Table 2.4.[20] In order for ATP to be generated by electron transport phosphorylation, the oxidation–reduction potential for the donor redox couple must be smaller (more negative) than the potential for the acceptor redox couple, there must be at least one site of proton translocation in the electron

Table 2.4 The Standard Oxidation–Reduction Potentials of Various Acceptor and Donor Redox Couples

Redox couple	E_0' (mV)
Acceptor	
$\frac{1}{2}O_2/H_2O$	+820
NO_3^-/NO_2^-	+433
NO_2^-/NO	+350
Fumarate/succinate	+ 33
SO_4^{2-}/SO_3^{2-}	− 60
CO_2/CH_4	−244
Donor	
$H_2/2H^+$	−420
$HCOOH/HCO_3^-$	−416
$NADH/NAD^+$	−320
Lactate/pyruvate	−197
Malate/oxaloacetate	−172
Succinate/fumarate	+ 33

From Ref. 20.

transport chain between the final acceptor and the point where the donor contributes its electrons, and the associated free energy change ($\Delta G^{0\prime}$) must exceed 44 kJ [$\Delta G^{0\prime} = -2F \cdot \Delta E_0'$, where F = 96.6 kJ/(V·mol)]. Nitrate and nitrite are important terminal electron acceptors in biochemical operations performing denitrification and the bacteria capable of using the nitrogen oxides as electron acceptors are biochemically and taxonomically diverse.[35] The enzyme nitrate reductase is responsible for the conversion of nitrate to nitrite. It is membrane bound and couples with the electron transport chain through a specific cytochrome b. The enzymes nitrite reductase, nitric oxide reductase, and nitrous oxide reductase are involved in the reduction of nitrite to nitrogen gas and appear to be linked to the electron transport chain through specific c-type cytochromes.[20,35] It is possible that all of the reactions are coupled to the generation of proton motive force, but the number of ATPs synthesized per electron transported is less than the number associated with oxygen as the terminal acceptor because the available free energy change is less. Consequently, bacteria growing with nitrate as the terminal electron acceptor exhibit lower yields than bacteria growing under aerobic conditions.[46]

Under strictly anaerobic conditions, i.e., when neither oxygen nor the nitrogen oxides are present, many Bacteria generate their ATP through substrate level phosphorylation associated with fermentation reactions in which the oxidation of one organic substrate is coupled to the reduction of another. The second substrate is generally a product of the catabolic pathway leading from the oxidized substrate with the result that the fermentation pathway is internally balanced, with neither a net production nor a net requirement for reducing power. Several types of fermentation reactions are listed in Table 2.5. Because ATP generation occurs only by substrate level phosphorylation and a large part of the available electrons in the original substrate end up in the reduced organic products, bacteria receive relatively little energy in this mode of growth, and thus have low yield per unit of substrate processed. As

Table 2.5 Types of Fermentations of Various Microorganisms

Type of fermentation	Products	Organisms
Alcoholic	Ethanol, CO_2	Yeast
Lactic acid	Lactic acid	*Streptococcus*, *Lactobacillus*
Mixed acid	Lactic acid, acetic acid, Ethanol, CO_2, H_2	*Escherichia*, *Salmonella*
Butanediol	Butanediol, ethanol, lactic acid, acetic acid, CO_2, H_2	*Aerobacter*, *Serratia*
Butyric acid	Butyric acid, acetic acid, CO_2, H_2	*Clostridium butyricum*
Acetone-butanol	Acetone, butanol, ethanol	*Clostridium acetobutylicum*
Propionic acid	Propionic acid	*Propionibacterium*

discussed in Section 2.3.2, however, the production of H_2 allows more oxidized products like acetate to be produced. As a result, more ATP can be produced by bacteria when they generate H_2, allowing them to have a higher biomass yield per unit of substrate processed.

Methanogens are obligate anaerobes that have very restricted nutritional requirements, with the oxidation of acetate and hydrogen being their main sources of energy. Even though methane is produced from the reduction of carbon dioxide during the oxidation of H_2, methanogens lack the components of a standard electron transport chain, and thus carbon dioxide does not function as a terminal electron acceptor in a manner analogous to nitrate or oxygen.[20] Rather, reduction of carbon dioxide to methane involves a complex sequence of events requiring a number of unique coenzymes.[84] However, there is a sufficient free energy change during methane formation for the theoretical production of two molecules of ATP and it appears that a normal chemiosmotic mechanism is involved,[20] although it involves a sodium motive force as well as a proton motive force.[84] Regardless of the exact mechanisms involved, it is important to recognize that ATP generation in Archaea is different from that associated with both respiration and fermentation in Bacteria and Eucarya. Furthermore, like bacteria growing in anaerobic environments, methanogens have low yields.

Factors Influencing Energy for Synthesis. Energy for synthesis represents the energy required by microorganisms to synthesize new cell material. In the absence of any other energy requirements, the energy required for synthesis is the difference between the energy available in the original substrate and the energy associated with the cell material formed, or in the common units of the environmental engineer, the difference between the COD of the original substrate and the COD of the biomass formed. Consequently, the energy for synthesis and the yield are intimately linked. If the efficiency of ATP generation were the same for all bacteria, it would be possible to theoretically predict the energy for synthesis, and hence the yield, from thermodynamic considerations.[41] However, as we saw above, the amount of ATP generated per electron transferred differs from microorganism to microorganism, which means

that the efficiency of energy generation differs. This, coupled with the fact that the pathways of synthesis and degradation are not the same in all microorganisms, makes it difficult to use exactly the thermodynamic approaches for predicting yields that have been presented in the environmental engineering literature. Nevertheless, there are many instances in which it would be advantageous to have a theoretical prediction of the energy for synthesis or the yield prior to experimental work and a technique based on the Gibbs energy dissipation per unit of biomass produced appears to be best.[23] Regardless, thermodynamic concepts are most useful for understanding why different substrates and different terminal electron acceptors have different energies of synthesis and yields associated with them.

During biomass growth, energy is required to synthesize the monomers needed to make the macromolecules that form the structural and functional components of the cell. This suggests that more energy would be required for a culture to grow in a minimal medium containing only a single organic compound as the carbon and energy source than in a complex medium in which all required monomers were supplied. Actually, such a conclusion is false.[69] For example, the energy needed to synthesize all of the amino acids needed by a cell amounts to only about 10% of the total energy needed to synthesize new cell material. This is because macromolecules are too large to be transported into the cell and must be formed inside even when all of the needed monomers are provided in the medium. Consequently, although the complexity of the growth medium has some effect on the energy required for synthesis, it is not large.

Of more importance are the oxidation state and size of the carbon source.[23] The oxidation state of carbon in biomass is roughly the same as that of carbon in carbohydrate.[69] If the carbon source is more oxidized than that, reducing power must be expended to reduce it to the proper level. If the carbon source is more reduced, it will be oxidized to the proper level during normal biodegradation and no extra energy will be required. Therefore, as a general rule, a carbon source at an oxidation state higher than that of carbohydrate will require more energy to be converted into biomass than will one at a lower oxidation state. Pyruvic acid occupies a unique position in metabolism because it lies at the end of many catabolic pathways and the beginning of many anabolic and amphibolic ones. As such, it provides carbon atoms in a form that can be easily incorporated into other molecules. Indeed, three-carbon fragments play an important role in the synthesis of many compounds. If the carbon source contains more than three carbon atoms it will be broken down to size without the expenditure of large amounts of energy. If it contains less than three carbon atoms, however, energy must be expended to form three-carbon fragments for incorporation. Consequently, substrates containing few carbon atoms require more energy for synthesis than do large ones.

Carbon dioxide, which is used by autotrophic organisms as their chief carbon source, is an extreme example of the factors just discussed, being a single-carbon compound in which the carbon is in the highest oxidation state. Consequently, the energy for synthesis for autotrophic growth is much higher than for heterotrophic growth. As a result, the amount of biomass that can be formed per unit of available electrons in the energy source is quite low.

True Growth Yield. The true growth yield (Y) is defined as the amount of biomass formed per unit of substrate removed when all energy expenditure is for synthesis. In this context, the substrate is usually taken to be the electron donor,

although it can be defined differently. If the electron donor is an organic compound, it is common in environmental engineering practice to express Y in terms of the amount of soluble COD removed from the wastewater. This is because wastewaters contain undefined, heterogenous mixtures of organic compounds and the COD is an easily determined measurement of their quantity. In addition, the COD is fundamentally related to available electrons, having an electron equivalent of eight grams of oxygen. Thus, a Y value expressed per gram of COD removed can be converted to a Y value per available electron by multiplying by eight. If the electron donor is an inorganic compound, such as ammonia or nitrite nitrogen, it is common to express Y in terms of the mass of the element donating the electrons. Furthermore, regardless of the nature of the electron donor, it has been common practice to express the amount of biomass formed on a dry weight basis, i.e., mass of suspended solids (SS), or on the basis of the dry weight of ash-free organic matter, i.e., mass of volatile suspended solids (VSS). When grown on a soluble substrate, microorganisms have an ash content of about 15%, and thus the value of Y when expressed as VSS will be slightly less than the value of Y when expressed as suspended solids. As will be discussed later, there are advantages to expressing biomass concentrations on a COD basis rather than on a SS or VSS basis, and thus yields are sometimes expressed as the amount of biomass COD formed per unit of substrate COD removed from the medium. This convention will be used throughout this book. If we assume an empirical formula for the organic, i.e., ash-free, portion of biomass of $C_5H_7O_2N$, the COD of that organic portion can be calculated to be 1.42 g COD/g VSS.[30] Furthermore, if we assume the ash content of biomass to be 15%, the theoretical COD of biomass is 1.20 g COD/g SS. These values can be used to convert between the various ways of expressing the yield.

The nature of the substrate influences the yield. Hadjipetrou et al.[19] summarized data from one species, *Aerobacter aerogenes*, which was grown in unrestricted batch growth in minimal media on a number of substrates, and found Y to vary from 0.40 to 0.56 mg biomass COD formed per mg substrate COD removed. (The values were not reported on a COD basis, but were converted to it for this book.) Recognizing that the yield expressed on the basis of cell COD formed per unit of substrate COD removed is a measure of the amount of energy available in the substrate that was conserved through cell synthesis, it can be seen that 40 to 56% of the available energy was conserved while 44 to 60% was expended.

The species of organism will also affect Y, although the effect will not be as great as the effect of substrate. Payne[56] collected Y values for eight bacterial species growing aerobically on glucose in minimal media and found them to vary from 0.43 to 0.59 mg biomass COD formed per mg substrate COD removed. The data were from a number of different published reports and thus some of the variation may be due to differences in experimental conditions, rather than to species. Nevertheless, they clearly show that the microbial species has an impact. (As above, the values were not originally reported on a COD basis, but were converted to it for this book.)

The growth environment, including media complexity, type of terminal electron acceptor, pH, and temperature will all affect Y.[23] As explained above, biomass grown in complex media will have only slightly higher Y values than biomass grown in minimal media, whereas biomass grown with oxygen as the terminal electron acceptor will exhibit significantly higher yields than biomass grown with nitrate as the acceptor. The yield from fermentations will depend on the reduced end products and

the method of expressing the yield. If Y is expressed on the basis of the amount of the original substrate removed, ignoring the COD returned to the medium as reduced end products, the value will be very small, on the order of 0.03 to 0.04 mg biomass COD formed per mg substrate COD removed. However, when expressed on the basis of the COD actually utilized (accounting for the COD remaining as reduced end products), the Y value is not much different from that obtained with aerobic cultures.[1] On the other hand, when methane is produced, so that most of the reduced end product is lost from the system as a gas, then the COD removed from solution is actually much higher than the COD utilized by the microorganisms, making the yield per unit of COD removed about an order of magnitude lower than for aerobic growth. The pH of the medium has long been known to affect microbial growth, but the quantitative effects are unclear. The yield is likely, however, to have a maximum around pH 7 because that is optimal for so many physiological functions. Temperature also affects Y, as shown in Figure 2.4.[49] Although the significance of temperature is apparent, no generalizations can be made, and most engineers assume that Y is constant over the normal physiological temperature range. A final factor that

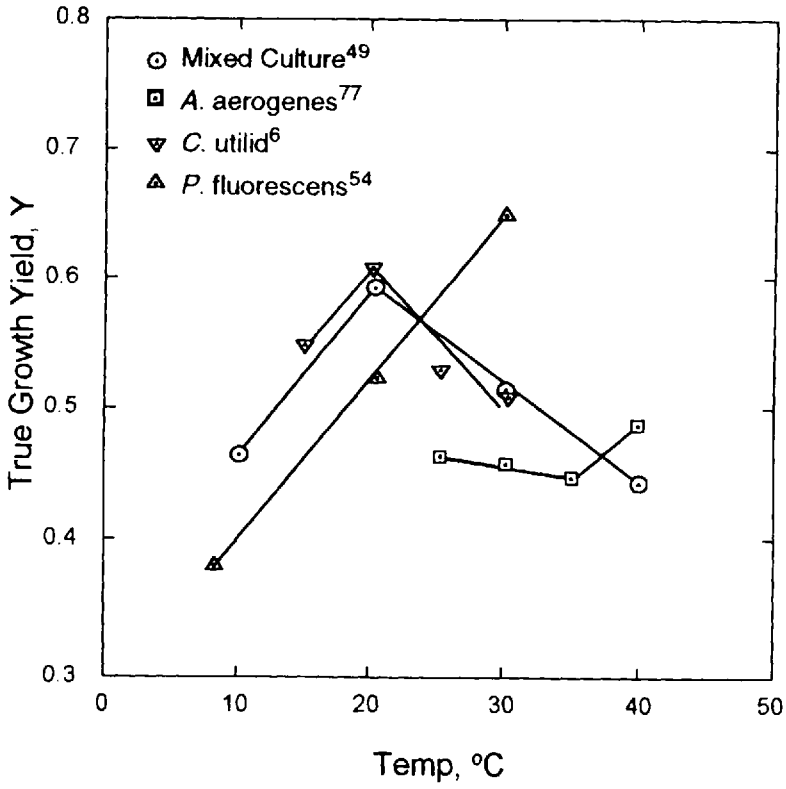


Figure 2.4 Effect of temperature on the true growth yield, Y. The units on Y are mg biomass formed per mg substrate COD removed. (From R. E. Muck and C. P. L. Grady, Jr., Temperature effects on microbial growth in CSTR's. *Journal of the Environmental Engineering Division, ASCE* **100**:1147–1163, 1974. Reprinted by permission of the American Society of Civil Engineers.).

may influence Y is the composition of the microbial community. When it is heterogeneous, the waste products from one species serve as growth factors for another, thereby converting a seemingly minimal medium into a complex one. Consequently, it might be anticipated that the yields from mixed microbial cultures would be slightly higher than those from pure cultures growing on the same medium. A comparison of the two revealed this to be the case.[28]

Constancy of Y *in Biochemical Operations.* Biochemical operations use mixed microbial communities to treat wastewaters containing mixtures of substrates. Thus, it is apparent that Y will depend on both the character of the wastewater and the particular community that develops on it. It is important that this variability be recognized by engineers designing biochemical operations, because then the estimated yield values will be interpreted in an appropriate way. As seen in Chapter 3, similar conclusions can be reached about the kinetic parameters associated with biochemical operations. This means that designers must utilize considerable judgement and allow for uncertainty. This situation does not prevent generalities from being made, however. For example, examination of a large number of yield values indicates that Y will generally lie within the range of 0.48–0.72 mg biomass COD formed per mg substrate COD utilized for aerobic heterotrophs degrading carbohydrates.[63] Under similar conditions, Y values for growth on a number of xenobiotic compounds, including substituted phenols, benzenes, and phthalate esters, lay within the range of 0.20–0.60 mg biomass COD formed per mg substrate COD removed.[10] One study[80] reported the range of yield values for autotrophs to be from 0.06 to 0.35 mg biomass COD per mg nitrogen oxidized, with values for *Nitrobacter* being lower than those for *Nitrosomonas*. Likewise, another study[85] reported the Y value for *Nitrobacter* to be 0.12 mg biomass COD per mg nitrogen oxidized and the value for *Nitrosomonas* to be 0.47. Although ranges such as these provide the engineer with an idea of the magnitudes to be expected, designs should only be based on estimates of Y obtained from laboratory and pilot-scale studies of the particular waste to be treated.

2.4.2 Maintenance, Endogenous Metabolism, Decay, Lysis, and Death

The yield values in the preceding section are those that result when all energy obtained by the biomass is being channeled into synthesis. Energy for synthesis is not the only energy requirement for microorganisms, however. They must also have energy for maintenance.[60]

Cellular processes, whether mechanical or chemical, require energy for their performance, and unless a supply is available these essential processes will cease and the cell will become disorganized and die. Mechanical processes include motility, osmotic regulation, molecular transport, maintenance of ionic gradients, and in the case of some Eucarya, cytoplasmic streaming. While it might be argued that motility can be dispensed with in some microorganisms, this argument would not hold for all because some require motility to find food. Osmotic regulation is quite important in all cells, even those protected by a rigid cell wall, and pump mechanisms, such as contractile vacuoles, exist in cells to counteract the normal tendency of osmotic pressure to pump water into them. Cell membranes are permeable to many small molecules, such as amino acids, and because of the high concentrations within the

cell these tend to diffuse into the medium. Active transport mechanisms operate to bring such molecules into the cell against the concentration gradient. Of a similar nature is the necessity for maintaining an ionic gradient across the cell membrane, which is closely linked to the proton motive force responsible for ATP synthesis. Maintenance of this gradient is thought to be a major consumer of maintenance energy.[73] Finally, cytoplasmic streaming and the movement of materials within Eucarya are often required for their proper functioning. They also require energy.

Chemical factors also contribute to maintenance energy needs. Microbial cells represent chemical organization and many of the components within them have higher free energies than the original compounds from which they were formed. In general, because of this organization, energy must be available to counteract the normal tendency toward disorder, i.e., to overcome entropy. The chemical processes contributing to the energy requirement for maintenance are those involved in resynthesis of structures such as the cell wall, flagella, the cell membrane, and the catabolic apparatus. For example, one study[39] suggested that energy for the resynthesis of proteins and nucleic acids was an important portion of the maintenance energy requirement for *Escherichia coli.*

A major point of controversy in the microbiological literature has concerned the impact on the maintenance energy requirement of the rate at which a culture is growing. Early investigations[60] suggested that the need for maintenance energy was independent of growth rate, but more recent research indicates the opposite.[73] Nevertheless, engineers generally consider maintenance energy needs to be independent of growth rate in biochemical operations for wastewater treatment and that is the approach that will be adopted in this book.

Given the existence of a need for maintenance energy, what energy sources can be used to supply it? The answer to that question depends on the growth conditions of the microorganisms. If an external (exogenous) energy supply is available, a portion of it will be used to meet the maintenance energy requirement and the remainder will be used for synthesis. As the rate of energy supply is decreased, less and less will be available for new growth and thus the net, or observed, yield will decline. When the point is reached at which the rate of energy supply just balances the rate at which energy must be used for maintenance, no net growth will occur because all available energy will be used to maintain the status quo. If the rate of energy supply is reduced still further, the difference between the supply rate and the maintenance energy requirement will be met by the degradation of energy sources available within the cell, i.e., by endogenous metabolism. This will cause a decline in the mass of the culture. Finally, if no exogenous energy source is available, all of the maintenance energy needs must be met by endogenous metabolism. When the point is reached at which all endogenous reserves have been exhausted, the cells deteriorate and die, or enter a resting state.

The nature of the materials serving as substrates for endogenous metabolism depends on both the species of the microorganism and the conditions under which the culture was grown. For example, when *E. coli* is grown rapidly in a glucose–mineral salts medium it stores glycogen.[39] If those cells are then placed in an environment devoid of exogenous substrate they will utilize the glycogen as an endogenous energy source. Amino acids and proteins show little net catabolism until the glycogen is gone. When grown in tryptone medium, on the other hand, *E. coli* accumulates little glycogen. As a result, endogenous metabolism utilizes nitrogenous

compounds immediately. Other organisms use still other compounds, including ribonucleic acid (RNA) and the lipid poly-β-hydroxybutyrate (PHB).

One question that has intrigued microbiologists concerns the route of energy flow when sufficient exogenous substrate is available to supply the maintenance energy requirements of the culture. Does endogenous metabolism continue under those circumstances so that part of the energy released from degradation of the substrate is used to resupply the energy reserves being degraded by endogenous metabolism? Or, alternatively, does endogenous metabolism cease so that the energy released from degradation of the exogenous substrate goes directly for maintenance functions? The evidence is still not conclusive. Actually, although such questions are of fundamental scientific significance, they have little bearing on the macroscopic energy balances used by engineers to mathematically model biochemical operations. In fact, as we see in Section 3.3.2, some models avoid the entire issue by introducing the concept of cell lysis and regrowth.

The amount of biomass actually formed per unit of substrate used in a biochemical operation, referred to as the observed yield (Y_{obs}), is always less than Y. One reason for this is the need for maintenance energy. The more energy that must be expended for maintenance purposes, the less available for synthesis and the smaller the quantity of biomass formed per unit of substrate degraded. Other factors also contribute to the difference, however. For example, consider the effect of predation. In a complex microbial community such as that found in the activated sludge process, protozoa and other Eucarya prey on the bacteria, reducing the net amount of biomass formed. To illustrate the effect of predation, assume that the value of Y for bacteria growing on glucose is 0.60 mg bacterial biomass COD formed per mg of glucose COD used. Thus, if 100 mg/L of glucose COD were used, 60 mg/L of bacterial biomass COD would result. Now assume that the value of Y for protozoa feeding on bacteria is 0.70 mg protozoan biomass COD formed per mg of bacterial biomass COD used. If the protozoa consumed all of the bacteria resulting from the glucose, the result would be 42 mg/L of protozoan biomass. As a consequence, if we observed only the net amount of biomass formed, without distinction as to what it was, we would conclude that 42 mg/L of biomass COD resulted from the destruction of 100 mg/L of glucose COD. Therefore, we would conclude that the observed yield was 0.42, which is less than the true growth yield for bacteria growing on glucose. Macroscopically, it is impossible to distinguish between the various factors acting to make the observed yield less than the true growth yield. Consequently, environmental engineers lump them together under the term "microbial decay," which is the most common way they have modeled their effect in biochemical operations.[36]

Another process leading to a loss of biomass in biochemical operations is cell lysis.[1] The growth of bacteria requires coordination of the biosynthesis and degradation of cell wall material to allow the cell to expand and divide. The enzymes responsible for hydrolysis of the cell wall are called autolysins and their activity is normally under tight regulation to allow them to act in concert with biosynthetic enzymes during cell division. Loss of that regulation, however, will lead to rupture of the cell wall (lysis) and death of the organism. When the cell wall is ruptured, the cytoplasm and other internal constituents are released to the medium where they become substrates for other organisms growing in the culture. In addition, the cell wall and cell membranes, as well as other structural units, begin to be acted upon

by hydrolytic enzymes in the medium, solubilizing them and making them available as substrates as well. Only the most complex units remain as cell debris, which is solubilized so slowly that it appears to be refractory in most biochemical operations.[45,47] The arguments for how lysis results in the loss of biomass are similar to those associated with predation, illustrated above. The yield exhibited by bacteria growing on the soluble products released by lysis is of the same magnitude as the yield associated with growth on other biogenic substrates. Consequently, if 100 mg/L of biomass is lysed, only 50–60 mg/L of new biomass will result from regrowth on the lysis products. Thus, the net effect of lysis and regrowth is a reduction in biomass within the system. In general, starvation itself does not initiate lysis, although the events that trigger it are not yet clear. Nevertheless, engineers seeking to model the decline in observed yield associated with situations in which the microbial community is growing slowly have focused on cell lysis as the primary mechanism.[12,26]

The final event impacting on the amount of active biomass in a biochemical operation is death. Traditionally, a dead cell has been defined as one that has lost the ability to divide on an agar plate[62] and studies based on this definition have shown that a large proportion of the microorganisms in slowly growing cultures are nonviable, or dead.[61,62,74] In addition, as summarized by Weddle and Jenkins,[80] a large number of studies using indirect evidence involving comparisons of substrate removal rates and enzyme activities have concluded that large portions of the MLSS in wastewater treatment systems are inactive. However, a later study,[40,41] using more sophisticated techniques for identifying dead bacteria, has suggested that a very low fraction of the cells present at low growth rates are actually dead. Instead, many are simply nonculturable by standard techniques, although they are still alive. Furthermore, the more recent work[40] suggests that dead cells do not remain intact for long, but rather lyse, leading to substrates and biomass debris, as discussed above. The presence of biomass debris acts to make the mass of viable microorganisms less than the mass of suspended solids in the system. Even though the predecessor of this book used a model[15] that explicitly considered cell death, it now appears that direct consideration of the phenomenon is not warranted.[40,41] Rather, the fact that only a portion of the MLSS in a biological wastewater treatment system is actually viable biomass can be attributed to the accumulation of biomass debris rather than to the presence of dead cells.

In summary, as a result of several mechanisms, biochemical reactors exhibit two important characteristics: (1) the observed yield is less than the true growth yield and (2) active, viable bacteria make up only a fraction of the "biomass." One simplified conceptualization of the events leading to these characteristics is that bacteria are continually undergoing death and lysis, releasing organic matter to the environment in which they are growing. Part of that organic matter is degraded very, very slowly, making it appear to be resistant to biodegradation and causing it to accumulate as biomass debris. As a consequence, only a portion of the "biomass" is actually viable cells. The remainder of the released organic matter is used by the bacteria as a food source, resulting in new biomass synthesis. However, because the true growth yield is always less than one, the amount of new biomass produced is less than the amount destroyed by lysis, thereby making the observed yield for the overall process less than the true growth yield on the original substrate alone.

2.4.3 Soluble Microbial Product Formation

Much of the soluble organic matter in the effluent from a biological reactor is of microbial origin and is produced by the microorganisms as they degrade the organic substrate in the influent to the bioreactor. The major evidence for this phenomenon has come from experiments in which single soluble substrates of known composition were fed to microbial cultures and the resulting organic compounds in the effluent were examined for the presence of the influent substrate.[64] The bulk of the effluent organic matter was not the original substrate and was of higher molecular weight, suggesting that it was of microbial origin. These soluble microbial products are thought to arise from two processes, one growth-associated and the other non–growth-associated. Growth-associated product formation results directly from biomass growth and substrate utilization. As such, it is coupled to those events through another yield factor, the microbial product yield, Y_{SMP}, and the biodegradation of one unit of substrate results in the production of Y_{SMP} units of products. Values of Y_{SMP} for a variety of organic compounds have been found to be less than 0.1.[16] Non–growth-associated product formation is related to decay and lysis and results in biomass-associated products. They are thought to arise from the release of soluble cellular constituents through lysis and the solubilization of particulate cellular components. Although little is known about the characteristics of these two types of soluble microbial products, they are thought to be biodegradable, although some at a very low rate. Compared to other aspects of biochemical operations, little research has been done on the production and fate of soluble microbial products and few researchers have attempted to model the contribution of such products to the organic matter discharged from wastewater treatment systems.[64,50] Nevertheless, an awareness of their existence is necessary for an accurate understanding of the response of biochemical operations.

2.4.4 Solubilization of Particulate and High Molecular Weight Organic Matter

Bacteria can only take up and degrade soluble organic matter of low molecular weight. All other organic material must be attacked by extracellular enzymes that release low molecular weight compounds that can be transported across cellular membranes. Many organic polymers, particularly those of microbial origin, such as cell wall components, proteins, and nucleic acids, are composed of a few repeating subunits connected by bonds that can be broken by hydrolysis. Consequently, the microbial process of breaking particulate and high molecular weight soluble organic compounds into their subunits is commonly referred to as hydrolysis, even though some of the reactions involved may be more complicated.

Hydrolysis reactions play two important roles in biochemical reactors for wastewater treatment. First, they are responsible for the solubilization of cellular components released as a result of cell lysis, preventing their buildup in the system. Because cell lysis occurs in all microbial systems, hydrolysis reactions are even important in bioreactors receiving only soluble substrate. Second, many biochemical operations receive particulate organic material, in which case hydrolysis is essential to bring about the desired biodegradation. In spite of its central position in the functioning of biochemical operations, relatively few studies have sought to under-

stand the kinetics and mechanisms of hydrolysis.[8,12,17] Nevertheless, it has important impacts on the outcome of biochemical operations and must be considered for a complete understanding of their functioning.

2.4.5 Ammonification

Ammonification is the name given to the release of ammonia nitrogen as amino acids and other nitrogen containing organic compounds undergo biodegradation. It occurs as a normal result of the biodegradation process, during which amino groups are liberated and excreted from the cell as ammonia. The rate of ammonification will depend on the rate of nitrogen containing substrate utilization and the carbon to nitrogen ratio of that substrate. Ammonification is very important in wastewater treatment processes for nitrogen control because organic nitrogen is not subject to oxidation by nitrifying bacteria. They can only oxidize nitrogen to nitrate after it has been converted to ammonia and released to the medium.

2.4.6 Phosphorus Uptake and Release

If a suspended growth bioreactor system is configured as two zones in series with the first zone anaerobic and the second aerobic, PAOs, which possess a special metabolic capability not commonly found in other bacteria, will proliferate and store large quantities of inorganic phosphate as polyphosphate, thereby allowing phosphorus removal from the wastewater via biomass wastage. Although PAOs are often present in significant numbers in totally aerobic suspended growth cultures, they only develop the ability to store large quantities of phosphate when they are subjected to alternating anaerobic and aerobic conditions by being recycled between the two zones.[37] This follows from their unique capability to store carbon at the expense of phosphate under anaerobic conditions and to store phosphate at the expense of carbon under aerobic conditions. Two scenarios have been postulated to explain the functioning of PAOs. One was developed independently by Comeau et al.[9] and Wentzel et al.,[82] whereas the other was developed by Arun et al.[2] The former is referred to as the Comeau–Wentzel model whereas the latter is called the Mino model.[81] The difference between the two models is the result of the metabolic diversity among PAOs, and since it is not yet known which model is the more generally applicable, both will be presented.

Comeau–Wentzel Model. We will first consider the events occurring in the anaerobic zone. Because of fermentations that occur in sewers, much of the soluble organic matter in domestic wastewater is in the form of acetate and other short chain fatty acids. Furthermore, when the wastewater enters an anaerobic bioreactor, additional quantities of fatty acids are formed by fermentative reactions performed by facultative heterotrophs. As indicated in Figure 2.5A, acetate is transported across the cell membrane by passive diffusion (as undissociated acetic acid), but once inside, it is activated to acetyl-CoA by coupled ATP hydrolysis, yielding ADP. Although not shown in the diagram, ATP is also used to maintain the proton motive force that has been lost by transport of the proton associated with the undissociated acetic acid. The cell responds to the decreasing ATP/ADP ratio by stimulating ATP resynthesis from stored polyphosphate (Poly-P_n). A portion of the acetyl-CoA is metabolized through the TCA cycle to provide the reducing power ($NADH + H^+$) required for

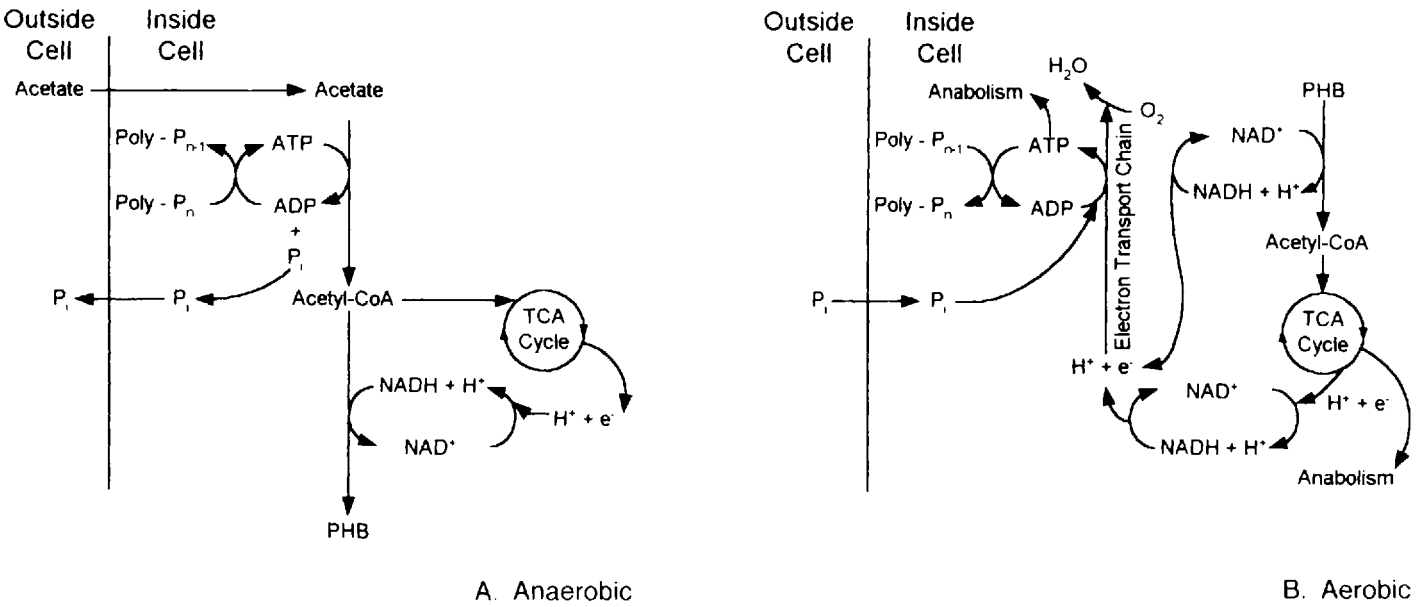


Figure 2.5 Schematic diagram depicting the Comeau–Wentzel model for the uptake and release of inorganic phosphate by PAOs: A. Anaerobic conditions; B. aerobic conditions. (Adapted from Wentzel et al.[51])

the synthesis of PHB. The remainder of the acetyl-CoA is converted into PHB, with about 90% of the acetate carbon being conserved in that storage polymer. Without the presence of the polyphosphate to provide energy for ATP resynthesis, acetate would build up in the cell, acetate transport would stop, and no PHB formation would occur. The hydrolysis of the polyphosphate to form ATP increases the intracellular concentration of inorganic phosphate, P_i, which is released to the bulk solution, along with cations (not shown) to maintain charge balance.

When the wastewater and the associated biomass enter the aerobic zone, the wastewater is low in soluble organic matter, but the PAOs contain large PHB reserves. Furthermore, the wastewater is rich in inorganic phosphate, while the PAOs have low polyphosphate levels. Because they have oxygen as an electron acceptor in the aerobic zone, the PAOs perform normal aerobic metabolism for growth by using the stored PHB as their carbon and energy source, generating ATP through electron transport phosphorylation, as illustrated in Figure 2.5B. Furthermore, as the ATP–ADP ratio increases, polyphosphate synthesis is stimulated, thereby removing phosphate and associated cations (not shown) from solution and regenerating the stored polyphosphate in the cells. Because of the large amount of energy provided by the aerobic metabolism of the stored PHB, the PAOs are able to take up all of the phosphate released in the anaerobic zone plus the phosphate originally present in the wastewater.

The continual cycling between the anaerobic and aerobic zones gives the PAOs a competitive advantage over ordinary heterotrophic bacteria, because without the capability to make and use polyphosphate, the ordinary heterotrophs are not able to take up organic matter in the anaerobic zone. It should be noted that while most systems that remove phosphate through the use of PAOs employ aerobic zones for the regeneration of the stored polyphosphate, some PAOs can use nitrate and nitrite as alternative electron acceptors,[47] allowing anoxic conditions to be used as well.

Mino Model. The Mino model, illustrated in Figure 2.6, is very similar to the Comeau–Wentzel model, the major difference being the role of glycogen, a carbohydrate storage polymer. In this case, in the anaerobic zone the reducing power required for synthesis of PHB from acetyl-CoA comes from the metabolism of glucose released from the glycogen. Glucose is oxidized to pyruvate through the Entner–Doudoroff (ED) or Embden–Meyerhof–Parnas (EMP) pathway, depending in the type of PAO, thereby providing some of the ATP required to convert acetate to acetyl-CoA and some of the reducing power needed for PHB synthesis. Pyruvate, in turn, is oxidatively decarboxylated to acetyl-CoA and carbon dioxide, with the electrons released also being used in the synthesis of PHB. Thus, all of the acetate taken up is stored as PHB, as is part of the carbon from the glycogen. In the aerobic zone, PHB is broken down as in the Comeau–Wentzel model to provide for biomass synthesis as well as for phosphate uptake and storage as polyphosphate. In addition, however, PHB is also used to replenish the stored glycogen.

2.4.7 Overview

A diagram depicting the overall sum of the events occurring in an aerobic bioreactor receiving a soluble substrate is shown in Figure 2.7. Bacteria consume the substrate (S_{S1}) and grow, leading to more bacteria, with the relationship between substrate consumption and biomass growth being given by the true growth yield, Y. There

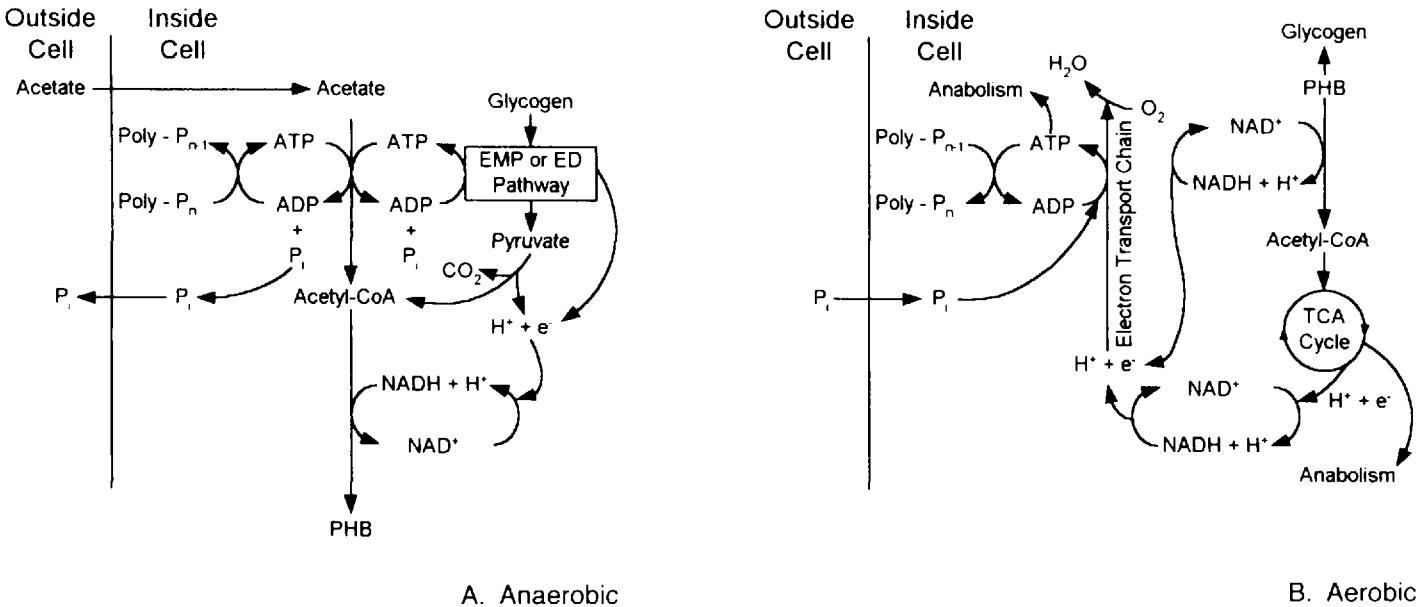


Figure 2.6 Schematic diagram depicting the Mino model for the uptake and release of inorganic phosphate by PAOs: A. Anaerobic conditions; B. aerobic conditions. (Adapted from Wentzel et al.[51])

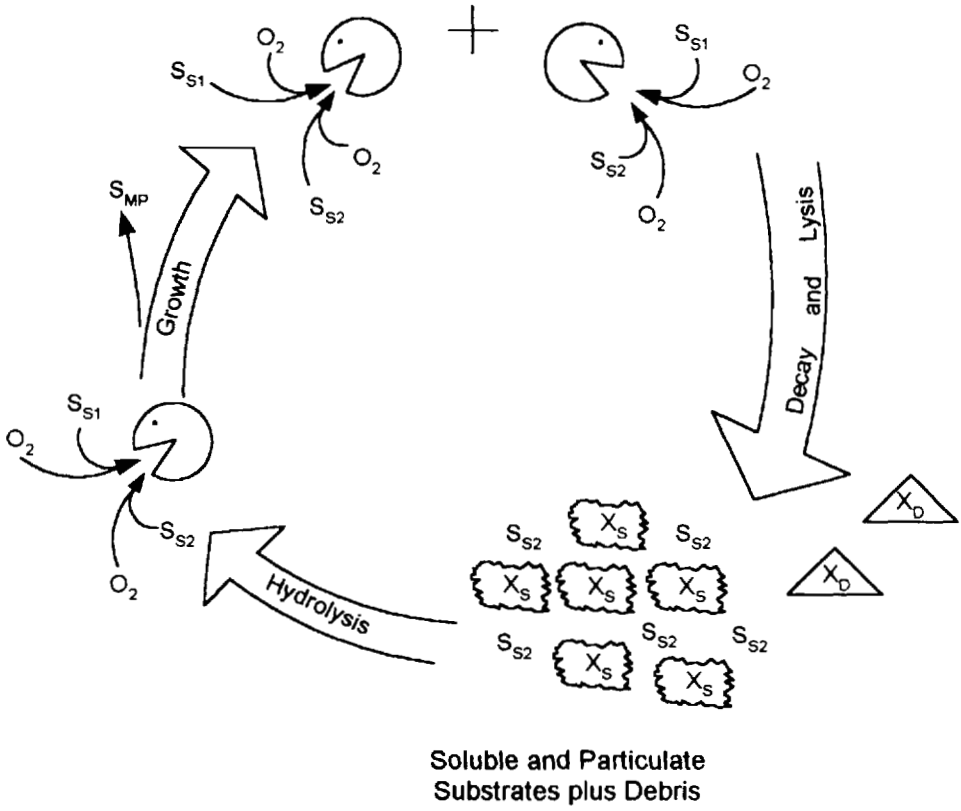


Figure 2.7 Overview of fundamental events occurring in an aerobic bioreactor receiving a soluble substrate (S_{S1}). (Adapted from Mason et al.[40])

will also be soluble microbial product (S_{MP}) formation associated with that substrate consumption and growth. Concurrently with growth, the biomass will be undergoing decay and lysis, releasing soluble (S_{S2}) and particulate (X_S) substrate to the medium. Cell debris (X_D), which is degraded so slowly that it appears to be nonbiodegradable, and biomass associated products (S_D) are also released. The particulate cell fragments (X_S) undergo hydrolysis, freeing more substrate (S_{S2}) that can be used by the cells. Part of the microbial products may undergo biodegradation, but others may be degraded so slowly that they appear inert. As might be imagined by the previous discussion in this section, more complicated conceptualizations could be depicted. However, this one contains the essential elements required to model biological processes and it will be used in later chapters for that purpose.

2.5 KEY POINTS

1. Biochemical operations use the carbon and nitrogen cycles to remove organic and nitrogenous pollutants from wastewaters.

2. The microorganisms in biochemical operations can be classified in several ways. Among the most important are: the type of electron donor used, the type of electron acceptor employed, their physical growth characteristics, and their function.
3. The microorganisms in aerobic/anoxic suspended growth bioreactors may be divided into five overlapping groups: (1) floc-forming organisms, (2) saprophytes, (3) nitrifying bacteria, (4) predators, and (5) nuisance organisms.
4. Attached growth bioreactors have more diverse microbial communities encompassing more trophic levels than suspended growth bioreactors.
5. Methanogenic anaerobic cultures are highly interdependent ecosystems with many complex interactions between Bacteria and Archaea. Acetic acid and H_2 play a central role in those interactions, being products of Bacteria and substrates for Archaea.
6. There are two major groups of methanogens: (1) those that oxidize H_2 and (2) those that cleave acetic acid. Both are essential to the proper functioning of anaerobic cultures receiving complex substrates.
7. In most situations, biomass growth and substrate utilization are coupled, with the true growth yield, Y, serving as the coupling factor. The yield is the amount of biomass formed per unit of substrate removed. Its value depends on the nature of the substrate, the organism involved, and the growth environment.
8. Heterotrophic bacteria obtain their energy from the oxidation of organic carbon. Hence, chemical oxygen demand (COD), which is a measure of available electrons, is a convenient way in which to express the concentration of organic matter in wastewaters. When an organic compound is mineralized, all of the electrons available in it must end up either in the biomass formed or in the terminal electron acceptor. Consequently, COD is also a convenient technique for expressing the concentration of biomass.
9. Yield values for heterotrophic biomass cover a very broad range, but seldom exceed 0.75 mg biomass COD formed per mg substrate COD removed because of the energy required for synthesis.
10. As a result of maintenance energy needs and decay, death, and lysis, biochemical reactors exhibit two characteristics: (1) the observed yield is less than the true growth yield, and (2) active, viable bacteria make up only a fraction of the "biomass."
11. Soluble microbial product formation is associated with substrate utilization and with biomass decay and lysis. As a consequence, much of the soluble organic matter leaving a biochemical operation is of microbial origin.
12. Hydrolysis reactions are important for the biodegradation of particulate substrates and cellular components released by biomass death and lysis.
13. Ammonification is the release of ammonia nitrogen as nitrogen containing organic compounds undergo biodegradation.
14. PAOs will only store large amounts of phosphorus as polyphosphate granules when they are cycled between substrate-rich anaerobic and substrate-poor aerobic environments.

2.6 STUDY QUESTIONS

1. Draw a sketch of the nitrogen cycle, labeling all reactions. Then explain the following terms and their importance in biochemical operations: ammonification, assimilation, nitrification, denitrification, and assimilative reduction.
2. Define or explain the following terms and their use in classifying the microorganisms in biochemical operations: electron donor, electron acceptor, heterotroph, autotroph, nitrifier, denitrifier, methanogen, obligate aerobe, obligate anaerobe, facultative anaerobe, biofloc, primary degrader, and secondary degrader.
3. Describe the roles of and give examples of microorganisms in each of the following groups commonly found in aerobic/anoxic suspended growth bioreactors: floc-forming organisms, saprophytes, nitrifying bacteria, predators, and nuisance organisms.
4. Draw a sketch depicting the multistep nature of methanogenic anaerobic cultures and use it to describe the roles of the major groups of microorganisms involved.
5. Why is the maintenance of a low partial pressure of H_2 necessary to the proper functioning of a methanogenic anaerobic culture? What is the role of methanogens in the maintenance of the required conditions?
6. There are two major groups of methanogens. Describe them, list their growth characteristics, and contrast their roles in anaerobic cultures.
7. Why does the value of the true growth yield, Y, depend on the nature of the substrate, the microorganism involved, and the growth environment?
8. Why is it convenient to express the concentrations of organic substrates and biomass in COD units?
9. Give a "typical" yield value for heterotrophic biomass growing on carbohydrates and then explain why there is considerable variability associated with Y in biochemical operations.
10. Explain why the observed yield in a biochemical reactor is less than the true growth yield. While so doing, explain what is meant by the term "decay."
11. Why does cell lysis in a biochemical operation make the observed yield less than the true growth yield and the viability less than 100%?
12. What is the difference between growth-associated and nongrowth-associated product formation?
13. Why are hydrolysis reactions important to the performance of all biochemical operations, even those receiving only soluble substrate?
14. Describe the scenarios that have been postulated to explain the functioning of phosphate accumulating bacteria.

REFERENCES

1. Andrews, J. F. and E. A. Pearson. Kinetics and characteristics of volatile acid production in anaerobic fermentation processes. *International Journal for Air and Water Pollution Research* **9**, 439–461, 1965.

2. Arun, V., T. Mino, and T. Matsuo. Biological mechanisms of acetate uptake mediated by carbohydrate consumption in excess phosphorus removal systems. *Water Research* **22**, 565–570, 1988.
3. Atlas, R. M. *Microbiology Fundamentals and Applications*. Macmillan, New York, NY, 1984.
4. Boone, D. R., Whitman, W. B., and P. Rouvière. Diversity and taxonomy of methanogens. In *Methanogenesis: Ecology, Physiology, Biochemistry & Genetics*, J. G. Ferry, ed., Chapman & Hall, New York, NY, pp. 35–80, 1993.
5. Brock, T. D., M. T. Madigan, J. M. Martinko, and J. Parker. *Biology of Microorganisms*. Seventh Edition, Prentice Hall, Englewood Cliffs, NJ, 1994.
6. Brown, C. M. and A. H. Rose. Effects of temperature on composition and cell volume of *Candida utilis*. *Journal of Bacteriology* **97**, 261–272, 1969.
7. Bruce, A. M. and H. A. Hawkes. Biological filters. In *Ecological Aspects of Used-Water Treatment*, Vol. 3, C. R. Curds and H. A. Hawkes, eds., Academic Press, New York, NY, pp. 1–111, 1983.
8. Bryers, J. D. and C. A. Mason. Biopolymer particulate turnover in biological waste treatment systems: a review. *Bioprocess Engineering* **2**, 95–109, 1987.
9. Comeau, Y., K. J. Hall, R. E. W. Hancock and W. K. Oldham. Biochemical model for enhanced biological phosphorus removal. *Water Research* **20**, 1511–1521, 1986.
10. Cooke, W. B. Trickling filter ecology. *Ecology* **40**, 273–291, 1959.
11. Curds, C. R. Protozoa. In *Ecological Aspects of Used-Water Treatment*, Vol. 1, C. R. Curds and H. A. Hawkes, eds., Academic Press, New York, NY, pp. 203–268, 1975.
12. Dold, P. L., G. A. Ekama and G. v. R. Marais. A general model for the activated sludge process. *Progress in Water Technology* **12(6)**, 47–77, 1980.
13. Eikelboom, D. H. Filamentous organisms observed in bulking activated sludge. *Water Research* **9**, 365–388, 1975.
14. Focht, D. D. and A. C. Chang. Nitrification and denitrification processes related to waste water treatment. *Advances in Applied Microbiology* **19**, 153–186, 1975.
15. Grady, C. P. L. Jr., and R. E. Roper Jr. A model for the bio-oxidation process which incorporates the viability concept. *Water Research* **8**, 471–483, 1974.
16. Grady, C. P. L. Jr., G. Aichinger, S. F. Cooper and M. Naziruddin. Biodegradation kinetics for selected toxic/hazardous organic compounds. In *Proceedings of the 1989 AWMA/EPA International Symposium on Hazardous Waste Treatment: Biosystems for Pollution Control*. Air and Waste Management Association, Pittsburgh, PA, pp. 141–153, 1989.
17. Gujer, W. The effect of particulate organic material on activated sludge yield and oxygen requirement. *Progress in Water Technology* **12(6)**, 79–95, 1980.
18. Gujer, W. and A. J. B. Zehnder. Conversion processes in anaerobic digestion. *Water Science and Technology* **15(8/9)**, 127–167, 1983.
19. Hadjipetrou, L. P., J. P. Gerrits, F. A. G. Teulings, and A. H. Stouthamer. Relation between energy production and growth of *Aerobacter aerogenes*. *Journal of General Microbiology* **36**, 139–150, 1964.
20. Hamilton, W. A. Microbial energetics and metabolism. In *Micro-Organisms in Action: Concepts and Applications in Microbial Ecology*, J. M. Lynch and J. E. Hobbie, eds., Blackwell Scientific Publications, Palo Alto, CA, pp. 75–100, 1988.
21. Hao, O. J., P. F. Strom and Y. C. Wu. A review of the role of *Nocardia*-like filaments in activated sludge foaming. *Water SA* **14**, 105–110, 1988.
22. Hawkes, H. A. The applied significance of ecological studies of aerobic processes. In *Ecological Aspects of Used-Water Treatment*, Vol. 3, C. R. Curds and H. A. Hawkes, eds., Academic Press, New York, NY, pp. 173–333, 1983.
23. Heijnen, J. J. and J. P. van Dijken. In search of a thermodynamic description of biomass yields for the chemotrophic growth of microorganisms. *Biotechnology and Bioengineering* **39**, 833–858, 1992.

24. Henze, M. Nitrate versus oxygen utilization rates in wastewater and activated sludge systems. *Water Science and Technology*, **18(6)**, 115–122, 1986.
25. Henze, M. The influence of raw wastewater biomass on activated sludge oxygen respiration rates and denitrification rates. *Water Science and Technology* **21(10/11)**, 603–607, 1989.
26. Henze, M., C. P. L. Grady Jr., W. Gujer, G. v. R. Marais and T. Matsuo. A general model for single-sludge wastewater treatment systems. *Water Research* **21**, 505–515, 1987.
27. Herbert, D. A. A theoretical analysis of continuous culture systems. In *Continuous Culture of Microorganisms*, Society of Chemical Industry, London, Monograph No. 12, pp. 21–53, 1960.
28. Hettling, L. J., D. R. Washington and S. S. Rao. Kinetics of the steady-state bacterial culture. II. Variation in synthesis. *Proceedings of the 19th Industrial Waste Conference*. Purdue University Engineering Extension Series No. 117, pp. 687–715, 1964.
29. Hobson, P. N. and B. G. Shaw. The bacterial population of piggery waste anaerobic digesters. *Water Research* **8**, 507–516, 1974.
30. Hoover, S. R. and N. Porges. Assimilation of dairy wastes by activated sludge. II. The equations of synthesis and rate of oxygen utilization. *Sewage and Industrial Wastes* **24**, 306–312, 1952.
31. Jenkins, D., M. G. Richard and G. T. Daigger. *Manual on the Causes and Control of Activated Sludge Bulking and Foaming*. 2nd ed. Lewis Publishers, Chelsea, MI, 1993.
32. Kelly, D. P. Autotrophy: Concepts of lithotrophic bacteria and their organic metabolism. *Annual Review of Microbiology* **25**, 177–210, 1971.
33. Kirsch, E. J. Studies on the enumeration and isolation of obligate anaerobic bacteria from digesting sewage sludge. *Developments in Industrial Microbiology* **10**, 170–176, 1969.
34. Knight, G. C., E. M. Seviour, R. J. Seviour, J. A. Soddell, K. C. Lindrea, W. Strachan, B. De Grey, and R. C. Bayly. Development of the microbial community of a full scale biological nutrient removal activated sludge plant during start-up. *Water Research* **29**, 2085–2093, 1995.
35. Knowles, R. Denitrification. *Microbiological Reviews* **46**, 43–70, 1982.
36. Lawrence, A. W. and P. L. McCarty. Unified basis for biological treatment design and operation. *Journal of the Sanitary Engineering Division, ASCE* **96**, 757–778, 1970.
37. Lötter, L. H., M. C. Wentzel, R. E. Loewenthal, G. A. Ekama and G. v. R. Marais. A study of selected characteristics of *Acinetobacter* ssp. isolated from activated sludge in anaerobic/anoxic/aerobic and aerobic systems. *Water SA* **12**, 203–208, 1986.
38. Mah, R. A. ESE Notes. University of North Carolina, **6**, 1, 1969.
39. Marr, A. G., E. H. Nilson and D. J. Clark. The maintenance requirement of *Escherichia coli*. *Annals of the New York Academy of Science* **102**, 536–548, 1963.
40. Mason, C. A., J. D. Bryers and G. Hamer. Activity, death and lysis during microbial growth in a chemostat. *Chemical Engineering Communications* **45**, 163–176, 1986.
41. Mason, C. A., G. Hamer and J. D. Bryers. The death and lysis of microorganisms in environmental processes. *FEMS Microbiology Reviews* **89**, 373–401, 1986.
42. McCarty, P. L. The methane fermentation. In *Principles and Applications of Aquatic Microbiology*, H. Heukelekian and N. C. Dondero, eds., John Wiley and Sons, New York, NY, pp. 314–343, 1964.
43. McCarty, P. L. Anaerobic waste treatment fundamentals. *Public Works* **95(9)**, 107–112; (10), 123–126; (11), 91–94; (12), 95–99, 1964.
44. McCarty, P. L. Energetics of organic matter degradation. In *Water Pollution Microbiology*, R. Mitchell, ed., John Wiley and Sons, Inc., New York, NY, pp. 91–118, 1972.
45. McCarty, P. L. and C. F. Brodersen. Theory of extended aeration activated sludge. *Journal, Water Pollution Control Federation* **34**, 1095–1103, 1962.

46. McClintock, S. A., J. H. Sherrard, J. T. Novak and C. W. Randall. Nitrate versus oxygen respiration in the activated sludge process. *Journal, Water Pollution Control Federation* **60**, 342–350, 1988.
47. McKinney, R. E. Mathematics of complete mixing activated sludge. *Journal of the Sanitary Engineering Division, ASCE* **88(SA3)**, 87–113, 1962.
48. Mobarry, B. K., M. Wagner, V. Urbain, B. E. Rittmann, and D. A. Stahl. Phylogenetic probes for analyzing abundance and spatial organization of nitrifying bacteria. *Applied and Environmental Microbiology* **62**, 2156–2162, 1996.
49. Muck, R. E. and C. P. L. Grady Jr. Temperature effects on microbial growth in CSTR's. *Journal of the Environmental Engineering Division, ASCE* **100**, 1147–1163, 1974.
50. Orhon, D. and N. Artan. *Modeling of Activated Sludge Systems*. Technomic Publishing, Lancaster, PA, 1994.
51. Pace, N. R. New perspectives on the natural microbial world: molecular microbial ecology. *ASM News* **62**, 463–470, 1996.
52. Painter, H. A. Microbial transformation of inorganic nitrogen. *Progress in Water Technology* **8(4/5)**, 3–29, 1977.
53. Painter, H. A. Metabolism and physiology of aerobic bacteria and fungi. In *Ecological Aspects of Used-Water Treatment*, Vol. 2, C. R. Curds and H. A. Hawkes, eds., Academic Press, New York, NY, pp. 11–75, 1983.
54. Palumbo, A. and L. D. Witter. Influence of temperature on glucose utilization by *Pseudomonas fluorescens*. *Applied Microbiology* **18**, 137–141, 1969.
55. Papen, H., R. von Berg, I. Hinkel, B. Thoene and H. Rennenberg. Heterotrophic nitrification by *Alcaligenes faecalis*: NO_2^-, NO_3^-, N_2O, and NO production in exponentially growing cultures. *Applied and Environmental Microbiology* **55**, 2068–2072, 1989.
56. Payne, W. J. Energy yield and growth of heterotrophs. *Annual Review of Microbiology* **24**, 17–52, 1970.
57. Pike, E. B. Aerobic bacteria. In *Ecological Aspects of Used-Water Treatment*, Vol. 1, C. R. Curds and H. A. Hawkes, eds., Academic Press, New York, NY, pp. 1–63, 1975.
58. Pike, E. B. and C. R. Curds. The microbial ecology of the activated sludge process. In *Microbial Aspects of Pollution*, G. Sykes and F. A. Skinner, eds., Academic Press, New York, NY, pp. 123–148, 1971.
59. Pipes, W. O. The ecological approach to the study of activated sludge. *Advances in Applied Microbiology* **8**, 77–103, 1966.
60. Pirt, J. S. Maintenance energy of bacteria in growing cultures. *Proceedings of the Royal Society (London), Series B* **163**, 224–231, 1965.
61. Postgate, J. R. Viability measurements and the survival of microbes under minimum stress. *Advances in Microbial Physiology* **1**, 1–23, 1967.
62. Postgate, J. R. and J. R. Hunter. The survival of starved bacteria. *Journal of General Microbiology* **29**, 233–263, 1962.
63. Ramanathan, M. and A. F. Gaudy Jr. Studies on sludge yield in aerobic systems. *Proceedings of the 26th Industrial Waste Conference*. Purdue University Engineering Extension Series No. 140, pp. 665–675, 1971.
64. Rittmann, B. E., W. Bae, E. Namkung and C.-J. Lu. A critical evaluation of microbial products formation in biological processes. *Water Science and Technology* **19(7)**, 517–528, 1987.
65. Rosselló-Mora, R. A., M. Wagner, R. Amann, and K.-H. Schleifer. The abundance of *Zoogloea ramigera* in sewage treatment plants. *Applied and Environmental Microbiology* **61**, 702–707, 1995.
66. Sahm, H. Anaerobic wastewater treatment. *Advances in Biochemical Engineering and Biotechnology* **29**, 83–115, 1984.
67. Sawyer, C. N., P. L. McCarty, and G. F. Parkin. *Chemistry for Environmental Engineering*, 4th ed., McGraw-Hill Book Company, New York, NY, 1995.

68. Scheifinger, C. C., B. Linehan, and M. J. Wolin. H_2 production by *Selenomonas ruminantium* in the absence and presence of methanogenic bacteria. *Applied Microbiology* **29**, 480–483, 1975.
69. Senez, J. C. Some considerations on the energetics of bacterial growth. *Bacteriological Reviews* **26**, 95–107, 1962.
70. Sezgin, M., D. Jenkins and D. S. Parker. A unified theory of filamentous activated sludge bulking. *Journal, Water Pollution Control Federation* **50**, 362–381, 1978.
71. Shea, T. G., W. A. Pretorius, R. D. Cole, and E. A. Pearson. Kinetics of hydrogen assimilation in the methane fermentation. *Water Research* **2**, 833–848, 1968.
72. Stephenson, T. *Acinetobacter*: its role in biological phosphate removal. In *Biological Phosphate Removal from Wastewaters*, R. Ramadori, ed., Pergamon Press, Elmsford, New York, NY, pp. 313–316, 1987.
73. Tempest, D. W. and O. M. Neijssel. The status of Y_{ATP} and maintenance energy as biologically interpretable phenomena. *Annual Review of Microbiology* **38**, 459–486, 1984.
74. Tempest, D. W., D. Herbert and P. J. Phipps. Studies on the growth of *Aerobacter aerogenes* at low dilution rates in a chemostat. In *Microbial Physiology and Continuous Culture*, edited by E. O. Powell et al., Her Majesty's Stationery Office, London, pp. 240–253, 1967.
75. Toerien, D. F. and W. H. J. Hattingh. Anaerobic digestion. I. The microbiology of anaerobic digestion. *Water Research* **3**, 385–416, 1969.
76. Tomlinson, T. G. and I. L. Williams. Fungi. In *Ecological Aspects of Used Water Treatment*, Vol. 1, C. R. Curds and H. A. Hawkes, eds., Academic Press, New York, NY, pp. 93–152, 1975.
77. Topiwala, H. and C. G. Sinclair. Temperature relationships in continuous culture. *Biotechnology and Bioengineering* **13**, 795–813, 1971.
78. Verstraete, W. and M. Alexander. Heterotrophic nitrification in samples of natural ecosystems. *Environmental Science and Technology* **7**, 39–42, 1973.
79. Wagner, M., R. Erhart, W. Manz, R. Amann, H. Lemmer, D. Wedi, and K.-H. Schleifer. Development of an rRNA-targeted oligonucleotide probe specific for the genus *Acinetobacter* and its application for in situ monitoring in activated sludge. *Applied and Environmental Microbiology* **60**, 792–800, 1994.
80. Weddle, C. L. and D. Jenkins. The viability and activity of activated sludge. *Water Research* **5**, 621–640, 1971.
81. Wentzel, M. C., L. H. Lötter, G. A. Ekama, R. E. Loewenthal, and G. v. R. Marais. Evaluation of biochemical models for biological excess phosphorus removal. *Water Science and Technology* **23**, 567–576, 1991.
82. Wentzel, M. C., L. H. Lötter, R. E. Loewenthal and G. v. R. Marais. Metabolic behavior of *Acinetobacter* ssp. in enhanced biological phosphorus removal—A biochemical model. *Water SA* **12**, 209–224, 1986.
83. Woese, C. R., O. Kandler, and M. L. Wheelis. Towards a natural system of organisms: proposal for the domains of Archaea, Bacteria, and Eukarya. *Proceedings of the National Academy of Science USA* **87**, 4576–4579, 1990.
84. Wolfe, R. S. 1776–1996: Alessandro Volta's combustible air. *ASM News* **62**, 529–534, 1996.
85. Yoshioka, T., H. Terai and Y. Saijo. Growth kinetics studies of nitrifying bacteria by the immunofluorescent counting method. *Journal of General and Applied Microbiology* **28**, 169–180, 1982.
86. Zinder, S. H. Microbiology of anaerobic conversion of organic wastes to methane: Recent developments. *ASM News* **50**, 294–298, 1984.
87. Zinder, S. H. Physiological ecology of methanogens. In *Methanogenesis: Ecology, Physiology, Biochemistry & Genetics*, J. G. Ferry, ed., Chapman & Hall, New York, NY, pp. 128–206, 1993.

This page intentionally left blank

3

Stoichiometry and Kinetics of Biochemical Operations

Stoichiometry is concerned with the relationships between the quantities of reactants and products in chemical reactions. Kinetics is concerned with the rates at which reactions take place. Because stoichiometry quantitatively relates a change in one reactant (product) to the change in another, when the reaction rate of one reactant (product) becomes known, stoichiometry may be used to determine the reaction rate of another in the reaction. In this chapter we will first examine these relationships on a generalized basis. Then we will apply them to the major biochemical events from Chapter 2 and examine the expressions that will be used to model the theoretical performance of biochemical operations in Parts II and IV.

3.1 STOICHIOMETRY AND GENERALIZED REACTION RATE

3.1.1 Alternative Bases for Stoichiometry

Stoichiometric equations are usually derived in molar units, but they are not the most convenient units for our purposes. This is because we must write mass balance equations for the various constituents being acted upon in a biochemical operation in order to model its performance. Thus, it would be more convenient if the stoichiometric equations for the reactions were written in mass units. Consequently, we need to know how to convert a molar-based stoichiometric equation into a mass-based one. Furthermore, we see in Chapter 2 that microorganisms gain their energy from oxidation/reduction reactions in which electrons are removed from the electron donor and passed ultimately to the terminal electron acceptor. This suggests that it would also be convenient to write electron balances. Unfortunately, as we saw earlier, we usually don't know the exact composition of the electron donor in a wastewater, making this difficult to do. However, we can experimentally determine the chemical oxygen demand (COD), which is a measure of available electrons, of the various constituents. Thus, we can accomplish the same thing by writing a mass balance on COD for each of the constituents that undergo a change in oxidation state. Consequently, we also need to know how to convert molar- or mass-based stoichiometric equations into COD-based equations.

The general formula for a stoichiometric equation can be written as:[n2]

$$a_1A_1 + a_2A_2 + \cdots + a_kA_k \rightarrow a_{k+1}A_{k+1} + a_{k+2}A_{k+2} + \cdots + a_mA_m \qquad (3.1)$$

where A_1 through A_k are the reactants and a_1 through a_k are their associated molar stoichiometric coefficients, and A_{k+1} through A_m are the products and a_{k+1} through a_m are their molar stoichiometric coefficients. Two characteristics allow recognition of a stoichiometric equation as molar-based. First, the charges are balanced. Second, the total number of moles of any given element in the reactants equals the number of moles of that element in the products.

When writing a mass-based stoichiometric equation it is common practice to normalize the stoichiometric coefficients relative to one of the reactants or products. Thus, each normalized mass-based stoichiometric coefficient represents the mass of the particular reactant used or product formed relative to the mass of the reference reactant used or product formed. If A_1 is the component that we want to use as the basis for our mass-based stoichiometric equation, its stoichiometric coefficient would be 1.0 and the new mass-based stoichiometric coefficient for every other component (referred to as a normalized stoichiometric coefficient, Ψ_i) would be calculated from:

$$\Psi_i = (a_i)(MW_i)/(a_1)(MW_1) \tag{3.2}$$

where a_i and MW_i are the molar stoichiometric coefficient and molecular weight, respectively, of component A_i, and a_1 and MW_1 have the same meanings for the reference component. Thus, the equation becomes:

$$A_1 + \Psi_2 A_2 + \cdots + \Psi_k A_k \rightarrow \Psi_{k+1} A_{k+1} + \Psi_{k+2} A_{k+2} + \cdots + \Psi_m A_m \tag{3.3}$$

Two characteristics can be used to identify this type of stoichiometric equation: (1) the charges do not appear to be balanced, and (2) the total mass of reactants equals the total mass of products. In other words, the sum of the stoichiometric coefficients for the reactants equals the sum of the stoichiometric coefficients for the products. The latter characteristic makes a mass-based stoichiometric equation well-suited for use in mass balance equations for biochemical reactors.

A similar approach can be used to write the stoichiometric equation in terms of compounds or components that change oxidation state by taking advantage of COD units.[3] In this case, the normalized stoichiometric coefficients are referred to as COD-based coefficients and are given the symbol Y. The COD-based coefficient, Y_i, for component A_i would be calculated from:

$$Y_i = (a_i)(MW_i)(COD_i)/(a_1)(MW_1)(COD_1) \tag{3.4}$$

$$Y_i = \Psi_i(COD_i)/(COD_1) \tag{3.5}$$

where COD_i and COD_1 are the COD per unit mass of component A_i and the reference component, respectively. They can be obtained by writing a balanced equation for the oxidation of the compound or component to carbon dioxide and water. Table 3.1 contains COD mass equivalents of several constituents that commonly change oxidation state in biochemical operations. Note that under oxidizing conditions, carbon dioxide has a COD of zero, since the carbon in it is already in the most oxidized state (+IV), as it is in bicarbonate and carbonate. Furthermore, oxygen is equivalent to negative COD since COD is oxygen demand, i.e., it represents loss of oxygen. Finally, it should be noted that any reactant or product containing only elements that do not change oxidation state during biochemical oxidation/reduction reactions will have a unit COD of zero, causing them to drop out of the COD-based stoichiometric equation.

Table 3.1 COD Mass Equivalents of Some Common Constituents

Constituent[a]	Change of oxidation state	COD equivalent[b]
Biomass, $C_5H_7O_2N$	C to +IV	1.42 g COD/g $C_5H_7O_2N$, 1.42 g COD/g VSS, 1.20 g COD/g TSS
Oxygen (as e acceptor)	O(0) to O(−II)	−1.00 g COD/gO_2[c]
Nitrate (as e acceptor)	N(+V) to N(0)	−0.646 g COD/g NO_3^-, −2.86 g COD/g N
Nitrate (as N source)	N(+V) to N(−III)	−1.03 g COD/g NO_3^-, −4.57 g COD/g N
Sulfate (as e acceptor)	S(+VI) to S(−II)	−0.667 g COD/g SO_4^{2-}, −2.00 g COD/g S
Carbon dioxide (as e acceptor)	C(+IV) to C(−IV)	−1.45 g COD/g CO_2, −5.33 g COD/g C
CO_2, HCO_3^-, $H_2CO_3^*$	No change in an oxidizing environment	0.00
Organic matter in domestic wastewater, $C_{10}H_{19}O_3N$	C to +IV	1.99 g COD/g organic matter
Protein, $C_{16}H_{24}O_5N_4$	C to +IV	1.50 g COD/g protein
Carbohydrate, CH_2O	C to +IV	1.07 g COD/g carbohydrate
Grease, $C_8H_{16}O$	C to +IV	2.88 g COD/g grease
Acetate, CH_3COO^-	C to +IV	1.08 g COD/g acetate
Propionate, $C_2H_5COO^-$	C to +IV	1.53 g COD/g propionate
Benzoate, $C_6H_5COO^-$	C to +IV	1.98 g COD/g benzoate
Ethanol, C_2H_5OH	C to +IV	2.09 g COD/g ethanol
Lactate, $C_2H_4OHCOO^-$	C to +IV	1.08 g COD/g lactate
Pyruvate, CH_3COCOO^-	C to +IV	0.92 g COD/g pyruvate
Methanol, CH_3OH	C to +IV	1.50 g COD/g methanol
$NH_4^+ \rightarrow NO_3^-$	N(−III) to N(+V)	3.55 g COD/g NH_4^+, 4.57 g COD/g N
$NH_4^+ \rightarrow NO_2^-$	N(−III) to N(+III)	2.67 g COD/g NH_4^+, 3.43 g COD/g N
$NO_2^- \rightarrow NO_3^-$	N(+III) to N(+V)	0.36 g COD/g NO_2^-, 1.14 g COD/g N
$S \rightarrow SO_4^{2-}$	S(0) to S(+VI)	1.50 g COD/g S
$H_2S \rightarrow SO_4^{2-}$	S(−II) to S(+VI)	1.88 g COD/g H_2S, 2.00 g COD/g S
$S_2O_3^{2-} \rightarrow SO_4^{2-}$	S(+II) to S(+VI)	0.57 g COD/g $S_2O_3^{2-}$, 1.00 g COD/g S
$SO_3^{2-} \rightarrow SO_4^{2-}$	S(+IV) to S(+VI)	0.20 g COD/g SO_3^{2-}, 0.50 g COD/g S
H_2	H(0) to H(+I)	8.00 g COD/g H

[a] Listed in the same order as the reactants in Table 3.2.
[b] A negative sign implies that the constitutent is receiving electrons.
[c] By definition, oxygen demand is negative oxygen.

Example 3.1.1.1

Consider a typical molar-based stoichiometric equation for bacterial growth on carbohydrate (CH_2O) with ammonia as the nitrogen source:

$$CH_2O + 0.290\ O_2 + 0.142\ NH_4^+ + 0.142\ HCO_3^- \rightarrow$$
$$0.142\ C_5H_7O_2N + 0.432\ CO_2 + 0.858\ H_2O \quad (3.6)$$

where $C_5H_7O_2N$ is the empirical formula for cell mass. Note that the charges are balanced and that the number of moles of each element in the reactants equals the number in the products. The molar-based stoichiometric equation tells us that the biomass yield is 0.142 moles of biomass formed per mole of carbohydrate used and that 0.290 moles of oxygen are required per mole of carbohydrate used to synthesize that biomass.

Convert this equation to a mass based stoichiometric equation. To do this, we need the molecular weight of each reactant and product. These are CH_2O, 30; O_2, 32; NH_4^+, 18; HCO_3^-, 61; $C_5H_7O_2N$, 113; CO_2, 44; and H_2O, 18. Using these with the stoichiometric coefficients from Eq. 3.6 in Eq. 3.2 gives:

$$CH_2O + 0.309\ O_2 + 0.085\ NH_4^+ + 0.289\ HCO_3^- \rightarrow$$
$$0.535\ C_5H_7O_2N + 0.633\ CO_2 + 0.515\ H_2O \quad (3.7)$$

In this case, the charges are no longer balanced, but the sum of the stoichiometric coefficients for the reactants equals the sum for the products. The mass-based stoichiometric equation tells us that the biomass yield is 0.535 grams of biomass formed per gram of carbohydrate used and that 0.309 grams of oxygen are required per gram of carbohydrate used to synthesize that biomass.

Now convert the molar-based equation to a COD-based equation. To do this, use must be made of the unit CODs given in Table 3.1. In this case, the unit COD of ammonia is taken as zero because the nitrogen in cell material is in the same oxidation state as the nitrogen in ammonia, i.e., −III; it does not undergo a change of oxidation state. Carrying out the conversion represented by Eq. 3.4 yields:

$$CH_2O\ COD + (-0.29)O_2 \rightarrow 0.71\ C_5H_7O_2N\ COD \quad (3.8)$$

Note that only three constituents remain because they are the only ones that can be represented by COD in this case. Also note that like the mass-based equation, the sum of the stoichiometric coefficients for the reactants equals the sum of the stoichiometric coefficients for the products. Finally, note that the stoichiometric coefficient for oxygen carries a negative sign even though it is a reactant. That is because it is being expressed as COD. Thus, the COD-based stoichiometric equation tells us that the biomass yield is 0.71 grams of biomass COD formed per gram of carbohydrate COD used and that 0.29 grams of oxygen are required per gram of carbohydrate COD used to synthesize that biomass.

3.1.2 Generalized Reaction Rate

Stoichiometric equations can also be used to establish the relative reaction rates for reactants or products. Because the sum of the stoichiometric coefficients in a mass-based stoichiometric equation equals zero, its general form may be rewritten in the following way:[62]

$$(-1)A_1 + (-\Psi_2)A_2 + \cdots + (-\Psi_k)A_k + \Psi_{k+1}A_{k+1} + \cdots + \Psi_m A_m = 0 \quad (3.9)$$

where components 1 through k are reactants, components k+1 through m are products, and reactant A_1 is the basis for the normalized stoichiometric coefficients. Note that the normalized stoichiometric coefficients are given negative signs for reactants and positive signs for products. Since there is a relationship between the masses of the different reactants used or products formed, it follows that there is also a relationship between the rates at which they are used or formed. If we let r_i represent the rate of formation of component i (where i = 1 → k), it follows that:

$$\frac{r_1}{(-1)} = \frac{r_2}{(-\Psi_2)} = \frac{r_k}{(-\Psi_k)} = \frac{r_{k+1}}{(\Psi_{k+1})} = \frac{r_m}{(\Psi_m)} = r \quad (3.10)$$

where r is called the generalized reaction rate. As above, the sign on Ψ_i signifies whether the component is being removed or formed. Consequently, if the stoichiometry of a reaction has been determined in mass units, and the reaction rate has been determined for one component, then the reaction rates in mass units are known for all other components.

Equations 3.9 and 3.10 also hold true for COD-based stoichiometric equations. The normalized stoichiometric coefficients (Ψ_i) are simply replaced with appropriate COD-based coefficients (Y_i).

Example 3.1.2.1

Biomass is growing in a bioreactor at a rate of 1.0 g/(L·h) and the growth conforms to the stoichiometry expressed by Eq. 3.7. At what rate are carbohydrate and oxygen being used in the bioreactor to support that growth?

Rewriting Eq. 3.7 in the form of Eq. 3.9 gives:

$$-CH_2O - 0.309\ O_2 - 0.085\ NH_3 - 0.289\ HCO_3 + 0.535\ C_5H_7O_2N + 0.633\ CO_2 + 0.515\ H_2O = 0$$

Use of Eq. 3.10 allows determination of the generalized reaction rate:

$$r = \frac{r_{C_5H_7O_2N}}{0.535} = \frac{1.0}{0.535} = 1.87\ g\ CH_2O/(L \cdot h)$$

Note that the generalized reaction rate is expressed in terms of the constituent that serves as the basis for normalization of the stoichiometric equation. The rates of carbohydrate and oxygen utilization can now also be determined from Eq. 3.10:

$$r_{CH_2O} = (-1.0)(1.87) = -1.87\ g\ CH_2O/(L \cdot h)$$

$$r_{O_2} = (-0.309)(1.87) = -0.58\ g\ O_2/(L \cdot h)$$

3.1.3 Multiple Reactions—The Matrix Approach

In Chapter 2 we learn that there are many important events occurring in biochemical operations. Consequently, multiple reactions will take place simultaneously, and all must be considered when mass balance equations are written for biochemical oper-

ations. Extension of the concepts above to multiple reactions simplifies the presentation of those mass balances and allows the fates of all reactants to be easily visualized.[5,6,7]

Consider a situation in which i components (where i = 1 → m) participate in j reactions (where j = 1 → n), in which case $\Psi_{i,j}$ represents the normalized mass-based stoichiometric coefficient for component i in reaction j. This situation gives a group of mass-based stoichiometric equations:

$$\begin{aligned}
(-1)A_1 &+ \cdots + (-\Psi_{k,1})A_k + (+\Psi_{k+1,1})A_{k+1} + \cdots + (+\Psi_{m,1})A_m = 0 \quad r_1 \\
(+\Psi_{1,2})A_1 &+ \cdots + (-1)A_k + (+\Psi_{k+1,2})A_{k+1} + \cdots + (+\Psi_{m,2})A_m = 0 \quad r_2 \\
&\vdots \qquad\qquad \vdots \qquad\qquad \vdots \qquad\qquad \vdots \\
(+\Psi_{1,n})A_1 &+ \cdots + (+\Psi_{k,n})A_k + (+\Psi_{k+1,n})A_{k+1} + \cdots + (-1)A_m = 0 \quad r_n
\end{aligned} \tag{3.11}$$

Note that A_1 does not necessarily represent the component chosen as the basis for the normalized stoichiometric coefficients. Rather, a different component may be selected for each reaction so that each resulting normalized stoichiometric coefficient has appropriate physical meaning. Nevertheless, because the equations are mass-based, the sum of the normalized stoichiometric coefficients in each equation must equal zero, as indicated in Eq. 3.11. This allows a continuity check to be made for each reaction. Furthermore, also note that any component A_i may be a reactant in one reaction and a product in another. This means that the overall rate of formation of that component will be the net rate obtained by considering the sum of the rates for all reactions in which it participates:

$$r_i = \sum_{j=1}^{n} \Psi_{i,j} r_j \tag{3.12}$$

If the net rate of formation is negative, the component is being consumed and if it is positive the component is being produced. The same approach can be used for COD-based stoichiometric equations by replacing $\Psi_{i,j}$ with $Y_{i,j}$. This approach will be applied in Part II when models are developed for biochemical reactors, and will be particularly useful when complex systems with several components and reactions are considered.

3.2 BIOMASS GROWTH AND SUBSTRATE UTILIZATION

3.2.1 Generalized Equation for Biomass Growth

It will be recalled from Section 2.4.1 that biomass growth and substrate utilization are coupled. Furthermore, we see in Section 2.4.2 that environmental engineers account for maintenance energy needs through the decay reaction. This means that as long as the production of soluble microbial products is negligible, the only use of substrate is for biomass growth. Consequently, when a stoichiometric equation for biomass growth is written with the substrate as the basis, the stoichiometric coefficient for the biomass term will be the biomass true growth yield. With this in mind, the generalized equation for microbial growth can be written as:

$$\text{Carbon source} + \text{energy source} + \text{electron acceptor} + \text{nutrients} \rightarrow$$
$$\text{biomass} + CO_2 + \text{reduced acceptor} + \text{end products} \quad (3.13)$$

For modeling purposes, it would be desirable to be able to write a quantitative equation in the same form for any situation, no matter what the carbon source, energy source, or electron acceptor. Using the concept of half-reactions, McCarty[77] has devised a technique whereby this may be done.

Half-Reaction Approach. In the absence of significant soluble microbial product formation, all nonphotosynthetic microbial growth reactions consist of two components, one for synthesis and one for energy. The carbon in the synthesis component ends up in biomass, whereas any carbon associated with the energy component becomes carbon dioxide. Such reactions are also oxidation–reduction reactions and thus involve the transfer of electrons from a donor to an acceptor. For heterotrophic growth the electron donor is an organic substrate, whereas for autotrophic growth the electron donor is inorganic. To allow consideration of all of these factors, McCarty[77] has written three types of half-reactions: one for cell material (R_c), one for the electron donor (R_d), and one for the electron acceptor (R_a). These are presented in Table 3.2 for a variety of substances. Reactions 1 and 2 represent R_c for the formation of biomass. Both are based on the empirical formula $C_5H_7O_2N$, but one uses ammonia nitrogen as the nitrogen source whereas the other uses nitrate. Reactions 3–6 are half-reactions R_a for the electron acceptors oxygen, nitrate, sulfate, and carbon dioxide, respectively. Reactions 7–17 are half-reactions R_d for organic electron donors. The first of these represents the general composition of domestic wastewater, while the next three are for wastes composed primarily of proteins, carbohydrates, and lipids, respectively. Reactions 11–17 are for specific organic compounds of interest in some biochemical operations. The last nine reactions represent possible autotrophic electron donors. Reactions 19–21 are for nitrification. To facilitate their combination, the half-reactions are all written on an electron equivalent basis, with the electrons on the right side.

The overall stoichiometric equation (R) is the sum of the half-reactions:

$$R = R_d - f_e \cdot R_a - f_s \cdot R_c \quad (3.14)$$

The minus terms mean that half-reactions R_a and R_c must be inverted before use. This is done by switching the left and right sides. The term f_s represents the fraction of the electron donor that is coupled with the electron acceptor, i.e., the portion used for energy, hence the subscript e, and f_s represents the fraction captured through synthesis. As such, they quantify the endpoint of the reaction. Furthermore, in order for Eq. 3.14 to balance:

$$f_e + f_s = 1.0 \quad (3.15)$$

This equation is equivalent to stating that all electrons originally in the electron donor end up either in the biomass synthesized (f_s) or in the electron acceptor (f_e). This is an important fundamental concept that we will return to later.

Empirical Formulas for Use in Stoichiometric Equations. As can be seen by examining Table 3.2, it was necessary to assume empirical formulas for biomass and alternative organic electron donors in order to write the half-reactions.

Various empirical formulas have been proposed to represent the organic composition of microbial cells. One of the oldest and most widely accepted in the field

Table 3.2 Oxidation Half-Reactions[a]

Reaction number	Half-reactions
Reactions for bacterial cell synthesis (R_c)	
Ammonia as nitrogen source:	
1.	$\frac{1}{20}C_5H_7O_2N + \frac{9}{20}H_2O = \frac{1}{5}CO_2 + \frac{1}{20}HCO_3^- + \frac{1}{20}NH_4^+ + H^+ + e$
Nitrate as nitrogen source:	
2.	$\frac{1}{28}C_5H_7O_2N + \frac{11}{28}H_2O = \frac{1}{28}NO_3^- + \frac{5}{28}CO_2 + \frac{29}{28}H^+ + e$
Reactions for electron acceptors (R_a)	
Oxygen:	
3.	$\frac{1}{2}H_2O = \frac{1}{4}O_2 + H^+ + e$
Nitrate:	
4.	$\frac{1}{10}N_2 + \frac{3}{5}H_2O = \frac{1}{5}NO_3^- + \frac{6}{5}H^+ + e$
Sulfate:	
5.	$\frac{1}{16}H_2S + \frac{1}{16}HS^- + \frac{1}{2}H_2O = \frac{1}{8}SO_4^{2-} + \frac{19}{16}H^+ + e$
Carbon dioxide (methanogenesis):	
6.	$\frac{1}{8}CH_4 + \frac{1}{4}H_2O = \frac{1}{8}CO_2 + H^+ + e$
Reactions for electron donors (R_d)	
Organic donors (heterotrophic reactions):	
Domestic wastewater:	
7.	$\frac{1}{50}C_{10}H_{19}O_3N + \frac{9}{25}H_2O = \frac{9}{50}CO_2 + \frac{1}{50}NH_4^+ + \frac{1}{50}HCO_3^- + H^+ + e$
Protein (amino acids, proteins, nitrogenous organics):	
8.	$\frac{1}{66}C_{16}H_{24}O_5N_4 + \frac{27}{66}H_2O = \frac{8}{33}CO_2 + \frac{2}{33}NH_4^+ + \frac{31}{33}H^+ + e$
Carbohydrate (cellulose, starch, sugars):	
9.	$\frac{1}{4}CH_2O + \frac{1}{4}H_2O = \frac{1}{4}CO_2 + H^+ + e$
Grease (fats and oils):	
10.	$\frac{1}{46}C_3H_8O + \frac{15}{46}H_2O = \frac{4}{23}CO_2 + H^+ + e$
Acetate:	
11.	$\frac{1}{8}CH_3COO^- + \frac{3}{8}H_2O = \frac{1}{8}CO_2 + \frac{1}{8}HCO_3^- + H^+ + e$
Propionate:	
12.	$\frac{1}{14}CH_3CH_2COO^- + \frac{5}{14}H_2O = \frac{1}{7}CO_2 + \frac{1}{14}HCO_3^- + H^+ + e$
Benzoate:	
13.	$\frac{1}{30}C_6H_5COO^- + \frac{13}{30}H_2O = \frac{1}{5}CO_2 + \frac{1}{30}HCO_3^- + H^+ + e$

Table 3.2 Continued

Reaction number	Half-reactions
Ethanol:	
14. $\frac{1}{12}\,CH_3CH_2OH + \frac{1}{4}\,H_2O$	$= \frac{1}{6}\,CO_2 + H^+ + e$
Lactate:	
15. $\frac{1}{12}\,CH_3CHOHCOO^- + \frac{1}{3}\,H_2O$	$= \frac{1}{6}\,CO_2 + \frac{1}{12}\,HCO_3^- + H^+ + e$
Pyruvate:	
16. $\frac{1}{10}\,CH_3COCOO^- + \frac{2}{5}\,H_2O$	$= \frac{1}{5}\,CO_2 + \frac{1}{10}\,HCO_3^- + H^+ + e$
Methanol:	
17. $\frac{1}{6}\,CH_3OH + \frac{1}{6}\,H_2O$	$= \frac{1}{6}\,CO_2 + H^+ + e$
Inorganic donors (autotrophic reactions):	
18. Fe^{++}	$= Fe^{+3} + e$
19. $\frac{1}{8}\,NH_4^+ + \frac{3}{8}\,H_2O$	$= \frac{1}{8}\,NO_3^- + \frac{5}{4}\,H^+ + e$
20. $\frac{1}{6}\,NH_4^+ + \frac{1}{3}\,H_2O$	$= \frac{1}{6}\,NO_2^- + \frac{4}{3}\,H^+ + e$
21. $\frac{1}{2}\,NO_2^- + \frac{1}{2}\,H_2O$	$= \frac{1}{2}\,NO_3^- + H^+ + e$
22. $\frac{1}{6}\,S + \frac{2}{3}\,H_2O$	$= \frac{1}{6}\,SO_4^= + \frac{4}{3}\,H^+ + e$
23. $\frac{1}{16}\,H_2S + \frac{1}{16}\,HS^- + \frac{1}{2}\,H_2O$	$= \frac{1}{8}\,SO_4^= + \frac{19}{16}\,H^+ + e$
24. $\frac{1}{8}\,S_2O_3^= + \frac{5}{8}\,H_2O$	$= \frac{1}{4}\,SO_4^= + \frac{5}{4}\,H^+ + e$
25. $\frac{1}{2}\,SO_3^= + \frac{1}{2}\,H_2O$	$= \frac{1}{2}\,SO_4^= + H^+ + e$
26. $\frac{1}{2}\,H_2$	$= H^+ + e$

[a]Adapted from McCarty.[77]

of wastewater treatment is the one introduced in Section 2.4.1 and used in Example 3.1.1.1, $C_5H_7O_2N$.[61] Other formulas consisting of the same elements have been used, but they all result in about the same COD per unit of biomass.[75] Another formula has been proposed that includes phosphorus, $C_{60}H_{87}O_{23}N_{12}P$.[76] While awareness of the need for phosphorus by biomass is essential, it is not necessary to include phosphorus in the empirical formula because the mass required is generally about one-fifth of the mass of nitrogen required. This allows the phosphorus requirement to be calculated even when the simpler empirical formula is used.

All empirical formulas for biomass seek to represent in a simple way material composed of a highly complex and integrated mixture of organic molecules. Fur-

thermore, because the relative quantities of those molecules change as the growth conditions of the culture change,[55] it would be purely fortuitous if a single chemical formula for biomass applied to all cases. An estimate of the constancy of the overall elemental composition can be obtained by measuring the COD and heat of combustion of biomass grown under various conditions, because constancy of those parameters would imply that the ratios of the elements C, H, O, and N were relatively constant. Investigations of that sort have indicated that the elemental composition is indeed a function of the growth conditions.[41] Thus, while an empirical formula can be written for biomass, its applicability to all situations is doubtful and one should view with caution equations said to depict "the biochemical reaction" exactly. Nevertheless, the concepts stated in Eq. 3.13 are still valid and many important relationships can be demonstrated through their use. Consequently, for illustrative purposes, the formula $C_5H_7O_2N$ will be used to represent biomass throughout this book. As discussed in Section 2.4.1, it has a COD of 1.42 mg COD/mg VSS, or 1.20 mg COD/mg SS.

In a laboratory or research situation, the exact composition of the electron donor is usually known. For example, if glucose were the energy source, its empirical formula $C_6H_{12}O_6$ would be used in the stoichiometric equation. Furthermore, if a synthetic medium contained several organic electron donors, the half-reaction for each could be written separately and then they could be combined to get R_d for the mixture by multiplying each half-reaction by the fractional contribution (on an electron equivalent basis) of its electron donor in the medium and adding them together.

An actual wastewater presents a more difficult situation because the chemical composition of the electron donor is seldom known. One approach would be to analyze the waste for its carbon, hydrogen, oxygen and nitrogen contents, and construct an empirical formula from the results. A half-reaction could then be written for that particular formula.[56] For example, as shown in Table 3.2, the empirical formula for the organic matter in domestic wastewater has been estimated to be $C_{10}H_{19}O_3N$. Alternatively, if the COD, organic carbon, organic nitrogen, and volatile solids content of a wastewater are known, they can be used to generate the half-reaction.[57] Finally, if a wastewater contains predominately carbohydrate, protein, and lipid, knowledge of their relative concentrations can be used to write the equation for microbial growth because each can be represented by a generalized empirical formula: CH_2O, $C_{16}H_{21}O_5N_4$, and $C_5H_{10}O$, respectively. As with other mixtures, the half-reaction for each is multiplied by the fraction of the component in the wastewater and the three are added to get R_d.

The nature of the electron acceptor depends on the environment in which the biomass is growing. If the environment is aerobic, the acceptor will be oxygen. If it is anaerobic, the acceptor will depend on the particular reaction taking place. For example, if lactic acid fermentation is occurring, pyruvic acid is the acceptor, whereas carbon dioxide is the acceptor for methanogenesis. Finally, nitrate can serve as the electron acceptor under anoxic conditions. Half-reactions have been written for all of these, as shown in Table 3.2.

Determination of f_s. Once the electron donor and the electron acceptor have been identified, either f_s or f_e must be determined before the balanced stoichiometric equation can be written. Generally, f_s is easier to estimate because it can be related to the true growth yield expressed on a COD basis. If f_s is the fraction of the electron donor transferred to the electron acceptor to provide the energy with which to syn-

thesize new biomass, conservation of energy and Eq. 3.15 tell us that the remainder of the electrons originally available in the donor must end up in the new biomass formed. If we accept $C_5H_7O_2N$ as being representative of biomass, we can see that carbon and nitrogen are the reduced elements that will house those electrons. Nitrogen in biomass is in the −III state, i.e., as amino nitrogen. If the nitrogen available for biomass synthesis is also in the −III state, as in ammonia, no electrons will be required to reduce it, and the electrons captured through synthesis will all be associated with the carbon. Consequently, the energy available in the carbon of the biomass is equal to the energy incorporated during synthesis, or f_s when expressed as a fraction of the electron donor. Thus, if we could measure the energy or electrons available in the biomass produced, we would have a measure of f_s.

In Section 2.4.1 the yield was defined as the amount of biomass formed per unit of substrate used. However, it was also pointed out that when the electron donor is an organic compound, it is often convenient to express the yield as mass of biomass COD formed per mass of substrate COD destroyed. The COD test is a measure of electrons available from carbon. Since COD is oxygen demand, and oxygen has an equivalent weight of eight, there are eight grams of COD per electron equivalent, as can be seen by examining half-reaction 3 in Table 3.2. This allows interconversion of COD and electron equivalents. Consequently, the yield is also the number of electrons available from carbon in the new biomass per unit of electrons removed from the substrate, or the fraction of the electron donor captured through synthesis, f_s. Thus, when ammonia nitrogen serves as the nitrogen source for heterotrophic biomass synthesis:

$$f_s = Y_H \text{ } (NH_4^+ \text{ as nitrogen source, organic electron donor}) \tag{3.16}$$

where Y_H is expressed on a COD basis and the subscript H indicates that the true growth yield is for heterotrophic biomass growth. The utility of Eq. 3.16 comes from the fact that the true growth yield, Y_H, can be determined in COD units from data collected with full-, pilot-, or lab-scale bioreactors, thereby giving f_s for the system under study. The techniques for doing this will be discussed in Chapter 8.

As long as ammonia or amino nitrogen is available to the microorganisms, they will use it preferentially for biomass synthesis. If it is not available, the microorganisms will use nitrate-N. (If no nitrogen is available, cell synthesis cannot occur because an essential reactant is missing.) When nitrate is the nitrogen source, the nitrogen must be reduced from the +V state to the −III state before it can be assimilated. This requires some of the electrons available in the substrate and they are part of the energy required for synthesis, i.e., part of f_s. However, the electrons required to reduce the nitrogen are not measured in the COD test because that test does not oxidize nitrogen, but leaves it in the −III state. In this case, the true growth yield expressed on a COD basis is not an accurate estimate of f_s. Rather, Y_H will be smaller than f_s. This artifact can be corrected, however, because we know the number of electrons required to reduce nitrate-N to the appropriate oxidation state. Assuming an empirical formula for biomass of $C_5H_7O_2N$, it can be shown that:

$$f_s = 1.40 \text{ } Y_H \text{ } (NO_3^- \text{ as nitrogen source, organic electron donor}) \tag{3.17}$$

Thermodynamics suggests that the true growth yield obtained for growth with nitrate as the nitrogen source will be smaller than the true growth yield obtained when ammonia is available.[52] For example, for carbohydrate as the electron and

carbon donor, the value of Y_H would be about 20 percent smaller with nitrate as the nitrogen source.

There are often circumstances in which one needs to establish the stoichiometry of biomass growth and substrate utilization before experimentally determined values of Y_H are available. Thus, it would be advantageous to have a theoretical basis for estimating f_s or Y_H. This has led a number of workers to seek a thermodynamic approach for predicting yield values.[51,75] However, as discussed in Section 2.4.1, this is a difficult task because of the large number of factors that influence the yield. The most successful approach to date is that of Heijnen et al.,[51,52] which is based on the Gibbs energy dissipation per C-mole of biomass produced, the degree of reduction of the carbon donor, the nature of the nitrogen source, and the available Gibbs energy per electron between the electron donor and acceptor. Their technique is capable of predicting true growth yield values for both heterotrophic and autotrophic biomass growth for a variety of situations. The error is approximately 13% when tabular values of the best estimates of the Gibbs energy dissipation per C-mole of biomass produced are used, but increases to 19% when that dissipation is estimated with a correlation equation that relates it to the carbon chain length and the degree of reduction of the carbon source. Because one should fully understand the technique of Heijnen et al.[52] before using it, and because the presentation required to establish that understanding is beyond the scope of this book, readers are referred to the original work if they desire to use such an approach.

3.2.2 Aerobic Growth of Heterotrophs with Ammonia as the Nitrogen Source

The best way to illustrate the use of half-reactions is by an example. We will develop the molar stoichiometric equation for aerobic growth of heterotrophs that was the starting point for Example 3.1.1.1.

Example 3.2.2.1

Write the stoichiometric equation for aerobic heterotrophic microbial growth on a carbohydrate using ammonia as the nitrogen source, under conditions such that the true growth yield (Y_H) is 0.71 mg of biomass COD formed per mg of carbohydrate COD removed.

To do this we must make use of Eqs. 3.14–3.16:

$$R = R_d - f_s \cdot R_s - f_e \cdot R_e$$

$$f_s = Y_H = 0.71$$

$$f_e = 1.00 - 0.71 = 0.29$$

Therefore

$$R = R_d - 0.29\ R_s - 0.71\ R_e$$

The electron donor is carbohydrate and the acceptor is oxygen. Thus, from Table 3.2:

$$R_d = 1/4\ CH_2O + 1/4\ H_2O = 1/4\ CO_2 + H^+ + e$$

$$R_e = 1/2\ H_2O = 1/4\ O_2 + H^+ + e$$

Since ammonia is the nitrogen source, R_c is:

$$R_c = 1/20\ C_5H_7O_2N + 9/20\ H_2O = 1/5\ CO_2 + 1/20\ HCO_3^- + 1/20\ NH_4^+ + H^+ + e$$

Applying Eq. 3.14 gives:

$$R_d = 0.25\ CH_2O + 0.25\ H_2O = 0.25\ CO_2 + H^+ + e$$
$$-0.29\ R_d = 0.0725\ O_2 + 0.29\ H^+ + 0.29\ e = 0.145\ H_2O$$
$$-0.71\ R_c = 0.142\ CO_2 + 0.0355\ HCO_3^- + 0.0355\ NH_4^+ + 0.71\ H^+ + 0.71\ e = 0.0355\ C_5H_7O_2N + 0.3195\ H_2O$$
$$R = 0.25\ CH_2O + 0.0725\ O_2 + 0.0355\ NH_4^+ + 0.0355\ HCO_3^- = 0.0355\ C_5H_7O_2N + 0.108\ CO_2 + 0.2145\ H_2O$$

This can be normalized to one mole of carbohydrate by dividing through by 0.25, giving Eq. 3.6, which was the starting point of Example 3.1.1.1:

$$CH_2O + 0.29\ O_2 + 0.142\ NH_4^+ + 0.142\ HCO_3^- \rightarrow 0.142\ C_5H_7O_2N + 0.432\ CO_2 + 0.858\ H_2O \qquad (3.6)$$

Equation 3.6 was converted to a COD-based stoichiometric equation in Example 3.1.1.1. If we rearrange that equation in the same form as Eq. 3.15, the result is:

$$0.29\ O_2 + 0.71\ C_5H_7O_2N\ COD = CH_2O\ COD \qquad (3.18)$$

We return to this equation to make three important points. First, note that the value of Y_H in Eq. 3.18 is 0.71 mg biomass COD formed/mg substrate COD used. This is the same as the Y_H value used to develop Eq. 3.6, as we would expect. Second, note that Eq. 3.18 expresses the same information as Eq. 3.15. In other words, since all of the electrons removed from the substrate must end up in either the electron acceptor or the biomass formed, we can state that the substrate COD removed must equal the biomass COD formed plus the oxygen used. Finally, since Eq. 3.18 expresses the same information as Eq. 3.15, we can see that the COD-based stoichiometric coefficient on oxygen is the same as f_e. The balance portrayed by Eqs. 3.15 and 3.18 is a very important one that we will make extensive use of throughout this book.

3.2.3 Aerobic Growth of Heterotrophs with Nitrate as the Nitrogen Source

As previously discussed, consideration must be given to the form of nitrogen available for cell synthesis when writing the stoichiometric equation for cell growth. Ammonia will be used preferentially, and thus half-reaction 1 in Table 3.2 should be used when ammonia is available, even if nitrate is serving as the terminal electron acceptor. Only when nitrate is present as the sole nitrogen source should half-reaction 2 be used. In that case, when expressing the stoichiometric equation on a COD basis, it must be recognized that nitrogen changes oxidation state from +V to −III. As an example, consider the case of the aerobic growth of heterotrophs on carbohydrate with nitrate as the nitrogen source. In this case, the true growth yield is 0.57 mg

biomass COD/mg carbohydrate COD removed, reflecting the energy that must be used to reduce the nitrogen. Applying Eq. 3.17 reveals that f_s is 0.80, giving the following molar stoichiometric equation:

$$CH_2O + 0.200\ O_2 + 0.114\ NO_3^- + 0.114\ H^+ \rightarrow 0.114\ C_5H_7O_2N + 0.429\ CO_2 + 0.657\ H_2O \quad (3.19)$$

After conversion to a mass basis by application of Eq. 3.2 this becomes:

$$CH_2O + 0.213\ O_2 + 0.236\ NO_3^- + 0.004\ H^+ \rightarrow 0.429\ C_5H_7O_2N + 0.629\ CO_2 + 0.395\ H_2O \quad (3.20)$$

Conversion of this equation to a mass of COD basis requires application of Eq. 3.5 using the unit CODs given in Table 3.1. Note that NO_3^- has a unit COD of -1.03 mg COD/mg NO_3^-. This is equivalent to saying that each mg of nitrate that is reduced to amino nitrogen in biomass accepts as many electrons as 1.03 mg of oxygen. Application of Eq. 3.4 to Eq. 3.20 gives:

$$CH_2O\ COD + (-0.200)O_2 + (-0.230)O_2 \text{ equivalents of } NO_3^- \rightarrow 0.570\ C_5H_7O_2N\ COD \quad (3.21)$$

Equation 3.21 shows clearly that the COD (electron) balance would not be correct if the change in oxidation state of the nitrogen was not considered. Failure to recognize this can lead to problems when COD balances are performed on operating bioreactors.

It is often convenient to express the COD equivalence of nitrate as a nitrogen source on the basis of the nitrogen utilized for biomass synthesis, rather than on the basis of nitrate. In that case, the conversion factor is -4.57 mg COD/mg N (or 4.57 mg O_2/mg N), as indicated in Table 3.1.

3.2.4 Growth of Heterotrophs with Nitrate as the Terminal Electron Acceptor and Ammonia as the Nitrogen Source

If nitrate were serving as the terminal electron acceptor under anoxic conditions, the amount needed could be calculated from the stoichiometric equation obtained when half-reaction No. 4 was used in place of No. 3 as R_a in Eq. 3.14. Exactly the same procedures would be followed for obtaining the molar and mass-based stoichiometric equations. Consider the case when ammonia serves as the nitrogen source for cell synthesis. Because it will allow illustration of an important point, we will assume that the true growth yield from carbohydrate is 0.71 mg biomass COD/mg substrate COD, which is the same as the Y_H value used to develop Eq. 3.6 for growth under aerobic conditions. Application of the appropriate techniques gives:

$$CH_2O + 0.232\ NO_3^- + 0.142\ NH_4^+ + 0.142\ HCO_3^- + 0.232\ H^+ \rightarrow 0.142\ C_5H_7O_2N + 0.432\ CO_2 + 0.116\ N_2 + 0.974\ H_2O \quad (3.22)$$

Converting this to a mass of carbohydrate basis by application of Eq. 3.2 gives:

$$CH_2O + 0.479\ NO_3^- + 0.085\ NH_4^+ + 0.289\ HCO_3^- + 0.008\ H^+ \rightarrow$$
$$0.535\ C_5H_7O_2N + 0.634\ CO_2 + 0.108\ N_2 + 0.584\ H_2O \qquad (3.23)$$

Because the true growth yield was assumed to be the same as in Example 3.1.1.1, the quantities of biomass formed in Eqs. 3.22 and 3.23 are the same as those in Eqs. 3.6 and 3.7, respectively.

Conversion of Eq. 3.23 to a COD basis requires inclusion of a conversion factor for the oxygen equivalence of nitrate nitrogen when it is being reduced to nitrogen gas, N_2, which is the case when nitrate serves as the terminal electron acceptor. Examination of Table 3.1 reveals that the unit COD for the reduction of NO_3^- to N_2 is -0.646 mg COD/mg NO_3^-. The sign is negative because the nitrate is accepting electrons. The source of this value may be seen from the half-reactions in Table 3.2 which reveal that 1/5 mole of nitrate is equivalent to 1/4 mole of oxygen. Conversion to a mass basis reveals that each gram of nitrate that is reduced to N_2 can accept as many electrons as 0.646 grams of oxygen. Applying Eq. 3.4 with the appropriate conversion factors to Eq. 3.23 gives:

$$CH_2O\ COD + (-0.29)O_2 \text{ equivalents of } NO_3^- \rightarrow$$
$$0.71\ C_5H_7O_2N\ COD \qquad (3.24)$$

Comparison of Eq. 3.24 to Eq. 3.8 reveals that they are the same. This follows from the fact that they are both expressed on a COD basis, that they both use ammonia as the nitrogen source for biomass synthesis, and that they both were derived for the same yield. Generally, however, the yield will be lower when nitrate serves as the terminal electron acceptor.[51,52,79]

Often it is convenient to express the oxygen equivalence of nitrate as an electron acceptor on the basis of nitrogen rather than nitrate. In that case, the conversion factor is -2.86 mg COD/mg N (or 2.86 mg O_2/mg N), as shown in Table 3.1.

It should be noted from the preceding that the COD conversion factor for nitrate as a nitrogen source is different from the COD conversion factor for nitrate as a terminal electron acceptor because the final oxidation state of nitrogen is different in the two cases. This becomes especially important when nitrate serves as both the nitrogen source and the terminal electron acceptor. The safest way to handle this situation is to keep the two uses of nitrate separate in writing the stoichiometric equation, and to apply the appropriate conversion factor for each when converting the equation to a COD basis.

3.2.5 Aerobic Growth of Autotrophs with Ammonia as the Electron Donor

Nitrifying bacteria are autotrophic microorganisms that obtain their energy from the oxidation of reduced nitrogen. As discussed previously, *Nitrosomonas* oxidizes ammonia-N to nitrite-N and *Nitrobacter* oxidizes nitrite-N to nitrate-N. The molar stoichiometric equations for their growth can be obtained by the half-reaction technique discussed previously, which requires knowledge of f_s. For autotrophic biomass growth, yield is often expressed as the mass of biomass COD formed per mass of inorganic element oxidized; for example, mg of biomass COD per mg of ammonia-

N removed for *Nitrosomonas*. To convert this yield value to an electron equivalent basis for determining f_s it is necessary to know that *Nitrosomonas* oxidizes ammonia-N (−III) to nitrite-N (+III), for a six electron change. Thus, the equivalent weight for nitrogen in this case is 14/6 = 2.33 grams/equivalent, which means that:

$$f_s = 0.291\ Y_{Nitrosomonas}\ (NH_4^+ \text{ as nitrogen source and electron donor}) \quad (3.25)$$

For *Nitrobacter*, nitrite-N (+III) serves as the electron donor and is oxidized to nitrate-N (+V), for a two electron change. Ammonia-N, however, serves as the nitrogen source. Consequently:

$$f_s = 0.875\ Y_{Nitrobacter}\ (NH_4^+ \text{ as nitrogen source, } NO_2^- \text{ as electron donor}) \quad (3.26)$$

for this organism, where $Y_{Nitrobacter}$ has units of mg biomass COD formed/mg nitrite-N removed. Often nitrifying bacteria are considered together as a group and nitrification is treated as a single reaction converting ammonia-N to nitrate-N. In that case, nitrogen undergoes an eight electron change so that:

$$f_s = 0.219\ Y_A\ (NH_4^+ \text{ as nitrogen source and electron donor}) \quad (3.27)$$

where Y_A represents the true growth yield for autotrophic nitrifying biomass and has units of mg biomass COD formed/mg ammonia-N removed.

Application of the half-reaction technique using typical yield values and Eq. 3.2 provides the mass-based stoichiometric equations for nitrification. For *Nitrosomonas*, when NH_4^+ is the basis, the equation is:

$$NH_4^+ + 2.457\ O_2 + 6.716\ HCO_3^- \rightarrow 0.114\ C_5H_7O_2N + 2.509\ NO_2^- + 1.036\ H_2O + 6.513\ H_2CO_3 \quad (3.28)$$

When NO_2^- is the basis, the equation for *Nitrobacter* is:

$$NO_2^- + 0.001\ NH_4^+ + 0.014\ H_2CO_3 + 0.003\ HCO_3^- + 0.339\ O_2 \rightarrow 0.006\ C_5H_7O_2N + 0.003\ H_2O + 1.348\ NO_3^- \quad (3.29)$$

Furthermore, combining the two reactions reveals that the overall stoichiometry is:

$$NH_4^+ + 3.300\ O_2 + 6.708\ HCO_3^- \rightarrow 0.129\ C_5H_7O_2N + 3.373\ NO_3^- + 1.041\ H_2O + 6.463\ H_2CO_3 \quad (3.30)$$

From these it can be seen that a large amount of alkalinity (HCO_3^-) is used during the oxidation of ammonia to nitrate: 6.708 mg HCO_3^-/mg NH_4^+ removed, which is equivalent to 8.62 mg HCO_3^-/mg NH_4^+-N removed. The vast majority of that alkalinity utilization is associated with neutralization of the hydrogen ions released during the oxidation of ammonia-N. Only a small part of the alkalinity is incorporated into the cell material. If the wastewater contains insufficient alkalinity and if pH control is not practiced, the pH will drop below the normal physiological range, retarding the activity of both the autotrophs and the heterotrophs, hurting system performance. The equations also tell us that considerable oxygen is required for nitrification: 3.30 mg O_2/mg NH_4^+ removed, which is equivalent to 4.33 mg O_2/mg of NH_4^+-N actually oxidized to nitrate-N. Of that amount 3.22 mg O_2 will be used by *Nitrosomonas* and 1.11 will be used by *Nitrobacter*. The oxygen requirement of the nitrifying bacteria can have a significant impact on the total amount of oxygen required by a biochem-

ical operation. Finally, it can be seen that relatively little biomass will be formed, reflecting the low yields associated with autotrophic growth. For every mg of NH_4^+ removed, only 0.129 mg of biomass will be formed, which is equivalent to 0.166 mg biomass/mg NH_4^+-N removed. Most of that, 0.146 mg biomass/mg NH_4^+-N removed, will be due to the growth of *Nitrosomonas*, and only 0.020 mg biomass/mg NH_4^+-N removed will be due to *Nitrobacter*. Overall, the growth of nitrifying bacteria will have little impact on the quantity of biomass in a biochemical operation, but will have a large impact on the oxygen and alkalinity requirements.

3.2.6 Kinetics of Biomass Growth

Equation 3.8 was the COD-based stoichiometric equation for aerobic growth of heterotrophic biomass with ammonia as the nitrogen source. Recognizing that the stoichiometric coefficient on biomass is the same as the true growth yield, Y_H, and that both substrate (S_S) and active heterotrophic biomass ($X_{B,H}$) are measured in COD units, it may be rewritten in terms of the true growth yield as:

$$(1)S_S + [-(1 - Y_H)]S_O \rightarrow Y_H X_{B,H} \tag{3.31}$$

where S_O is oxygen, which is expressed in COD units, and thus carries a negative sign as indicated in Table 3.1.[a] Putting this in the form of Eq. 3.9, while retaining COD units, gives:

$$(-1)S_S + (-1)[-(1 - Y_H)]S_O + Y_H X_{B,H} = 0 \tag{3.32}$$

This equation is based on substrate as the reference constituent. Alternatively, it could be rewritten with active heterotrophic biomass as the reference constituent, and that is the convention used herein:

$$\left(-\frac{1}{Y_H}\right) S_S + (-1)\left[-\left(\frac{1 - Y_H}{Y_H}\right)\right] S_O + X_{B,H} = 0 \tag{3.33}$$

Application of Eq. 3.27 gives:

$$\frac{r_{SS}}{\left(-\frac{1}{Y_H}\right)} = \frac{r_{SO}}{(-1)\left[-\left(\frac{1 - Y_H}{Y_H}\right)\right]} = \frac{r_{XB}}{1} = r \tag{3.34}$$

where [r] = mg COD/(L·hr). Thus, once r_{XB} has been defined, the other rates are also known.

Similar equations can be written for the growth of heterotrophs with nitrate as the terminal electron acceptor and for the aerobic growth of autotrophs. The derivation of such equations is left as an exercise for the reader.

Bacteria divide by binary fission. Consequently, the reaction rate for bacterial growth can be expressed as first order with respect to the active biomass concentration:

$$r_{XB} = \mu \cdot X_B \tag{3.35}$$

---

[a] S represents soluble constituents and X represents particulate constituents, with the subscript denoting the particular constituent involved.

where μ is the specific growth rate coefficient (hr^{-1}). It is referred to as a specific rate coefficient because it defines the rate of biomass growth in terms of the concentration of active biomass present, i.e., the mass of biomass COD formed per unit time per unit of active biomass COD present. Equation 3.35 holds for any type of bacterial growth, regardless of the nature of the electron donor or acceptor, although much of the following is written in terms of heterotrophic biomass growth on an organic substrate. Consequently, subscripts are not used at this point to distinguish between heterotrophic and autotrophic biomass, although they will be used later when it is necessary to make that distinction. Substitution of Eq. 3.35 into Eq. 3.34 defines the rates of substrate removal and oxygen (electron acceptor) utilization associated with biomass growth. It is important to note that the equation for oxygen utilization is also true for other electron acceptors, such as nitrate, as long as the quantity is expressed in oxygen equivalents.

3.2.7 Effect of Substrate Concentration on μ

The Monod Equation. Originally, exponential growth of bacteria was considered to be possible only when all nutrients, including the substrate, were present in high concentration. In the early 1940s, however, it was found that bacteria grow exponentially even when one nutrient is present only in limited amount.[53] Furthermore, the value of the specific growth rate coefficient, μ, was found to depend on the concentration of that limiting nutrient, which can be the carbon source, the electron donor, the electron acceptor, nitrogen, or any other factor needed by the organisms for growth. Since that time, the generality of this observation has been substantiated often, so that it can now be considered to be a basic concept of microbial kinetics.[54] Let us first consider the situation when only an organic substrate is growth limiting.

Figure 3.1 illustrates the relationship that is obtained when μ is measured as a function of a single limiting substrate concentration. A number of different types of experiments can be performed to develop such a relationship and they are discussed in Chapter 8. The important thing to note at this time is that μ initially rises rapidly as the substrate concentration is increased, but then asymptotically approaches a maximum, which is called the maximum specific growth rate, $\hat{\mu}$.

The question of the best mathematical formula to express the relationship shown in Figure 3.1 has been the subject of much debate. No one yet knows enough about the mechanisms of biomass growth to propose a mechanistic equation that will characterize growth exactly. Instead, experimenters have observed the effects of various factors on growth and have then attempted to fit empirical equations to their observations. Consequently, all equations that have been proposed are curve-fits and the only valid arguments for use of one over another are goodness of fit, mathematical utility, and broad acceptance.

The equation with historical precedence and greatest acceptance is the one proposed by Monod (mo nō′).[53] Although his original work was done in batch reactors, it was later extended and refined by workers using continuous cultures of single bacterial species growing on defined media and it was concluded that the curve could be approximated adequately by the equation for a rectangular hyperbola.[53] Consequently, Monod proposed the equation:

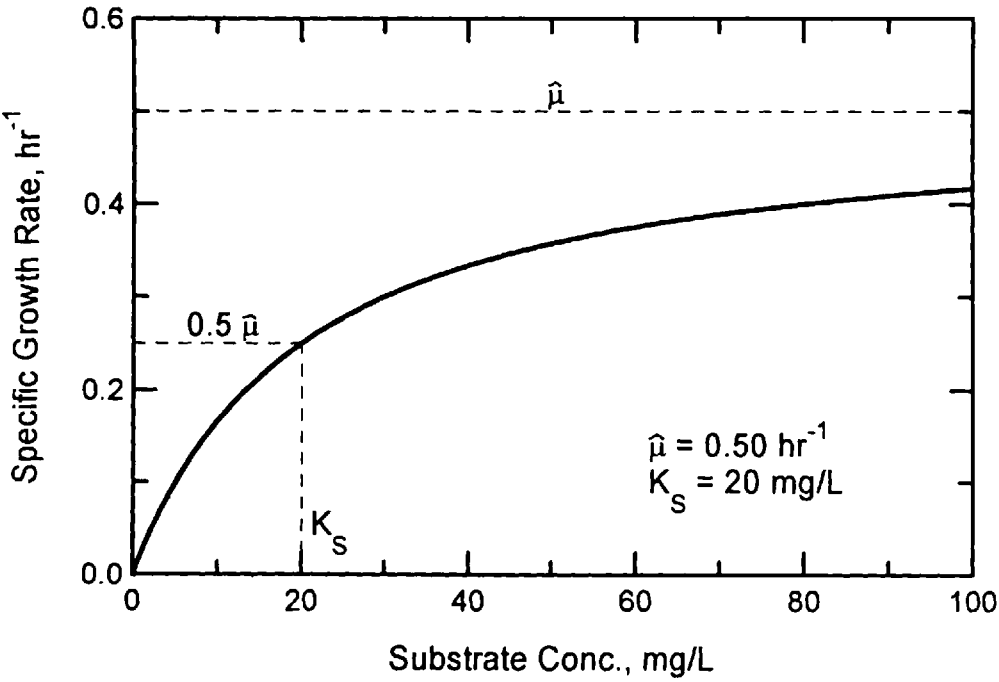


Figure 3.1 Typical plot of the relationship between the specific growth rate coefficient and the concentration of a noninhibitory substrate. The parameter values given were used to construct the curve with the Monod equation (3.36).

$$\mu = \hat{\mu} \frac{S_S}{K_S + S_S} \tag{3.36}$$

where K_S is the half-saturation coefficient. K_S determines how rapidly μ approaches $\hat{\mu}$ and is defined as the substrate concentration at which μ is equal to half of $\hat{\mu}$, as shown in Figure 3.1. The smaller it is, the lower the substrate concentration at which μ approaches $\hat{\mu}$. Because of his pioneering efforts in defining the kinetics of microbial growth, Eq. 3.36 is generally referred to as the Monod equation.

Because of the similarity of Eq. 3.36 to the Michaelis–Menten equation in enzyme kinetics, many people have erroneously concluded that Monod proposed it on mechanistic grounds. While the Michaelis–Menten equation can be derived from consideration of the rates of chemical reactions catalyzed by enzymes, and has a mechanistic basis, the Monod equation is strictly empirical. In fact, Monod himself emphasized its empirical nature.[83]

The Monod equation has been found to fit the data for many pure cultures growing on single substrates, both organic and inorganic, and has been used extensively in the development of models describing the continuous cultivation of microorganisms. It has not been blindly accepted, however, and other workers have proposed alternative equations that fit their data better.[84,97,108] Nevertheless, it is still the most widely used equation.

Because the Monod equation was developed for pure cultures of bacteria growing on single organic substrates, two significant questions arise when its adoption is

considered for modeling biochemical operations for wastewater treatment. The first concerns whether it can be used to express removal of a substrate that is really a mixture of hundreds of organic compounds measured by a nonspecific test like COD, since that is the nature of the organic matter in wastewater. Can the Monod equation adequately describe the effect of biodegradable COD on the specific growth rate of bacteria? The second question arises from consideration of the microbial communities present in wastewater treatment operations. As seen in Chapter 2, those communities are highly complex, containing not only many bacterial species, but higher life forms as well. Can the growth of such a heterogeneous assemblage be expressed simply as "biomass" by the Monod equation? Many researchers have investigated these questions, and it is generally agreed that the answer to both is yes.[8,15,31,36,71] Nevertheless, it should be recognized that the manner in which the culture is grown will have a strong impact on its community structure, and that the values of $\hat{\mu}$ and K_S obtained from mixed culture systems are in reality average values resulting from many interacting species.[19,36,38] Consequently, it has been recommended that $\hat{\mu}$ and K_S be characterized by ranges, rather than by single values, just as was recommended for Y. It can be concluded, however, that the Monod equation is a reasonable model with which to describe the kinetics of microbial growth on complex organic substrates in wastewater treatment systems, and consequently, it is widely used. There are situations, however, in which it would be desirable to model the effects on microbial growth rates of individual organic compounds in complex mixtures. This situation is very complicated,[72] however, and will be covered in Chapter 22.

Simplifications of the Monod Equation. Examination of Eq. 3.36 reveals that two simplifications can be made, and this is often done in the modeling of wastewater treatment systems. First, it can be seen that if S_S is much larger than K_S, the equation may be approximated as:

$$\mu \approx \hat{\mu} \tag{3.37}$$

This is called the zero-order approximation because under that condition the specific growth rate coefficient is independent of the substrate concentration, i.e., it is zero order with respect to S_S, and equal to the maximum specific growth rate coefficient. In other words, the bacteria will be growing as rapidly as possible. Second, if S_S is much smaller than K_S, the term in the denominator may be approximated as K_S and the equation becomes:

$$\mu \approx \frac{\hat{\mu}}{K_S} S_S \tag{3.38}$$

This is called the first-order approximation because μ is first order with respect to S_S. Although Eq. 3.38 is often easier to use than the Monod equation, care should be exercised in its use because serious error can result if S_S is not small relative to K_S. When COD is used as a measure of the total quantity of biodegradable organic matter, K_S can be relatively large, with the result that S_S in activated sludge reactors is often less than K_S. Consequently, Eq. 3.38 is sometimes used to model such systems.

Garrett and Sawyer[35] were the first to propose the use of Eqs. 3.37 and 3.38 because they had observed that the specific growth rate coefficient for bacteria was directly proportional to the substrate concentration at low values and independent of

it at high values. Although they recognized that these two conditions were special cases of the Monod equation, others who adopted their first-order equation incorrectly considered it to be an alternative expression.

Inhibitory Substrates. On occasion, particularly in the treatment of synthetic (xenobiotic) organic compounds in industrial wastewaters, situations are encountered in which the specific growth rate of the microorganisms reaches a maximum and then declines as the substrate concentration is increased, as illustrated in Figure 3.2. Obviously, the Monod equation is not adequate for depicting this situation, and consequently, considerable effort has been expended to determine an appropriate equation.[32,86,104] As with normal, naturally-occurring, noninhibitory (biogenic) substrate, many different models could be used to represent the observed relationship between the substrate concentration and μ, and from a statistical point of view there is little to recommend one over another.[32,104] Consequently, as with the Monod equation, it has been argued that model selection should be based on familiarity and ease of use, leading to a recommendation that an equation based on the enzymatic model of Haldane[46] should be used. Andrews[2] was the first to propose general use of such a function for depicting the effects of inhibitory organic substrates on bacterial growth rates, and thus it will be called the Andrews equation herein. Its form is:

$$\mu = \hat{\mu} \frac{S_S}{K_S + S_S + S_S^2/K_I} \tag{3.39}$$

Examination of Eq. 3.39 reveals that it is similar to the Monod equation, containing only one additional parameter, K_I, the inhibition coefficient. Note that when K_I is very large the Andrews equation simplifies to the Monod equation, demonstrating that $\hat{\mu}$ and K_S have the same meaning in both equations. Unlike the situation for a noninhibitory substrate, however, $\hat{\mu}$ cannot actually be observed and is a hypothetical maximum specific growth rate that would be attained if the substrate were not inhibitory. Furthermore, since $\hat{\mu}$ cannot be observed, K_S also takes on a hypothetical meaning. The most outstanding characteristic of the curve in Figure 3.2 is that μ passes through a maximum, μ^*, at substrate concentration S_S^*, where

$$\mu^* = \frac{\hat{\mu}}{2(K_S/K_I)^{0.5} + 1} \tag{3.40}$$

and

$$S_S^* = (K_S \cdot K_I)^{0.5} \tag{3.41}$$

Equation 3.40 is important because it demonstrates that the degree of inhibition is determined by K_S/K_I, and not just by K_I alone. The larger K_S/K_I, the smaller μ^* is relative to $\hat{\mu}$, and thus, the greater the degree of inhibition. Furthermore, because they are measurable, μ^* and S_S^* are important in the determination of the kinetic parameters for inhibitory substrates. Equation 3.39 has been used widely in the modeling of various wastewater treatment systems, and will be adopted herein for depicting the effect of an inhibitory substrate on the specific growth rate of bacteria degrading it.

Effects of Other Inhibitors. Sometimes one compound may act to inhibit microbial growth on another compound. For example, some organic chemicals are known to inhibit the growth of nitrifying bacteria,[59,122] whereas others inhibit the

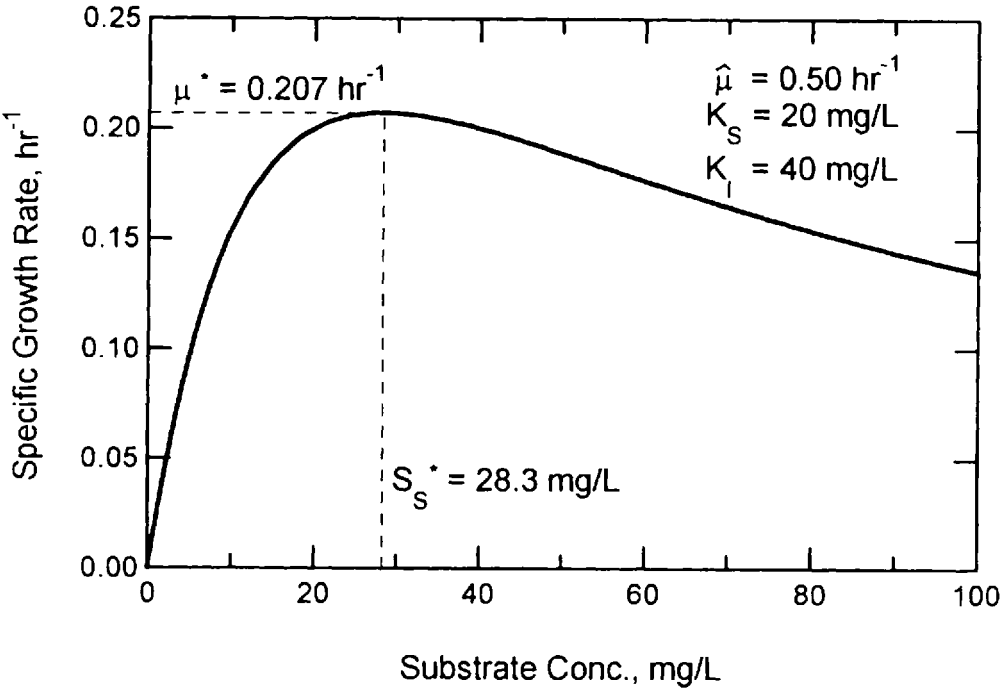


Figure 3.2 Typical plot of the relationship between the specific growth rate coefficient and the concentration of an inhibitory substrate. The parameter values given were used to construct the curve with the Andrews equation (3.39). Note that the values of $\hat{\mu}$ and K_S are the same as in Figure 3.1.

growth of heterotrophic bacteria on biogenic organic matter.[126] In those cases it is necessary for the kinetic expression to depict the effect of the concentration of the inhibitor (S_I) on the relationship between μ and S_S. If the Monod equation can be used to relate μ to S_S in the absence of the inhibitor, then the effect of the inhibitor can be expressed as an effect on $\hat{\mu}$ and/or K_S.[50,125] Several types of inhibitors have been defined by analogy to enzyme inhibition, but all can be modeled by an extension of the Monod model proposed by Han and Levenspiel:[48]

$$\mu = \hat{\mu}\left(1 - \frac{S_I}{S_I^*}\right)^n \left[\frac{S_S}{S_S + K_S(1 - S_I/S_I^*)^m}\right] \tag{3.42}$$

where S_I^* is the inhibitor concentration that causes all microbial activity to cease and m and n are exponents that reflect the impact of increasing inhibitor concentrations on K_S and $\hat{\mu}$, respectively. Equation 3.42 has been used successfully to model the effects of various xenobiotic compounds on the removal of biogenic organic matter.[126] Its use will be discussed in Chapter 22.

3.2.8 Specific Substrate Removal Rate

In earlier sections it was stated that the basis for writing stoichiometric equations was arbitrary and that the reference component was the choice of the investigator.

Thus, it is not surprising that many investigators[71,82,124] have selected substrate removal, rather than biomass growth, as their basic event and have written their rate equations accordingly. Combining Eqs. 3.34 and 3.35 yields:

$$r_{SS} = -(\mu/Y)X_B \tag{3.43}$$

The term μ/Y has been called the specific substrate removal rate and given the symbol q.[43] (Note that the subscript H has been dropped from Y and X_B to emphasize the general nature of Eq. 3.43.) Obviously, q will be influenced by S_S in exactly the same way as μ, and Eqs. 3.37 through 3.42 can be written in terms of it. When this is done, the maximum specific substrate removal rate, $\hat{q}$, is used in place of $\hat{\mu}$, where:

$$\hat{q} = \hat{\mu}/Y \tag{3.44}$$

Both first- and zero-order approximations have been used for the relationship between q and S_S, just as they have for μ. In fact, the ratio of $\hat{q}$ over K_S has been called the mean reaction rate coefficient and given the symbol k_c:[29]

$$k_c = \hat{q}/K_S \tag{3.45}$$

All restrictions that apply to the approximate expressions for the effect of S_S on μ also apply to q.

3.2.9 Multiple Limiting Nutrients

In the broad sense, nutrients can be divided into two categories: complementary and substitutable.[9] Complementary nutrients are those that meet entirely different needs by growing microorganisms. For example, ammonia provides the nitrogen needed for protein synthesis while glucose provides carbon and energy. If either was missing from the growth medium and no substitute was provided, no growth would occur. Substitutable nutrients, on the other hand, are those that meet the same need. For example, ammonia and nitrate can both provide nitrogen whereas glucose and phenol can both provide carbon and energy. Thus, ammonia and nitrate are substitutable for each other, as are glucose and phenol. In this section, we will consider simultaneous limitation of specific growth rate by two complementary nutrients. As stated previously, consideration of the effects of multiple carbon sources, i.e., multiple substitutable nutrients, is very complex,[72] and will be covered in Chapter 22.

In spite of its potential importance in the environment, relatively little is known about how microorganisms respond to simultaneous limitation by two or more complementary nutrients.[9] Because the uncertainty increases greatly as the number of nutrients involved increases, we will limit our considerations to only two.

Interactive and Noninteractive Relationships. Consider two complementary nutrients, S_{S1} and S_{S2}. Both are required for biomass growth and are present at low concentration in the environment in which the biomass is growing. Which will control the specific growth rate? Two different philosophies have been developed to answer this question, and the models representing them have been classified as interactive and noninteractive.[6]

An interactive model is based on the assumption that two complementary nutrients can both influence the specific growth rate at the same time. If both are

required for growth and are present at concentrations equal to their half-saturation coefficients, then each alone can reduce μ to one-half of $\hat{\mu}$. However, since both effects are occurring simultaneously, the result would be to reduce μ to one-fourth of $\hat{\mu}$. The most common type of interactive model in use is the multiple Monod equation:[7,113]

$$\mu = \hat{\mu}\left(\frac{S_{S1}}{K_{S1} + S_{S1}}\right)\left(\frac{S_{S2}}{K_{S2} + S_{S2}}\right) \tag{3.46}$$

Any time the concentrations of S_{S1} and S_{S2} are such that both $S_{S1}/(K_{S1} + S_{S1})$ and $S_{S2}/(K_{S2} + S_{S2})$ are less than one, they both act to reduce μ below $\hat{\mu}$. This has two impacts. First, for a given value of S_{S1}, μ will be lower when S_{S2} is also limiting than it would be if S_{S2} were present in excess. Second, there is not a unique value of μ associated with a given value of S_{S1} or S_{S2}, as there was with Eq. 3.36. Rather, it depends on both.

A noninteractive model is based on the assumption that the specific growth rate of a microbial culture can only be limited by one nutrient at a time. Therefore, μ will be equal to the lowest value predicted from the separate single-substrate models:[7,113]

$$\mu = \min\left(\frac{\hat{\mu}S_{S1}}{K_{S1} + S_{S1}}, \frac{\hat{\mu}S_{S2}}{K_{S2} + S_{S2}}\right) \tag{3.47}$$

If $S_{S1}/(K_{S1} + S_{S1}) < S_{S2}/(K_{S2} + S_{S2})$, nutrient S_{S1} is rate limiting, and vice versa. If $S_{S1}/(K_{S1} + S_{S1}) = S_{S2}/(K_{S2} + S_{S2})$, then both are rate limiting, but that occurs only under special conditions. In the noninteractive conceptualization, the normal Monod equation (Eq. 3.36) would apply for whichever nutrient was rate limiting and the concentration of the other would have no impact on μ.

Only limited experimental evidence is available to support one model over the other. Bae and Rittmann[7] have shown both theoretically and experimentally that the interactive model is more appropriate when the two limiting constituents are the electron donor and acceptor. Furthermore, Bader[6] has compared the mathematical characteristics of the two expressions. The noninteractive model, by its very nature, causes a discontinuity at the transition from one nutrient limitation to another. It also predicts significantly higher growth rates in the region where S_{S1}/K_{S1} and S_{S2}/K_{S2} are small. The interactive model does not cause discontinuities, but may err on the side of predicting lower growth rates when S_{S1}/K_{S1} and S_{S2}/K_{S2} are both small. Both functions become asymptotically the same if either nutrient is present in excess. Finally, the interactive model is mathematically preferable for modeling dynamic situations because it is continuous.

Equation 3.46, the interactive model, will be adopted for use herein. There are three reasons for this choice. First is the evidence provided by Bae and Rittmann.[7] Second, for the type of situation likely to be encountered in biochemical operations for wastewater treatment, the interactive model is more conservative. Third, it works well when one nutrient is the electron donor (the substrate) and the other is the electron acceptor (oxygen or nitrate),[105,113] a common occurrence in wastewater treatment systems.

A special case of multiple nutrients occurs when an increase in the concentration of one nutrient acts to diminish microbial activity. For example, consider the

growth of heterotrophic bacteria under anoxic conditions. Because nitrate reduction can serve as an alternative to aerobic respiration, the enzymes involved in the transfer of electrons to nitrate and its reduced products are influenced negatively by the dissolved oxygen concentration and consideration must be given to this fact when expressing the kinetics of growth under anoxic conditions. Oxygen can have two effects: (1) it can repress the synthesis of denitrifying enzymes, and (2) it can inhibit their activity.[24,67,94,118] Although there are exceptions, as a general rule the presence of oxygen in the medium (and/or its active utilization as the terminal electron acceptor) represses the synthesis of the nitrate reducing enzyme system. When oxygen is absent, or is present in amounts that are insufficient to meet the needs of the culture, derepression occurs and the enzymes are synthesized. Complications occur, however, when the biomass is cycled between aerobic and anoxic conditions and this appears to alter the regulatory system so that some enzyme synthesis can continue at diminished rates even in the presence of dissolved oxygen.[112] The effect of oxygen on the activity of the enzymes depends on the bacterial species involved. In some, the activities are diminished in the presence of oxygen, whereas in others they are not. Nevertheless, it appears that inhibition of enzyme activity by oxygen is the primary mechanism influencing nitrate reduction rates in systems in which the bacteria are continually cycled between aerobic and anoxic conditions,[112] and that prior growth under anoxic conditions will provide enzymes which can function at diminished rate even in the presence of dissolved oxygen. One factor complicating the determination of the effects of oxygen on nitrate reduction in wastewater treatment systems is the necessity to grow the bacteria as flocculent cultures or as biofilms. Because diffusion is the only mechanism supplying oxygen to the bacteria in the interior of a floc particle or biofilm, some bacteria may be in an environment completely devoid of oxygen even when oxygen is present in the bulk liquid.[68]

Because of the complexity associated with the effects of dissolved oxygen on anoxic growth of heterotrophic bacteria, and because all effects have not been clearly defined, relatively simple models have been used to express them.[11,54,55] A popular approach has been to use Eq. 3.46 to depict the simultaneous effects of organic substrate (S_S), and nitrate (S_{NO}) on μ and to add a third term which diminishes μ as the dissolved oxygen concentration, S_O, increases:

$$\mu = \hat{\mu}\left(\frac{S_S}{K_S + S_S}\right)\left(\frac{S_{NO}}{K_{NO} + S_{NO}}\right)\left(\frac{K_{IO}}{K_{IO} + S_O}\right) \tag{3.48}$$

The third term is the function most commonly used to depict the effects of a classical noncompetitive inhibitor as modeled in enzyme kinetics.[125] The parameter K_{IO} is the inhibition coefficient for oxygen.

Implications of Multiple Nutrient Limitation. Biochemical operations are designed on the premise that there is a functional relationship between the specific growth rate of biomass and the concentration of the growth-limiting nutrient in a bioreactor. Because of that relationship, if engineering control can be exerted over the specific growth rate, it will be possible to control the concentration of the growth-limiting nutrient leaving the bioreactor. This can only be achieved, however, if the nutrient the engineer wishes to control is the growth-limiting one. If the design objective is the removal of soluble organic matter, then all other nutrients must be supplied in excess. Or, if the goal is to remove nitrate-N by allowing it to serve as

the terminal electron acceptor, then it should be made rate limiting at the appropriate place in the process. A clear definition of the objective to be met must be combined with knowledge of the concentrations of the various constituents in the wastewater to ensure that the resultant biochemical operation can indeed meet that objective.

Because oxygen is a gas of very low solubility, it must be supplied continuously to aerobic systems and the concentration in solution will depend on the relative rates of supply and utilization. Furthermore, because oxygen transfer is one of the major costs associated with aerobic wastewater treatment, it is uneconomic to oversize the oxygen delivery system. As a consequence, it is not uncommon for the oxygen concentration to decrease sufficiently to make $S_O/(K_O + S_O) < 1.0$ (where K_O is the half-saturation coefficient for dissolved oxygen). Thus, it would be instructive to examine the impact of this occurrence. Figure 3.3 illustrates the simultaneous limitation of the specific growth rate of autotrophic nitrifying bacteria by ammonia (the electron donor) and oxygen (the electron acceptor) using typical parameter values. These bacteria were chosen because they are more sensitive to dissolved oxygen concentration than heterotrophic bacteria, ($K_{O,H} < K_{O,A}$, where the subscripts H and A signify heterotrophic and autotrophic bacteria, respectively). Examination of Figure 3.3 reveals two things. First, if we could operate a bioreactor in a way that maintained a constant specific growth rate, decreasing the oxygen concentration in the bioreactor would cause the ammonia concentration to increase. Second, decreasing the oxygen concentration is analogous to decreasing $\hat{\mu}$ for the bacteria. This can also be seen

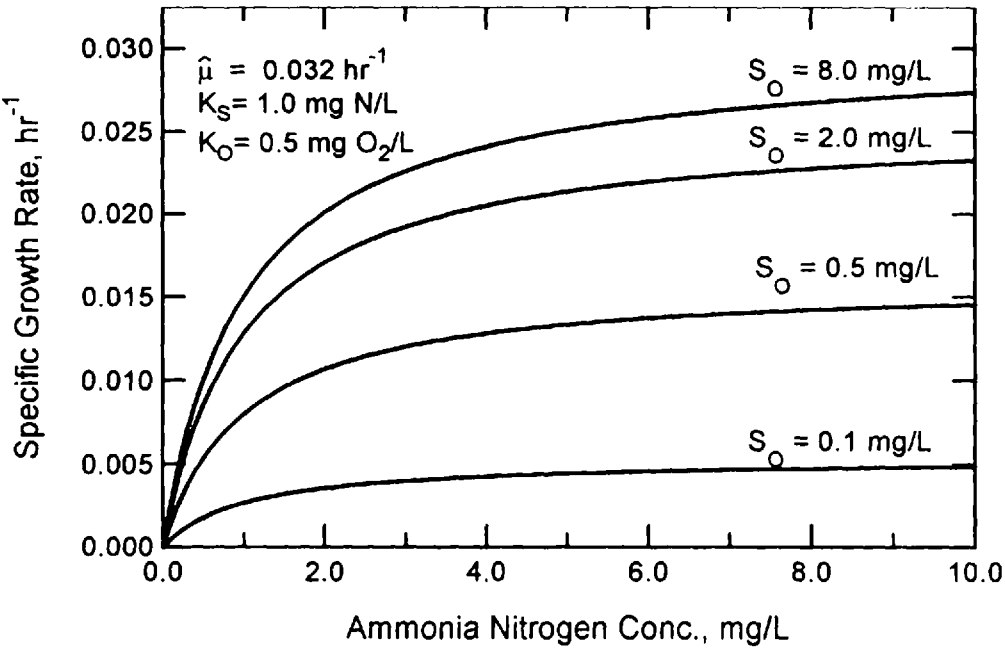


Figure 3.3 Double Monod plot showing the effects of both ammonia nitrogen and dissolved oxygen concentrations on the specific growth rate of autotrophic nitrifying bacteria. The parameter values given were used to construct the curves with Eq. 3.46.

by examining Eq. 3.46. The consequence of this is discussed in more detail in Chapter 6, but suffice it to say now that a decrease in $\hat{\mu}$ makes it more difficult for the autotrophic bacteria to compete for space in the bioreactor.

Both nitrogen and phosphorus are required for the synthesis of new biomass. If those proper quantities are not present, balanced biomass growth cannot occur and treatment performance will be impaired. Thus, care must be exercised to provide sufficient quantities. We have just seen, however, that if the concentrations of essential nutrients are very low in a bioreactor they can become rate limiting, which is undesirable when the treatment objective is removal of organic matter. This means that the concentration of nitrogen or phosphorus supplied to a bioreactor must be sufficiently high to meet the synthesis needs of the biomass as defined by stoichiometry while leaving enough residual in solution to prevent their concentrations from being rate limiting. Goel and Gaudy[30] determined that K_S for ammonia nitrogen during normal heterotrophic growth lies between 1.5 and 4.0 mg/L as N. Using 0.50 hr^{-1} as a representative value for $\hat{\mu}$, it can be shown that if the influent nitrogen concentration exceeds the stoichiometric requirement by 1.0 mg/L as N, nitrogen will not be rate limiting to heterotrophic biomass at the specific growth rates normally employed in wastewater treatment. Although some work has been done on kinetic limitation of heterotrophs by phosphorus, the results are not as clear as those with nitrogen. Attempts to measure the limiting phosphorus concentration in both pure and mixed microbial cultures found it to be too low to detect with the techniques available at the time.[107] Consequently, if the concentration of phosphorus in the influent exceeds the stoichiometric amount by a few tenths of a mg/L as P, phosphorus should not be rate limiting. In some biochemical operations, the microorganisms pass through a growth cycle, and nutrients will be taken up in one phase and released in another. To prevent nutrient limitation during the phase of nutrient uptake, the amounts presented above should be in excess of the maximum quantity removed, not the net amount as determined by the final effluent.

3.2.10 Representative Kinetic Parameter Values for Major Microbial Groups

Aerobic Growth of Heterotrophic Bacteria. The values of the parameters $\hat{\mu}_H$ and K_S are very dependent on the organism and substrate employed. If an axenic bacterial culture is grown on each of several substrates under fixed environmental conditions, the values of $\hat{\mu}_h$ and K_S will vary from substrate to substrate. Likewise, if the same substrate is fed to each of several pure cultures, the values of $\hat{\mu}_H$ and K_S will depend on the species of organism. This makes it very difficult to generalize about parameter values and care should be exercised in the use of values considered to be typical. It can be stated, however, that readily biodegradable substrates are characterized by high values of $\hat{\mu}_H$ and low values of K_S, whereas slowly biodegradable substrates have low $\hat{\mu}_H$ values and high K_S values. For example, benzoic acid had $\hat{\mu}_H$ values between 0.61 and 0.64 hr^{-1} and K_S values between 4.2 and 5.8 mg/L as COD, whereas 2-chlorophenol had values of 0.020–0.025 hr^{-1} and 16–17 mg/L as COD for the two parameters.[23] Even lower K_S values have been reported for very easily degradable substrates, such as biogenic materials like carbohydrates and amino acids, with values as low as 0.2 mg/L for galactose and 0.5 mg/L for

glutamic acid.[21] This means that degradation of many biogenic substrates may behave in a zero-order manner over a broad range of substrate concentrations.

Wastewaters usually contain complex mixtures of organic compounds and the total concentration of biodegradable soluble organic matter is commonly characterized by the COD concentration. When K_S is measured on such mixtures using the COD concentration, the values are generally one to two orders of magnitude higher than they are for single substrates expressed as COD. For example, poultry and soybean processing wastewater have been reported to have K_S values of 500 and 350 mg/L as five day biochemical oxygen demand (BOD_5), which is another measure of biodegradable organic matter.[64] Thus, as a whole, overall removal of organic matter in wastewater treatment systems may behave in a first-order manner even though the removal of individual constituents may be zero-order.[121]

Domestic wastewater is perhaps the most common example of a complex substrate, and because of its ubiquity there has been considerable interest in characterizing its biodegradation kinetics. As one might expect from the discussion above, considerable variation in the parameter values has been reported, with $\hat{\mu}_H$ ranging from 0.12 to 0.55 hr^{-1} and K_S from 10 to 180 mg/L as COD.[54,55] An important characteristic of domestic wastewater that has only recently been recognized is that the organic component can be divided into readily and slowly biodegradable fractions, greatly improving the ability of mathematical models to mimic process performance.[26] Use of this division should decrease the range of values observed. As a consequence, values of 0.25 hr^{-1} and 20 mg/L as COD have been adopted as representative of the $\hat{\mu}_H$ and K_S values for the readily biodegradable fraction.[55]

The microbial communities in wastewater treatment systems are complex, containing many microbial species, and the relative predominance of the species depends on the physical configuration of the system. Therefore, since the values of $\hat{\mu}_H$ and K_S are species dependent, it follows that their values in mixed culture systems will depend on the bioreactor configuration. For example, reactors that subject the microorganisms to variations in substrate concentrations from very high to very low tend to select species that can grow rapidly (higher $\hat{\mu}_H$), whereas reactors which maintain a low, uniform substrate concentration throughout select microorganisms that are good scavengers of substrate (low K_S).[21,25] This complicates kinetic analysis and requires that experiments to determine kinetic parameters be conducted with systems that mimic the physical configuration to be employed in the full-scale facility. This topic is discussed in more detail in Chapter 8.

The biodegradation kinetics for many xenobiotic compounds can best be characterized by the Andrews equation (Eq. 3.39). Dividing both the numerator and denominator by K_S yields:

$$\mu = \hat{\mu} \frac{S_S/K_S}{1 + S_S/K_S + (S_S/K_S)^2(K_S/K_I)} \tag{3.49}$$

Expressing the equation in this manner emphasizes that the degree of substrate inhibition is determined by the ratio of K_S/K_I, rather than by K_I alone, as we saw with Eq. 3.40. Furthermore, Eq. 3.49 also makes it easy to see that the larger the ratio, the more inhibitory the substrate. 1,3- and 1,4-Dichlorobenzene are both moderately inhibitory compounds and have ratios of 0.14 and 0.08, respectively.[42]

In Section 3.2.9, the undesirability of oxygen being rate limiting was discussed, suggesting that knowledge of the oxygen half-saturation coefficient for heterotrophs,

$K_{O,H}$, is important. In spite of that, relatively little work has been done to estimate $K_{O,H}$ values for mixed microbial cultures, probably because population shifts occur in the community in response to changes in the dissolved oxygen concentration, making estimation of the value difficult. Nevertheless, limited pure culture data suggests that $K_{O,H}$ is very low. For example, values of 0.01, 0.08, and 0.15 mg O_2/L have been reported for *Sphaerotilus natans*[70] (a filamentous bacterium), *Candida utilis*[113] (a yeast), and *Citrobacter* sp.[70] (a floc-forming bacterium), respectively. This suggests that dissolved oxygen concentrations must be very low before they have serious impacts on the growth of heterotrophic bacteria, although they may influence the competition between filamentous and floc-forming bacteria. For depicting the impacts of dissolved oxygen on general heterotrophic biomass growth, one group adopted a value of 0.2 mg O_2/L for $K_{O,H}$.[55]

It will be recalled from Section 3.2.8 that many investigators use substrate removal, rather than biomass growth, as the primary event with which to characterize biochemical operations. In that case, the primary kinetic parameter is the maximum specific substrate removal rate, $\hat{q}$, rather than the maximum specific growth rate. Equation 3.44 defined $\hat{q}$ as $\hat{\mu}/Y$. Thus, $\hat{q}$ will be influenced by variations in Y as well as variations in $\hat{\mu}$. Like $\hat{\mu}$, Y is influenced both by the substrate being degraded and the microorganism performing the degradation (see Section 2.4.1). It should be noted, however, that Y is a reflection of the energy available in a substrate whereas $\hat{\mu}$ is a reflection of how rapidly a microorganism can process that energy and grow. Because they represent different characteristics, there is no correlation between the two parameters. For example, some substrates that are degraded very slowly (low $\hat{\mu}$) provide more energy to the degrading culture (i.e., higher Y) than do substrates that are degraded rapidly.[42] This suggests that deductions about the variability in $\hat{q}$ cannot be made from data on $\hat{\mu}$ alone, and vice versa. Knowledge of the true growth yield is also important. Typical Y values are discussed in Section 2.4.1.

Anoxic Growth of Heterotrophic Bacteria. As seen in Chapter 2, the only difference between aerobic and anoxic growth of heterotrophic bacteria on many substrates, such as biogenic organic matter, is the nature of the terminal electron acceptor and its impact on the amount of ATP that the cells can generate. Thus, for substrates for which this is true, we might expect the kinetic parameters describing growth under the two conditions to be very similar, and that is exactly what has been observed. When mixed microbial cultures were grown with excess oxygen or nitrate as the terminal electron acceptors and peptone as the rate-limiting substrate, the values of $\hat{\mu}_H$ and K_S were very similar, being 0.14 hr^{-1} and 67 mg/L as COD, respectively, under aerobic conditions and 0.13 hr^{-1} and 76 mg/L as COD under anoxic conditions.[79] Furthermore, as expected from the lower potential ATP formation under anoxic conditions, the anoxic yield was lower, being only 0.39 mg biomass COD/mg substrate COD versus a value of 0.71 aerobically. Consequently, $\hat{q}_H$ was almost twice as large under anoxic conditions. Although data directly comparing kinetic parameters under aerobic and anoxic conditions are limited, experience with treatment systems suggest that these findings are generally true.[79]

Anoxic growth conditions are generally imposed in biochemical operations for the purpose of reducing the nitrate concentration to low levels. Thus, there is a possibility that the terminal electron acceptor concentration will become rate limiting. Proper modeling of this situation requires knowledge of K_{NO}, the half-saturation

coefficient for nitrate. As with oxygen, the half-saturation coefficient for nitrate as the terminal electron acceptor has been found to be low, with values around 0.1 to 0.2 mg/L as N being reported.[20,33,93] Consequently, values in that range have been adopted by investigators conducting modeling studies.[11,25]

Another parameter required to fully define the kinetics of microbial growth under anoxic conditions is K_{IO}, the oxygen inhibition coefficient used in Eq. 3.48. If the cells are growing in a dispersed state so that all are exposed to the oxygen concentration in the bulk liquid, it appears that they do not denitrify when the dissolved oxygen concentration is above 0.1 to 0.2 mg/L.[101] However, when they grow as aggregates or films, the requirement for oxygen transport by diffusion allows biomass in the interior to be free of oxygen even when the bulk liquid contains it. Consequently, anoxic growth will occur even when the dissolved oxygen concentration in the bulk liquid exceeds 0.2 mg/L.[101] Thus, modelers have assumed values for K_{IO} ranging from 0.2[55] to 2.0[11] mg/L.

Aerobic Growth of Autotrophic Bacteria. The nitrifying bacteria are the most important aerobic autotrophs and for the nitrogen levels normally found in domestic wastewater the kinetics of their growth can be adequately represented by the Monod equation (Eq. 3.36). Because only a limited number of genera and species are involved, the variability in the values of the kinetic parameters is less than that associated with heterotrophs. The maximum specific growth rate coefficient for *Nitrosomonas* has been reported to lie between 0.014[71] and 0.092[109] hr^{-1}, with a value of 0.032 hr^{-1} considered to be typical at 20°C.[102] The half-saturation coefficient for ammonia has been reported to be between 0.06 and 5.6 mg/L as N,[109] but a commonly accepted value is 1.0 mg/L.[55,102] The maximum specific growth rate coefficient for *Nitrobacter* is similar to that for *Nitrosomonas*, having been reported to lie between 0.006[71] and 0.060[109] hr^{-1}. Likewise, the value considered to be typical,[102] 0.034 hr^{-1}, is similar to that for *Nitrosomonas*. The reported range of the half-saturation coefficient for *Nitrobacter* is slightly larger than that for *Nitrosomonas*, being 0.06 to 8.4 mg/L as nitrite-N,[109] as is the value thought to be typical, 1.3 mg/L.[102] The maximum specific growth rate coefficients for the autotrophic bacteria are considerably less than those for heterotrophic bacteria, reflecting their more restricted energy yielding metabolism and the fact that they must synthesize all cell components from carbon dioxide. This suggests that special consideration must be given to their requirements during design of reactors in which both carbon oxidation and nitrification are to occur. Although the half-saturation coefficients for the autotrophs are less than the reported values for heterotrophs growing on complex substrates, they are similar to the values reported for heterotrophs growing on single organic compounds. As a consequence of their small size, the kinetics of nitrification will behave in a zero-order manner over a broad range of ammonia and nitrite concentrations. As will be seen later, this has a significant impact on bioreactor performance.

A major difference in the growth characteristics of heterotrophic and autotrophic biomass is the greater sensitivity of the latter to the concentration of dissolved oxygen. Whereas the value of the half-saturation coefficient for oxygen is very low for heterotrophs, the values for the two genera of autotrophs are sufficiently high in comparison to typical dissolved oxygen concentrations that dual nutrient limitation as expressed by Eq. 3.46 should be considered to be the norm. For example, values of $K_{O,A}$ for both *Nitrosomonas* and *Nitrobacter* have been reported to lie between 0.3 and 1.3 mg/L.[109] Measurements which considered the effects of diffusional re-

sistance on half-saturation coefficients have suggested that the true values lie near the lower end of the range,[111,120] and values of 0.50 and 0.68 have been adopted as typical for *Nitrosomonas* and *Nitrobacter*, respectively, in systems in which some diffusional resistance will occur.[102]

Another difference between heterotrophic and autotrophic biomass is the greater sensitivity of the latter to changes in pH. Although all bacteria grow poorly outside of the normal physiological pH range of 6.0 to 8.0, nitrifying bacteria are particularly sensitive to pH, especially *Nitrosomonas*, as shown in Figure 3.4.[98] There it can be seen that the rate reaches a maximum at a pH of about 8, and declines sharply for lower pH values. A wide range of pH optima has been reported,[116] but most workers agree that as the pH becomes more acid the rate of ammonia oxidation declines.[93] Furthermore, if a culture is acclimated to a low pH the effect is less severe than if the pH is suddenly shifted. Siegrist and Gujer[111] have modeled the effect in Figure 3.4 with Eq. 3.50:

$$\hat{\mu}_A = \hat{\mu}_{Am}[1 + 10^{(0.5\ pH)}]^{-1} \tag{3.50}$$

where $\hat{\mu}_{Am}$ is the maximum specific growth rate at the optimum pH. It should be

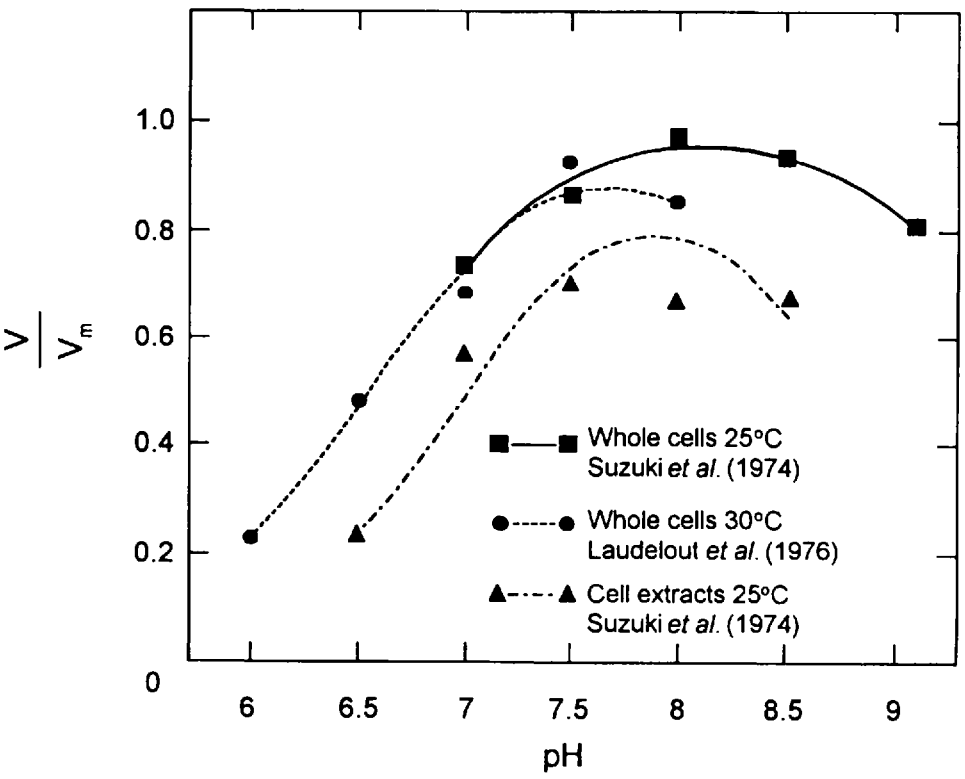


Figure 3.4 Effect of pH on the maximal activity of *Nitrosomonas*. The listed references are cited in 98. (From A. V. Quinlan, Prediction of the optimum pH for ammonia-N oxidation by *Nitrosomonas europaea* in well-aerated natural and domestic-waste waters. *Water Research* **18**:561–566, 1984. Copyright © Elsevier Science Ltd.; reprinted with permission.)

noted that this equation only predicts the decline in rate at low pH and does not predict the observed drop-off at pH above 8.5. This is not generally a problem, however, because the release of hydrogen ions during nitrification acts to depress the pH so that values in excess of 8.5 are seldom encountered. There is less agreement concerning the effects of pH on *Nitrobacter*. For example, Boon and Laudelot[13] have suggested that their maximum specific growth rate is independent of pH over the range between 6.5 and 9, whereas others[115] have shown a strong pH dependence. Because of this, and because the growth of *Nitrosomonas* is generally thought to be rate controlling, most investigators do not model the effect of pH on *Nitrobacter*.

The necessity for employing equations like 3.50 is due in part to the way in which the Monod equation is normally written for nitrifying bacteria. Although, ammonia and nitrite are both ionizable species, the Monod equation is normally written in terms of the total ammonia or nitrite concentration, without regard for the ionization state. However, the nonionized form of ammonia (free ammonia) is thought to be the actual substrate for *Nitrosomonas*,[98] and it is possible that undissociated nitrous acid is the substrate for *Nitrobacter*. For a given total ammonia concentration, the concentration of the nonionized form will change as the pH is changed, thereby making $\hat{\mu}$ as normally defined an apparent function of pH. A more direct approach would be to write the kinetic expression directly in terms of the true substrate and this has been done for *Nitrosomonas*.[98] However, because this approach is more complex than combining Eq. 3.36 with Eq. 3.50 to reflect the effect of pH, the latter is more commonly used at nitrogen concentrations normally found in domestic wastewaters.

Free ammonia and undissociated nitrous acid become more of a problem at high nitrogen concentrations because they both act as inhibitory substrates as their concentrations are increased.[1,4,13] Furthermore, free ammonia can also inhibit nitrite oxidation to nitrate.[4] This suggests that there are complex relationships between the total ammonia and nitrite concentrations, the pH, and the activity of both groups of nitrifying bacteria. Although these relationships become very important when wastewaters containing high concentrations of nitrogen are being treated, no kinetic relationships are available to depict all of the effects, although some are available for ammonia oxidation.[88] Nevertheless, it is important to recognize that the simple Monod equation is not adequate to depict the kinetics of nitrification when the concentration of ammonia exceeds that normally found in domestic wastewater (around 30 to 40 mg/L as N) and that alternative expressions should be sought.

Because of the autotrophic nature of nitrifying bacteria the concept developed that organic compounds display a general toxicity toward them. That this concept is fallacious has been demonstrated in pure[100] and mixed[58,60] cultures. Nitrification can proceed at rapid rates in the presence of organic matter, provided that other environmental factors, such as pH and dissolved oxygen concentration, are adequate. In fact, under some circumstances, the presence of biogenic organic matter can even enhance the rate of nitrification.[60] There are some organic compounds that are inhibitory, however, and act to decrease the specific growth rate of nitrifying bacteria. The most potent specific inhibitors of nitrification are compounds that chelate metals[60] and contain amine groups,[59] some of which are capable of decreasing the nitrification rate by 50% at concentrations of less than 1.0 mg/L. Furthermore, it appears that *Nitrosomonas* is the weak link in the nitrification chain, being more susceptible than *Nitrobacter* to organic inhibitors.[59] Many inhibitors have been shown to act in

a noncompetitive manner against nitrifiers,[91,92] allowing an equation like Eq. 3.48 to be used to depict their effect:

$$\mu_A = \hat{\mu}_A \left(\frac{S_{NH}}{K_{NH} + S_{NH}}\right)\left(\frac{K_I}{K_I + S_I}\right) \tag{3.51}$$

where S_{NH} is the ammonia-N concentration, K_{NH} is the half-saturation coefficient for ammonia-N, S_I is the concentration of the inhibitor, and K_I is the inhibition coefficient. As might be expected, K_I is very small for some compounds,[92] denoting extreme inhibition. Although many inhibitors of *Nitrosomonas* act in a noncompetitive manner, methane and ethylene act as competitive inhibitors.[95] This is because they are similar in size to ammonia and compete directly with it for the active site on the enzyme that initiates ammonia oxidation. Halogenated hydrocarbons act in a noncompetitive manner, but many are also reactive with the enzyme and can lead to products that damage the cell, thereby making their effects worse than simple inhibition.

There have also been suggestions in the literature that the presence of heterotrophic bacteria is deleterious to the activity of nitrifying bacteria, but this has been shown to be false.[12,60] Any effect of heterotrophs is indirect, such as a decrease in dissolved oxygen concentration or an alteration of pH. Because of the sensitivity of autotrophs to these factors, care must be given to the design of facilities in which autotrophs and heterotrophs share the same space.

Anaerobic Cultures. As discussed in Section 2.3.2 and seen in Figure 2.3, in anaerobic operations three groups of bacteria are involved in acidogenesis and two in methanogenesis. Fermentative bacteria convert amino acids and simple sugars to acetic acid, volatile acids, and a minor amount of H_2. Bacteria performing anaerobic oxidation convert long chain fatty acids and volatile acids to acetic acid and major amounts of H_2. Finally H_2-oxidizing acetogens form acetic acid from carbon dioxide and H_2, but they are considered to be of minor importance in anaerobic wastewater treatment operations and will not be considered here. The two groups of methanogens are aceticlastic methanogens, which split acetic acid into methane and carbon dioxide, and H_2-oxidizing methanogens, which reduce carbon dioxide.

To have a complete picture of the kinetics of microbial growth and substrate utilization in anaerobic systems, the kinetic parameters for all groups should be characterized. Unfortunately, because of the role of H_2 in regulating microbial activity and the close association between H_2-producing and H_2-consuming bacteria, this is not an easy task. For this reason and because the complex interactions among the microbial groups have only recently been recognized, most kinetic studies of anaerobic treatment processes have measured rates associated with entire communities rather than individual groups. That literature is too extensive to include here, but reviews[53,114] provide good summaries and the reader is encouraged to consult them for overall kinetic information.

As our understanding of the interactions in anaerobic processes has increased, engineers have sought to model anaerobic systems on a more fundamental level by including reaction steps for each important microbial group.[14,48] Although those efforts represent first attempts at expressing the kinetics of these complex systems, they provide information that is helpful in developing an appreciation of the kinetic characteristics of anaerobic bacteria. Because a temperature of 35°C is commonly

used for anaerobic operations, the following parameter values are for that temperature range. Fermentative bacteria (group 2 in Figure 2.3) grow relatively rapidly on amino acids and simple sugars, and their kinetics can be represented by the Monod equation (Eq. 3.36) with a $\hat{\mu}$ value on the order of 0.25 hr^{-1} and a K_S value around 20 to 25 mg/L as COD. Review of available data suggests that this reaction does not limit system performance.[45] The bacteria which oxidize long chain fatty acids (group 3 in Figure 2.3) grow more slowly than the fermentative bacteria and are subject to inhibition by H_2. The values of $\hat{\mu}$ and K_S depend on the degree of saturation of the fatty acid serving as growth substrate, with saturated acids having lower $\hat{\mu}$ and K_S values than unsaturated ones.[45] Nevertheless, Bryers[14] has adopted a $\hat{\mu}$ value of 0.01 hr^{-1} and a K_S value of 500 mg/L as COD as being representative of the entire group. Reaction 4 in Figure 2.3 represents the bacteria that degrade short chain fatty acids, such as propionic and butyric acids. Butyric acid appears to be degraded in a manner similar to that of the long chain fatty acids and bacteria growing on it have kinetic parameters similar to those in group 3. Propionic acid, on the other hand, is degraded by more specialized bacteria which grow more slowly. Gujer and Zehnder[45] reported $\hat{\mu}$ and K_S values of 0.0065 hr^{-1} and 250 mg/L as COD, respectively, whereas Bryers[14] chose values of 0.0033 hr^{-1} and 800 mg/L as COD based on other studies. Although the two sets of values differ somewhat in magnitude, they both suggest that growth on propionic acid is much slower than growth on other fatty acids. Aceticlastic methanogenesis (reaction 6 in Figure 2.3) is a very important reaction in anaerobic operations because it produces about 70% of the methane. Two major types of aceticlastic methanogenic bacteria can be present in anaerobic systems, but the one which predominates will depend on the bioreactor conditions imposed because their growth kinetics are quite different. *Methanosarcina* can grow rapidly, but do not have a high affinity for acetic acid. Representative parameter values for them are 0.014 hr^{-1} for $\hat{\mu}$ and 300 mg/L as COD of acetic acid for K_S.[106] *Methanosaeta* (formerly *Methanothrix*), on the other hand, grow more slowly, but have a higher affinity for acetic acid, as shown by a $\hat{\mu}$ value of 0.003 hr^{-1} and a K_S value of 30 to 40 mg/L as COD of acetic acid.[106] Finally, the H_2-oxidizing methanogens produce methane from H_2, thereby keeping the H_2 concentration low and allowing the H_2-producing reaction to proceed as discussed in Section 2.3.2. The kinetic parameters for their growth have been reported to be $\hat{\mu}$ = 0.06 hr^{-1} and K_S = 0.6 mg/L as COD of dissolved H_2,[45] although others have reported K_S to be in the range of 0.03–0.21 mg/L as COD of dissolved H_2.[128]

The pH of an anaerobic system has a strong impact on $\hat{\mu}$, with an optimum around pH 7. Just as with nitrifying bacteria, this is probably because the nonionized form of the substrate (fatty acids in this case) serves as the actual substrate for growth and the amount of nonionized form will depend on the pH. As a consequence, relationships between $\hat{\mu}$ and pH are needed for the major groups of bacteria. On the other hand, some[2,27] have modeled acetic acid utilization with the Andrews equation using nonionized acetic acid as the substrate, but it is unclear whether that equation should be used if $\hat{\mu}$ is made an explicit function of pH. In addition, the role of H_2 in regulating the utilization of propionic and butyric acids and the activity of the H_2-producing bacteria is very important, but is not reflected in the parameter values reported above, which are all for low H_2 levels. Bryers[14] has argued that the H_2 effect is based on thermodynamics, and as such, does not translate directly into kinetic expressions. Labib et al.,[69] on the other hand, have demonstrated inhibition of butyric

acid utilization by H_2 separate from the thermodynamic effects. Thus, even though information on the kinetic impacts of H_2 is very limited, it appears to be important, suggesting that additional studies are needed to allow development of appropriate rate expressions. In spite of these limitations, however, the kinetic parameters above provide a good sense of the relative capabilities of the microorganisms involved in anaerobic operations.

3.3 MAINTENANCE, ENDOGENOUS METABOLISM, DECAY, LYSIS, AND DEATH

As discussed in Section 2.4.2, a number of complex events interact to make the observed yield in biochemical operations less than the true growth yield and to cause only a fraction of the suspended solids to be active biomass. Even if our knowledge of all of those events was sufficient to allow mechanistically accurate kinetic models to be written, it is doubtful that they would be used in engineering practice because of their complexity. Consequently, as is common in engineering, simplified models have been adopted because of their utility and adequacy, and two will be reviewed in this section. The traditional approach has been in use for many years and has found many applications.[20,31,40,71,80] Its main attributes are its simplicity and familiarity. Its main weakness, however, is its inability to easily handle situations in which the nature of the terminal electron acceptor is changing. The second model addresses that situation.[25,26,54,55,73] It is called the lysis:regrowth approach.[26]

3.3.1 The Traditional Approach

In the traditional approach, all of the events leading to the reduction in yield and viability are expressed by the following stoichiometry:

$$\text{Biomass} + \text{electron acceptor} \rightarrow CO_2 + \text{reduced acceptor} + \text{nutrients} + \text{biomass debris} \tag{3.52}$$

The important concepts incorporated into this expression are that active biomass is destroyed as a result of "decay" and that the electrons removed as a result of the oxidation of the carbon to carbon dioxide pass to the electron acceptor. Furthermore, not all of the biomass is totally oxidized and a portion is left as biomass debris.[63,78,80] Although the debris is ultimately biodegradable,[37,90] its rate of biodegradation is so low that for all practical purposes it is inert to further biological attack in most biochemical operations, causing it to accumulate, reducing the fraction of active biomass in the suspended solids. Finally, nitrogen is released as ammonia-N, although some remains in the biomass debris. Figure 3.5 illustrates how these events are related to microbial growth in an aerobic environment.

If Eq. 3.52 is rewritten as a COD balance the result is:

$$\text{Biomass COD} + [-(1 - f_D)]O_2 \text{ equivalents of electron acceptor} \rightarrow f_D \text{ biomass debris COD} \tag{3.53}$$

where f_D is the fraction of the active biomass contributing to biomass debris, X_D. For the type of biomass normally found in biochemical operations for wastewater

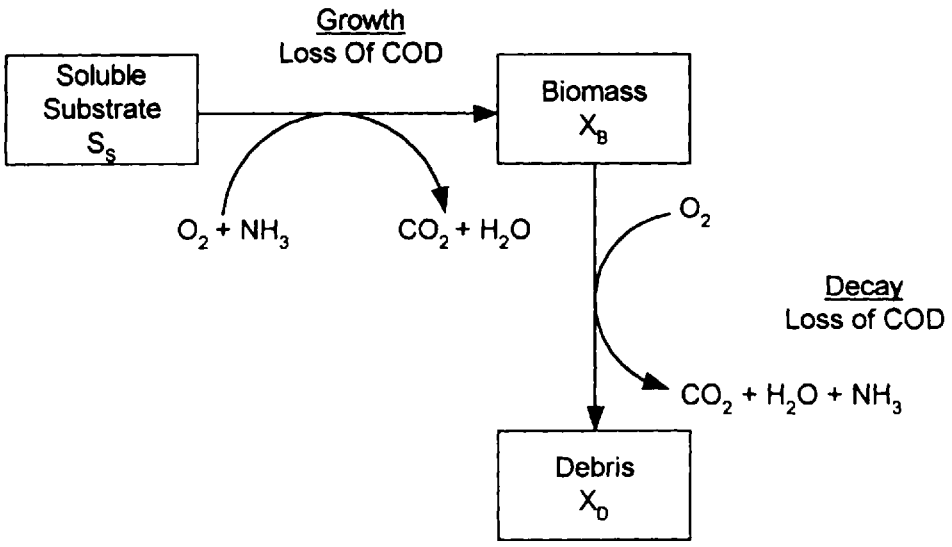


Figure 3.5 Schematic representation of the traditional approach to modeling biomass decay and loss of viability.

treatment it has a value of around 0.2.[25,78,81] Equation 3.53 shows that the utilization of oxygen or nitrate due to decay must equal the loss of active biomass COD minus the production of biomass debris COD.

Another important concept inherent in Eq. 3.52 is that nitrogen is released as ammonia as biomass is destroyed. If Eq. 3.52 were reformulated as a nitrogen based stoichiometric equation it would read:

$$\text{Biomass N} \rightarrow \text{NH}_3\text{-N} + \text{biomass debris N} \tag{3.54}$$

Since we have used biomass COD as the basic measurement of biomass, it would be convenient to write the nitrogen based stoichiometric equation in a way which linked it to biomass COD. This can be done by introducing two conversion factors, $i_{N\,XB}$ and $i_{N\,XD}$, which are respectively, the mass of nitrogen per mass of COD in active biomass and the mass of nitrogen per mass of COD in biomass debris. Their use leads to:

$$i_{N\,XB} \cdot \text{Biomass COD} \rightarrow \text{NH}_3\text{-N} + i_{N\,XD} \cdot \text{biomass debris COD} \tag{3.55}$$

Because the destruction of a unit mass of biomass COD leads to the generation of f_D units of biomass debris COD (Eq. 3.53), Eq. 3.55 tells us that the amount of ammonia-N released from the destruction of a unit mass of biomass COD is ($i_{N\,XB} - i_{N\,XD}f_D$). If $C_5H_7O_2N$ is representative of biomass, then $i_{N\,XB}$ has a value of 0.087 mg N/mg biomass COD. The nature of biomass debris is less well characterized than active biomass and thus there is no generally accepted empirical formula from which $i_{N\,XD}$ can be calculated. However, because many nitrogenous compounds serve as energy reserves that are destroyed during endogenous metabolism, it is likely that the nitrogen content of biomass debris is less than that of biomass. As a result, a value of 0.06 mg N/mg COD has been recommended for $i_{N\,XD}$.[54,55]

The rate expression for decay of biomass is first order with respect to the biomass concentration:

$$r_{XB} = -b \cdot X_B \tag{3.56}$$

where b is the decay coefficient, with units of hr^{-1}. Employing the concept in Eq. 3.10, the rate of production of biomass debris can be seen to be:

$$r_{XD} = b \cdot f_D \cdot X_B \tag{3.57}$$

and the rate of oxygen (electron acceptor) utilization associated with biomass decay is:

$$r_{SO} = (1 - f_D)b \cdot X_B \text{ (COD units)} = -(1 - f_D)b \cdot X_B \text{ (O}_2\text{ units)} \tag{3.58}$$

The same equation would hold for utilization of nitrate expressed as oxygen equivalents, although the numerical value of the decay coefficient may well be different with alternative electron acceptors. Finally, the rate of ammonia-N release is:

$$r_{SNH} = (i_{N\,XB} - i_{N\,XD} \cdot f_D)b \cdot X_B \tag{3.59}$$

As might be expected from the discussion of parameter values in Section 3.2.10, the value of b is very dependent on both the species of organism involved and the substrate on which it is grown. The latter effect is probably due to the nature of the energy reserves synthesized during growth. Because Eq. 3.56 is an approximation describing very complex events, the value of b also depends to some extent on the rate at which the biomass is grown. Dold and Marais[25] have reviewed the literature concerning b and have concluded that in aerobic and anoxic wastewater treatment systems a typical value for heterotrophic biomass is 0.01 hr^{-1}. Others[10] have reported values as low as 0.002 hr^{-1} as being common in similar systems. Thus, it can be seen that quite a large range can exist. A large range of b values has also been reported for autotrophic nitrifying bacteria,[22] with values ranging from 0.0002 to 0.007 hr^{-1}. A value of 0.003 hr^{-1} is considered typical at 20°C.[55]

Decay also occurs in anaerobic systems, but the b values for such systems are lower than those for aerobic systems because the bacteria have much lower $\hat{\mu}$ values, and the two parameters appear to be correlated. For example, Bryers[14] has reported b values around 0.0004 hr^{-1} for bacteria carrying out anaerobic oxidations and methanogenesis and values around 0.001 hr^{-1} for fermentative bacteria.

3.3.2 The Lysis:Regrowth Approach

The most complete model depicting the loss of viability and biomass in biochemical operations was devised by Mason et al.[73] after an extensive review of the literature.[74] In that model, viable biomass can either die or be inactivated, leading to dead and nonviable biomass, respectively. Furthermore, all biomass can undergo lysis, although at different rates for different types, leading to soluble and particulate organic matter. The particulate organic matter is hydrolyzed to soluble organic matter, and the soluble organic matter from either source can be used by the viable biomass for new growth. Loss of viability is accounted for because the presence of dead biomass and particulate organic matter reduces the number of viable bacteria per unit mass of particulate material. Loss of biomass, i.e., decay, results from the fact that yield values are less than one so that the amount of biomass grown from the soluble

substrate released is always less than the amount destroyed by lysis, as discussed in Section 2.4.1.

A conceptually similar, but less complex, model was developed by Dold et al.[26] for use in modeling wastewater treatment systems containing both aerobic and anoxic zones. Only one type of biomass is considered to be present: active, viable biomass. However, it is viewed as continually undergoing death and lysis, yielding particulate substrate and biomass debris. As in the model of Mason et al.,[73] particulate substrate is hydrolyzed to soluble substrate, and the soluble substrate is used by the viable biomass for growth, yielding new cell material. However, as above, because biomass yield values are always less than one, the amount of new biomass formed is always less than the amount destroyed by death and lysis, resulting in a net loss of biomass from the system (i.e., decay). A loss of viability results from the accumulation of biomass debris and particulate substrate.

The model of Dold et al.[26] is simpler than that of Mason et al.,[73] yet appears to be adequate for modeling many important wastewater treatment systems.[52] Furthermore, it can account for differences in decay observed as bacteria are cycled through aerobic, anoxic, and anaerobic conditions, whereas those differences cannot be accounted for by the traditional decay approach.[123] Finally, it has been adopted for use in a general model of single-sludge processes[54,55] that has been shown to adequately represent the dynamic performance of full-scale systems.[5] Thus, it will be used herein as an alternative to the traditional approach. The events in it are depicted in Figure 3.6.

The COD-based stoichiometry of the lysis:regrowth approach of Dold et al.[26] is:

$$\text{Biomass COD} \rightarrow (1 - f'_D)\ \text{particulate substrate COD} + f'_D\ \text{biomass debris COD} \tag{3.60}$$

where f'_D is the fraction of active biomass contributing to biomass debris. No COD is lost during death and lysis. Rather active biomass COD is simply converted into an equivalent amount of COD due to biomass debris and particulate substrate. As a

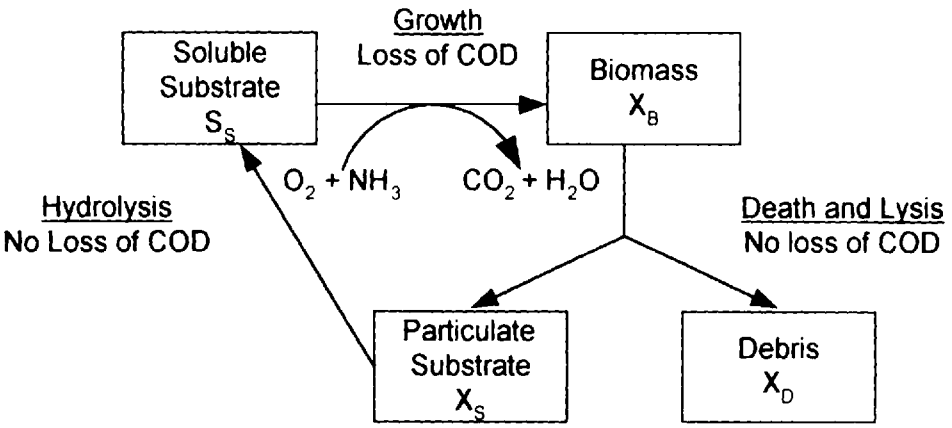


Figure 3.6 Schematic representation of the lysis:regrowth approach to modeling biomass decay and loss of viability.

consequence, no use of electron acceptor is directly associated with the loss of biomass, i.e., decay. Electron acceptor utilization occurs as soluble substrate, which arises from hydrolysis of particulate substrate, is used by active biomass for growth. As with the traditional approach, cell debris is assumed to be resistant to microbial attack within the time constraints of biochemical operations.

The nitrogen in the biomass is divided between biomass debris and particulate substrate, with the latter being called particulate biodegradable organic nitrogen. The nitrogen based stoichiometric equation depicting this is:

$$\text{Biomass N} \rightarrow \text{particulate biodegradable organic N} + \text{biomass debris N} \tag{3.61}$$

Giving the same meanings to $i_{N\,XB}$ and $i_{N\,XD}$ as given above, Eq. 3.61 can be rewritten in terms of biomass COD and biomass debris COD:

$$i_{N\,XB} \cdot \text{Biomass COD} \rightarrow \text{particulate biodegradable organic N} + i_{N\,XD} \cdot \text{biomass debris COD} \tag{3.62}$$

Thus, each unit of biomass COD lost to decay yields $(i_{N\,XB} - i_{N\,XD} \cdot f'_D)$ units of particulate biodegradable organic nitrogen. This differs from the traditional approach which leads directly to soluble ammonia nitrogen.

As in the traditional approach, the rate of loss of biomass COD to death and lysis is considered to be first order with respect to the active biomass concentration:

$$r_{XB} = -b_L \cdot X_B \tag{3.63}$$

where b_L has units of hr^{-1}, just as b does. In a manner similar to the traditional approach, the rate of production of biomass debris COD is:

$$r_{XD} = b_L \cdot f'_D \cdot X_B \tag{3.64}$$

And the rate of production of particulate substrate COD (X_S) is:

$$r_{XS} = (1 - f'_D) b_L \cdot X_B \tag{3.65}$$

Note the similarity of this equation to Eq. 3.58, the equation for oxygen consumption in the traditional approach. This similarity arises from the retention in the particulate substrate of all electrons lost from active biomass, rather than their transfer to oxygen. Finally, the rate of production of particulate, biodegradable organic nitrogen (X_{NS}) is:

$$r_{XNS} = (i_{N\,XB} - i_{N\,XD} \cdot f'_D) b_L \cdot X_B \tag{3.66}$$

It is important to realize that b_L is conceptually and numerically different from b and that f'_D is numerically different from f_D. This follows from the cycling of COD that occurs in the lysis:regrowth approach. Biomass COD is lost, releasing particulate substrate COD, which is hydrolyzed to soluble substrate COD, which is degraded by active biomass yielding new biomass, which is lost by death and lysis giving particulate substrate COD, etc. The net effect of the two approaches is the same because a given amount of biomass will be lost from a bioreactor regardless of how we conceptualize the actual events occurring. Since it is necessary for carbon to cycle around the system several times in the lysis:regrowth conceptualization to achieve the same loss of biomass that the traditional approach achieves in one pass,

b_L must be numerically larger than b. Likewise, since the same amount of biomass debris is ultimately formed from the loss of a given amount of biomass by decay, f'_D must be numerically smaller than f_D. In fact, the values of the four parameters are related:[26]

$$f'_D \cdot b_L = f_D \cdot b \tag{3.67}$$

Furthermore,

$$f'_D = \left(\frac{1 - Y}{1 - Y \cdot f_D}\right) f_D \tag{3.68}$$

It was stated above that f_D has a value around 0.2. Given the Y values associated with the biomass for which f_D was estimated, Eq. 3.68 suggests that the value of f'_D is around 0.08.[25] The values of f_D and f'_D are not likely to vary greatly, and thus those values will be adopted herein. It should be noted, however, that the relationship between b_L and b also depends on Y:[25]

$$b_L = \frac{b}{[1 - Y(1 - f'_D)]} \tag{3.69}$$

Although it is common during parameter evaluation studies to measure both Y and b, neither f_D nor f'_D is commonly measured. Since Y can influence the relationship between b_L and b, it is recommended that Eq. 3.69 be used instead of Eq. 3.67 to convert measured b values to b_L values.[55]

An important assumption implicit in the lysis:regrowth approach is that within a given culture, cell lysis occurs all of the time with the same value of the rate coefficient b_L, regardless of the rate at which the bacteria are growing. The validity of this assumption has been confirmed by measuring the release of nucleic acids as direct evidence of cell lysis.[95]

For autotrophic growth, the relationship between b_L and b is different.[54,55] This is because autotrophic organisms do not use organic matter for growth. Thus, death and lysis will not lead to additional autotrophic biomass growth. (The amount of autotrophic biomass that will grow from the nitrogen released is negligible.) Rather, heterotrophic biomass will grow on the organic matter released. As a consequence, the lysis:regrowth and traditional approaches are the same for autotrophic biomass; the result is that the two parameter values are equal.

3.4 SOLUBLE MICROBIAL PRODUCT FORMATION

As discussed in Section 2.4.3, soluble microbial products are thought to arise from two processes, one growth associated and the other non-growth-associated.[103]

Growth associated product formation results directly from biomass growth and substrate utilization. If soluble microbial product formation was occurring in appreciable amount, it would be necessary to modify the stoichiometric equation for microbial growth to account for it. Letting S_{MP} represent the concentration of soluble microbial products in COD units and Y_{MP} the product yield in units of product COD formed per unit of substrate COD used, Eq. 3.31 can be rewritten to account for soluble microbial product formation:

$$(1)S_S + [-(1 - Y_H - Y_{MP})]S_O \rightarrow Y_H X_{B,H} + Y_{MP} S_{MP} \quad (3.70)$$

This shows that less electron acceptor is used when soluble products are formed because part of the COD of the substrate remains in the medium as those products. Rewriting this equation in the form of Eq. 3.9 with biomass as the reference constituent gives:

$$\left(-\frac{1}{Y_H}\right) S_S + (-1)\left[-\left(\frac{1 - Y_H - Y_{MP}}{Y_H}\right)\right] S_O + X_{B,H} + \frac{Y_{MP}}{Y_H} S_{MP} = 0 \quad (3.71)$$

This tells us:

$$r_{SMP} = (Y_{MP}/Y_H) r_{XB} \quad (3.72)$$

Combining Eq. 3.72 with Eq. 3.35 for r_{XB} gives:

$$r_{SMP} = (Y_{MP}/Y_H)\mu \cdot X_{B,H} \quad (3.73)$$

The fact that r_{SMP} is proportional to μ shows that it is growth-associated.

Non-growth-associated product formation, also called biomass-associated product formation,[103] occurs as a result of cell lysis and decay. Rewriting Eq. 3.53 to incorporate soluble product formation into the COD-based stoichiometry of the traditional approach to decay gives:

$$\text{Biomass COD} + [-(1 - f_D - f_{MP})]\text{O}_2 \text{ equivalents of electron acceptor} \rightarrow f_D \text{ biomass debris COD} + f_{MP} \text{ soluble product COD} \quad (3.74)$$

where f_{MP} is the fraction of active biomass contributing to biomass-associated products. Using this with Eq. 3.56 gives the rate of production of biomass-associated product:

$$r_{SMP} = b \cdot f_{MP} \cdot X_{B,H} \quad (3.75)$$

By analogy to biomass debris formation, a similar approach could be used to account for soluble microbial product formation in the lysis:regrowth approach, giving a parameter f'_{MP} which is smaller than f_{MP} in the same way that f'_D is smaller than f_D.

Combining Eqs. 3.73 and 3.75 suggests that the specific rate of soluble microbial product formation is linearly related to the specific growth rate. While such a relationship may be adequate for slowly growing cultures like those found in activated sludge systems, it is not adequate for more rapidly growing systems[49] and thus Eqs. 3.73 and 3.75 cannot be considered to be of general applicability to all systems. Although a relatively large body of research on soluble microbial product formation has been conducted,[49,103] it is still not sufficient to allow consensus on the rate expressions to be used. Thus, in spite of its known importance, soluble microbial product formation will not be incorporated into the models in Parts II and IV.

Insufficient information is available to provide typical values for f_{MP} and f'_{MP}, but, as indicated in Section 2.4.3, Y_{MP} values have been found to be less than 0.1.[42]

3.5 SOLUBILIZATION OF PARTICULATE AND HIGH MOLECULAR WEIGHT ORGANIC MATTER

The conversion of particulate and high molecular weight organic matter into forms small enough for bacteria to take up and degrade is an important step in biochemical operations for wastewater treatment because such materials are commonly present in wastewaters and also arise from lysis reactions as discussed previously. In spite of that, relatively few studies have focused on those reactions. Perhaps this is because many types of particulate materials are attacked by distinctly different mechanisms, even though they are collectively referred to as hydrolysis.

The stoichiometry of hydrolysis is thought to be very simple, with organic material simply changing form. Consequently, most investigators have assumed that COD is conserved, i.e., that no energy is consumed. This is indicated in Figure 3.6. Because no energy is consumed, no electrons are removed and no terminal electron acceptor is used. Thus, the stoichiometric equation is simply:

$$\text{Particulate substrate COD} \rightarrow \text{soluble substrate COD} \tag{3.76}$$

This means that the rate of formation of soluble substrate COD is equal to the rate of loss of particulate substrate COD.

In the face of complex situations in which reactions are ill defined, it is common for engineers to choose the simplest possible reaction rate expression, and that is what a number of investigators have done, assuming that hydrolysis is first order with respect to the concentration of particulate substrate, X_S.[14,15,28,44,73] This approach, however, ignores the effect that the biomass concentration will have on the rate.

One group[55] performed an extensive literature survey before adopting a kinetic expression for the hydrolysis of particulate organic matter patterned after that of Dold et al.,[26] which is based on the work of Stenstrom:[117]

$$r_{XS} = -k_h \left[\frac{X_S/X_{B,H}}{K_X + (X_S/X_{B,H})} \right] X_{B,H} \tag{3.77}$$

In this expression k_h is the hydrolysis coefficient (hr^{-1}) and K_X is a half-saturation coefficient (mg particulate substrate COD/mg active biomass COD). An important characteristic of this expression is that even though the rate is first order with respect to the heterotrophic biomass concentration, it is controlled by the ratio of particulate substrate concentration to heterotrophic biomass concentration, rather than by the particulate substrate concentration alone. This is necessary because the reaction is thought to be surface mediated, depending on the presence of extracellular enzymes whose quantity will be proportional to the biomass concentration.[26]

Data on the values of k_h and K_X are very limited. Based primarily on the recommendations of Dold and Marais,[25] one group[54] adopted a value of 0.092 hr^{-1} for k_h and a value of 0.15 for K_X.

The rate of hydrolysis will also be influenced by the electron acceptor concentration,[25,54,55] even though no electron acceptor is used in the reaction. Under aerobic conditions, an interactive, dual nutrient limitation expression has been adopted in a manner similar to that in Eq. 3.46. Under anoxic conditions, an expression similar to that in Eq. 3.48 has been found to be appropriate, with nitrate stimulating anoxic hydrolysis and oxygen inhibiting it. In both expressions, the effect of the particulate substrate should be given by Eq. 3.77. Under anaerobic conditions of short duration,

hydrolysis is assumed to stop. While this would not be true for long term anaerobic conditions, it is consistent with observations in biochemical operations that cycle bacteria between aerobic and anoxic conditions.[25] It would probably be satisfactory to use Eq. 3.77 alone to describe hydrolysis under fully anaerobic conditions, but with a k_h value much smaller than that used for aerobic conditions.

As seen in Eq. 3.61, biomass decay results in the formation of particulate biodegradable organic nitrogen. In addition, organic nitrogen will be associated with the particulate organic matter in the wastewater. All of this material will be converted into soluble, biodegradable organic nitrogen, S_{NS}, (i.e., the nitrogen associated with amino acids and other soluble, nitrogen containing organic substrates) as the particulate substrate is hydrolyzed. The rate of generation of S_{NS} (r_{SNS}) is numerically equivalent to the rate of loss of particulate organic nitrogen (r_{XNS}), which is proportional to the hydrolysis rate of particulate organic matter:[25]

$$r_{SNS} = -r_{XNS} = -(X_{NS}/X_S)r_{XS} \tag{3.78}$$

where X_{NS} is the concentration of particulate, biodegradable organic nitrogen.

3.6 AMMONIFICATION AND AMMONIA UTILIZATION

Ammonification is the conversion of soluble organic nitrogen into ammonia-N that occurs as bacteria consume soluble organic matter containing nitrogen. Actually, the true rate of ammonification is difficult to measure because ammonia-N is being consumed by the bacteria as they grow, and the only measurable event is the net accumulation or loss of ammonia in the medium. If the amount of nitrogen available in the organic substrate is just sufficient to meet the biosynthetic needs of the new biomass, there will be no net change in the ammonia-N concentration in the medium. On the other hand, if that amount exceeds the need, the ammonia concentration in the medium will increase, whereas if that amount is less than the need, the ammonia concentration will decrease. It should be recognized, however, that whether organic nitrogen is incorporated directly into new biomass depends on its form. The nitrogen in simple compounds like amino acids may be incorporated directly as the amino acids are used for protein synthesis, while nitrogen in complex synthetic organic chemicals may be released to the medium as ammonia.

In an effort to make this complex situation mathematically tractable, most modelers assume that all nitrogen goes through the medium before being used. Thus, ammonification is assumed to release all organic nitrogen to the medium as ammonia, and nitrogen utilizing reactions are assumed to obtain their ammonia from the medium. Whether ammonia accumulates or is removed depends on the relative rates of its production and utilization.

Because ammonification occurs as heterotrophic biomass destroys nitrogen containing soluble organic matter, it is likely that its rate is proportional to the rate of soluble substrate removal. Relatively little work has been done to investigate the rate of ammonification in complex substrates in which only a part of the soluble organic matter contains nitrogen, and it is uncertain whether a direct proportionality can be assumed between soluble substrate removal and ammonification. Consequently, ammonification has been represented as a reaction that is first order with respect to both

the heterotrophic biomass concentration and the concentration of soluble, biodegradable organic nitrogen:[25,54,55]

$$r_{SNS} = -k_a \cdot S_{NS} \cdot X_{B,H} \tag{3.79}$$

where k_a is the ammonification rate coefficient (L/(mg biomass COD · hr)). Very little information is available about its value. As discussed above, the assumed stoichiometry of ammonification is such that all nitrogen removed from nitrogen containing soluble organic matter is released as ammonia, although some may ultimately be used for biomass synthesis. Thus, the rate of production of ammonia nitrogen (S_{NH}) through ammonification is:

$$r_{SNH} = -r_{SNS} \tag{3.80}$$

Ammonia is removed from solution by two reactions. First, it is used in the synthesis of new biomass as seen in Eq. 3.6 and others presented in Section 3.2. Second, it is used as a substrate by autotrophic biomass. The rate expression for the second use is the same as any other substrate, as discussed in Section 3.2.6. The rate expression for the first use can be determined from the generalized rate expression and the stoichiometry of growth. Since $i_{N\,XB}$ is the mass of nitrogen per unit of biomass COD, the rate of ammonia removal through biomass growth is simply:

$$r_{SNH} = -i_{N\,XB} \cdot \mu \cdot X_B \tag{3.81}$$

Equation 3.81 is true for both heterotrophic and autotrophic growth, and thus the general symbol for biomass, X_B, has been used in it.

3.7 PHOSPHORUS UPTAKE AND RELEASE

Biological phosphorus removal is a complex process that is dependent on the growth of specialized phosphate accumulating organisms (PAOs), which store phosphorus as polyphosphate (poly-P), as discussed in Section 2.4.6. Because biological phosphorus removal is still the subject of active experimental investigation, there is little consensus concerning all of the rate expressions describing it. Furthermore, the microbial events involved are subject to complex control through the concentrations of several constituents. As a consequence, a discussion beyond the scope of this chapter would be needed to fully define the kinetics and stoichiometry of biological phosphorus removal. There are, however, two events that should be considered briefly because they are distinctly different from the events discussed above. These are the uptake and release of phosphorus by the PAOs.

It will be recalled from Section 2.4.6 that two conceptual models exist for biological phosphorus removal, the Comeau/Wentzel model and the Mino model, illustrated in Figures 2.5 and 2.6, respectively. The major difference between them is that the Mino model incorporates glycogen formation and utilization, whereas the Comeau/Wentzel model does not. In developing a mathematical model for biological phosphorus removal, one group decided that the introduction of a detailed mechanistically based mathematical model for the process was premature.[56] Rather, they preferred to recommend the simplest mathematical model that allows adequate prediction of biological phosphorus removal. Consequently, they basically followed the Comeau/Wentzel model. Furthermore, they assumed that PAOs cannot use nitrate as

a terminal electron acceptor and that they can only grow on the PHB stored in the cell. Although these assumptions are severe restrictions, we have chosen to use activated sludge model (ASM) No. 2 as the basis for the rate expressions presented herein because it represents a consensus among several investigators in the field.[56] It should be recognized, however, that more complete expressions are likely to be developed as more research is done.

Under anaerobic conditions, PAOs do not grow, but store acetic acid as PHB through the cleavage of Poly-P with the associated release of soluble phosphate. (In the following equations, all phosphate concentrations are expressed as phosphorus, P, and all organic materials are expressed as COD). The rate of removal of acetic acid, r_{SA}, can be modeled with an interactive, dual limiting nutrient expression:

$$r_{SA} = -\hat{q}_A \left(\frac{S_A}{K_A + S_A}\right)\left[\frac{X_{PP}/X_{B,P}}{K_{PP} + (X_{PP}/X_{B,P})}\right] X_{B,P} \tag{3.82}$$

where $\hat{q}_A$ is the maximum specific rate of acetic acid uptake, hr^{-1}, S_A is the acetic acid concentration in COD units, K_A is a half-saturation coefficient for acetic acid, X_{PP} is the poly-P concentration in the biomass, expressed as a liquid phase P concentration, K_{PP} is a half-saturation coefficient for poly-P in the same units, and $X_{B,P}$ is the concentration of PAO biomass in COD units. The stoichiometry of the reaction is such that each mg/L of acetic acid COD removed from the medium forms a mg/L of PHB COD in the biomass. Thus, the rate expression for PHB (X_{PHB}) formation, r_{XPHB}, is:

$$r_{XPHB} = -r_{SA} \tag{3.83}$$

Furthermore, Y_P units of soluble phosphate, S_P, are released for each unit of acetic acid stored as COD, increasing S_P by an amount equal to the decrease in stored Poly-P concentration. Thus:

$$r_{SP} = -r_{XPP} = Y_P \cdot r_{XPHB} \tag{3.84}$$

The value of Y_P selected by Henze et al.[56] was 0.40 mg P/mg COD, reflecting the average stoichiometry for the process. The values of K_{SA} and K_{PP} were chosen to be small to make the parenthetical terms in Eq. 3.82 serve as switching functions that change rapidly from one to zero, thereby turning the reaction on and off. The value chosen for K_{SA} was 4.0 mg COD/L whereas the value chosen for K_{PP} was 0.01 mg P/mg PAO COD.

Under aerobic conditions, the PAOs grow by using the stored PHB as a carbon and energy source. This is assumed to be their only substrate for growth, even though they are capable of growth on soluble substrates. Because little soluble substrate is likely to be present in the aerobic portion of a biological phosphorus removal process, Henze et al.[56] ignored it to simplify the model. Furthermore, because the process can only occur under aerobic conditions, a switching function for oxygen was included to make the rate go to zero when oxygen is absent. Considering all of these factors, the rate of PAO growth can be described by:

$$r_{XBP} = \hat{\mu}_P \left[\frac{X_{PHB}/X_{B,P}}{K_{PHB} + (X_{PHB}/X_{B,P})}\right]\left(\frac{S_P}{K_P + S_P}\right)\left(\frac{S_O}{K_O + S_O}\right) X_{B,P} \tag{3.85}$$

where $\hat{\mu}_P$ is the maximum specific growth rate coefficient for PAOs, X_{PHB} is the stored PHB concentration in mg/L as COD, S_P is the soluble phosphate concentration

in mg/L as P, K_P is the half-saturation coefficient for soluble phosphate, S_O is the dissolved oxygen concentration, and K_O is the half-saturation coefficient for dissolved oxygen. It should be noted that the expression for the effect of PHB concentration on biomass growth is written in terms of the amount of PHB available per unit of biomass COD because the PHB is not free in the medium, but is stored in the biomass. As a result, K_{PHB} has units of mg PHB COD/mg PAO COD. Because of biomass lysis, phosphate will continually be released to the medium. Consequently, S_P will never reach a zero concentration and phosphorus will always be available for growth. The values chosen for the half-saturation coefficients by Henze et al.[56] were 0.01 mg PHB COD/mg PAO COD, 0.20 mg P/L, and 0.20 mg O_2/L, for K_{PHB}, K_P, and K_O, respectively.

The stoichiometry of the aerobic growth reaction on a COD basis is the same as that in Eq. 3.33, except that PHB is the growth substrate. Consequently, the relationship between r_{XBP}, r_{XPHB}, and r_{SO} (with all in COD units) will be the same as the relationship between r_{XB}, r_{SS}, and r_{SO} in Eq. 3.34, or:

$$\frac{r_{XPHB}}{\left(-\frac{1}{Y_{PAO}}\right)} = \frac{r_{SO}}{(-1)\left[-\left(\frac{1-Y_{PAO}}{Y_{PAO}}\right)\right]} = \frac{r_{XBP}}{1} \tag{3.86}$$

where Y_{PAO} is the yield coefficient for PAOs growing on stored PHB. The value assumed for it in ASM No. 2 is 0.63 mg PAO COD/mg PHB COD.[56] It should be noted that the rates of PHB loss and oxygen consumption expressed by Eq. 3.86 are that associated only with PAO growth.

Storage of polyphosphate also occurs under aerobic conditions and the energy for it also comes from PHB utilization. The rate expression includes all of the parenthetical terms in Eq. 3.85. It has been observed, however, that storage of poly-P stops if its content in the PAOs becomes too high.[56] It is necessary to include a term that decreases the rate of Poly-P storage as the Poly-P concentration per unit of PAOs approaches a maximum value of K_{PMAX}. Considering these factors, the rate of Poly-P storage, r_{XPP}, can be expressed as:

$$r_{XPP} = \hat{q}_{PP}\left[\frac{X_{PHB}/X_{B,P}}{K_{PHB} + (X_{PHB}/X_{B,P})}\right]\left(\frac{S_P}{K_P + S_P}\right)\left(\frac{S_O}{K_O + S_O}\right) \cdot \left[\frac{K_{PMAX} - (X_{PP}/X_{B,P})}{K_{IPP} + K_{PMAX} - (X_{PP}/X_{B,P})}\right] X_{B,P} \tag{3.87}$$

where $\hat{q}_{PP}$ is the maximum specific rate of Poly-P storage, which has a typical value of 0.06 mg P/(mg PAO COD·h) at 20°C. K_{IPP} is the inhibition coefficient for Poly-P storage, with an assumed value of 0.02 mg P/mg PAO COD. All other terms were defined following Eq. 3.85. Soluble phosphate is removed from the medium in direct proportion to the amount incorporated into Poly-P. Furthermore, PHB is lost and oxygen is utilized proportionally as well. The relationship between the rates is determined from the stoichiometry as:

$$\frac{r_{XPHB}}{(-Y_{PHB})} = \frac{r_{SO}}{(-1)(-Y_{PHB})} = \frac{r_{SP}}{-1} = \frac{r_{XPP}}{1} \tag{3.88}$$

where Y_{PHB} is the PHB requirement for poly-P storage, which has a typical value of 0.20 mg PHB COD/mg P.[56] The rates of PHB loss and oxygen consumption in this

expression are those associated with only Poly-P storage. The total rates of each under aerobic conditions must be obtained by adding the expressions from Eqs. 3.86 and 3.88. Furthermore, Eq. 3.88 does not give the total rate of soluble phosphate loss since polyphosphate formation is not the only mechanism for removing soluble phosphate from the liquid. Rather, phosphorus is also a required nutrient for biomass synthesis. If $i_{P\,XB}$ is the mass of phosphorus incorporated into cell material per unit of PAO COD formed, the total rate of removal of soluble phosphorus by the PAOs will be:

$$r_{SP} = -(i_{P\,XB} \cdot r_{XBP}) - r_{XPP} \tag{3.89}$$

where r_{XBP} is given by Eq. 3.85. Cellular biomass contains about 2.5 percent phosphorus on a mass basis, so on a biomass COD basis, $i_{P\,XB}$ has a value around 0.02 mg P/mg biomass COD. If heterotrophs and autotrophs are growing in the system, they will also consume soluble phosphate for incorporation into biomass with the same stoichiometry.

3.8 SIMPLIFIED STOICHIOMETRY AND ITS USE

In Chapter 5, we use the concepts developed in Section 3.1.3 to construct mathematical models that incorporate the various events discussed in this chapter. There are many circumstances, however, in which the use of stoichiometric concepts would be very useful even without the development of rigorous equations. For example, examination of Eq. 3.13 expressing biomass growth and Eq. 3.52 expressing biomass decay by the traditional approach reveals that they could be combined into a single equation that incorporates both reactions. Since biomass is a product in Eq. 3.13 and a reactant in Eq. 3.52, the effect would be to reduce the net amount of biomass formed. Likewise, since nutrients are reactants in Eq. 3.13 and products in Eq. 3.52, the net amount of nutrients used would also be reduced. The electron acceptor, on the other hand, is a reactant in both equations, so the effect of combining them would be to increase the amount of electron acceptor required. Consideration of what is occurring when the equations are combined, in combination with the discussion of yield in Section 2.4.1, reveals that the net stoichiometric coefficient on biomass in a mass-based combined equation is the observed yield. In other words, it is the yield when maintenance energy needs and decay are taken into account. By making use of the fact that the observed yield is a function of the growth conditions imposed on the biomass, the combined equation may be used to show how the nutrient and electron acceptor requirements change as the growth conditions are changed.[110]

3.8.1 Determination of the Quantity of Terminal Electron Acceptor Needed

Although other stoichiometric equations can be used, the COD-based equation is the most useful for determining the quantity of terminal electron acceptor required for growth of heterotrophs. Writing the combined stoichiometric equation for heterotrophic biomass growth in COD units illustrates a very important point that will be used throughout this book. When ammonia serves as the nitrogen source, the sum of the oxygen (or oxygen equivalents of nitrate) used and the biomass (active plus

debris) formed (in COD units) must equal the COD removed from solution. This follows from the fact that COD is a measure of available electrons. In other words, all of the electrons available in a substrate being biodegraded are either removed and transferred to the terminal electron acceptor or they are incorporated into the biomass formed. As discussed in Section 3.2.1, when ammonia serves as the nitrogen source, no electrons are transferred to nitrogen during biomass synthesis. When nitrate serves as the nitrogen source, however, some of those electrons must be used to reduce nitrogen from the +V state to the −III state, and thus those electrons are incorporated into the biomass even though they will not be measured in the COD test. This is because nitrogen does not accept or give up electrons in the COD test. Consequently, if biomass is represented by $C_5H_7O_2N$, its COD must be multiplied by 1.4 for the balance to work, as suggested by Eq. 3.17. Thus, to generalize:

$$\text{COD removed} = O_2 \text{ equivalents of terminal electron acceptor used} + \alpha_N(\text{COD of biomass formed}) \tag{3.90}$$

where:

α_N = 1.0 NH_4^+ as nitrogen source

α_N = 1.4 NO_3^- as nitrogen source

Equation 3.90 is generally applicable and is much easier to use for determining the amount of terminal electron acceptor required than the writing of a molar or mass-based stoichiometric equation. Thus, it is widely employed and will be used frequently herein. Ammonia will be assumed to be the nitrogen source throughout this book, unless specifically stated otherwise. Thus, α_N will generally be set equal to 1.0.

3.8.2 Determination of Quantity of Nutrient Needed

The amount of nitrogen required for heterotrophic biomass growth can also be calculated from the combined stoichiometric equation. Since the only use of nitrogen in the equation is for synthesis of biomass, the equation may be used to establish a relationship that is very useful for estimating nutrient requirements. If the ammonium ion requirement is expressed per unit of biomass COD formed, it is found to be 0.112 mg of NH_4^+ per mg of biomass COD formed. Or, expressed as the amount of nitrogen required, it is 0.087 mg of N per mg of biomass COD formed. Actually, this can be considered to be a generality that is independent of the source of the nitrogen, provided that $C_5H_7O_2N$ represents the composition of biomass. This suggests that once the observed yield has been determined, the amount of nitrogen required can be estimated easily, allowing adequate amounts to be provided if they are not naturally present in the wastewater. Likewise, each time a mg of biomass COD is destroyed, 0.087 mg of nitrogen will be released to the medium, and this fact must be considered in operations such as aerobic digestion which are designed to destroy biomass. As with determination of the electron acceptor requirement, the main purpose of traditional stoichiometric equations has been to provide simplified relationships such as these for engineering use. Thus, the conversion factor is generally used in lieu of writing a new balanced stoichiometric equation for each situation.

Table 3.3 Approximate Micronutrient Requirements for Bacterial Growth

Micronutrient	Approximate requirement,* μg/mg biomass COD formed
Potassium	10
Calcium	10
Magnesium	7
Sulfur	6
Sodium	3
Chloride	3
Iron	2
Zinc	0.2
Manganese	0.1
Copper	0.02
Molybdenum	0.004
Cobalt	<0.0004

* Estimates based on the judgment of the authors after considering information in Refs. 30, 99, and 127.

As mentioned in Section 3.2.1, the phosphorus requirement for normal microbial growth can be estimated as one-fifth of the nitrogen requirement on a mass basis. Consequently, about 0.017 mg of phosphorus will be required for each mg of heterotrophic or autotrophic biomass COD formed, and an equal amount will be released for each mg destroyed. If PAOs are in the system, the amount released by destruction of the biomass will be different and will depend on the amount of poly-P stored.

The provision of sufficient nutrients is essential if efficient wastewater treatment is to be achieved, because without them the microorganisms will not be able to perform their synthesis reactions. Although nitrogen and phosphorus are the nutrients needed in greatest quantity (macronutrients), many other elements are required by the microorganisms but are not normally included in the stoichiometric equation because of the complicating effect they would have. The need for them should not be ignored nor should their presence be taken for granted because severe problems can result if sufficient quantities are not available.[16,127] Table 3.3 lists the major micronutrients required for bacterial growth.[87] There is little agreement in the literature concerning their quantities in biomass. One reason is that different bacteria have different requirements. Another is that bacteria tend to adsorb cations, thereby making it difficult to determine exactly the quantity actually incorporated into biomass. The values listed in Table 3.3 are the authors' best estimates of the quantities required based on examination of several sources.[30,99,127]

3.9 EFFECTS OF TEMPERATURE

Temperature can exert an effect on biological reactions in two ways: by influencing the rates of enzymatically catalyzed reactions and by affecting the rate of diffusion

of substrate to the cells. The importance of both has not always been recognized and this has led to some confusion in the quantification of temperature effects. For example, temperature effects observed in the laboratory are often more pronounced than those observed in the field. This is due in part to the fact that full-scale reactors are apt to be diffusion controlled. Consequently, the temperature coefficients given below are provided simply to give an idea of the importance of temperature to various microbial processes. For system design, actual temperature effects should always be measured in prototype systems that simulate the anticipated mixing regime.

3.9.1 Methods of Expressing Temperature Effects

There are three techniques commonly used to quantify the effects of temperature on biochemical operations. The oldest is that of Arrhenius,[5] who first applied it in 1889 to quantify the effects of temperature on the enzymatic hydrolysis of sugar. It is:

$$k = A \cdot e^{-u/RT} \tag{3.91}$$

where k is the temperature dependent rate coefficient, A is a constant, u is the temperature coefficient, R is the gas constant, and T is the absolute temperature. The value of u may be obtained by plotting ln k versus 1/T and determining the slope. For normal SI units, the units of u are kJ/mole and a positive value means that k increases as the temperature is increased.

Although microorganisms have been found in extreme environments that can grow at temperatures approaching either the freezing point or the boiling point of water, most microorganisms exhibit a relatively narrow temperature range over which they can function. Within that range, most reaction rate coefficients increase as the temperature is increased, but then eventually decrease as the heat begins to inactivate cellular enzymes. The Arrhenius equation, as well as the others to be discussed below, are only applicable over the range where the coefficient increases with increasing temperature. Microorganisms are grouped into three categories depending on that temperature range. Of chief concern in biochemical operations are mesophilic organisms, which grow well over the range of 10–35°C. The two other groups, psychrophilic and thermophilic, have ranges on either side and find use under special conditions. Unless otherwise specified, all parameter values given in this book will be for mesophilic microorganisms.

If a rate coefficient is known at one temperature, it may be calculated at another through rearrangement of the Arrhenius equation:

$$\ln(k_1/k_2) = \frac{u(T_1 - T_2)}{(R \cdot T_1 \cdot T_2)} \tag{3.92}$$

Because the mesophilic temperature range is small when T is expressed in K, the term $(R \cdot T_1 \cdot T_2)$ does not vary appreciably and may be considered to be constant. Consequently, a more commonly used expression is:[59]

$$k_1 = k_2 \cdot e^{C(T_1 - T_2)} \tag{3.93}$$

where

$$C = \frac{u}{(R \cdot T_1 \cdot T_2)} \approx 0.0015\ u \tag{3.94}$$

for the normal mesophilic temperature range. Note that when Eq. 3.93 is used, the temperature may be expressed in °C because only the temperature difference enters into the equation. In that case the units of C are °C $^{-1}$. The value of C may be determined by plotting ln(k) versus T, giving a slope equal to C.

Finally, a third equation has found considerable use in the environmental engineering literature:[96]

$$k_1 = k_2 \cdot \theta^{(T_1 - T_2)} \tag{3.95}$$

Actually, Eqs. 3.93 and 3.95 are the same since:

$$C = \ln(\theta) \tag{3.96}$$

Thus, the coefficient θ may also be estimated by plotting ln(k) versus T, giving a slope equal to $\ln(\theta)$. θ is dimensionless.

The temperature coefficients for the three equations may be interconverted by:

$$\ln(\theta) = C \approx 0.0015\ u \tag{3.97}$$

in which the temperature is expressed in °C or K.

3.9.2 Effects of Temperature on Kinetic Parameters

Biomass Growth and Substrate Utilization. It will be recalled from Eqs. 3.35 and 3.43 that biomass growth and substrate utilization are proportional to each other, with the yield being the proportionality coefficient. It will also be recalled from Figure 2.4 that temperature can influence the value of the yield. This suggests that temperature can influence growth and substrate utilization in quantitatively different ways. Nevertheless, because of the uncertainty associated with the impact of temperature on Y, most engineers assume it to be independent of temperature, thereby allowing the same temperature coefficient to be used for both growth and substrate utilization.

Two parameters are required to characterize biomass growth, $\hat{\mu}$ and K_S. The first is clearly a rate coefficient, and as such, its value increases with increasing temperature. The second describes how substrate concentration influences the specific growth rate, and thus the impact of temperature on it is less clear, with it increasing under some circumstances and decreasing under others. Consequently, there is no consensus about its relationship to temperature, and each situation must be experimentally determined.

Most studies of the impact of temperature have been done on the aerobic growth of heterotrophs. Two studies[17,85] have reviewed the literature, and have reported values of u for $\hat{\mu}$ ranging from 21.3 to 167.4 kJ/mole. The average value for the larger data base[17] (18 values) was 59.8 kJ/mole, which converts to C and θ values of 0.090 °C $^{-1}$ and 1.094, respectively. Very few studies reporting the effects of temperature on K_S were cited, and there was no consensus among them as to whether it increased or decreased with increasing temperature.

Very few studies have been done to quantify the effects of temperature on microbial growth under anoxic conditions. van Haandel et al.[123] recommend that a θ value of 1.20 (C = 0.182 °C, $^{-1}$ u = 121 kJ/mole) be used for $\hat{q}$. This value is near the upper range for the aerobic values reported above, which suggests that it may be high. Until more data are available, it may be prudent to adopt a value more

consistent with aerobic growth and substrate utilization since the two processes are mechanistically similar. No values have been reported for the effect on K_S under anoxic conditions.

Temperature is a critical consideration for nitrifying bacteria because their $\hat{\mu}$ values are low even under the best of circumstances. Characklis and Gujer[17] reported four temperature coefficients for $\hat{\mu}$ for nitrification, with an average u value of 71.8 kJ/mole (C = 0.108 °C $^{-1}$, θ = 1.114). However, there appears to be little consensus about the relative effects of temperature on the two major genera of nitrifiers. For example, Characklis and Gujer[17] reported an average u of 74.3 kJ/mole (C = 0.111 °C $^{-1}$, θ = 1.118) for $\hat{\mu}$ of *Nitrosomonas* and 44.0 kJ/mole (C = 0.066 °C $^{-1}$, θ = 1.068) for *Nitrobacter*. In contrast, Hall and Murphy[47] reported a u value of 62.4 kJ/mole (C = 0.094 °C $^{-1}$, θ = 1.098) for $\hat{q}$ for *Nitrosomonas* and 71.1 kJ/mole (C = 0.107 °C $^{-1}$, θ = 1.112) for *Nitrobacter*. Nevertheless, there still seems to be a general consensus that the temperature coefficient for *Nitrobacter* is smaller than it is for *Nitrosomonas*. In contrast to heterotrophs, for which temperature appears to have variable effects on K_S, increases in temperature cause the half-saturation coefficient for nitrifiers to increase. The most widely cited data is that of Knowles et al.,[96] for which u associated with the K_S for *Nitrosomonas* was 78.7 kJ/mole (C = 0.118 °C $^{-1}$, θ = 1.125) and u associated with the K_S for *Nitrobacter* was 97.3 kJ/mole (C = 0.146 °C $^{-1}$, θ = 1.157).

Temperature is also known to play an important role in anaerobic operations. Most studies, however, have looked at overall system performance rather than at the impact on each of the groups of microorganisms discussed in Section 3.2.6. For example, Henze and Harremoës[53] combined data from seven studies to estimate the temperature coefficient for methanogenesis and found u to be 66.7 kJ/mole (C = 0.10 °C $^{-1}$, θ = 1.105) for a temperature range of 10–30°C. The methane production rate was constant from 30–40°C, and decreased for higher temperatures. Characklis and Gujer[17] used data from the literature to estimate that the value of u associated with K_S for acetic acid was −132.9 kJ/mole (C = −0.199 °C $^{-1}$, θ = 0.819), showing that K_S decreases as the temperature is increased for this process.

Maintenance, Endogenous Metabolism, Decay, Lysis, and Death. Most studies have used the traditional decay concept to quantify the impacts of maintenance, etc. on microbial systems, and thus all temperature data are available in terms of the rate coefficient b (Eq. 3.56). However, because b and b_l are proportional to each other (Eq. 3.67), the resulting temperature coefficients should also be applicable to b_l.

Because the factors contributing to decay of heterotrophs are the same as those contributing to growth, it is logical to expect temperature to have similar effects on b and $\hat{\mu}$, and that has been observed, with data from three studies giving u values for b equal to 1.1 times the u values for $\hat{\mu}$ for a given culture.[85] Thus, from the effects of temperature on $\hat{\mu}$ reported earlier, a typical u value for b might be expected to be 65.8 kJ/mole (C = 0.120 °C $^{-1}$, θ = 1.104). Others,[26] however, have used much smaller values for the effects of temperature on decay, with a u value of 19.1 kJ/mole (C = 0.029 °C $^{-1}$, θ = 1.029).

In spite of the importance of temperature to nitrification, few studies have systematically studied the effects of temperature on the decay coefficient for nitrifying bacteria. Dold et al.[26] used the same u value for autotrophic decay that was used for heterotrophic decay, although no data were presented.

Solubilization of Particulate and High Molecular Weight Organic Matter. As might be anticipated from the discussion in Section 3.5, relatively little work has been done on the effects of temperature on the hydrolysis of particulate substrate. However, because it is an enzymatic step, the hydrolysis coefficient, k_h, is likely to rise as the temperature is increased. From comparison of experimental data to simulation results from a complex system model, van Haandel et al.[123] concluded that a u value of 38.8 kJ/mole ($C = 0.058\ °C^{-1}$, $\theta = 1.060$) was appropriate for both aerobic and anoxic environments. No information was given for the effect of temperature on K_X, the half-saturation coefficient for hydrolysis.

Other Important Microbial Processes. Insufficient data are available to allow quantification of the effects of temperature on other processes, such as phosphorus release, but it is likely that appropriate temperature coefficients will be developed for them in the future.

3.10 KEY POINTS

1. Stoichiometric equations may be written on a mass basis rather than a molar basis. When this is done the total mass of reactants equals the total mass of products. When a stoichiometric equation is written on a mass of COD basis, only constituents containing elements that change oxidation state are included. The COD of the reactants must equal the COD of the products.
2. When nitrate serves as the terminal electron acceptor, nitrogen changes oxidation state from +V to 0. Consequently, the oxygen equivalence of nitrate is −2.86 mg COD/mg N (2.86 mg O_2/mg N). When nitrate serves as the nitrogen source for biomass growth, the nitrogen is reduced to the amino level; i.e., from the +V to the −III state. In that case the oxygen equivalence is −4.57 mg COD/mg N (4.57 mg O_2/mg N).
3. If the general form of the mass-based stoichiometric equation is written as:

$$(-1)A_1 + (-\Psi_2)A_2 + \cdots + (-\Psi_k)A_k + \Psi_{k+1}A_{k+1} + \cdots + (\Psi_m)A_m = 0$$

then, r, the generalized reaction rate is given by:

$$r = \frac{r_1}{(-1)} = \frac{r_2}{(-\Psi_2)} = \frac{r_k}{(-\Psi_k)} = \frac{r_{k+1}}{(\Psi_{k+1})} = \frac{r_m}{(\Psi_m)}$$

Furthermore, if there are j reactions (where $j = 1 \rightarrow n$) involving i components (where $i = 1 \rightarrow m$), the overall reaction rate for component i will be given by:

$$r_i = \sum_{j=1}^{n} \Psi_{i,j} \cdot r_j$$

If r_i is negative, component i is being consumed, whereas if it is positive, the component is being produced.

4. Knowledge of the yield is required before the stoichiometric equation for microbial growth can be written. If McCarty's half-reaction approach is used to write the stoichiometric equation, f_S, the fraction of the electron donor captured through synthesis, is directly related to the yield.
5. When the electron donor is an organic compound, ammonia serves as the nitrogen source, and the yield is expressed as biomass COD formed per unit of substrate COD used, f_S and Y are equal. For other circumstances, f_S may be either greater than or smaller than Y.
6. Bacteria divide by binary fission. Thus, their rate of growth is first order with respect to the concentration of active biomass present:

$$r_{XB} = \mu X_B$$

The rate coefficient, μ, is called the specific growth rate coefficient. It is influenced by the substrate concentration. If the substrate is noninhibitory, the most commonly used expression is that of Monod:

$$\mu = \hat{\mu} \frac{S_S}{K_S + S_S}$$

If the substrate is inhibitory to its own biodegradation, the Andrews equation is commonly used:

$$\mu = \hat{\mu} \frac{S_S}{K_S + S_S + S_S^2/K_I}$$

7. When the substrate concentration is large relative to K_S, the Monod equation may be simplified to an expression that is zero order with respect to the substrate concentration. When the substrate concentration is small relative to K_S, the specific growth rate coefficient is approximately first order with respect to the substrate concentration.
8. Complementary nutrients are those that meet different needs by growing microorganisms whereas substitutable nutrients are those that meet the same need. The effects of limitation by two complementary nutrients may be depicted by interactive and noninteractive models. The interactive approach is more appropriate for modeling wastewater treatment systems.
9. Biochemical operations can be designed most easily when the nutrient the system is being designed to control acts as the growth limiting nutrient for the biomass in the system.
10. The kinetic parameters in the Monod and Andrews equations depend strongly on the species of microorganism and the substrate upon which the microorganisms are growing. Since wastewater treatment operations use mixed cultures, and wastewaters contain many compounds, the parameters used to describe such operations should be characterized by ranges rather than by single values.
11. Nitrifying bacteria have lower maximum specific growth rate coefficients than heterotrophic bacteria and are more sensitive to pH and to low dissolved oxygen concentrations.

12. The kinetic parameters describing the growth characteristics of the different types of anaerobic microorganisms are difficult to assess because of the strong interactions within the microbial communities. Nevertheless, it can be stated that the aceticlastic methanogens have the lowest maximum specific growth rate coefficient, and thus represent the weak link in the chain.
13. Two approaches have been used to model the loss of viability and biomass in biochemical operations: the traditional decay approach and the lysis:regrowth approach. In the traditional approach, loss of active biomass leads directly to the use of electron acceptor and the production of biomass debris, which accumulates and acts to reduce the viability. In the lysis:regrowth approach, active biomass is lost by lysis, which releases particulate substrate and biomass debris. Electron acceptor consumption occurs only after soluble substrate, which is formed by hydrolysis of the particulate substrate, is used for new biomass growth. Because the yield is always less than one, the amount of new biomass formed is always less than the biomass lost by lysis, leading to a loss of biomass in the bioreactor.
14. Both the traditional and the lysis:regrowth approaches to modeling decay and loss of viability depict the rate of active biomass loss as being first order with respect to the active biomass concentration, as is the generation of biomass debris. However, the decay coefficient in the traditional approach is smaller than the coefficient in the lysis:regrowth approach, although the fraction of the biomass leading to debris is larger.
15. Although soluble microbial product formation is known to occur in biochemical operations, there has been insufficient research on the subject to allow consensus concerning the rate expressions to be used.
16. For modeling purposes, solubilization of particulate and high molecular weight organic matter is assumed to occur by hydrolysis, with conservation of COD. The rate expression adopted to describe hydrolysis is similar to the Monod equation, except that it is controlled by the particulate substrate to biomass ratio rather than by the particulate substrate concentration:

$$r_{XS} = -k_h \left[\frac{X_S/X_{B,H}}{K_X + (X_S/X_{B,H})} \right] X_{B,H}$$

This is necessary because the reaction is thought to be surface mediated.
17. Even though organic nitrogen may be used directly in biomass synthesis, it is simpler to model the flow of nitrogen in biochemical operations by assuming that nitrogen is released to the medium as ammonia and then taken up for biomass synthesis as needed. The release as ammonia, called ammonification, is assumed to be first order with respect to both the biomass and soluble, biodegradable organic nitrogen concentrations. The uptake of ammonia for growth is assumed to be proportional to the rate of growth.
18. During biological phosphorus removal, the uptake of acetic acid, the formation of PHB, and the release of soluble phosphate by PAOs under

anaerobic conditions are all coupled. The rate of acetic acid uptake is controlled by both the acetic acid concentration in solution, S_A, and the polyphosphate concentration in the biomass, X_P:

$$r_{SA} = -\hat{q}_A \left(\frac{S_A}{K_A + S_A}\right)\left[\frac{X_{PP}/X_{B,P}}{K_{PP} + (X_{PP}/X_{B,P})}\right] X_{B,P}$$

Under aerobic conditions, the PAOs grow by using the stored PHB as a carbon and energy source, storing polyphosphate in the process:

$$r_{XBP} = \hat{\mu}_P \left[\frac{X_{PHB}/X_{B,P}}{K_{PHB} + (X_{PHB}/X_{B,P})}\right]\left(\frac{S_P}{K_P + S_P}\right)\left(\frac{S_O}{K_O + S_O}\right) X_{B,P}$$

The rate of phosphorus storage is coupled to the rate of biomass growth and thus is expressed by a similar equation. An additional term is required, however, to reflect the fact that there is a limit to the amount of polyphosphate that the PAOs can accumulate.

19. The COD-based stoichiometric equation states that the COD removed by a biological reaction must equal the oxygen equivalents of the terminal electron acceptor used plus α times the COD of the biomass formed. The value of α depends on the nature of the nitrogen source. It is 1.0 when ammonia is the source and 1.4 when nitrate is.
20. If $C_5H_7O_2N$ can be considered to be representative of the elemental composition of biomass, then 0.087 mg of nitrogen is required to synthesize a mg of biomass COD. Conversely, each time a mg of biomass COD is destroyed by decay, 0.087 mg of nitrogen is released. Although not shown in the empirical equation for biomass, approximately 0.017 mg of phosphorus will be required (released) each time a mg of biomass COD is formed (destroyed).
21. Within a relatively narrow physiological range, the maximum specific growth rate coefficient, $\hat{\mu}$, increases as the temperature is increased. Furthermore, for rapidly growing cultures the effect of temperature on the traditional decay coefficient, b, appears to be closely correlated with the effect on $\hat{\mu}$. No conclusions can be drawn about the effects of temperature on the half-saturation coefficient, K_S.
22. Three expressions are commonly used to relate the rate coefficients in biological operations (k) at different temperatures (T):

$$k = A \cdot e^{-u/RT}$$
$$k_1 = k_2 \cdot e^{(T_1 - T_2)}$$
$$k_1 = k_2 \cdot \theta^{(T_1 - T_2)}$$

The temperature coefficients for the three equations may be interconverted by:

$$\ln(\theta) = C \approx 0.0015\ u$$

3.11 STUDY QUESTIONS

1. Why must the yield be known before the stoichiometric equation for microbial growth can be written? Which type of yield, the true growth yield or the observed yield, is most appropriate for doing this? Why? How is knowledge of the yield used to write the stoichiometric equation using McCarty's half-reaction approach?
2. Using the half-reaction-technique, write the molar stoichiometric equation for microbial growth for each of the following situations:
   a. Aerobic growth on domestic wastewater with ammonia nitrogen as the nitrogen source. The yield is 0.60 mg biomass COD formed/mg substrate COD removed.
   b. Growth on a carbohydrate with nitrate as the terminal electron acceptor and ammonia as the nitrogen source. The yield is 0.50 mg biomass COD formed/mg substrate COD used.
   c. Growth on a carbohydrate with nitrate as the terminal electron acceptor and nitrogen source. The yield is 0.40 mg biomass COD formed/mg substrate COD used.
   d. Normalize them with respect to the electron donor.
3. Convert the molar stoichiometric equation from Study Question 2a into a mass based equation with the electron donor as the reference component.
4. Convert the molar stoichiometric equation from Study Question 2a into a COD based equation with the electron donor as the reference component.
5. Write the rate expression for bacterial growth and relate it to the rates of substrate and oxygen utilization for heterotrophic biomass growth on an organic substrate. Then state the Monod and Andrews equations relating the specific growth rate coefficient to the substrate concentration. Finally, draw sketches depicting the effects represented by both equations and use them to define the parameters in the equations.
6. State the zero- and first-order approximations of the Monod equation. Under what circumstances may they be used?
7. Explain the difference between complementary and substitutable nutrients. Then differentiate between interactive and noninteractive models for describing the effects of two complementary nutrients. Finally, state why the interactive approach was adopted herein.
8. Draw a sketch depicting the effects of two interactive, complementary nutrients on the specific growth rate of biomass and use it to explain why it is easier to design a biochemical operation to achieve a desired concentration of a given nutrient if that nutrient serves as the sole growth limiting nutrient for the biomass.
9. Even though it is best to characterize the kinetic parameters in the Monod and Andrews equations by ranges rather than by unique values, it is possible to state several generalities about the sizes of those parameters. Use such generalities to contrast and compare the growth characteristics of heterotrophic and autotrophic biomass.

10. Discuss the effects that organic compounds and heterotrophic biomass can have on the growth of nitrifying bacteria.
11. Describe the major groups of microorganisms participating in anaerobic operations and contrast their growth characteristics as described by their kinetic parameters.
12. Describe in detail the traditional and lysis:regrowth approaches to modeling the loss of biomass and viability observed in biochemical operations. In your description, contrast the routes of carbon, nitrogen and electron flow, and explain how they influence the magnitudes of the kinetic parameters used to characterize the events.
13. Write the rate equations for loss of active biomass as depicted by the traditional and lysis:regrowth approaches. Then explain the relationships between the kinetic and stoichiometric parameters used in the two approaches.
14. Write the rate equation for the hydrolysis of particulate substrate, compare it to the Monod equation, and explain any differences.
15. Discuss the fate of nitrogen in biochemical operations and state the rate equations used to model that fate.
16. State the rate equations that have been proposed to represent acetic acid uptake, PHB formation, and phosphorus release by PAOs under anaerobic conditions in a biological phosphorus removal system. Then state the rate equations depicting PAO growth, soluble phosphorus uptake, and polyphosphate formation under aerobic conditions. Use those equations in a discussion of the events occurring in such systems and explain why the various terms were included in the rate expressions.
17. An aerobic culture is growing on a mixture of organic matter, such as that found in domestic wastewater, with ammonia as the nitrogen source. How many mg of nitrogen (N), phosphorus (P) and oxygen (O_2) must be provided per mg of COD removed for each of the following situations? What quantities of micronutrients will be required? Do not derive the stoichiometric equations. Rather, answer the question using generalizations presented in the text.
    a. Y_{Hobs} = 0.70 mg biomass COD formed/mg substrate COD removed.
    b. Y_{Hobs} = 0.57 mg biomass COD formed/mg substrate COD removed.
    c. Y_{Hobs} = 0.36 mg biomass COD formed/mg substrate COD removed.
18. Demonstrate why the value of α_N in Eq. 3.90 is 1.40 when nitrate serves as the nitrogen source and biomass is represented by $C_5H_7O_2N$.
19. Three techniques are often used to describe the effects of temperature on microbial cultures. Describe each of them and tell how you would plot data to determine the values of the temperature coefficients in the equations.
20. The data on the following page describe the effects of temperature on the traditional decay coefficient, b. Use that data to determine the temperature coefficient by each of the three techniques. Use 20°C as the reference temperature. Discuss the utility of each technique for describing the effects of temperature on this parameter.

T	b
°C	hr^{-1}
10	0.0037
20	0.0095
30	0.0229
40	0.0372

REFERENCES

1. Aleem, M. I. H., The physiology and chemoautotrophic metabolism of *Nitrobacter agilis*, Ph.D. Thesis, Cornell University, Ithaca, NY, 1959.
2. Andrews, J. F., A mathematical model for the continuous culture of microorganisms utilizing inhibitory substrates. *Biotechnology and Bioengineering* **10**:707–723, 1968.
3. Andrews, J. F., Kinetic models of biological waste treatment. *Biotechnology and Bioengineering Symposium 2*, pp. 5–33, 1971.
4. Anthonisen, A. C., R. C. Loehr, T. B. S. Prakasam and E. G. Srinath, Inhibition of nitrification by ammonia and nitrous acid. *Journal, Water Pollution Control Federation* **48**:835–852, 1976.
5. Arrhenius, S., Über die reaktionsgeschwindigkeit bei der inversion von rohrzucker durch sauren. *Zeitschrift für Physikalische Chemie* **4**:226–248, 1889.
6. Bader, F. G., Kinetics of double-substrate limited growth. In *Microbial Population Dynamics*, M. J. Bazin ed., CRC Press, Boca Raton, FL, pp. 1–32, 1982.
7. Bae, W. and B. E. Rittmann, A structured model of dual-limitation kinetics. *Biotechnology and Bioengineering* **49**:683–689, 1996.
8. Baillod, C. R., Oxygen utilization in activated sludge plants: Simulation and model calibration. U. S. EPA Report No. EPA/600/S2-88/065, 1989.
9. Baltzis, B. C. and A. G. Fredrickson, Limitation of growth rate by two complementary nutrients: Some elementary but neglected considerations. *Biotechnology and Bioengineering* **31**:75–86, 1988.
10. Barnard, J. L., A consolidated approach to activated sludge process design: discussion. *Progress in Water Technology* **7**(1):73–90, 1975.
11. Batchelor, B., Kinetic analysis of alternative configurations for single-sludge nitrification/denitrification. *Journal, Water Pollution Control Federation* **54**:1493–1504, 1982.
12. Blanc, J., J. M. Audic and G. M. Faup, Enhancement of *Nitrobacter* activity by heterotrophic bacteria. *Water Research* **20**:1375–1381, 1986.
13. Boon, B. and H. Laudelot, Kinetics of nitrite oxidation by *Nitrobacter winogradsky*. *Biochemistry Journal* **85**:440–447, 1962.
14. Bryers, J.D., Structured modeling of the anaerobic digestion of biomass particulates. *Biotechnology and Bioengineering* **27**:638–649, 1985.
15. Bryers, J. D. and C. A. Mason, Biopolymer particulate turnover in biological waste treatment systems: a review. *Bioprocess Engineering* **2**:95–109, 1987.
16. Carter, J. L. and R. E. McKinney, Effects of iron on activated sludge treatment. *Journal of the Environmental Engineering Division, ASCE* **99**:135–152, 1973.
17. Characklis, W. G. and W. Gujer, Temperature dependency of microbial reactions. In *Kinetics of Wastewater Treatment*, S. H. Jenkins ed., Pergamon Press:Elmsford, NY, pp. 111–130, 1979.

18. Chiu, S. Y., L. E. Erickson, L. T. Fan and I. C. Kao, Kinetic model identification in mixed populations using continuous culture data. *Biotechnology and Bioengineering* **14**:207–231, 1972.
19. Chiu, S. Y., L. T. Fan, I. C. Kao and L. E. Erickson, Kinetic behavior of mixed populations of activated sludge. *Biotechnology and Bioengineering* **14**:179–199, 1972.
20. Christensen, M. J. and P. Harremoës, Biological denitrification of sewage: a literature review. *Progress in Water Technology*, **8**(4/5):509–555, 1977.
21. Chudoba, J., J. S. Cech, J. Farkac and P. Grau, Control of activated sludge filamentous bulking: experimental verification of a kinetic selection theory. *Water Research* **19**: 191–196, 1985.
22. Cobb, J. B. and K. L. Murphy, Estimation of the active nitrifying biomass in activated sludge. *Water Research* **29**:1855–1862, 1995.
23. Dang, J. S., D. M. Harvey, A. Jobbagy and C. P. L. Grady Jr., Evaluation of biodegradation kinetics with respirometric data. *Research Journal, Water Pollution Control Federation* **61**:1711–1721, 1989.
24. Delwiche, C. C. and B. A. Bryan, Denitrification. *Annual Review of Microbiology* **30**: 241–262, 1976.
25. Dold, P. L. and G. v. R. Marais, Evaluation of the general activated sludge model proposed by the IAWPRC task group. *Water Science and Technology* **18**(6):63–89, 1986.
26. Dold, P. L., G. A. Ekama and G. v. R. Marais, A general model for the activated sludge process. *Progress in Water Technology* **12**(6):47–77, 1980.
27. Duarte, A. C. and G. K. Anderson, Inhibition modelling in anaerobic digestion. *Water Science and Technology* **14**(4/5–6/7):749–763, 1982.
28. Eastman, J. A. and J. F. Ferguson, Solubilization of particulate organic carbon during the acid phase of anaerobic digestion. *Journal, Water Pollution Control Federation* **53**: 352–366, 1981.
29. Eckenfelder, W. W. Jr., *Industrial Water Pollution Control*. 2nd ed, McGraw-Hill:New York, N. Y., 1989.
30. Eckenfelder, W. W. Jr. and J. L. Musterman, Activated sludge treatment of industrial waters. In *Activated Sludge Process Design and Control: Theory and Practice*, W. W. Eckenfelder and P. Grau eds., Technomic Publishing:Lancaster, PA, pp. 127–266, 1992.
31. Eckhoff, D. W. and D. Jenkins, Activated sludge systems, kinetics of the steady and transient states. Report No. 67–12 of the Sanitary Engineering Research Laboratory, University of California, Berkeley, 1967.
32. Edwards, V. H., The influence of high substrate concentrations on microbial kinetics. *Biotechnology and Bioengineering* **12**:679–712, 1970.
33. Engberg, D. J. and E. D. Schroeder, Kinetics and stoichiometry of bacterial denitrification as a function of cell residence time. *Water Research* **9**:1051–1054, 1975.
34. Fencl, Z., Theoretical analysis of continuous culture systems. In *Theoretical and Methodological Basis of Continuous Culture of Microorganisms*, I. Malek and Z. Fencl eds., Academic Press, New York, NY, pp. 67–153, 1966.
35. Garrett, M. T. and C. N. Sawyer, Kinetics of removal of soluble BOD by activated sludge. *Proceedings of the 7th Industrial Waste Conference* Purdue University Engineering Extension Series No. 79, pp. 51–77, 1952.
36. Gaudy A. F. Jr. and E. T. Gaudy, Biological concepts for design and operation of the activated sludge process. *Environmental Protection Agency Water Pollution Research Series*, Report #17090 FQJ 09/71, Sept. 1971.
37. Gaudy, A. F., M. Ramanathan, P. Y. Yang and T. V. DeGeare, Studies on the operational stability of the extended aeration process. *Journal, Water Pollution Control Federation* **42**:165–179, 1970.

38. Ghosh, S. and F. G. Pohland, Population dynamics in continuous cultures of heterogeneous microbial populations. *Developments in Industrial Microbiology* **12**:295–311, 1971.
39. Goel, K. C. and A. F. Gaudy Jr., Studies on the relationship between specific growth rate and concentration of nitrogen source for heterogeneous microbial population of sewage origin. *Biotechnology and Bioengineering* **11**:67–78, 1969.
40. Goodman, B. L. and A. J. Englande, Jr., A unified model of the activated sludge process. *Journal, Water Pollution Control Federation* **46**:312–332, 1974.
41. Grady, C. P. L. Jr., P. L. Findley and R. E. Muck, Effects of growth conditions on the oxygen equivalence of microbial cells. *Biotechnology and Bioengineering* **17**:859–872, 1975.
42. Grady, C. P. L. Jr., G. Aichinger, S. F. Cooper and M. Naziruddin, Biodegradation kinetics for selected toxic/hazardous organic compounds. In *Proceedings of the 1989 AWMA/EPA International Symposium on Hazardous Waste Treatment: Biosystems for Pollution Control*, Air and Waste Management Association, Pittsburgh, PA, pp. 141–153, 1989.
43. Grau, P., P. M. Sutton, M. Henze, S. Elmaleh, C. P. L. Grady Jr., W. Gujer and J. Koller, Notation for use in the description of wastewater treatment processes. *Water Research* **21**:135–139, 1987.
44. Gujer, W., The effect of particulate organic material on activated sludge yield and oxygen requirement. *Progress in Water Technology* **12**(6):79–95, 1980.
45. Gujer, W. and A. J. B. Zehnder, Conversion processes in anaerobic digestion. *Water Science and Technology* **15**(8/9):127–167, 1983.
46. Haldane, J. B. S., *Enzymes*. Longmans, London, 1930.
47. Hall, E. R. and K. L. Murphy, Sludge age and substrate effects on nitrification kinetics. *Journal, Water Pollution Control Federation* **57**:413–418, 1985.
48. Han, K. and O. Levenspiel, Extended Monod kinetics for substrate, products and cell inhibition. *Biotechnology and Bioengineering* **32**:430–437, 1988.
49. Hao, O. J. and A. O. Lau, Kinetics of microbial by-product formation in chemostat pure cultures. *Journal of Environmental Engineering* **114**:1097–1115, 1988.
50. Hartmann, L., and G. Laubenberger, Toxicity measurements in activated sludge. *Journal of the Sanitary Engineering Division, ASCE* **94**:247–255, 1968.
51. Heijnen, J. J. and J. P. van Dijken, In search of a thermodynamic description of biomass yields for the chemotrophic growth of microorganisms. *Biotechnology and Bioengineering* **39**:833–858, 1992.
52. Heijnen, J. J., M. C. M. van Loosdrecht, and L. Tijhuis, A black box mathematical model to calculate auto- and heterotrophic yields based on Gibbs energy dissipation. *Biotechnology and Bioengineering* **40**:1139–1154, 1992.
53. Henze, M. and P. Harremoës, Anaerobic treatment of wastewater in fixed film reactors—a literature review. *Water Science and Technology* **15**(8/9):1–101, 1983.
54. Henze, M., C. P. L. Grady Jr., W. Gujer, G. V. R. Marais, and T. Matsuo, A general model for single-sludge wastewater treatment systems. *Water Research* **21**:505–515, 1987.
55. Henze, M., C. P. L. Grady Jr., W. Gujer, G. v. R. Marais and T. Matsuo, Activated sludge model No. 1. *IAWPRC Scientific and Technical Reports*, No. 1, 1987.
56. Henze, M., W. Gujer, T. Mino, T. Matsuo, M. C. Wentzel, and G. v. R. Marais, Activated sludge model No. 2. *IAWQ Scientific and Technical Reports*, No. 3, 1995.
57. Herbert, D., The chemical composition of microorganisms as a function of their environment. In *Microbial Reaction to Environment*, 11th Symposium of the Society for General Microbiology, Cambridge University Press, England, pp. 391–416, 1961.
58. Heukelekian, H., The effect of carbonaceous materials on nitrification. *Proceedings of the 4th International Congress on Microbiology*, 460, 1947.

59. Hockenbury, M. R. and C. P. L. Grady Jr., Inhibition of nitrification—Effects of selected organic compounds. *Journal, Water Pollution Control Federation* **49**:768–777, 1977.
60. Hockenbury, M. R., G. T. Daigger and C. P. L. Grady Jr. Factors affecting nitrification. *Journal of the Environmental Engineering Division, ASCE*, **103**:9–19, 1977.
61. Hoover, S. R. and N. Porges, Assimilation of dairy wastes by activated sludge. II. The equations of synthesis and rate of oxygen utilization. *Sewage and Industrial Wastes* **24**:306–312, 1952.
62. Irvine, R. L. and J. D. Bryers, Stoichiometry and Kinetics of Waste Treatment. In *Comprehensive Biotechnology, Vol 4, The Principles, Applications and Regulations of Biotechnology in Industry, Agriculture and Medicine*, C. W. Robinson and J. A. Howell eds., Pergamon Press, New York, NY, pp. 757–772, 1985.
63. Jewell, W. J. and P. L. McCarty, Aerobic decomposition of algae. *Environmental Science and Technology* **5**:1023–1031, 1971.
64. Jorden, W. L., F. G. Pohland and B. H. Kornegay, Evaluating treatability of selected industrial wastes. *Proceedings of the 26th Industrial Waste Conference*, Purdue University Engineering Extension Series No. 140, pp. 514–529, 1971.
65. Keener, W. K. and D. J. Arp, Kinetic studies of ammonia monooxygenase inhibition in *Nitrosomonas europaea* by hydrocarbons and halogenated hydrocarbons in optimized whole-cell assay. *Applied and Environmental Microbiology* **59**:2501–2510, 1993.
66. Knowles, G., A. L. Downing, and M. J. Barrett, Determination of kinetic constants for nitrifying bacteria in mixed culture with the aid of an electronic computer. *Journal of General Microbiology* **38**:263–276, 1965.
67. Knowles R., Denitrification. *Microbiological Reviews* **46**:43–70, 1982.
68. Krul, J. M., The relationship between dissimilatory nitrate reduction and oxygen uptake by cells of an *Alcaligenes* strain in flocs and in suspension and by activated sludge flocs. *Water Research* **10**:337–341, 1976.
69. Labib, F., J. F. Ferguson, M. M. Benjamin, M. Merigh, and N. L. Ricker, Anaerobic butyrate degradation in a fluidized-bed reactor: Effects of increased concentrations of H_2 and acetate. *Environmental Science and Technology* **26**:369–376, 1992.
70. Lau, A. O., P. F. Strom and D. Jenkins, Growth kinetics of *Sphaerotilus natans* and a floc former in pure and dual continuous culture. *Journal, Water Pollution Control Federation* **56**:41–51, 1984.
71. Lawrence, A. W. and P. L. McCarty, Unified basis for biological treatment design and operation. *Journal of the Sanitary Engineering Division, ASCE* **96**:757–778, 1970.
72. Machado, R. J. and C. P. L. Grady Jr., Dual substrate removal by an axenic bacterial culture. *Biotechnology and Bioengineering* **33**:327–337, 1989.
73. Mason, C. A., J. D. Bryers and G. Hamer, Activity, death and lysis during microbial growth in a chemostat. *Chemical Engineering Communications* **45**:163–176, 1986.
74. Mason, C. A., G. Hamer and J. D. Bryers, The death and lysis of microorganisms in environmental processes. *FEMS Microbiology Reviews* **39**:373–401, 1986.
75. McCarty, P. L., Thermodynamics of biological synthesis and growth. *Proceedings of the Second International Conference on Water Pollution Research*, Pergamon Press, New York, NY, pp. 169–199, 1965.
76. McCarty, P. L., Phosphorus and nitrogen removal by biological systems. *Proceedings of the Wastewater Reclamation and Reuse Workshop*, Lake Tahoe, CA, 226, June 1970.
77. McCarty, P. L., Stoichiometry of biological reactions. *Progress in Water Technology* **7**(1):157–172, 1975.
78. McCarty, P. L. and C. F. Brodersen, Theory of extended aeration activated sludge. *Journal, Water Pollution Control Federation* **34**:1095–1103, 1962.

79. McClintock, S. A., J. H. Sherrard, J. T. Novak and C. W. Randall, Nitrate versus oxygen respiration in the activated sludge process. *Journal, Water Pollution Control Federation* **60**:342–350, 1988.
80. McKinney, R. E., Mathematics of complete mixing activated sludge. *Journal of the Sanitary Engineering Division, ASCE* **88**(SA3):87–113, 1962.
81. McKinney, R. E. and R. J. Ooten, Concepts of complete mixing activated sludge. *Transactions of the 19th Annual Conference on Sanitary Engineering*, University of Kansas, 32–59, 1969.
82. McLellan, J. C. and A. W. Busch, Hydraulic and process aspects of reactor design. I. Basic concepts in steady-state analysis. *Proceedings of the 22nd Industrial Waste Conference*, Purdue University Engineering Extension Series No. 129, pp. 537–552, 1967.
83. Monod, J., The growth of bacterial cultures. *Annual Review of Microbiology* **3**:371–394, 1949.
84. Moser, H. The dynamics of bacterial populations maintained in the chemostat. *Carnegie Institute of Washington, Publication No. 614*, 1958.
85. Muck, R. E. and C. P. L. Grady Jr., Temperature effects on microbial growth in CSTR's. *Journal of the Environmental Engineering Division, ASCE* **100**:1147–1163, 1974.
86. Mulchandani A. and J. H. T. Luong, Microbial inhibition kinetics revisited. *Enzyme and Microbial Technology* **11**:66–73, 1989.
87. Neidhardt, F. C., J. L. Ingraham and M. Schaechter, *Physiology of the Bacterial Cell: A Molecular Approach*, Sinauer Associates, Inc., Sunderland, MA, 1990.
88. Neufeld, R. D., A. J. Hill and D. O. Adekoya, Phenol and free ammonia inhibition to *Nitrosomonas* activity. *Water Research* **14**:1695–1703, 1980.
89. Novak, J. T., Temperature-substrate interactions in biological treatment. *Journal, Water Pollution Control Federation* **46**:1984–1994, 1974.
90. Obayashi, A. W. and A. F. Gaudy, Jr., Aerobic digestion of extracellular microbial polysaccharides. *Journal, Water Pollution Control Federation* **45**:1584–1594, 1973.
91. Oslislo, A. and Z. Lewandowski, Inhibition of nitrification in the packed bed reactors by selected organic compounds. *Water Research* **19**:423–426, 1985.
92. Pantea-Kiser, L., R. F. Wukasch and J. E. Alleman, The effect of inhibitory compounds on biological nitrification. *Proceedings of the 44th Industrial Waste Conference*, 1989, Purdue University, Lewis Publishers, Chelsea, MI, pp. 465–474, 1989.
93. Parker, D. S., R. W. Stone and R. J. Stenquist, *Process Design Manual for Nitrogen Control*, U.S. Environmental Protection Agency, Technology Transfer, Oct. 1975.
94. Payne, W. J., Reduction of nitrogenous oxides by microorganism. *Bacteriological Reviews* **37**:409–452, 1973.
95. Perez-Padilla, G. G. III, Effects of toxic organic chemicals on microbial death and lysis, Ph.D. Dissertation, Clemson University, Clemson, SC, 1996.
96. Phelps, E. B., *Stream Sanitation*, John Wiley & Sons, Inc., New York, NY, 71–75, 1944.
97. Powell, E.O., The growth rate of microorganisms as a function of substrate concentration. In *Microbial Physiology and Continuous Culture*, E. O. Powell, *et al.* eds., Her Majesty's Stationery Office, London, pp. 34–55, 1967.
98. Quinlan, A. V., Prediction of the optimum pH for ammonia-N oxidation by *Nitrosomonas europaea* in well-aerated natural and domestic-waste waters. *Water Research* **18**:561–566, 1984.
99. Ribbons, D. W., Quantitative relationships between growth media constituents and cellular yields and composition. In *Methods in Microbiology*, Vol. 3A, J.R. Norris and D.W. Ribbons eds., Academic Press, New York, pp. 297–304, 1970.
100. Rittenberg, S. C., The roles of exogenous organic matter in the physiology of chemolithotrophic bacteria. *Advances in Microbial Physiology* **3**:159–196, 1969.

101. Rittmann, B. E. and W. E. Langeland, Simultaneous denitrification with nitrification in single-channel oxidation ditches. *Journal, Water Pollution Control Federation* **57**: 300–308, 1985.
102. Rittmann, B. E. and V. L. Snoeyink, Achieving biologically stable drinking water. *Journal, American Water Works Association* **76**(10):106–114, 1984.
103. Rittmann, B. E., W. Bae, E. Namkung, and C.-J. Lu, A critical evaluation of microbial product formation in biological processes. *Water Science and Technology* **19**(7):517–528, 1987.
104. Rozich, A. F., A. F. Gaudy Jr. and P. C. D'Adamo, Selection of growth rate model for activated sludges treating phenol. *Water Research* **19**:481–490, 1985.
105. Ryder, D. N. and C. G. Sinclair, Model for the growth of aerobic microorganisms under oxygen limiting conditions. *Biotechnology and Bioengineering* **14**:787–798, 1972.
106. Sahm, H., Anaerobic wastewater treatment. *Advances in Biochemical Engineering and Biotechnology* **29**:83–115, 1984.
107. Schaezler, D. J., A. W. Busch and C. H. Ward, Kinetic and stoichiometric limitations of phosphate in pure and mixed bacterial cultures. *Proceedings of the 24th Industrial Waste Conference*, Purdue University Engineering Extension Series No. 135, pp. 507–533, 1969.
108. Schulze, K. L. and R. S. Lipe, Relationship between substrate concentration, growth rate, and respiration in *Escherichia coli* in continuous culture. *Archiv fur Mikrobiologie* **48**:1–20, 1964.
109. Sharma, B. and R. C. Ahlert, Nitrification and nitrogen removal. *Water Research* **11**: 897–925, 1977.
110. Sherrard, J. H. Kinetics and stoichiometry of completely mixed activated sludge. *Journal, Water Pollution Control Federation* **49**:1968–1975, 1977.
111. Siegrist, H. and W. Gujer, Demonstration of mass transfer and pH effects in a nitrifying biofilm. *Water Research* **21**:1481–1487, 1987.
112. Simpkin, T. J. and W. C. Boyle, The lack of repression by oxygen of the denitrifying enzymes in activated sludge. *Water Research* **22**:201–206, 1988.
113. Sinclair, C. G. and D. N. Ryder, Models for the continuous culture of microorganism under both oxygen and carbon limiting conditions. *Biotechnology and Bioengineering* **17**:375–398, 1975.
114. Speece, R. E., Environmental requirements for anaerobic digestion of biomass. *Advances in Solar Energy* **2**:51–123, 1985.
115. Srinath, E. G., R. C. Loehr and T. B. S. Prakasam, Nitrifying organism concentration and activity. *Journal of the Environmental Engineering Division, ASCE* **102**:449–463, 1976.
116. Stankewich, M. J. Jr., Biological nitrification with the high purity oxygenation process. *Proceedings of the 27th Industrial Waste Conference*, Purdue University Engineering Extension Series No. 141, pp. 1–23, 1972.
117. Stenstrom, M. K., A dynamic model and computer compatible control strategy for wastewater treatment plants. Ph.D. Dissertation, Clemson University, Clemson, SC, 1975.
118. Stouthamer, A. H., Biochemistry and genetics of nitrate reductase in bacteria. *Advances in Microbial Physiology* **14**:315–375, 1976.
119. Sykes, R. M., Identification of the limiting nutrient and specific growth rate. *Journal, Water Pollution Control Federation* **45**:888–895, 1973.
120. Tanaka, H., S. Uzman and I. J. Dunn, Kinetics of nitrification using a fluidized sand bed reactor with attached growth. *Biotechnology and Bioengineering* **23**:1683–1702, 1981.

121. Tischler, L. F. and W. W. Eckenfelder, Linear substrate removal in the activated sludge process. In *Advances in Water Pollution Research, Proceedings of the Fourth International Conference, Prague*, S. H. Jenkins, ed., Pergamon Press, Oxford, pp. 361–374, 1969.
122. Tomlinson, T. G., A. G. Boon and G. N. A. Trotman, Inhibition of nitrification in the activated sludge process of sewage disposal. *Journal of Applied Bacteriology* **29**: 266–291, 1966.
123. van Haandel, A. C., G. A. Ekama and G. v. R. Marais, The activated sludge process. 3. Single sludge denitrification. *Water Research* **15**:1135–1152, 1981.
124. van Uden, N., Kinetics of nutrient-limited growth. *Annual Review of Microbiology* **23**: 473–486, 1969.
125. Volskay, V. T. Jr. and C. P. L. Grady Jr., Toxicity of selected RCRA compounds to activated sludge microorganisms. *Journal, Water Pollution Control Federation* **60**: 1850–1856, 1988.
126. Volskay, V. T., H. H. Tabak and C. P. L. Grady Jr., Effect of selected RCRA compounds on activated sludge activity. *Research Journal, Water Pollution Control Federation* **62**: 654–664, 1990.
127. Wood, D. K. and G. Tchobanoglous, Trace elements in biological waste treatment. *Journal, Water Pollution Control Federation* **47**:1933–1945, 1975.
128. Zinder, S. H., Physiological ecology of methanogens. In *Methanogenesis: Ecology, Physiology, Biochemistry & Genetics*, J. G. Ferry ed., Chapman & Hall, New York, NY, pp. 128–206, 1993.

This page intentionally left blank

Part II

Theory: Modeling of Ideal Suspended Growth Reactors

The primary function of a mathematical model is to reduce a complex system to the minimum terms essential for its description so that those terms may be manipulated, thereby helping us to understand how the system will respond under a variety of conditions. Generally, mathematical models do not describe a system completely, but if the terms are chosen with care, the model response will be qualitatively similar to the real system. In Part I we considered in detail the major events occurring in biochemical operations. Now the mathematical descriptions of those events will be incorporated into mass balance equations for the major reacting components in order to develop mathematical models describing a number of reactor configurations representing suspended growth systems. Chapter 4 presents the techniques for describing both ideal and nonideal reactors in mathematical terms. Chapter 5 establishes several fundamental principles governing the performance of suspended growth biochemical operations by considering the situation of heterotrophic microbial growth on a soluble substrate in a single ideal reactor. Chapter 6 extends the concepts of Chapter 5 by adding additional reactions, such as autotrophic growth of nitrifying bacteria, in a single ideal reactor. In Chapter 7, other reactor configurations are included to demonstrate how the engineer can control system performance through selection of the appropriate reaction environment. Finally, Chapter 8 describes techniques whereby the kinetic and stoichiometric parameters used in the models may be evaluated. In investigating reactor performance through modeling, it will be assumed that the reactors are ideal, with respect to both fluid flow and the response of the microbial culture. In other words, we will investigate how the reactors would respond if the mathematical models were absolutely correct. In Part III, any significant deviations from ideality are discussed and incorporated into the application of the models to design.

This page intentionally left blank

4

Modeling Suspended Growth Systems

In Chapter 1, we saw that many types of biochemical operations are used in wastewater treatment. Understanding how each performs would be a difficult task if we had to approach each as a unique entity. Fortunately, there is a great deal of commonality among them because only a finite number of events occur within them, as seen in Chapter 2. This suggests that the major differences among them arise from their reactor configurations. Because the kinetics of the events within biochemical operations are reasonably well established, reactor engineering principles can be applied to see how reactor configuration influences their outcome, thereby setting the stage for understanding how the various types of biochemical operations perform.

Reactor engineering is the application of mathematical modeling to the analysis and design of chemical and biochemical reactors. In this chapter we will review briefly some of the basic concepts of reactor engineering. For more complete coverage, the reader should consult texts on the subject.[7,8]

4.1 MODELING MICROBIAL SYSTEMS

Microbial systems are extremely complex and the models describing them can be very complicated. Luckily, relatively simple models have proven to be satisfactory for describing the performance of many biochemical operations and we will focus on them because our intent is to gain an appreciation for the basic manner in which the systems function. For that purpose, additional complexity is more likely to confuse than to clarify. All of the models that we will consider are transport phenomena models in that they are based on the conservation of mass, momentum, and energy.[2] Furthermore, they are primarily phenomenological because the rate expressions in them, which are presented in Chapter 3, seek to depict, in simple terms, the basic mechanisms involved. At the same time, the models are also empirical because the ultimate justification for use of those rate expressions is observation and experience rather than derivation from first principles.

Many simplifying assumptions have been made by the developers of the rate expressions and models used in this text. Although many are implicit, it is important that the more common ones be stated explicitly.[2,6] The first assumption is that all of the organisms of a given species are the same within a given reactor in a system. In reality, individual microorganisms will differ in their physiological states be-

cause they will be at different points in their lifecycles. However, little is known about the impact of those states and the mathematics of considering them would be complex. Thus, they will not be considered. The second assumption is that stochastic phenomena can be neglected, i.e., that any random differences among the cells of a given type can be ignored. This assumption causes few problems because the enormous number of cells present in most biochemical operations causes the random deviations to be canceled out. The third assumption is that within a functional class (e.g., aerobic heterotroph, autotroph, etc.), all microorganisms are treated as if they were the same species. While this is never true in wastewater treatment systems, the problems associated with modeling dozens (if not hundreds) of distinct species are obvious. However, distinctions will be made between the important groups discussed in Chapters 2 and 3. The fourth assumption is an extension of the third in that within a given class of microorganism, individuals are ignored. That is, our focus will be on the mass of organisms present, rather than on the individual members forming that mass. This assumption works reasonably well as long as growth is balanced because then changes in mass and numbers are proportional. Furthermore, the organisms will be considered to be distributed evenly throughout the culture within a given vessel. Finally, even though the microorganisms form a distinct phase within the reactor, the reactions within suspended growth reactors will be treated as if they were homogeneous. This assumption allows the transport of reactants from the liquid phase to the solid (biomass) phase to be ignored, thereby greatly simplifying the modeling task. Although transport is indeed important, particularly in flocculent systems like activated sludge,[1] imposition of this assumption causes no great difficulties as long as it is recognized that parameters like the half-saturation coefficient in the Monod equation (Eq. 3.36) are influenced by transport effects,[10] thereby making them dependent on the physical characteristics of the biomass, such as floc size.

4.2 MASS BALANCE EQUATION

Transport phenomena models are based on the conservation of mass, momentum, and energy. However, for most suspended growth bioreactor models, only mass balances are required and thus our focus will be on them. Furthermore, because the elemental composition of many reactants and products in biochemical operations are unknown, it is usually more convenient to work in mass units than in molar units (see Section 3.1.1).

The starting point for a mass balance on any system is the specification of the control volume, or the boundary of the system. When the reaction conditions, including the composition, are uniform over the whole reactor volume, then the entire reactor may be taken as the control volume. Otherwise, a differential element should be used. Having picked an appropriate control volume, a mass balance must be written for each reactant or product around the control volume by keeping track of the amounts entering, leaving, being generated, and being consumed.

The mass balance for a given constituent takes the form:

Net rate of accumulation in the control volume
= rate of flow into the control volume
− rate of flow out of the control volume
+ net rate of generation in the control volume (4.1)

or simply

Accumulation = input − output + generation (4.2)

Each term in the mass balance equation has the units of mass/time. The generation term represents the sum of all reactions in which the constituent of interest participates, and incorporates the reaction terms seen in Chapter 3, as expressed in Eq. 3.12. If it is positive, the constituent is being produced in the control volume; if it is negative, the constituent is being destroyed. Furthermore, as seen in Chapter 3, the reaction term for any given constituent may be a function of the concentrations of several constituents. If so, it will be necessary to solve several mass balance equations simultaneously to determine the concentrations of the constituents in the control volume.

4.3 REACTOR TYPES

As seen in Chapter 1, suspended growth biochemical operations employ a number of different types of reactors. Most of them are continuous flow, which means that liquid flows through them continuously, bringing in reactants and carrying away products. On occasion, however, environmental engineers use batch reactors, which have no flow through them while the reaction is occurring. Rather, they are loaded, allowed to react, and then unloaded.

4.3.1 Ideal Reactors

Continuous Stirred Tank Reactor. A continuous stirred tank reactor (CSTR), also known as a continuous-flow stirred tank reactor (CFSTR), backmix reactor, or completely mixed reactor, is used frequently, particularly for experimental studies. As shown in Figure 4.1, it has a feed stream called the influent and an exit stream called the effluent. It is usually equipped with baffles and is mixed sufficiently to make mixing perfect, i.e., mixing is homogeneous and instantaneous so that any reactant carried into the reactor by the feed is dispersed evenly throughout the reactor without any time delay. Thus, samples taken from all parts of the reactor have the same composition. In addition, the effluent composition is the same as the reactor composition.

Under the above assumptions, it is logical to pick the entire reactor volume as the control volume for writing mass balance equations. Application of Eq. 4.2 on reactant A around the entire reactor volume, V, yields:

$$V \cdot \frac{dC_A}{dt} = F_0 \cdot C_{A0} - F \cdot C_A + r_A \cdot V \tag{4.3}$$

where F_0 and F are the volumetric flow rates of the influent and effluent, and C_{A0}

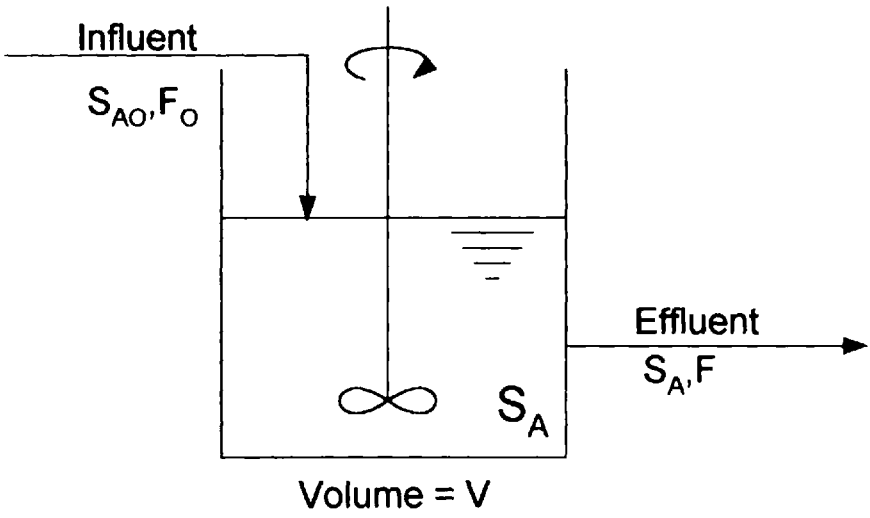


Figure 4.1 Continuous stirred tank reactor (CSTR).

and C_A are the concentrations of A in the influent and effluent (or reactor), respectively. The generalized concentration symbol C has been used here to emphasize the point that the mass balance equation may be written for any constituent, regardless of its state, i.e., soluble or particulate. For the systems with which we deal it is safe to assume that the flow rates of the influent and effluent are equal, i.e., $F_0 = F$. If the influent flow rate and concentration are constant, it is common for such reactors to achieve steady-state, which means that the time rate of change of any constituent is zero. This allows the accumulation term to be set equal to zero, thereby simplifying Eq. 4.3:

$$-r_A = \frac{F}{V}(C_{AO} - C_A) \tag{4.4}$$

Since the right side of Eq. 4.4 is the difference between the mass flow rates of A in the influent and effluent per unit volume of reactor, and the left side is the net generation rate of A in the reactor, Eq. 4.4 states that the difference in the mass flow rates of A into and out of the reactor is due to the generation of A in the reactor. Note that if A is consumed, C_A will be less than C_{AO}, and the generation rate will be negative. A negative generation rate is often called a consumption rate.

A steady-state mass balance equation like Eq. 4.4 can be used for several things. First, because it allows calculation of the generation rate of A, it may be used to obtain a reaction rate expression experimentally. By varying the influent flow rate, F, the reactor volume, V, or the influent concentration, C_{AO}, and measuring the corresponding effluent concentration it is possible to determine how the reaction rate varies with the concentration of A (as well as the concentration of any other reactants). On the other hand, if the reaction rate expression is known, a rearranged form of Eq. 4.4 can be used to determine the reactor volume required to achieve a desired effluent concentration:

$$V = \frac{F(C_{A0} - C_A)}{-r_A} \tag{4.5}$$

Finally, Eq. 4.4 can also be rearranged to allow calculation of the flow rate that can be handled by a reactor of volume V:

$$F = \frac{-r_A \cdot V}{C_{A0} - C_A} \tag{4.6}$$

Thus, it can be seen that the mass balance equation for a CSTR is a powerful tool.

Plug–Flow Reactor. A plug–flow reactor (PFR) can be either a simple tube or one packed with a catalyst or some other type of packing. A feed containing the reactants is fed continuously to the reactor inlet while the effluent containing the products and unreacted reactants is removed from the outlet. The name of the reactor comes from the assumption that the flow pattern inside has a uniform velocity and concentration in the radial direction, and no axial mixing, so that each fluid element moves through the reactor in the same order relative to all other elements, just like ping-pong balls rolling through a tube. The PFR is also known as a tubular or piston-flow reactor.

Because of the assumption of plug flow and because the reaction takes place all along the reactor length, the concentrations of reactants and products vary with the axial distance only. Therefore, it is appropriate to consider as the control volume an infinitesimal volume, ΔV, in which the concentration may be considered uniform. This control volume is shown in Figure 4.2.

The mass balance on component A around the control volume is

$$\frac{\partial(A_c \cdot \Delta x \cdot C_A)}{\partial t} = F \cdot C_A|_x - F \cdot C_A|_{x+\Delta x} + r_A \cdot A_c \cdot \Delta x \tag{4.7}$$

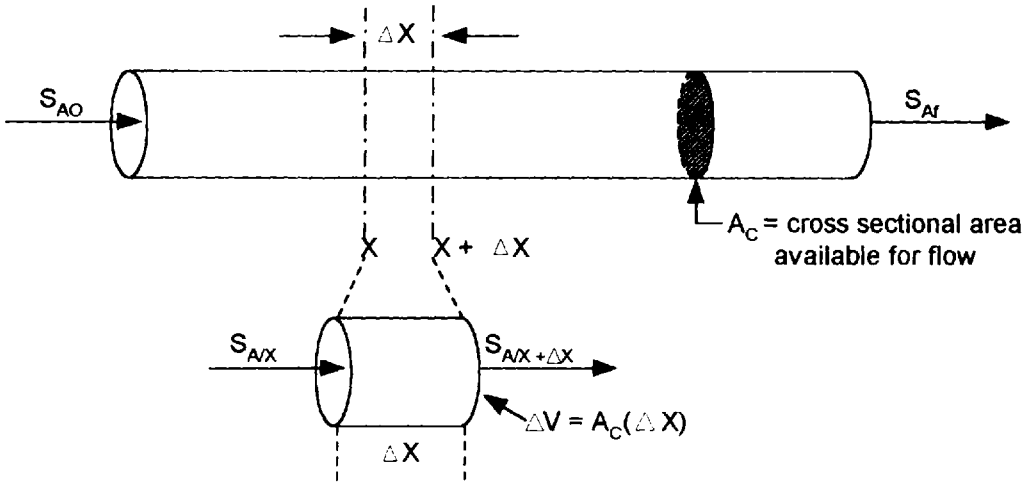


Figure 4.2 Plug-flow reactor (PFR).

or

$$A_c \frac{\partial C_A}{\partial t} = -\frac{F \cdot C_A|_{x+\Delta x} - F \cdot C_A|_x}{\Delta x} + r_A \cdot A_c \tag{4.8}$$

where A_c is the cross-sectional area of the reactor, x the distance from the reactor entrance, Δx the length of the infinitesimal volume, and $F \cdot C_A|_x$ and $F \cdot C_A|_{x+\Delta x}$ the mass flow rates of A evaluated at the distances x and $x + \Delta x$ from the reactor entrance. In the limit as $\Delta x \to 0$, the first term on the right side in Eq. 4.8 becomes the partial derivative of $F \cdot C_A$ with respect to distance and Eq. 4.8 reduces to

$$A_c \frac{\partial C_A}{\partial t} = -\frac{\partial (F \cdot C_A)}{\partial x} + r_A \cdot A_c \tag{4.9}$$

At constant flow rate, the reactor will achieve steady-state, at which there is no change with respect to time in the concentration at any point within the reactor, reducing the equation to

$$0 = -F \frac{dC_A}{dx} + r_A \cdot A_c \tag{4.10}$$

or

$$r_A = \frac{F}{A_c} \cdot \frac{dC_A}{dx} \tag{4.11}$$

Hence, it is theoretically possible to calculate the rates by determining the concentration gradient through the reactor. In practice, however, this is difficult to do so plug–flow reactors are seldom used to generate rate data in this fashion.

The main purpose for which the mass balance equation is used is to determine the size of reactor required to achieve a desired conversion from a feed stream of given composition and flow rate. If the rate expression is known, Eq. 4.11 can be rearranged and integrated over the length of the reactor, L, to give

$$\int_0^L \frac{A_c \cdot dx}{F} = \frac{A_c \cdot L}{F} = \frac{V}{F} = \int_{C_{A0}}^{C_A} \frac{dC_A}{r_A} \tag{4.12}$$

Hence, the required ratio of reactor volume to feed flow rate, V/F, can be obtained.

Batch Reactor. The CSTR and the PFR are both continuous flow reactors. As stated previously, however, environmental engineers sometimes use batch reactors which do not receive flow throughout the entire operational cycle. For the simplest cycle, they are rapidly charged with feed, allowed to react, and the treated effluent removed. Under such conditions a batch reactor has no in or out terms in the mass balance equation during the reaction period. As in a CSTR, a batch reactor is assumed to be perfectly mixed so that the entire reactor content is homogeneous at any given time. Hence, the rate is independent of position in the reactor and the concentration varies only with time. Therefore, it is appropriate to take the whole reactor volume as the control volume. The mass balance on reactant A around a batch reactor is

$$\frac{d(C_A \cdot V)}{dt} = 0 - 0 + r_A \cdot V \tag{4.13}$$

For constant volume, this simplifies to

$$r_A = \frac{dC_A}{dt} \tag{4.14}$$

which states that the rate is equal to the time rate of change of the reactant concentration.

Equation 4.14 suggests that by running a batch reactor experiment, measuring the concentrations of appropriate reactants over time, and taking the time derivatives of the concentration history, one can obtain numerical values of the rate, r_A, associated with various concentrations of reactants. Having obtained the data of rate versus reactant concentration, it is then possible to deduce the reaction rate expression, which can be used in the mass balance equations for any other batch or continuous flow reactor to determine the holding time or size required to achieve a given degree of reaction.

An analogy between a batch and a plug–flow reactor can be seen by comparing Eq. 4.14 to Eq. 4.11. Noting that the term F/A_x in Eq. 4.11 is the axial velocity through the PFR and that distance divided by velocity is time, it can be seen that Eqs. 4.14 and 4.11 are the same when distance x is translated into the time required to reach it. Thus, it can be seen that each element of fluid passing through a PFR may be thought of as an infinitesimal batch reactor.

4.3.2 Nonideal Reactors

Ideal continuous flow reactors are useful for experimental purposes and for understanding how factors like the flow rate and reactor volume influence performance. In fact, a significant portion of this book will be devoted to modeling the performance of ideal reactors as a way of gaining understanding about biochemical operations. It is important to recognize, however, that few wastewater treatment reactors are ideal. There are several reasons for this. One is size. The larger reactors are, the harder it is to achieve ideal mixing conditions. Another is the effect of other requirements. For example, the need for oxygen transfer to aerobic systems imparts considerable axial mixing in reactors that have been designed to approximate plug flow. Yet another is practicality. Site requirements and other such factors may prevent the economical construction of a reactor with a configuration that even approximates ideality.

Residence Time Distribution. Since most full-scale reactors in biochemical operations are nonideal, how might their mixing characteristics be identified and represented? One method is through measurement of the residence time distribution (RTD).

Equations 4.4 and 4.12 both contain the term V/F. This term represents the average length of time that an element of fluid (and therefore a dissolved constituent) stays in a reactor of constant volume receiving a constant flow rate of fluid with constant density. Hence, it has been given the name mean hydraulic residence time, or simply hydraulic residence time. It will be given the symbol τ for use in equations and the acronym HRT for use in the text. It represents the time required to process one reactor volume of feed:

$$\tau = V/F \tag{4.15}$$

Although the HRT is the mean residence time of fluid elements in a reactor, it is not the actual residence time of all elements. Rather, different elements of fluid reside in a reactor for different lengths of time, depending on the routes they follow, and the distribution of those times depends on the reactor's mixing characteristics.[5,12] Let the function, $F(t)$, be the fraction of the elements in the effluent stream having residence times less than t. With this definition, it is apparent that $F(0) = 0$ and $F(\infty) = 1$. In other words, none of the fluid can pass through the reactor in zero time and all must come out eventually. This function, which is shown in Figure 4.3, is known as the cumulative distribution function, or F curve. Another function is the point distribution function, $E(t)$, which is related to the cumulative distribution function by:

$$E(t) = dF(t)/dt \tag{4.16}$$

Thus, it follows that $E(t)dt$ is the fraction of the effluent that has a residence time between t and $t + dt$, and thus is the RTD function of the fluid in the reactor. The area under the RTD curve (also called the E curve) between the limits of 0 and ∞ is unity since the entire fraction of fluid must have residence times between 0 and ∞. A typical RTD (E) curve is shown in Figure 4.4.

The mixing characteristics of the two ideal continuous flow reactors represent the extremes between which all others lie. In a perfect PFR, all fluid elements stay in the reactor for exactly the mean residence time. Thus, there is no distribution of residence times; they are all the same. In contrast, in a CSTR all fluid elements have equal probability of leaving the reactor at any moment, regardless of how long they have been in it. This means that the residence time of a fluid element in a CSTR is not fixed but is subject to statistical fluctuations.[6] In particular, the RTD is a negative exponential:

$$E(t) = \frac{1}{\tau} e^{-t/\tau} \tag{4.17}$$

Integrating Eq. 4.17 from $t = 0 \rightarrow \infty$ confirms that the area under the $E(t)$ curve for a CSTR is indeed unity. Furthermore, Eq. 4.17 shows that the most probable residence time for fluid elements in a CSTR is zero. The mean residence time, τ, is also the standard deviation of the distribution of residence times.

Experimental Determination of Residence Time Distribution. The RTD for a given reactor and flow rate may be determined experimentally by introducing an inert tracer into the reactor input and observing the time response of the tracer concentration in the reactor effluent. The two most convenient types of tracer inputs are step and impulse signals.[8]

Imagine a reactor receiving influent at constant flow rate, F. At time zero a continuous flow of a soluble tracer is added to that stream, instantly changing the tracer concentration from zero to S_{TO} and maintaining it at that concentration thereafter, i.e., a step change in concentration. We immediately start measuring the output tracer concentration, S_T, and plot S_T/S_{TO} versus time. How does the resulting curve relate to the RTD? To answer that, imagine that we could divide the effluent stream into two fractions, one that has spent less time than t in the reactor, $F(t)$, and one that has spent more, $1 - F(t)$, where t is measured from the time that the tracer was introduced. Any flow that has been in the reactor less than t will have the tracer in

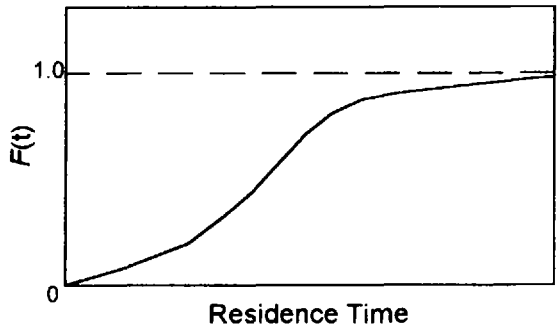


Figure 4.3 Cumulative residence time distribution function, $F(t)$.

it, and thus, the fraction $F(t)$ will have a tracer concentration $S_T = S_{TO}$. Conversely, any fluid that has been in the reactor more than t was present before the tracer was introduced and thus has no tracer in it. Consequently, the fraction $1 - F(t)$ has $S_T = 0$. Since mass flow rate is the product of liquid flow rate and concentration, the total mass flow rate of tracer in the effluent at time t, $F \cdot S_T(t)$, will be:

$$F \cdot S_T(t) = F(t) \cdot F \cdot S_{TO} + [1 - F(t)]F \cdot 0 \tag{4.18}$$

That is:

$$F(t) \cdot F \cdot S_{TO} = F \cdot S_T(t) \tag{4.19}$$

or

$$F(t) = S_T(t)/S_{TO} \tag{4.20}$$

Equation 4.20 states that the cumulative distribution function, $F(t)$, is identical to the normalized tracer concentration response due to the step input. Thus, imposition of a step input of tracer to a reactor is a convenient way of experimentally determining the cumulative distribution function. According to Eq. 4.16, it may be differentiated to obtain the RTD curve.

Differentiation of experimental data is risky, and thus it would be better to have a direct way of determining the RTD function. A similar analysis can be used to

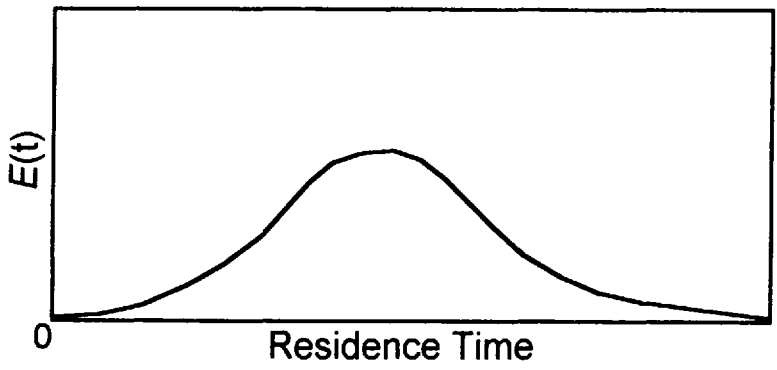


Figure 4.4 Point residence time distribution function, $E(t)$.

establish that the point distribution function, $E(t)$, is identical to the normalized effluent concentration curve resulting from an impulse input into the feed.[11] An impulse input is one in which a slug of tracer is added instantaneously to the reactor feed. Consequently, in this case, the measured output concentration is normalized by dividing each measured concentration, S_T, by the total area under the curve of S_T versus time. The area can be obtained by Simpson's rule or by graphical integration. Furthermore, the area is equivalent to the total mass of tracer added divided by the flow rate to the reactor, if both of those quantities are known. Equation 4.16 also suggests that the $F(t)$ function can be obtained from the impulse input response by integration. Hence, either a step input or an impulse input may be used to obtain both $E(t)$ and $F(t)$. However, the more commonly used technique is to obtain the $E(t)$ function from an impulse input, and then obtain the $F(t)$ function by integration if it is needed.

Use of a tracer test on a reactor will allow determination of its mixing characteristics. If the RTD function resulting from the test conforms to either of the ideal reactor types, the performance of the reactor can be predicted or simulated by application of mass balance equations for all relevant constituents with appropriate reaction rate expressions. However, if the RTD function deviates from the ideal, more involved techniques must be used.

4.4 MODELING NONIDEAL REACTORS

The use of RTDs for the prediction of reactor performance is a complex subject and the reader is encouraged to consult other sources for more complete coverage.[7,8,11] The techniques commonly used in environmental engineering are relatively straightforward, however, and thus we will look at them briefly.

4.4.1 Continuous Stirred Tank Reactors in Series Model

The simplest way to model a reactor with a nonideal flow pattern is as a series of CSTRs and this technique will be used extensively in this book. The basis for doing this may be seen by considering the response of such a system to a step input of tracer. Consider a chain of N CSTRs, each with volume V, receiving a flow F, giving each CSTR a mean hydraulic residence time τ (Figure 4.5). At time zero the feed to the first tank is switched to one with a tracer concentration S_{T0}. The response

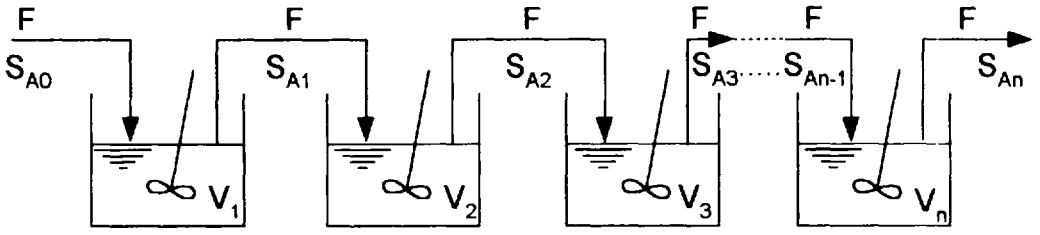


Figure 4.5 CSTRs in series.

from the first tank is the feed to the second tank, the response of the second is the feed for the third, etc. If we write and solve the mass balance equations for each (with no reaction term since the tracer is assumed to be inert), we obtain the following expression:

$$\frac{S_{TN}}{S_{TO}} = 1 - \left[1 + t/\tau + \frac{(t/\tau)^2}{2!} + \cdots + \frac{(t/\tau)^{N-1}}{(N-1)!}\right] e^{-t/\tau} \qquad (4.21)$$

Tracer concentration profiles, i.e., $F(t)$ curves, for various numbers of tanks in series are shown in Figure 4.6 while the corresponding $E(t)$ curves are shown in Figure 4.7. The curve for N = 1 is the classical response of a single CSTR. More significantly, however, the curve for N = ∞ is the classical response for a PFR, i.e., a step change in effluent concentration after one HRT. This suggests that the step response for N CSTRs in series will lie somewhere between that of a single CSTR and a PFR, with the pattern depending on the number of tanks in the chain. Furthermore, this implies that a real reactor which has the response of neither a CSTR nor a PFR can be simulated as N CSTRs in series. The easiest way of determining the appropriate value for N, which is sufficiently accurate in many cases, is to plot either the $F(t)$ or the $E(t)$ curve for the reactor in question and compare it to the curves in Figures 4.6 and 4.7, thereby selecting the value of N which corresponds most closely.

The tanks in series model has been used frequently in environmental engineering practice. For example, Murphy and Boyko[9] found that many conventional activated sludge systems had RTDs that were equivalent to three to five CSTRs in series.

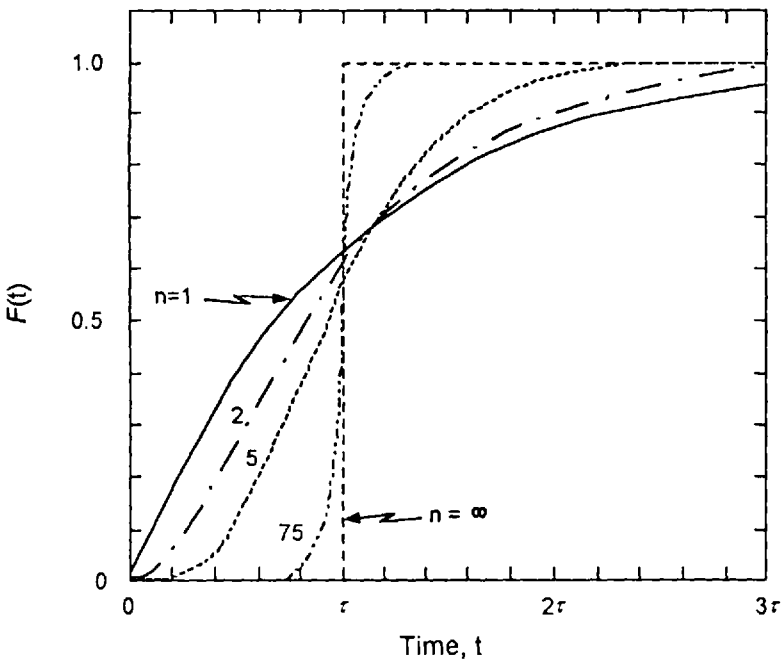


Figure 4.6 Responses of N CSTRs in series to a step tracer input. τ in the abscissa is the total HRT for N CSTRs in series.

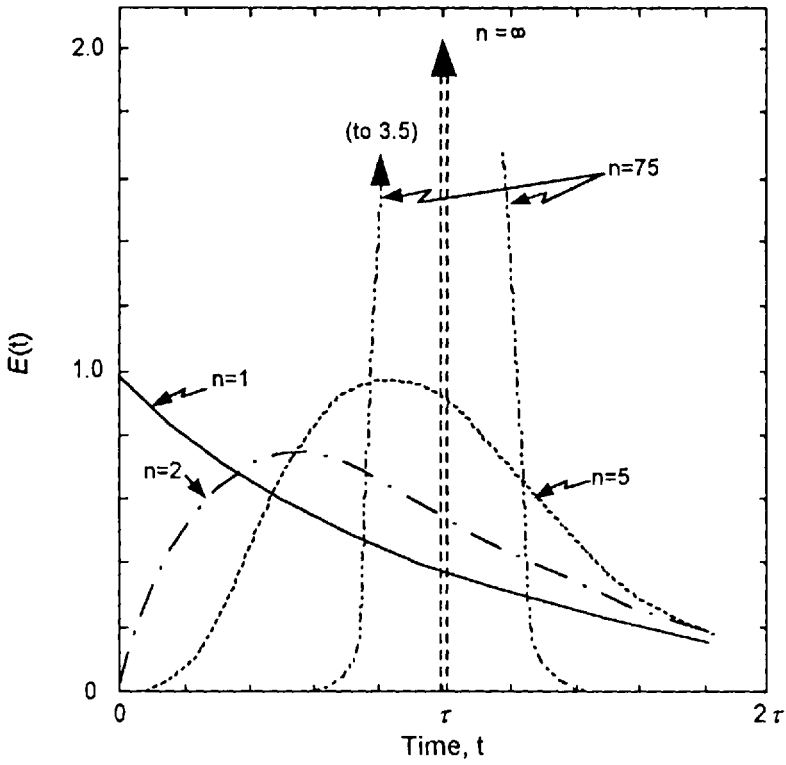


Figure 4.7 Responses of N CSTRs in series to an impulse tracer input. τ in the abscissa is the total HRT for N CSTRs in series.

4.4.2 Axial Dispersion Model

An alternative approach is to superimpose some degree of backmixing upon a plug flow of fluid. The magnitude of the backmixing is assumed to be independent of the position in the reactor and is expressed by the axial dispersion coefficient, D_L, which is analogous to the coefficient of molecular diffusion in Fick's law of diffusion. Modeling the superimposed backmixing by axial dispersion requires adding another transport term to the mass balance equation for the differential element of the PFR in Figure 4.2. In addition to the advective transport term used in Eq. 4.7, a term for transport by axial dispersion must be included, thereby increasing the number of terms in the resulting partial differential equation:

$$\frac{\partial S_A}{\partial t} = D_L \frac{\partial^2 S_A}{\partial x^2} - \frac{F}{A_c} \frac{\partial S_A}{\partial x} + r_A \tag{4.22}$$

The equation is usually rewritten to include the term D_L/vL (the dispersion number):

$$\frac{\partial S_A}{\partial \theta} = \frac{D_L}{vL} \frac{\partial^2 S_A}{\partial z^2} - \frac{\partial S_A}{\partial z} + \tau r_A \tag{4.23}$$

in which v is the longitudinal velocity through the basin (F/A_c), L is the basin length, z is dimensionless distance along the basin (x/L) and θ is dimensionless time (t/τ).

When the dispersion number is zero, there is no axial dispersion and therefore plug flow, whereas when it is infinitely large, complete backmixing exists and the reactor behaves as a CSTR.

The effect of the value of the dispersion number on the RTD may be seen by solving Eq. 4.23 with the appropriate initial and boundary conditions for a step input of an inert tracer.[8] The solution, expressed in the form of an $F(t)$ curve, is shown in Figure 4.8. To characterize the mixing pattern in a reactor by this technique, we need a way to select the appropriate value for the dispersion number. One way is to evaluate the derivative of the $F(t)$ function at one mean residence time:

$$\left.\frac{dF(t)}{dt}\right|_{t=\tau} = \left(\frac{1}{2\tau(\pi)^{0.5}}\right)\left(\frac{vL}{D_L}\right)^{0.5} \tag{4.24}$$

or

$$\frac{D_L}{vL} = \frac{1}{4\pi\tau^2\left(\left.\frac{dF(t)}{dt}\right|_{t=\tau}\right)^2} \tag{4.25}$$

Thus, the dispersion number for a nonideal reactor can be approximated by determining the slope of the $F(t)$ curve at the time equal to one HRT and using Eq. 4.25. There are also several ways of determining the dispersion number using the $E(t)$ curve and they are discussed in detail elsewhere.[8]

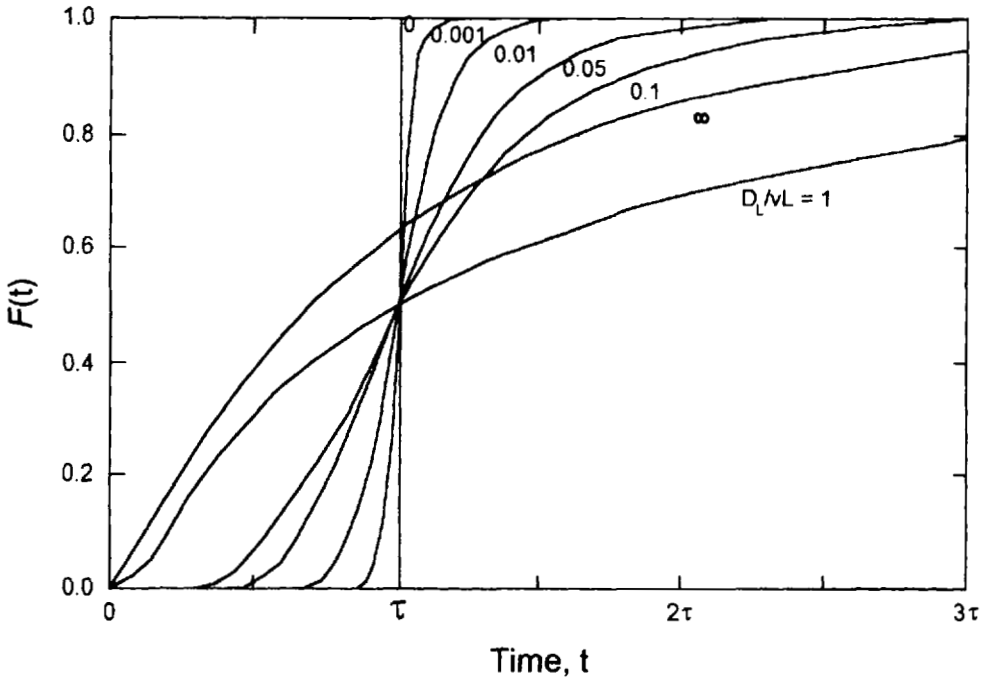


Figure 4.8 Step input tracer response for the axial dispersion model.

After the dispersion number has been determined, the performance of the reactor can be determined by using Eq. 4.23 with the appropriate reaction rate expressions. This often requires numerical techniques. Because of the complexity of such an approach, CSTRs in series have been used more frequently to model system performance in environmental engineering practice. However, the dispersion number is often used to characterize the flow patterns in reactors for other purposes.[1]

4.4.3 Representation of Complex Systems

The reactor systems used for some suspended growth biochemical operations are quite simple, whereas those used for others are complex, as seen in Table 1.2. Luckily, the concepts presented in this chapter suggest that many of the complex systems can be modeled simply as CSTRs in series, although those CSTRs may be of different size or may contain different biochemical environments (i.e., aerobic or anoxic). In Chapter 7 we will use this technique to investigate the theoretical performance of several systems.

Sometimes the flow patterns in reactor systems are so complex that the RTDs cannot be modeled by the techniques discussed above. In that case, it may be necessary to model the system as a complex network of flow regions with various modes of flow between and around them. This technique was proposed by Cholette and Cloutier[3] and is based on the premise that just as adding different numbers of tanks in series changes the RTD, so will adding different types of mixers in series and parallel. The main difference is that by adding different types of regions and flows it is possible to reproduce almost any RTD. As one might expect, however, the techniques for doing this can be quite complex. The reader is referred to other sources for information about their use.[3,8]

4.5 KEY POINTS

1. The models used to depict suspended growth biochemical operations are phenomenological mass transport models that contain a number of implicit assumptions.
2. The mass balance equation for any constituent in the control volume of a suspended growth reactor is:

   $$\text{Accumulation} = \text{Input} - \text{Output} + \text{Generation}$$

3. Reactors may be characterized by the distribution of residence times within them, with continuous stirred tank reactors (CSTRs) and plug-flow reactors (PFRs) representing the two extremes of ideality. Most reactors are nonideal, however, with residence time distributions between the two extremes.
4. The residence time distribution for a reactor may be determined directly from the normalized concentration response of the reactor to the imposition of an impulse input of tracer to the feed. It may also be obtained by differentiating the normalized concentration response to the imposition of a step input of tracer.

Table SQ4.1 Step-Input Tracer Response of a Reactor

t min	S_t/S_{t0}	t min	S_t/S_{t0}
0	0	45	0.88
5	0.006	50	0.96
10	0.017	55	0.98
15	0.084	60	0.99
20	0.220	65	0.99
25	0.380	70	0.99
30	0.550	75	0.99
35	0.700	80	0.99
40	0.800	90	1.00

5. Within environmental engineering practice, the two most common techniques for modeling the performance of a nonideal reactor is as a series of CSTRs or as a PFR with axial dispersion.

4.6 STUDY QUESTIONS

1. List and discuss the significance of the six assumptions commonly made during the modeling of suspended growth reactors.
2. What is meant by the terms "hydraulic residence time" and "residence time distribution?" Describe a continuous stirred tank reactor (CSTR) and a plug flow reactor (PFR) in terms of their hydraulic residence times and residence time distributions. What are the physical characteristics of each reactor that are responsible for the nature of its residence time distribution?
3. A reactor with a volume of 4.0 L receives a constant flow of water at a rate of 3.0 L/hr. At time zero an impulse of dye is added to the reactor which instantaneously places a dye concentration of 1200 mg/L into it. Effluent samples taken 1, 2, 3, 4, and 6 hr after addition of the dye have concentrations of 567, 268, 126, 60, and 13 mg/L, respectively. Is the reactor a perfect CSTR?
4. Explain how you would use the data from a tracer study to choose the appropriate model for characterizing the performance of a nonideal reactor.
5. Given in Table SQ4.1 are the response data from a reactor (τ = 30 min) to which a step input of tracer was applied. If you desired to model this reactor as a series of CSTRs, how many CSTRs would you use? If you desired to model the reactor with an axial dispersion model, what dispersion number would you use?

REFERENCES

1. Baillod, C. R. and W. C. Boyle, Mass transfer limitations in substrate removal. *Journal of the Sanitary Engineering Division, ASCE* **96**:525–545, 1970.

2. Characklis, W. G., Process analysis. In *Biofilms*, W. G. Characklis and K. C. Marshall, eds. John Wiley and Sons, New York, pp. 17–54, 1990.
3. Cholette, A. and L. Cloutier, Mixing efficiency determinations for continuous flow systems. *Canadian Journal of Chemical Engineering* **37**:105–112, 1959.
4. Chudoba, J., V. Ottová and V. Madêra, Control of activated sludge filamentous bulking. I. Effect of the hydraulic regime or degree of mixing in aeration tank. *Water Research* **7**:1163–1182, 1973.
5. Danckwerts, P. V., Continuous flow systems—Distribution of residence times. *Chemical Engineering Science* **2**:1–13, 1953.
6. Fredrickson, A. G., R. D. Megee, III, and H. M. Tsuchiya, Mathematical models for fermentation processes. *Advances in Applied Microbiology* **13**:419–465, 1970.
7. Himmelblau, D. M. and K. B. Bischoff, *Process Analysis and Simulation*. John Wiley and Sons, New York, 1968.
8. Levenspiel, O., *Chemical Reaction Engineering*, 2nd ed. John Wiley and Sons, New York, 1972.
9. Murphy, K. L. and B. I. Boyko, Longitudinal mixing in spiral flow aeration tanks. *Journal of the Sanitary Engineering Division, ASCE* **96**:211–221, 1970.
10. Powell, E. O., The growth rate of microorganisms as a function of substrate concentration. In *Microbial Physiology and Continuous Culture*, E. O. Powell, et al., eds. Her Majesty's Stationery Office, London, pp. 34–55, 1967.
11. Smith, J. M., *Chemical Engineering Kinetics*, 3rd ed. McGraw-Hill, New York, 1981.
12. Zwietering, T. N., The degree of mixing in continuous flow systems. *Chemical Engineering Science* **11**:1–15, 1959.

5

Aerobic Growth of Heterotrophs in A Single Continuous Stirred Tank Reactor Receiving Soluble Substrate

By returning to Chapter 1 and studying Table 1.2 it can be seen that the single continuous stirred tank reactor (CSTR) is the simplest reactor configuration used in biochemical operations, finding application in activated sludge, aerated lagoons, aerobic digestion, anaerobic digestion, and biological nutrient removal. It has also found extensive use in microbiological and environmental engineering research; much of our knowledge about microbial growth has come from it.

Because of its simplicity, the CSTR provides an ideal system with which to study the modeling of microbial growth. In this chapter we will develop models describing the growth of microbial cultures in a single CSTR, either with or without biomass recycle, and use them to gain an understanding of how such systems behave. For simplicity, the models will be confined to the growth of aerobic heterotrophic biomass in an environment that contains ample nutrients so that the soluble, biodegradable organic substrate (expressed in chemical oxygen demand [COD] units) is the growth limiting material. Furthermore, the traditional approach to modeling biomass decay will be used because it is the approach found most often in the literature, making it important that the reader be familiar with it. With minor modifications, the same models can be applied to anaerobic and anoxic heterotrophic growth, as well as to aerobic autotrophic growth, making the principles illustrated of general importance.

5.1 BASIC MODEL FOR A CONTINUOUS STIRRED TANK REACTOR

A schematic diagram of a single CSTR is shown in Figure 5.1. A bioreactor with volume V receives a flow at rate F containing only soluble, noninhibitory, biodegradable organic substrate at concentration S_{SO} (in COD units) and sufficient inorganic nutrients to make the organic substrate the growth limiting material. The influent flow and concentrations are constant, as are pH, temperature, and other environmental conditions. Within the bioreactor the heterotrophic biomass uses the substrate as its food source, thereby growing to concentration $X_{B,H}$ while reducing the substrate concentration to S_S. Biomass decay accompanies the growth so that

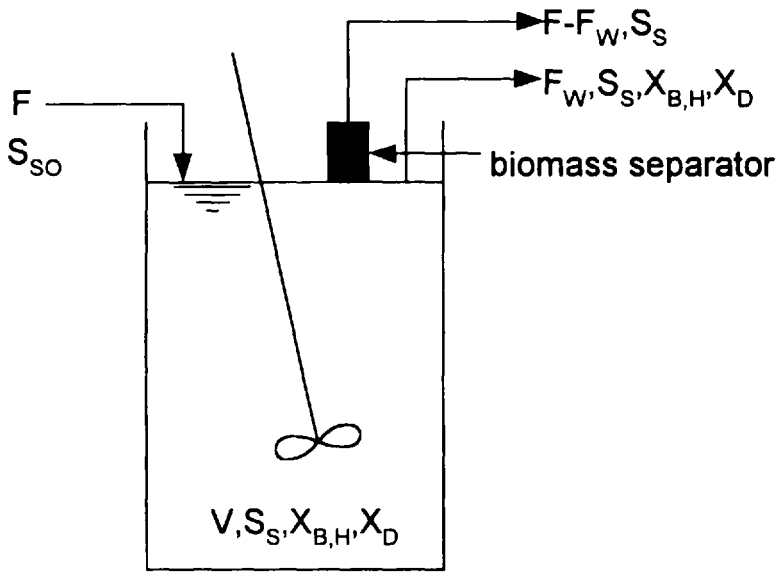


Figure 5.1 Schematic diagram of a CSTR. The biomass separator returns biomass from the effluent to the reactor. The stream F_w removes biomass from the reactor at a concentration equal to that in the reactor.

microbial debris at concentration X_D is also present. Two effluent streams are discharged from the bioreactor, but because the bioreactor is completely mixed, the concentrations of all soluble constituents in them are the same as the concentrations in the bioreactor. One stream, with flow rate F_w, flows directly from the bioreactor and carries biomass and cell debris at concentrations equal to those in the bioreactor. The other, with flow $F - F_w$, passes through a biomass separator before discharge, making it free of suspended material. All particulate material removed by that separator is returned to the bioreactor.

5.1.1 Definitions of Residence Times

A residence time defines the average amount of time a constituent stays in a system. Two types of constituents are present in the CSTR in Figure 5.1: (1) soluble, denoted by the symbol S, and (2) particulate, denoted by the symbol X. Because their residence times are not necessarily the same, one must be defined for each.

Dissolved constituents are intimately associated with the fluid and cannot be easily separated from it. Thus, their residence time in a reactor is equal to the mean hydraulic residence time, which is defined by Eq. 4.15:

$$\tau = V/F \tag{4.15}$$

Particulate constituents can be separated from a fluid by physical means, such as filtration or settling, and engineers use this characteristic to control their discharge from a bioreactor. The bioreactor shown in Figure 5.1 has two discharge streams, one containing particulate material at a concentration equal to that in the bioreactor and one free of it. The stream containing particulate material, F_w, is called the wast-

age stream because it provides the means for wasting particulate material from the bioreactor. The other stream, called the main effluent stream, is free of particulate material because it has passed through a separator such as a filter or settler. Although in reality, such streams will contain small concentrations of particulate material due to inefficiencies in the separation device, we will consider them to be totally free of such material for now. In Part III, we consider how to account for the presence of particulate material in the effluent from real systems.

The importance of the biomass separator is that it makes the residence time of particulate materials greater than the residence time of soluble materials. Thus, we must define a second residence time, called the solids retention time or mean cell residence time, which represents the average length of time a particulate constituent stays in a bioreactor. It will be given the symbol Θ_c for use in equations and the acronym SRT for use in the text. By analogy to Eq. 4.15, the SRT is defined as the mass of a particulate constituent contained in the bioreactor divided by the mass discharged from the bioreactor per unit time:

$$\Theta_c = V \cdot X / F_w \cdot X_w \tag{5.1}$$

For the case illustrated in Figure 5.1, in which the concentration of particulate material in the wastage stream, X_w, is equal to the concentration in the bioreactor, X, Eq. 5.1 may be simplified to:

$$\Theta_c = V/F_w \tag{5.2}$$

It should be remembered, however, that the basic definition of SRT involves mass flow rates rather than volumetric flow rates. Comparison of Eq. 5.2 to Eq. 4.15 reveals that:

$$\Theta_c \geq \tau \tag{5.3}$$

In other words, the closer F_w approaches F, the closer the SRT approaches the HRT, so that in the limiting case where no biomass separator is employed, the SRT and the HRT are the same.

5.1.2 Format for Model Presentation

To describe any reactor, mass balance equations must be written around a control volume for all constituents of importance. As seen in Section 4.3.1, the appropriate control volume for a CSTR is the entire reactor, since it is homogeneous throughout. If we let C_A represent the mass-based concentration of constituent A in the reactor in Figure 5.1 and consider the reactor to have constant volume, the mass balance equation is:

$$(dC_A/dt)V = F \cdot C_{A0} - F_w \cdot C_A - (F - F_w)C'_A + r_A \cdot V \tag{5.4}$$

where C_{A0} is the concentration of A in the influent stream and C'_A is its concentration in the stream passing through the biomass separator. For soluble constituents, C'_A is the same as the concentration in the CSTR, whereas for particulate constituents, C'_A is zero. The reaction term r_A represents the sum of all reactions in which constituent A participates, as seen in Eq. 3.12. Furthermore, as seen in Chapter 3, r_A may be a function of the concentrations of several constituents. If so, it will be

necessary to solve several mass balance equations simultaneously to determine the concentration of any single constituent in a bioreactor.

For the situation depicted in Figure 5.1, mass balance equations must be written for at least three constituents: S_S, $X_{B,H}$, and X_D. In addition, we will be interested in the amount of oxygen that must be supplied through an oxygen transfer system. We will also need a mass balance equation for it, making a total of four.

Consideration of the number of mass balance equations required for this simple situation, and reflection on the number of events and constituents that could be considered, as discussed in Chapters 2 and 3, makes it clear that a system is needed for providing the required information. The matrix format[4, 6] discussed in Section 3.1.3 provides such a system.

Table 5.1 presents all of the information required to compile the reaction rate terms, r_j, for insertion into the mass balance equations for the situation described above. It contains the information in Eq. 3.11, with the entries in the body of the table representing the stoichiometric coefficients in COD units for each constituent participating in each reaction (recall that oxygen is negative COD). The entries in the right column represent the process rates for the reactions, and the subscript H on the coefficients signifies that they are applicable to heterotrophic biomass. Only two reactions are considered in this case, growth and decay. The stoichiometric equation for growth with active biomass as the reference constituent was given in COD units by Eq. 3.33, and examination of it reveals that the coefficients in Table 5.1 correspond to the coefficients in it, and their sum should equal zero, as it does. The term, process rate, r_j, in Table 5.1 refers to the generalized reaction rate for process j as defined in Eq. 3.10. For the growth process it can be obtained by substituting Eq. 3.35 into Eq. 3.34. As we saw in Section 3.2.2, the specific growth rate coefficient, μ_H, is a function of the substrate concentration, S_S, but to simplify upcoming explanations, that substitution will be made later. The stoichiometric equation for the traditional approach of modeling decay is given by Eq. 3.53 and inspection of it reveals its correspondence to the coefficients in Table 5.1. Again, summing the coefficients reveals that continuity has been maintained. The process rate expression for decay may be obtained by multiplying Eq. 3.56 by the stoichiometric coefficient for biomass in Table 5.1, giving the result shown in the table.

As stated by Eq. 3.12, the overall rate expression for each constituent to be inserted into its mass balance equation is obtained by summing the products of the process rate expressions times the stoichiometric coefficients appearing in the column

Table 5.1 Process Kinetics and Stoichiometry for Aerobic Growth of Heterotrophic Bacteria. Traditional Model for Biomass Decay

	Component[a]				Process rate,
Process	$X_{B,H}$	X_D	S_S	S_O[b]	r_j
Growth	1		$-(1/Y_H)$	$\left(\frac{1 - Y_H}{Y_H}\right)$	$\mu_H \cdot X_{B,H}$
Decay	-1	f_D		$(1 - f_D)$	$b_H \cdot X_{B,H}$

[a] All components and coefficients are expressed as COD.
[b] Coefficients must be multiplied by -1 to express them as oxygen.

under the constituent. For more complex situations, it is more convenient to express this in matrix format as illustrated by Mason et al.,[7] but for this simple case, the overall rate expression for each constituent can be obtained by examination of Table 5.1. Both reactions influence active biomass, $X_{B,H}$. Consequently, the overall reaction rate for it is obtained by multiplying the generalized reaction rate for each process by the corresponding stoichiometric coefficient in the column under $X_{B,H}$, and summing them:

$$r_{XB} = \mu_H \cdot X_{B,H} - b_H \cdot X_{B,H} \tag{5.5}$$

Only one of the processes influences biomass debris, and that is decay, which generates it. Consequently:

$$r_{XD} = b_H \cdot f_D \cdot X_{B,H} \tag{5.6}$$

Likewise, soluble organic substrate is only influenced by one process (growth) so that:

$$r_{SS} = -(\mu_H/Y_H)X_{B,H} \tag{5.7}$$

Finally, oxygen is influenced by both processes. Thus, its reaction rate includes two terms, which in COD units gives:

$$r_{SO} = \left(\frac{1 - Y_H}{Y_H}\right) \mu_H \cdot X_{B,H} + (1 - f_D) b_H \cdot X_{B,H} \qquad \text{(as COD)} \tag{5.8}$$

Note that as obtained from application of Eq. 3.12, the overall rate expression for oxygen carries a positive sign. This does not mean that oxygen is generated. Remember that oxygen is expressed in units of oxygen demand. Thus, oxygen demand is generated, which means that oxygen is used. To convert the rate expression to more familiar units of oxygen, it must be multiplied by -1, giving:

$$r_{SO} = -\left[\left(\frac{1 - Y_H}{Y_H}\right) \mu_H \cdot X_{B,H} + (1 - f_D) b_H \cdot X_{B,H}\right] \qquad \text{(as } O_2\text{)} \tag{5.9}$$

This makes it clearer that oxygen is being utilized in the system.

5.1.3 Concentrations of Soluble Substrate and Biomass

At steady-state the derivative in Eq. 5.4 is zero, allowing the generalized mass balance equation to be simplified:

$$F \cdot C_{AO} - F_w \cdot C_A - (F - F_w)C'_A + r_A \cdot V = 0 \tag{5.10}$$

Substituting the appropriate terms for biomass and noting that the influent and main effluent streams contain none (i.e., C_{AO} and C'_A are both zero) gives:

$$-F_w \cdot X_{B,H} + (\mu_H \cdot X_{B,H} - b_H \cdot X_{B,H})V = 0 \tag{5.11}$$

Substitution and simplification yield:

$$\mu_H = \frac{1}{\Theta_c} + b_H \tag{5.12}$$

Several things are important about this equation. Because $X_{B,H}$ has canceled out of it, the mass balance on active biomass does not lead to an equation for calculating its concentration. Rather, it leads to an equation that shows that at steady-state the specific growth rate of the active biomass, μ_H, is determined by its rate of loss from the bioreactor, as reflected by the SRT and the loss to decay. Because the engineer can control the SRT through manipulation of the wastage rate, F_w, he/she has control over how rapidly the biomass grows. Furthermore, we saw in Section 3.2.7 that μ_H is related to the substrate concentration. This suggests that control of the bioreactor SRT also allows the engineer to control the concentration of substrate being discharged from it. Finally, because the HRT does not appear in Eq. 5.12, the specific growth rate of the biomass is independent of it. The only time the HRT has any impact on the steady-state specific growth rate of biomass is when there is no biomass separator and all effluent leaves via the wastage stream, making the SRT equal to the HRT.

In order to calculate the substrate concentration in the CSTR and its effluent, the functional relationship between μ_H and S_S must be known. Since this is a non-inhibitory substrate and it is the sole rate limiting constituent, the Monod equation, as given by Eq. 3.36 would be appropriate:

$$\mu_H = \hat{\mu}_H \frac{S_S}{K_S + S_S} \tag{3.36}$$

Substitution of it for μ_H in Eq. 5.12 and rearrangement gives:

$$S_S = \frac{K_S(1/\Theta_c + b_H)}{\hat{\mu}_H - (1/\Theta_c + b_H)} \tag{5.13}$$

Because μ_H is a function of only the SRT, so is the substrate concentration. Furthermore, Eq. 5.13 states that the substrate concentration in a CSTR is independent of the concentration in the influent, since S_{S0} does not appear in the equation. The reason for this will become clear shortly. This has been shown to be the case for a pure culture growing on a single substrate when S_S represents the concentration of that substrate in COD units.[5] It should be recognized, however, that even though this simple model did not include the reaction, the biomass will form soluble microbial products as it metabolizes the substrate, as discussed in Section 2.4.3. Since the amount of product formed will depend on the amount of substrate used, the total amount of soluble organic matter in the bioreactor and in the effluent, as measured by the COD test, will be roughly proportional to the influent COD.[1,9] This effect should be recognized and remembered, even though it is not explicitly stated in models and design equations.

Examination of Eq. 5.12 reveals that as the SRT becomes very large (so that $1/\Theta_c \rightarrow 0$), the specific growth rate of the active biomass approaches the specific rate of decay. This means that in a single CSTR, substrate must always be present to drive the required growth reaction. Consequently, there is a minimum substrate concentration that can be achieved and it is given by the limit of Eq. 5.13 as Θ_c approaches infinity:

$$S_{S,min} = \frac{K_S \cdot b_H}{\hat{\mu}_H - b_H} \tag{5.14}$$

Equation 5.14 shows that the minimum attainable substrate concentration depends on the kinetic parameters describing the biodegradation, and therefore on both the nature of the substrate undergoing biodegradation and the type of culture degrading it. If a CSTR is being considered for treatment of a wastewater to a desired concentration, that concentration should be compared to S_{Smin}. If it is less than S_{Smin}, some other bioreactor configuration must be used because a single CSTR will be insufficient.

The maximum rate at which biomass can grow on a given substrate is when the surrounding substrate concentration is equal to the concentration in the influent to the bioreactor:

$$\mu_{Hmax} = \hat{\mu}_H \frac{S_{SO}}{K_S + S_{SO}} \tag{5.15}$$

Consequently, the minimum SRT (Θ_{cmin}), at which biomass can grow on a given influent can be calculated by setting μ_H in Eq. 5.12 equal to μ_{Hmax} in Eq. 5.15 and rearranging:

$$\Theta_{cmin} = \frac{K_S + S_{SO}}{S_{SO}(\hat{\mu}_H - b_H) - K_S \cdot b_H} \tag{5.16}$$

The minimum SRT is also called the point of washout because at shorter SRTs all organisms will be washed out of the bioreactor and no growth will occur. At washout no biomass is produced and no substrate is utilized because the substrate concentration in the bioreactor (and its effluent) equals the concentration in the influent, i.e., the process has failed.

In theory, there is no minimum HRT for a CSTR because as long as biomass can be separated from the effluent and returned to the bioreactor to keep the SRT greater than Θ_{cmin}, growth can be maintained. In practice, however, it is dangerous to make the HRT less than Θ_{cmin}. If the HRT were less than Θ_{cmin} and something happened to the biomass separator so that all effluent contained biomass at a concentration equal to that in the bioreactor, the SRT would then equal the HRT and the process would fail. Furthermore, the bioreactor would be impossible to restart once failure occurred, unless biomass were added from an external source. To be safe, the HRT should be kept larger than Θ_{cmin}.

Example 5.1.3.1

A culture of microorganisms is being grown in a CSTR with a volume of 8 liters. The medium contains m-cresol as the sole carbon source and all inorganic nutrients are provided in excess. When the concentration of m-cresol in the feed is 200 mg/L as COD, the kinetic parameters have the values shown in Table E5.1.

a. What is the maximum permissible flow rate through the bioreactor?

When the flow is maximum, the HRT will be minimum, but τ_{min} should not be less than Θ_{cmin}. Thus, calculate Θ_{cmin} with Eq. 5.16:

$$\Theta_{cmin} = \frac{3.5 + 200}{200(0.20 - 0.01) - (3.5)(0.01)}$$

$$\Theta_{cmin} = 5.36 \text{ hrs.}$$

Table E5.1 Kinetic Parameters and Stoichiometric Coefficients for a Culture Growing on m-Cresol at S_{SO} = 200 mg as COD

Symbol	Units	Value
$\hat{\mu}_H$	hr^{-1}	0.20
K_S	mg/L as COD of m-cresol	3.5
Y_H	mg biomass COD/mg m-cresol COD	0.34
b_H	hr^{-1}	0.01
f_D	mg debris COD/mg biomass COD	0.20

Therefore,

$$\tau_{min} = 5.36 \text{ hrs}$$

and the maximum permissible flow is

$$F = 8.0 \text{ L}/5.36 \text{ hrs} = 1.49 \text{ L/hr}$$

b. The flow through the bioreactor is 1.0 L/hr and the wastage flow is 0.05 L/hr. What is the concentration of m-cresol (in COD units) in the effluent?

First, we must calculate the SRT using Eq. 5.2:

$$\Theta_c = 8.0 \text{ L}/(0.05 \text{ L/hr}) = 160 \text{ hrs}$$

Then we must use Eq. 5.13 to find the m-cresol concentration:

$$S_S = \frac{3.5(1/160 + 0.01)}{0.20 - (1/160 + 0.01)}$$

$$S_S = 0.31 \text{ mg/L as COD}$$

c. What is the minimum m-cresol concentration (in COD units) that can be obtained from a CSTR?

This may be determined by using Eq. 5.14:

$$S_{Smin} = \frac{(3.5)(0.01)}{0.20 - 0.01}$$

$$S_{Smin} = 0.18 \text{ mg/L as COD}$$

If a lower concentration were desired, some other bioreactor configuration would have to be used.

The suspended solids in a bioreactor receiving only soluble substrate will contain two components, active biomass, $X_{B,H}$, and biomass debris, X_D. Thus, we must be able to calculate the concentration of each.

The only source of active biomass in the system is from growth due to substrate utilization and its concentration may be calculated from a mass balance on substrate. Substituting the appropriate terms into Eq. 5.10 gives:

$$F \cdot S_{SO} - F_w \cdot S_S - (F - F_w)S_S - (\mu_H/Y_H)X_{B,H} \cdot V = 0 \qquad (5.17)$$

Rearrangement gives:

$$X_{B,H} = \frac{Y_H(S_{SO} - S_S)}{\mu_H \cdot \tau} \tag{5.18}$$

Substitution of Eq. 5.12 for μ_H gives:

$$X_{B,H} = \frac{Y_H(S_{SO} - S_S)}{(1/\Theta_c + b_H)\tau} = \left(\frac{\Theta_c}{\tau}\right) \frac{Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c} \tag{5.19}$$

This states that the active biomass concentration depends on both the SRT and the HRT. Furthermore, rearrangement of Eq. 5.19 shows that for a fixed SRT (which determines S_S), the product $X_{B,H} \cdot \tau$ is constant:

$$X_{B,H} \cdot \tau = \frac{\Theta_c \cdot Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c} \tag{5.20}$$

In other words, at a fixed SRT and a fixed flow rate, a fixed mass of substrate will be removed per unit time, generating a fixed mass of microorganisms. A bioreactor with a small volume (short HRT) will contain a higher concentration of active biomass than one with a large volume (large HRT), although they will both contain the same mass. Likewise, if the bioreactor volume is fixed and the flow is increased at constant SRT, the mass of microorganisms in the bioreactor must increase to keep it consistent with the mass of substrate being removed. As a consequence, the biomass concentration must increase proportionally. Equation 5.20 demonstrates why the steady-state performance of a CSTR equipped with a biomass separator is independent of the HRT. If the HRT is changed for any reason, the concentration of biomass will also change to maintain the mass of organisms sufficient to produce an effluent substrate concentration consistent with the SRT. Likewise, if the influent substrate concentration, S_{SO}, is changed, the concentration of active biomass will change until the effluent substrate concentration is consistent with the SRT, thereby making S_S independent of S_{SO}.

The situation concerning the active biomass concentration is different when the bioreactor has no biomass separator so that all effluent contains biomass and the SRT is equal to the HRT. In that case, the biomass concentration depends solely on the HRT, as can be seen by substituting τ for Θ_c in Eq. 5.19:

$$X_{B,H} = \frac{Y_H(S_{SO} - S_S)}{1 + b_H \cdot \tau} \tag{5.21}$$

The major disadvantage of a CSTR without a biomass separator is that its performance depends on its HRT, and this follows directly from its inability to maintain a constant mass of microorganisms as the HRT is changed. However, S_S is still independent of S_{SO} because a change in S_{SO} will cause the active biomass concentration to change until S_S is consistent with the HRT (SRT).

The units used to express the biomass concentration will depend on the units used to define the true growth yield, Y_H. In this part of the book, we are using the quantity of biomass COD formed per unit of substrate COD utilized as the units for yield. Consequently, the biomass concentration will also be in COD units. It can be converted to a mass of solids basis by dividing by 1.20 g COD/g SS and to a mass

of volatile solids basis by dividing by 1.42 g COD/g VSS as discussed in Section 2.4.1 and given in Table 3.1.

The concentration of biomass debris in the bioreactor can be obtained from a mass balance on the debris, recognizing that its concentration in the influent and the effluent from the biomass separator is zero:

$$-F_w \cdot X_D + f_D \cdot b_H \cdot X_{B,H} \cdot V = 0 \tag{5.22}$$

Rearrangement gives:

$$X_D = f_D \cdot b_H \cdot \Theta_c \cdot X_{B,H} = \left(\frac{\Theta_c}{\tau}\right)\left[\frac{f_D \cdot b_H \cdot \Theta_c \cdot Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{5.23}$$

The units of X_D will be the same as the units of $X_{B,H}$. Debris adds to the total suspended solids concentration in the bioreactor, but does not add to the degradative capability because it has no biological activity associated with it. If the bioreactor has no biomass separator, the SRT will equal the HRT and Eq. 5.23 should be modified appropriately in calculating X_D.

The total biomass concentration in a bioreactor, X_T, is the sum of the active biomass and biomass debris concentrations. Adding Eqs. 5.19 and 5.23 gives:

$$X_T = \left(\frac{\Theta_c}{\tau}\right)\frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c} \tag{5.24}$$

Examination of it reveals that like the $X_{B,H} \cdot \tau$ product, the $X_T \cdot \tau$ product is fixed if the SRT is fixed. As far as theory is concerned, once the SRT has been chosen to give a desired effluent substrate concentration, any combination of bioreactor size and total biomass concentration may be used as long as it gives the proper $X_T \cdot \tau$ product. There are practical limits, of course, and these are discussed in Chapter 10.

As discussed above for the active biomass concentration, when a bioreactor has no biomass separator, thereby making the SRT equal to the HRT, the total biomass concentration depends solely on the HRT as can be seen by substituting τ for Θ_c in Eq. 5.24:

$$X_T = \frac{(1 + f_D \cdot b_H \cdot \tau)Y_H(S_{SO} - S_S)}{1 + b_H \cdot \tau} \tag{5.25}$$

The active fraction of the biomass, f_A, is defined as the concentration of active biomass divided by the total biomass concentration. Division of Eq. 5.19 by Eq. 5.24 and rearrangement yields:

$$f_A = \frac{1}{(1 + f_D \cdot b_H \cdot \Theta_c)} \tag{5.26}$$

Examination of it shows that the active fraction declines as the SRT is increased due to the build up of biomass debris in the bioreactor.

As discussed in Sections 2.4.2 and 3.8, the observed yield in a biochemical operation is always less than the true growth yield because some of the energy in the substrate must go to meet the maintenance energy needs of the culture. The observed yield associated with a bioreactor is equal to the actual net mass of biomass formed per unit mass of substrate destroyed, taking into consideration the amount of biomass lost to decay. When engineers attempt to measure the concentration of

biomass formed it is difficult to distinguish the active biomass from the biomass debris. The total biomass concentration is generally used for purposes of defining the observed yield. At steady-state the total biomass formed in the bioreactor must equal the mass wasted from it. Thus, the observed yield, Y_{Hobs}, is given by:

$$Y_{Hobs} = \frac{F_w \cdot X_T}{F(S_{SO} - S_S)} \tag{5.27}$$

Substitution of Eq. 5.24 for X_T and simplification gives:

$$Y_{Hobs} = \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H}{1 + b_H \cdot \Theta_c} \tag{5.28}$$

It can be seen that the larger the SRT of the bioreactor, the smaller the observed yield will be. This is because longer SRTs provide greater opportunity for biomass decay and greater need for maintenance energy, leaving less energy for synthesis of new biomass.

Example 5.1.3.2

Continue with the problem begun in Example 5.1.3.1.

a. What is the active biomass concentration in the bioreactor when the influent flow is 1.0 L/hr and the wastage flow is 0.05 L/hr?

From Example 5.1.3.1, for this condition the SRT is 160 hr and S_S is 0.31 mg/L as COD. Furthermore, the HRT is 8 hr. Thus, using Eq. 5.19:

$$X_{BH} = \frac{0.34(200 - 0.31)}{(1/160 + 0.01)8}$$

X_{BH} = 523 mg/L as COD = 436 mg/L as suspended solids

b. What is the total biomass concentration under the same conditions?

Use Eq. 5.24 to calculate this:

$$X_T = \left(\frac{160}{8}\right)\left\{\frac{[1 + (0.20)(0.01)(160)](0.34)(200 - 0.31)}{1 + (0.01)(160)}\right\}$$

X_T = 690 mg/L as COD = 575 mg/L as suspended solids

c. What is the active fraction of the biomass?

This may be calculated from its definition or from Eq. 5.26. Using the definition gives:

f_A = 523/690 = 0.76

Using Eq. 5.26 gives:

f_A = 1/[1 + (0.2)(0.01)(160)] = 0.76

d. What is the observed yield?

This may be calculated from Eq. 5.28:

$$Y_{Hobs} = \frac{[1 + (0.20)(0.01)(160)]0.34}{1 + (0.01)(160)}$$

Y_{Hobs} = 0.17 mg biomass COD produced/mg substrate COD destroyed

This is only 50% of the true growth yield, showing the impact of decay and maintenance energy requirements on the net production of biomass and debris.

5.1.4 Excess Biomass Production Rate, Oxygen Requirement, and Nutrient Requirements

The two major costs associated with the treatment of wastewaters in aerobic CSTRs are from the disposal of the excess biomass produced and the provision of ample oxygen. It is important to be able to determine the amount of excess biomass produced and the quantity of oxygen that must be supplied. In addition, because of the negative impacts of nutrient limitations it is important to be able to determine the nutrient requirements as well.

Excess biomass is removed from the bioreactor via the wastage stream and the mass that must be disposed of per unit time is simply the concentration in that stream times its flow rate. At steady-state, this must equal the net production rate. Letting W_T represent the total biomass wastage rate gives:

$$W_T = F_w \cdot X_T \tag{5.29}$$

Combining the equation relating F_w to the SRT (Eq. 5.2) with Eq. 5.24 for X_T gives:

$$W_T = F\left[\frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{5.30}$$

Since S_S depends only on the SRT, it can be seen that W_T depends on the SRT, the flow rate of the wastewater, and the concentration of substrate in it. Furthermore, it can be seen that the excess biomass wastage rate will decrease as the SRT is increased. This is due to the increased importance of decay at long SRTs. One use of biochemical operations is the stabilization of insoluble organic matter (see Section 1.2.1). The decrease in the amount of excess biomass brought about by decay is one example of stabilization. As the SRT is increased, more and more of the active biomass is oxidized and converted to debris, meaning that less excess biomass must be disposed of.

Comparison of Eq. 5.30 with Eq. 5.28, and substitution of the latter into the first reveals that the excess biomass wastage rate is just the observed yield times the substrate removed, which is consistent with the fact that the mass of biomass wasted must equal the mass produced at steady-state:

$$W_T = F \cdot Y_{Hobs}(S_{SO} - S_S) \tag{5.31}$$

It can be seen that knowledge of the observed yield makes it easy to estimate the amount of excess biomass that must be disposed of.

The units on W_T as given by Eqs. 5.29, 5.30, or 5.31 are mass of total biomass COD per time. Often, however, it is desirable to know the actual mass of dry solids that must be disposed of. This may be obtained by dividing by 1.20 g COD/g SS, just as was done to convert the biomass COD concentration to mass units.

The rate at which the microorganisms utilize oxygen in the bioreactor is equal to the overall rate expression for oxygen as developed from Table 5.1 and expressed in Eq. 5.9. Consequently, oxygen must be supplied at the same rate. If we multiply

that rate by the bioreactor volume to give the mass per unit time required (RO), the result is:

$$RO = \left[\left(\frac{1 - Y_H}{Y_H}\right)\mu_H + (1 - f_D)b_H\right] X_{B,H} \cdot V \tag{5.32}$$

Substitution of Eq. 5.12 for μ_H, Eq. 5.19 for $X_{B,H}$ and Eq. 4.15 for the HRT, and simplification gives:

$$RO = F(S_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H}{1 + b_H \cdot \Theta_c}\right] \tag{5.33}$$

Because the stoichiometric coefficients in Table 5.1 are from a COD balance for each reaction, Eq. 5.33 represents a COD balance across the bioreactor. This can be seen in the following way. A COD balance states that the amount of oxygen which must be supplied to a bioreactor must equal the total COD in minus the total COD out, including the COD of the biomass and the biomass debris:

$$RO = F \cdot S_{SO} - F \cdot S_S - F_w \cdot X_I \tag{5.34}$$

Examination of Eq. 5.34 reveals that the last term is just W_I, the excess biomass wastage rate. Substitution of Eq. 5.30 for W_I yields an equation identical to Eq. 5.33 after rearrangement. Furthermore, substitution of Eq. 5.31 for W_I gives an equation for the oxygen requirement in terms of only the observed yield and the amount of substrate removed:

$$RO = F(S_{SO} - S_S)(1 - Y_{Hobs}) \tag{5.35}$$

Substitution of Eq. 5.28 for Y_{Hobs} also gives an expression identical to Eq. 5.33. Thus, if the observed yield is known, the oxygen requirement is also known.

All of the above follows directly from the COD-based stoichiometry discussed in Sections 3.1.1 and 3.8.1. Equation 3.90 stated that the COD removed equals the oxygen equivalents of terminal electron acceptor used plus the COD of the biomass formed, which is the same as Eq. 5.34. Thus, all of the above equations have their roots in the basic stoichiometry discussed earlier. The ability to calculate the steady-state oxygen requirement directly from such simple equations is the key advantage to expressing yields and biomass concentrations in COD units.

Equation 5.33 clearly shows that the oxygen requirement in a CSTR increases as the SRT is increased. This too, is indicative of the increased stabilization that occurs as the SRT is increased. Increased stabilization implies that more of the electrons in a material end up being transferred to the terminal electron acceptor. Thus, as less excess biomass is produced, more oxygen must be used.

The amount of nutrient required can also be determined directly from the stoichiometry of biomass growth as discussed in Section 3.8.2. It was seen that the amount of nitrogen required to form biomass that can be represented by the empirical formula $C_5H_7O_2N$ is 0.087 mg N/mg biomass COD. If we assume that the nitrogen content of biomass debris is the same as that of active biomass, then the amount of nitrogen required per unit of substrate COD removed, NR, is just 0.087 times the observed yield, or:

$$NR = \frac{0.087(1 + f_D \cdot b_H \cdot \Theta_c)Y_H}{1 + b_H \cdot \Theta_c} \tag{5.36}$$

Furthermore, as seen earlier, the phosphorus requirement is about one-fifth of the nitrogen requirement on a mass basis and thus it may be calculated by replacing 0.087 in Eq. 5.36 with 0.017. The requirements for micronutrients may be determined in a similar manner by using appropriate factors from Table 3.3. Nutrients should be added in slight excess of the theoretical amounts to ensure that the organic substrate is rate limiting, as discussed in Section 3.2.9.

Example 5.1.4.1

Continue with the problem begun in Example 5.1.3.1.

a. How many mg/hr of dry solids would have to be disposed of when the flow through the bioreactor is 1.0 L/hr and the wastage rate is 0.05 L/hr?

Using Eq. 5.30

$$W_1 = 1.0 \left\{ \frac{[1 + (0.20)(0.01)(160)](0.34)(200 - 0.31)}{1 + (0.01)(160)} \right\}$$

$W_1 = 34.5$ mg/hr as COD

To convert this to a dry solids basis, divide by 1.20 g COD/g dry solids:

$W_1 = 28.7$ mg/hr as dry solids

b. How many mg/hr of oxygen must be supplied to the bioreactor?

This can be obtained from use of Eq. 5.34:

$RO = 1.0(200 - 0.31) - 34.5 = 165.2$ mg/hr

It can also be obtained from the fundamental system parameters by using Eq. 5.33:

$$RO = (1.0)(200 - 0.31) \left\{ 1 - \frac{[1 + (0.20)(0.01)(160)]0.34}{1 + (0.01)(160)} \right\}$$

$RO = 165.2$ mg/hr

Finally, it can be obtained directly from the observed yield through use of Eq. 5.35:

$RO = 1.0(200 - 0.31)(1.0 - 0.17)$

$RO = 165.2$ mg/hr

Since the oxygen demand associated with the influent substrate is 200 mg/hr, these calculations show that it is necessary to supply sufficient oxygen to meet 83% of that demand. The remainder is associated with the excess biomass formed and wasted from the system.

c. How many mg/L of nitrogen and phosphorus should the influent contain?

The nitrogen requirement can be calculated with Eq. 5.36

$$NR = \frac{(0.087)[1 + (0.20)(0.01)(160)](0.34)}{1 + (0.01)(160)}$$

$NR = 0.015$ mg N/mg substrate COD removed

Since the substrate COD removed was $200 - 0.31 = 199.69$ mg/L, the biomass will require 2.95 mg/L of nitrogen. If we allow an extra 0.5 mg/L to prevent

nitrogen from being rate limiting, the influent should contain approximately 3.5 mg/L as N.

The phosphorus requirement will be about one-fifth of the nitrogen requirement, and the biomass will use about 0.59 mg/L. If we allow an extra 0.25 mg/L to prevent phosphorus from being rate limiting, the influent should contain approximately 0.85 mg/L as P.

5.1.5 Process Loading Factor

Before the widespread use of SRT as the basic independent variable for design and control of a CSTR, most designers used the process loading factor, also called the food to microorganism (or F/M) ratio. The process loading factor, U, is defined as the mass of substrate applied per unit time divided by the mass of microorganisms contained in the bioreactor.[10] Because it is difficult to distinguish active biomass from biomass debris, the mass of microorganisms has generally been defined in terms of the total biomass concentration, X_T:

$$U \equiv \frac{F \cdot S_{SO}}{V \cdot X_T} \tag{5.37}$$

The SRT is an important design parameter because at steady-state it is related in a simple way to μ_H, the specific growth rate coefficient of the biomass, which in turn controls the substrate concentration in the bioreactor and its effluent. The relationship of the process loading factor to μ_H can be obtained by rearranging the mass balance on substrate (Eq. 5.17) and substituting into it the fact that the active biomass concentration is equal to the total biomass concentration times the active fraction, giving:

$$\frac{f_A \cdot \mu_H}{Y_H} = \frac{F(S_{SO} - S_S)}{V \cdot X_T} \tag{5.38}$$

For the conditions generally found in bioreactors used in wastewater treatment, $S_S << S_{SO}$. Therefore:

$$U \approx \frac{f_A \cdot \mu_H}{Y_H} = f_A \cdot q_H \tag{5.39}$$

This shows that the active fraction must be known before the specific growth rate or the specific substrate removal rate of the biomass can be determined exactly from the process loading factor. Conversely, Eq. 5.12 showed that at steady-state the specific growth rate can be determined directly from the SRT without requiring such knowledge. Since the active fraction depends on the characteristics of the bioreactor and requires knowledge of the SRT (Eq. 5.26), it is simpler to work directly with SRT as the fundamental design and operational parameter for a CSTR at steady-state. Furthermore, when a wastewater contains particulate substrates, the active fraction cannot be calculated easily from an expression like Eq. 5.26 and thus its determination becomes a major problem. Consequently, the relationship of the process loading factor to the specific growth rate of the biomass is more difficult to determine than it is for the simplified situation under consideration here. This has led to the use of the SRT instead of the process loading factor by many designers. Nevertheless, because the process loading factor is related to the specific growth and substrate

removal rates, there are situations in which it provides valuable information, particularly in tanks-in-series systems where the applied substrate varies from tank to tank but the active fraction does not. In Chapters 10 and 11 we consider such situations.

5.1.6 First-Order Approximation

As discussed in Section 3.2.2, occasions arise in which the steady-state substrate concentration in a CSTR is much less than the half-saturation coefficient so that the Monod equation may be simplified into a first-order equation:

$$\mu_H \approx \left(\frac{\hat{\mu}_H}{K_S}\right) S_S \tag{3.38}$$

This situation often arises when the SRT is long, thereby making it difficult to evaluate $\hat{\mu}_H$ and K_S independently from steady-state data. In that case, the mean reaction rate coefficient, k_c, is often used. Recalling Eqs. 3.44 and 3.45 and substituting them into Eq. 3.38 shows that:

$$\mu_H = Y_H \cdot k_c \cdot S_S \tag{5.40}$$

The use of Eq. 5.40 does not alter any of the mass balances nor does it alter the fact that μ_H is controlled by the SRT as expressed in Eq. 5.12. All it does is simplify the relationship between S_S and the SRT. Substitution of Eq. 5.40 into Eq. 5.12 and rearrangement gives:

$$S_S = \frac{1/\Theta_c + b_H}{Y_H \cdot k_c} \tag{5.41}$$

Comparison of Eq. 5.41 with Eq. 5.13 shows the effect that the first-order approximation has. None of the other equations are affected, except through the effect on S_S. It should be emphasized that care should be exercised in the application of the first-order approximation of the Monod equation to ensure that the basic assumption (i.e., $S_S \ll K_S$) is valid.

5.2 EXTENSIONS OF BASIC MODEL

The simple model developed in Section 5.1 was for a system receiving only soluble substrate. However, most wastewaters contain soluble organic matter that is nonbiodegradable. Furthermore, all domestic and many industrial wastewaters contain suspended matter that escapes removal by sedimentation prior to entrance of the wastewater into the biochemical operation, and the impacts of those solids must be accounted for in any models depicting fully the operation of CSTRs.

Suspended material may be classified in many ways, and one of the most commonly used methods is to split it into organic and inorganic, which have traditionally been measured as volatile and nonvolatile suspended solids, respectively. Unfortunately, this division is not the best for describing biochemical operations, in part because around 40% of the volatile suspended solids in domestic wastewater are nonbiodegradable, and therefore inert to biological attack.[8] A more appropriate division would be inert, biodegradable, and biomass, because each influences biochemical operations in a different way. By definition, biodegradable suspended mat-

ter and biomass are both organic. Inert suspended matter may be either organic or inorganic. We will consider only inert organic particulate matter herein, although it should be recognized that inorganic material will behave in the same manner. In keeping with the convention adopted earlier, the concentrations of all organic particulate constituents will be reported in COD units. It should be noted, however, that the conversion factor from COD to mass units will depend on the nature of the suspended matter and may well be different for each. Even though a conversion factor of 1.20 g COD per g of dry solids has been adopted herein for biomass and microbial debris, it is impossible to generalize about the value of the conversion factor for other constituents and they must be determined on a case by case basis.

5.2.1 Soluble, Nonbiodegradable Organic Matter in Influent

Soluble, nonbiodegradable organic matter will not be acted on by the biomass in a biochemical operation, although its concentration can be influenced by physical/chemical phenomena such as adsorption and volatilization. If the material is not adsorbed into biomass and if it is not volatilized, then its concentration in the effluent from a bioreactor, S_I, will be the same as its concentration in the influent, S_{IO}:

$$S_I = S_{IO} \tag{5.42}$$

If the material is volatile and/or adsorbable, appropriate reaction terms must be inserted into the mass balance equation, yielding an equation for the effluent concentration. Such expressions are discussed in Chapter 22 where the fate of synthetic organic chemicals will be considered. The total concentration of soluble organic matter (SOM) will be the sum of S_I and S_S, plus any soluble microbial products, as discussed earlier. Except in special cases, the presence of soluble, nonbiodegradable organic matter will be ignored in this book since it has no impact on the systems to be discussed.

5.2.2 Inert Organic Solids in Influent

Like soluble, nonbiodegradable organic matter, inert organic solids undergo no reaction in a biological reactor. Consequently, the mass leaving the bioreactor must equal the mass entering it if steady-state is to be achieved. Unlike soluble, nonbiodegradable organic matter, however, the concentration of inert organic solids in the bioreactor depends on the magnitude of the SRT relative to the HRT, as can be seen by performing a mass balance across the bioreactor. This follows from the fact that the solids only leave the bioreactor through the wastage stream. Reference to Figure 5.1 and construction of the mass balance in which X_I represents the concentration of inert organic solids yields:

$$F \cdot X_{IO} - F_w \cdot X_I = 0 \tag{5.43}$$

or

$$X_I = (F/F_w)X_{IO} \tag{5.44}$$

Division of both F and F_w by V and invocation of the definitions of HRT and SRT reveals:

$$X_I = (\Theta_c/\tau)X_{IO} \tag{5.45}$$

The concentration of inert organic solids in the bioreactor is greater than the concentration in the influent, with the concentration factor being Θ_c/τ. If no effluent is removed through the biomass separator so that all leaves through the wastage flow, then the SRT and HRT are the same and the concentration of inert organic solids in the bioreactor is the same as the concentration in the influent.

Although derived for inert organic solids, Eq. 5.45 holds for all inert solids, regardless of whether they are organic or inorganic, but the concentration of inorganic solids must be expressed in mass units rather than as COD. Generally it is best to consider each type separately because that will allow easy determination of how the fraction of volatile solids in a sludge will change during treatment. Because consideration of inert inorganic solids is simply a bookkeeping exercise, we will consider only organic solids herein. Nevertheless, the impact of inorganic solids should be kept in mind because they contribute to the total concentration of suspended material that must be handled in a treatment system.

If a treatment system receives inert solids, the suspended solids in the bioreactor will include them in addition to active biomass and biomass debris. This combination of suspended solids is called mixed liquor suspended solids (MLSS) and will be given the symbol X_M. The concentration of MLSS is the sum of X_I, as given by Eq. 5.45, and X_I, as given by Eq. 5.24:

$$X_M = \left(\frac{\Theta_c}{\tau}\right)\left[X_{IO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{5.46}$$

Looking at the bracketed term it can be seen that the contribution of biomass related solids (right term) decreases as the SRT is increased, whereas the contribution of inert solids (left term) does not. As a consequence, the percentage of active biomass and biomass debris in the MLSS decreases as the SRT is increased.

The active fraction of the MLSS is the active biomass concentration divided by the MLSS concentration. Use of the appropriate equations in this definition gives:

$$f_A = \frac{1}{1 + f_D \cdot b_H \cdot \Theta_c + X_{IO}\left[\dfrac{(1 + b_H \cdot \Theta_c)}{Y_H(S_{SO} - S_S)}\right]} \tag{5.47}$$

This reduces to Eq. 5.26 when X_{IO} is zero. As one might expect, it tells us that the active fraction will be smaller when X_{IO} is larger, relative to S_{SO}. It also tells us that the active fraction is not affected by the SRT to HRT ratio, even though the MLSS concentration is. Thus, maintenance of a high active fraction requires minimization of the amount of inert suspended solids entering a biological reactor.

In Section 5.1.5, it was seen that the process loading factor is less convenient than the SRT as a design and control parameter for CSTRs at steady-state because of the necessity of knowing the active fraction of biomass in the MLSS before determining the microbial specific growth rate. This is particularly true when a wastewater contains inert solids because of the impact of those solids on the active fraction, as defined in Eq. 5.47.

As discussed in Section 5.1.3, the observed yield is defined as the mass of biomass formed per unit mass of substrate removed. As such, the presence of inert solids has no impact on it and it is still given by Eq. 5.28.

The mass rate at which solids must be disposed of will be increased by the presence of inert solids. Since nothing happens in the biochemical operation to reduce the amount of inert solids present, the mass rate of solids disposal will just be increased by the rate at which inert solids enter the system. Letting W_M represent the wastage rate of MLSS, often referred to as the solids wastage rate, gives:

$$W_M = F \cdot X_{IO} + W_I \tag{5.48}$$

$$W_M = F\left[X_{IO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{5.49}$$

Since nothing happens to inert solids in the bioreactor, they will have no impact on the oxygen requirement. Thus, it is still given by Eqs. 5.33–5.36.

Example 5.2.2.1

Continue with the problem begun in Example 5.1.3.1.

a. What will be the MLSS concentration in the bioreactor expressed as COD if the influent contains 25 mg/L as COD of inert organic solids.

From the previous examples, we know that τ = 8 hr, Θ_c = 160 hr, S_{SO} = 200 mg/L as COD, and S_S = 0.31 mg/L as COD. Insertion of these into Eq. 5.46 gives:

$$X_M = \left(\frac{160}{8}\right)\left\{25 + \frac{[1 + (0.20)(0.01)(160)](0.34)(200 - 0.31)}{1 + (0.01)(160)}\right\}$$

$$X_M = 1190 \text{ mg/L as COD}$$

Comparison of this value to the total biomass concentration in Example 5.1.3.2 shows that the inert solids increased the suspended solids concentration by 500 mg/L as COD. This could also have been seen through application of Eq. 5.45:

$$X_I = (160/8)(25) = 500 \text{ mg/L as COD}$$

b. For the situation in question a, what will be the MLSS concentration as suspended solids if the nature of the inert organic solids is such that they have a COD of 1.0 g COD/g SS?

From Example 5.1.3.2, we know that the total biomass concentration is 575 mg/L as suspended solids. Because the COD of the inert organic solids is 1.0 g COD/g SS, their concentration in the bioreactor is 500 mg/L as suspended solids. Thus, the MLSS concentration, expressed as suspended solids, is:

$$X_M = X_i + X_I$$

$$= 575 + 500 = 1075 \text{ mg/L as suspended solids}$$

Note that it was necessary to apply the conversion to each type of solids separately because each has a different value of COD/SS.

c. What fraction of the MLSS is made up of active biomass?

When the MLSS contains inert organic solids that have a unit COD different from that of biomass, the answer to this question depends on the way in which the concentration is measured. If it is measured in COD units, the active biomass concentration is 523 mg/L (from Example 5.1.3.2), and thus the active

fraction is:

$$f_A = 523/1190 = 0.44$$

The same value may be obtained from application of Eq. 5.47. If the concentration is measured and expressed in suspended solids units, the active biomass concentration is 436 mg/L (from Example 5.1.3.2), and the active fraction is:

$$f_A = 436/1075 = 0.41$$

Equation 5.47 may also be used to obtain this value, but X_{IO} must be expressed in suspended solids units and Y_H must have units of mg biomass suspended solids formed per unit of substrate COD removed.

The dependence of the active fraction on the unit system used to measure the components makes it important to specify that unit system. Regardless of the unit system, however, it is clear that the presence of an apparently insignificant concentration of inert organic solids in the influent to a bioreactor can have a significant impact on the active fraction of the MLSS, in this case decreasing it from 0.76 to 0.44 (when the concentration is expressed as COD).

d. What is the solids wastage rate?

Again, the answer to this question depends on the unit system used. Using Eq. 5.48 with W_I in COD units as given in Example 5.1.4.1:

$$W_M = (1.0)(25) + 34.5 = 59.5 \text{ mg/hr as COD}$$

Using the same equation with X_{IO} and W_I in suspended solids units:

$$W_M = 25 + 28.7 = 53.7 \text{ mg/hr as dry solids}$$

The same values can be obtained from application of Eq. 5.49 provided X_{IO} and Y_H are expressed in the appropriate units. The significant point, however, is that the addition of 25 mg/hr (as either COD or suspended solids) of inert organic solids to the bioreactor caused an increase of 25 mg/hr in the amount of solids wasted.

5.2.3 Biomass in Influent

The effect of the presence of active biomass in the influent to a CSTR can be seen by performing a new mass balance on biomass using Eq. 5.10, and including influent biomass at concentration $X_{B,HO}$. Performance of the steps that led to Eq. 5.12 gives:

$$\mu_H = \frac{1}{\Theta_c} + b_H - \frac{X_{B,HO}}{\tau \cdot X_{B,H}} \tag{5.50}$$

which reduces to Eq. 5.12 when $X_{B,HO}$ is zero. It is important to recognize that by definition, the SRT is given by Eq. 5.1. The presence of active biomass in the influent to the bioreactor does not change that definition. Rather, Eq. 5.50 shows that the presence of active biomass in the influent reduces the specific growth rate of the biomass in the bioreactor relative to the SRT, and that greater amounts reduce it more. In other words, if biomass is present in the influent, the biomass in the reactor does not have to grow as fast to maintain itself as it does when the influent contains no biomass. This means that if two bioreactors have the same HRT and SRT, but one receives active biomass in the influent, it will produce an effluent with a lower substrate concentration.

The effluent substrate concentration cannot be found by substituting the equation for μ_H into the Monod equation as was done before because the result will contain $X_{B,H}$, which is unknown. Thus, we must use a different approach. The mass balance on substrate is unchanged, and thus Eq. 5.18 is still valid. Substitution of Eq. 5.50 for μ_H into it gives:

$$X_{B,H} = \left(\frac{\Theta_c}{\tau}\right)\left[\frac{X_{B,HO}}{1 + b_H \cdot \Theta_c} + \frac{Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{5.51}$$

Comparison of Eq. 5.51 with Eq. 5.20 reveals that the active biomass concentration will be higher than in a bioreactor receiving no biomass in the influent. Furthermore, it can be seen that the term in the brackets has been divided into two components. The one on the right is the contribution of new growth to the active biomass whereas the term on the left is the contribution of the influent biomass. Note that the latter is less than the input biomass concentration because of decay. Substitution of Eq. 5.51 into Eq. 5.50, with subsequent substitution of the resulting equation into Eq. 3.36 yields a quadratic equation for the substrate concentration:

$$[\hat{\mu}_H - (1/\Theta_c + b_H)]S_S^2 - [\hat{\mu}_H(X_{B,HO}/Y_H + S_{SO}) + (K_S - S_{SO})(1/\Theta_c + b_H)]S_S + S_{SO} \cdot K_S(1/\Theta_c + b_H) = 0 \tag{5.52}$$

The easiest way to use the equations is to calculate the substrate concentration first and then use that result to calculate the biomass concentration.

Any biomass debris in the influent will behave as inert solids, thereby making its concentration in the bioreactor greater than that in the influent by a factor equal to the ratio of the SRT to HRT. In addition, biomass debris will be generated from decay of the active biomass in the influent as well as by decay of the active biomass formed in the bioreactor. Consequently, the total biomass debris concentration will be:

$$X_D = \left(\frac{\Theta_c}{\tau}\right)\left[X_{DO} + \frac{f_D \cdot b_H \cdot \Theta_c \cdot X_{B,HO}}{1 + b_H \cdot \Theta_c} + \frac{f_D \cdot b_H \cdot \Theta_c \cdot Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{5.53}$$

where X_{DO} is the concentration of biomass debris in the influent. The first term in the brackets is the contribution of biomass debris in the influent, the second term represents the debris formed through decay of the active biomass in the influent, and the last term is the formation of biomass debris from growth and decay of new biomass in the bioreactor.

The MLSS concentration is the sum of the active biomass and the debris:

$$X_M = \left(\frac{\Theta_c}{\tau}\right)\left[X_{DO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)X_{B,HO}}{1 + b_H \cdot \Theta_c} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{5.54}$$

The first term in the brackets is the contribution of influent biomass debris, the second term is the contribution of the active biomass in the influent, and the third is the contribution of new biomass growth on the soluble substrate. Note that Eqs. 5.51, 5.53, and 5.54 all reduce to the equations in Section 5.1.3 when the influent is free of biomass.

One significant impact of having biomass in the influent to a CSTR is to reduce $S_{S,min}$, the minimum substrate concentration attainable. As was done in Section 5.1.3,

S_{Smin} can be calculated by letting the SRT become very large so that $1/\Theta_c$ approaches zero. If the assumption is made that S_{Smin} is negligible with respect to S_{SO}, which will generally be the case, then it is possible to show that:

$$S_{Smin} = \frac{K_S \cdot b_H}{\hat{\mu}_H \left(1 + \frac{X_{B,HO}}{Y_H \cdot S_{SO}}\right) - b_H} \tag{5.55}$$

Comparison of Eq. 5.55 to Eq. 5.14 shows clearly that S_{Smin} will be smaller when the influent contains biomass. Furthermore, if we let Ω represent the value of S_{Smin} when the influent contains active biomass expressed as a fraction of the value in the absence of active biomass, it can be shown that:

$$\Omega = \frac{1 - \frac{b_H}{\hat{\mu}_H}}{1 - \frac{b_H}{\hat{\mu}_H} + \frac{X_{B,HO}}{Y_H \cdot S_{SO}}} \tag{5.56}$$

For many situations this can be further simplified by noting that $b_H/\hat{\mu}_H$ is often much less than one, allowing it to be dropped from the equation:

$$\Omega \approx \frac{1}{1 + \frac{X_{B,HO}}{Y_H \cdot S_{SO}}} \tag{5.57}$$

Equations 5.56 and 5.57 show clearly that the degree of reduction in S_{Smin} will depend on the magnitude of the influent biomass concentration relative to the influent substrate concentration, with larger values producing lower S_{Smin} values. This suggests that one way to meet a desired S_{Smin} concentration when a normal CSTR cannot is to add active biomass to the influent.

Another impact of having biomass in the influent to a CSTR is the prevention of washout, because if the influent contains active biomass, so will the bioreactor, no matter how small the SRT is made. Thus, a minimum SRT can no longer be defined in the same sense that it was defined for a bioreactor without biomass in the influent. However, the degree of substrate removal that will occur at very small SRTs depends on the influent biomass concentration, as can be seen by examination of Eq. 5.52. Consider the special situation in which the SRT has been selected so that

$$\frac{1}{\Theta_c} + b_h = \hat{\mu}_H \tag{5.58}$$

Under that condition, Eq. 5.52 reduces to

$$\left[\frac{X_{B,HO}}{Y_H} + S_{SO} + K_S - S_{SO}\right] S_S = S_{SO} \cdot K_S \tag{5.59}$$

or

$$S_S = \frac{S_{SO} \cdot K_S}{X_{B,HO}/Y_H + K_S} \tag{5.60}$$

Consequently, the larger the influent biomass concentration, the greater the degree

of substrate removal, even when the SRT is very small. Equations 5.52 and 5.57 can be used to evaluate the potential impacts of purposeful addition of biomass (bioaugmentation) on process performance.

In Eq. 5.47, we saw how the presence of inert solids in the influent to a CSTR influenced the active fraction of the suspended solids. Thus, it would be instructive to see how the entrance of biomass into a CSTR influences it. Application of the definition of active fraction gives:

$$f_A = \frac{1}{1 + f_D \cdot b_H \cdot \Theta_c + X_{DO}\left[\dfrac{1 + b_H \cdot \Theta_c}{X_{B,HO} + Y_H(S_{SO} - S_S)}\right]} \tag{5.61}$$

Comparison of it to Eqs. 5.26 and 5.47 reveals some interesting things. First, if the influent biomass is all active, so that X_{DO} is zero, the active fraction is the same as that in a bioreactor that does not receive any solids. In other words, Eq. 5.61 reduces to Eq. 5.26. Second, if the influent biomass is all debris and none is active, Eq. 5.61 reduces to Eq. 5.47, as we would expect. Finally, if the influent biomass contains both active biomass and debris, which would be the more usual case, the active fraction in the bioreactor will depend on the active fraction of the influent biomass, but will be different from it.

The impact of influent biomass on the active fraction is another drawback to the use of the process loading factor as a design tool for bioreactors. As when inert solids are present, SRT is more easily related to the biomass specific growth rate and therefore to process performance.

The presence of biomass in the influent to a CSTR increases the mass of solids that must be disposed of. The basic definition of the MLSS mass wastage rate is the wastage flow rate, F_w, multiplied by the MLSS concentration. Using Eq. 5.54 for X_M and using the definitions for SRT and HRT gives:

$$W_M = F\left[X_{DO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)X_{B,HO}}{1 + b_H \cdot \Theta_c} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{5.62}$$

Alternatively, if the MLSS concentration is known, the wastage rate can be determined by multiplying the wastage flow rate by the MLSS concentration, i.e., by invoking the definition used to derive Eq. 5.62. The cell debris entering the bioreactor is unaffected by microbial activity, so the mass of influent debris leaving in the waste solids equals the mass entering. However, the active biomass entering undergoes decay so that the mass discharged in the waste biomass stream is less than that in the influent. If we could mark the influent biomass in order to distinguish it from biomass generated in the bioreactor, we would find that its contribution was given by the second term in the brackets. Finally, new biomass is formed by growth and substrate utilization, and it too must be wasted from the bioreactor. Its contribution is given by the last term in the brackets.

As discussed in Section 5.1.3, the observed yield is defined as the mass of new biomass formed per unit mass of substrate removed. Examination of the last term in Eq. 5.62 reveals that the amount of new biomass formed when biomass is in the influent to the bioreactor is given by the same expression as that in a bioreactor not receiving biomass. Thus, the presence of biomass in the influent has no impact on the observed yield, which is still given by Eq. 5.28. It should be noted, however,

that the observed yield is difficult to measure in such a bioreactor because it is impossible to distinguish new biomass from that entering in the influent. In Chapter 8, we discuss techniques for measuring kinetic and stoichiometric coefficients in the presence of influent solids, including biomass.

The basic equation for the oxygen requirement in a CSTR receiving both soluble substrate and biomass is the same as Eq. 5.32 because that equation came from the overall rate expression for oxygen as given by Eq. 5.9, which is independent of the influent characteristics. The effects of the influent characteristics are incorporated through substitution of the appropriate equations for μ_H and $X_{B,H}$. Using Eq. 5.50 for the former and Eq. 5.51 for the latter yields:

$$RO = F\left\{\frac{(1 - f_D)b_H \cdot \Theta_c \cdot X_{B,HO}}{1 + b_H \cdot \Theta_c} + (S_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H}{1 + b_H \cdot \Theta_c}\right]\right\} \qquad (5.63)$$

Comparison of this expression to Eq. 5.33 reveals that the second term within the braces is the oxygen requirement associated with soluble substrate removal. The first term is the requirement associated with the input of active biomass. When $X_{B,HO}$ is zero, Eq. 5.63 reduces to Eq. 5.33. The input of biomass debris has no effect on the oxygen requirement because the debris passes through the bioreactor without reaction. The oxygen requirement can also be calculated by performing a mass balance on COD across the bioreactor, as discussed previously for the soluble substrate case, with proper consideration of the input of active biomass on that balance.

Example 5.2.3.1

Continue with the problem begun in Example 5.1.3.1. The conditions are the same as those used previously (τ = 8 hr, Θ_c = 160 hr, S_{SO} = 200 mg/L as COD), except that 25 mg/L as COD of active biomass is added to the influent. No biomass debris is added.

a. What is the effluent m-cresol concentration?

To determine this we must use Eq. 5.52 with the kinetic parameters and stoichiometric coefficients given in Table E5.1:

$$[0.20 - (1/160 + 0.01)]S_S^2 - 0.20[(25/0.34) + 200]$$
$$+ (3.5 - 200)(1/160 + 0.01)S_S + (200)(3.5)(1/160 + 0.01) = 0$$

S_S = 0.21 mg/L as COD

Thus, the presence of 25 mg/L as COD of active biomass in the influent to the CSTR decreased the effluent m-cresol concentration by 0.10 mg/L. This is approaching the minimum that could be obtained in such a bioreactor if no biomass was entering in the influent (as shown in Example 5.1.3.1).

b. What is the minimum m-cresol concentration that could be obtained in this bioreactor?

This must be calculated with Eq. 5.55:

$$S_{Smin} = \frac{(3.5)(0.01)}{0.20\{1 + [25/(0.34)(200)]\} - 0.01}$$

S_{Smin} = 0.13 mg/L as COD

If it was necessary to decrease S_S below this value, either another bioreactor configuration would have to be used or more active biomass would have to be added to the influent.

c. What is the MLSS concentration?

Insertion of the appropriate values into Eq. 5.54 gives:

$$X_M = \left(\frac{160}{8}\right)\left\{0 + \frac{[1 + (0.20)(0.01)(160)]25}{1 + (0.01)(160)} + \frac{[1 + (0.20)(0.01)(160)](0.34)(200 - 0.21)}{1 + (0.01)(160)}\right\}$$

$$X_M = 944 \text{ mg/L}$$

Thus, the addition of 25 mg/L as COD of active biomass to the influent increased the suspended solids concentration in the bioreactor by 254 mg/L as COD. This is less than the impact of 25 mg/L of inert solids considered in Example 5.2.2.1 because the active biomass underwent decay in the bioreactor, whereas the inert solids did not undergo any reaction.

d. What is the wastage rate of MLSS from the bioreactor?

This may be calculated with Eq. 5.62 or by multiplying the MLSS concentration by the wastage flow rate. Since we already know the MLSS concentration, the latter approach is easier:

$$W_M = (0.05)(944) = 47.2 \text{ mg/hr as COD}$$

The addition of 25 mg/hr as COD of active biomass to the bioreactor increased the mass of solids to be disposed of by 12.7 mg/hr as COD. The remainder of the influent biomass was destroyed by decay in the bioreactor.

e. How much oxygen must be supplied to the bioreactor?

This may be determined with Eq. 5.63:

$$RO = 1.0\left\{\frac{(1 - 0.20)(0.01)(160)(25)}{1 + (0.01)(160)} + (200 - 0.21)\left[1 - \frac{[1 + (0.20)(0.01)(160)]0.34}{1 + (0.01)(160)}\right]\right\}$$

$$RO = 177.6 \text{ mg/hr}$$

Comparison of this value to the oxygen requirement calculated in Example 5.1.4.1 shows that the oxygen requirement was increased by 12.4 mg/hr. Of this, 0.1 mg/hr was due to increased substrate removal and 12.3 mg/hr was due to decay of the added biomass. This latter fact could have been predicted by noting that 25 mg/hr as COD of biomass was added to the system, but only 12.7 mg/hr as COD of additional solids was wasted. The remainder (12.3 mg/hr) was oxidized and increased the oxygen requirement. This illustrates the advantage of expressing the concentrations of all organic constituents in COD units.

Frequently the waste solids stream from a bioreactor is directed to a CSTR to allow stabilization of the wasted biomass before ultimate disposal. This represents an extreme case of a CSTR receiving biomass in its influent because the influent MLSS concentration is generally high whereas the concentration of influent soluble

substrate is very low. In that case, we are seldom concerned with the concentration of soluble substrate in the effluent and the term $(S_{SO} - S_S)$ can be considered to be negligible compared to the other terms in the performance equations, allowing their simplification. If this is done to Eqs. 5.51, 5.53, 5.54, 5.62, and 5.63, the resulting expressions are:

$$X_{B,H} = \left(\frac{\Theta_c}{\tau}\right)\left(\frac{X_{B,HO}}{1 + b_H \cdot \Theta_c}\right) \tag{5.64}$$

$$X_D = \left(\frac{\Theta_c}{\tau}\right)\left(X_{DO} + \frac{f_D \cdot b_H \cdot \Theta_c \cdot X_{B,HO}}{1 + b_H \cdot \Theta_c}\right) \tag{5.65}$$

$$X_M = \left(\frac{\Theta_c}{\tau}\right)\left[X_{DO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)X_{B,HO}}{1 + b_H \cdot \Theta_c}\right] \tag{5.66}$$

$$W_M = F\left[X_{DO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)X_{B,HO}}{1 + b_H \cdot \Theta_c}\right] \tag{5.67}$$

$$RO = F\left[\frac{(1 - f_D)b_H \cdot \Theta_c \cdot X_{B,HO}}{1 + b_H \cdot \Theta_c}\right] \tag{5.68}$$

The equation for the active fraction, Eq. 5.61 is unchanged. As usual, the term F in Eqs. 5.64–6.68 (either directly or through the HRT, τ) is the flow rate entering the bioreactor. It should be noted, however, that for this situation, the flow often arises from another bioreactor, e.g., as its waste solids stream. Furthermore, bioreactors receiving only biomass in their feed often do not employ a biomass separator, but discharge their entire effluent stream to a solids dewatering device. In that case the SRT and HRT are equal, allowing further simplification of Eqs. 5.64–5.68.

Example 5.2.3.2

Continue with the problem begun in Example 5.1.3.1. If the waste solids stream from the bioreactor receiving only the soluble substrate is directed to a CSTR with an SRT of 480 hr and an HRT of 24 hr, what will be the MLSS concentration, the active fraction, and the oxygen requirement in the CSTR? In addition, how many mg/hr as COD of solids must be sent to ultimate disposal from it?

a. What is the MLSS concentration?

This may be determined from Eq. 5.66. The value of $X_{B,HO}$ is the same as the value of $X_{B,H}$ calculated in Example 5.1.3.2, which was 523 mg/L as COD. The value of X_{DO} is the same as the biomass debris concentration in the bioreactor in Example 5.1.3.2. This value was not calculated in the example, but is the difference between X_I and $X_{B,H}$, or 167 mg/L as COD. These values must be substituted into Eq. 5.66, along with the other appropriate values

$$X_M = \left(\frac{480}{24}\right)\left\{167 + \frac{[1 + (0.20)(0.01)(480)]523}{1 + (0.01)(480)}\right\}$$

$$X_M = 6875 \text{ mg/L as COD}$$

b. What is the active fraction?

This may be determined in either of two ways: from its definition or from Eq.

5.61, which is applicable in this case also. By definition, the active fraction is the concentration of active biomass divided by the MLSS concentration. The active biomass concentration may be calculated with Eq. 5.64:

$$X_{B,H} = \left(\frac{480}{24}\right)\left[\frac{523}{1 + (0.01)(480)}\right]$$

$$X_{B,H} = 1803 \text{ mg/L as COD}$$

Thus, the active fraction is 0.26. Most of the MLSS is biomass debris that accumulated as the active biomass underwent decay.

c. What is the oxygen requirement?

The oxygen requirement can be calculated with Eq. 5.68. For this equation, the flow rate, F, into the bioreactor must be known. Since the flow entering the CSTR in this example is the wastage flow from the bioreactor in Example 5.1.3.1, its flow rate is 0.05 L/hr. Therefore:

$$RO = 0.05\left[\frac{(1 - 0.20)(0.01)(480)(523)}{1 + (0.01)(480)}\right]$$

$$RO = 17.3 \text{ mg/hr}$$

The input rate of COD into the bioreactor is (0.05 L/hr)(690 mg/L) = 34.5 mg/hr. Thus, approximately half of the original oxygen demand was satisfied in the bioreactor. This represents 50% stabilization of the waste solids. The remainder of the oxygen demand remains in the sludge that goes to ultimate disposal, but much of it is in the form of biomass debris, which degrades very slowly.

d. How many mg/hr as COD of solids go to ultimate disposal?

Equation 5.67 may be used to determine this:

$$W_M = 0.05\left\{167 + \frac{[1 + (0.20)(0.01)(480)]523}{1 + (0.01)(480)}\right\}$$

$$W_M = 17.2 \text{ mg/hr as COD}$$

Note that the sum of this value and the oxygen requirement equals the input rate of COD into the bioreactor. This follows from the requirement for a COD balance across the bioreactor. All electrons in the waste solids going to the bioreactor must either be transferred to oxygen or remain in the unreacted solids. This serves as a convenient continuity check on the computations.

5.2.4 Biodegradable Solids in Influent

With the exception of some industrial wastes, most wastewaters contain particulate organic matter, much of which is biodegradable. This is true even when the biochemical operation is preceded by a sedimentation basin because much of the particulate organic matter is colloidal in size, making it too small for removal by settling. As a consequence, consideration must be given to the fate of particulate, biodegradable organic matter in order for models to accurately reflect the responses of biochemical operations treating many wastewaters.

An important characteristic of particulate organic matter is that it is too large to be transported across cell membranes. Thus, it must be acted on by extracellular

enzymes to release soluble constituents that can be taken up and used as substrate by the biomass. As discussed in Sections 2.4.4 and 3.5, the solubilization reactions, commonly referred to as hydrolysis, are quite complex and have received little research attention. It is clear, however, that a specific reaction term must be included for conversion of particulate substrate into soluble substrate. The basic model presented in Section 5.1 does not include such a reaction term, and it is not adequate for considering the impact of particulate biodegradable organic matter. Furthermore, inclusion of terms for hydrolysis complicate the situation sufficiently that explicit equations of the type presented in this chapter are difficult to obtain. Consequently, the impacts of particulate substrate will not be considered further here. Rather, Chapter 6 presents a more complex model that considers the fate of particulate substrate. Nevertheless, the basic concepts presented in this chapter, such as the importance of SRT, the buildup of inert solids, etc. are valid for all wastewaters and all suspended growth cultures, regardless of the nature of the electron donor. Thus, the concepts presented herein serve as a foundation for consideration of more complex situations.

5.3 METHODS OF BIOMASS RECYCLE AND WASTAGE

In developing the models in the preceding sections a portion of the flow from the bioreactor passed through a biomass separator, as shown in Figure 5.1. However, the characteristics of that separator were not specified.

On a laboratory scale, reactors have been used which mimic closely the situation depicted in Figure 5.1. In one design, the biomass is contained in a vessel with porous walls through which clear liquid flows, leaving the biomass inside. The wastage flow is taken directly from the vessel. In another design, tangential-flow membrane filtration is used as a biomass separator, with a high recirculation flow being removed from the bioreactor, passed over the filter, and returned to the bioreactor while a small percentage passes through the filter for discharge as the biomass-free effluent. Again, biomass wastage is through a stream removed directly from the bioreactor.

Almost all full-scale bioreactor systems use sedimentation as the means of biomass separation, and as a consequence so do many lab-scale bioreactors. The technique that most closely mirrors the ideal situation pictured in Figure 5.1 is the internal, upflow clarifier. It employs a quiescent zone in the bioreactor which flow enters from the bottom. By designing the system so that the upflow velocity of the effluent fluid is less than the settling velocity of the biofloc, it is possible to discharge a clarified effluent while retaining the biomass in the reactor. Wastage is taken directly from the mixed bioreactor. The most common technique in practice is to use an external clarifier with return of a concentrated solids stream, called solids (or biomass) recycle, to the bioreactor. Two such systems are shown in Figure 5.2. The system shown in Figure 5.2a is called the Garrett[2] configuration. Its distinguishing characteristic is that it wastes biomass directly from the bioreactor, just as was done in Figure 5.1. The system shown in Figure 5.2b is the configuration most often found in practice. Thus, it is called the conventional configuration. Its distinguishing characteristic is that biomass is wasted from the concentrated solids recycle stream.

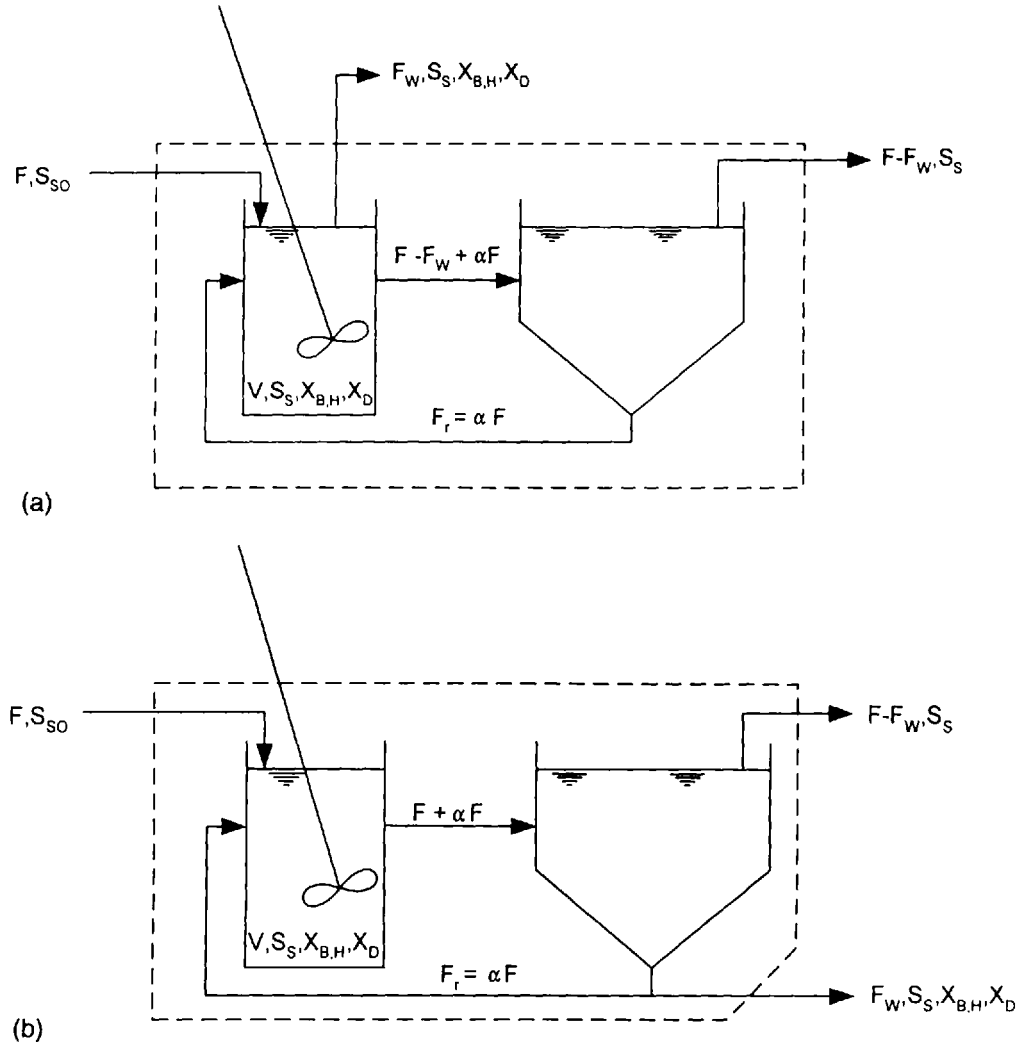


Figure 5.2 Schematic diagrams of two CSTRs with biomass recycle from sedimentation basins. (a) The Garrett configuration in which biomass is wasted directly from the reactor. (b) Conventional configuration in which biomass is wasted from the sludge recycle flow.

5.3.1 Garrett Configuration

The key features of the Garrett configuration are that for a CSTR of fixed volume, the SRT is controlled solely by the flow rate of the waste solids stream, F_w, and the performance of the bioreactor is independent of the solids recycle flow rate, F_r. This means that the recycle flow rate can be chosen to give proper operation of the settler, thereby ensuring that all biomass is returned to the bioreactor. In Chapter 10, we consider the question of the choice of that flow rate.

The rationale behind the Garrett configuration can be seen by considering the system boundary to be the dashed line in Figure 5.2a. The flows in and out of that

boundary are the same as those in Figure 5.1, and thus the Garrett configuration corresponds to the ideal bioreactor configuration used to derive the equations in Sections 5.1 and 5.2. Because the bioreactor is completely mixed, the concentration of the constituents in the effluent stream leaving it and flowing to the settler are the same as those in the bioreactor. Furthermore, if the settler is operated properly, the solids blanket will be small so that the mass of solids in the settler is small relative to the mass in the bioreactor. This has two effects. First, there will be little reaction in the settler so that the concentrations of soluble constituents in the recycle stream are the same as those in the bioreactor. Because all soluble concentrations are the same, the recycle of soluble constituents around the system has no impact on system performance. This can be shown by performing mass balances around both the bioreactor and the settler. Furthermore, since the settler is perfect, all biomass entering it is returned to the bioreactor, and thus the recycle stream has no impact on the biomass concentration in the bioreactor either. (In Chapter 10, we see how to correct system operation for the fact that the effluent stream from all real-world settlers contains a small amount of biomass.) Second, the SRT is still defined by Eq. 5.1, and since the concentration of biomass in the wastage stream is the same as the concentration in the bioreactor, Eq. 5.2 is also true. This means that the Garrett configuration conforms to the ideal situation considered in Sections 5.1 and 5.2, and the equations developed there are directly applicable. Because the recycle flow rate does not appear in any of those equations and because the SRT is not influenced by the recycle flow rate, it can be seen that bioreactor performance is independent of it.

5.3.2 Conventional Configuration

The conventional configuration shown in Figure 5.2b can also be defined by a dashed system boundary and by use of the arguments above, the recycle flow rate need not appear in the system descriptive equations, provided they are written in terms of the system SRT. However, because the solids in the recycle flow (and hence in the wastage flow) are at a higher concentration than the solids in the bioreactor, Eq. 5.2 is no longer valid, although Eq. 5.1 is valid. Furthermore, the waste solids concentration in Eq. 5.1 is a function of the recycle flow rate. Thus, the key to understanding the impact of the recycle flow rate on the performance of a bioreactor with the conventional configuration is an understanding of the effect that the recycle flow rate has on the waste solids concentration.

Examination of Figure 5.2b reveals that the concentration of any constituent in the wastage flow will be the same as the concentration in the solids recycle stream since the wastage flow comes from it. The MLSS in a bioreactor can generally be considered to be homogeneous and to settle without segregation so that the ratio of active biomass concentration to MLSS concentration in the recycle stream is the same as the ratio in the bioreactor. Thus, we may simply perform a mass balance on MLSS across the settler to determine the value of X/X_u for use in Eq. 5.1 for defining the SRT in terms of the recycle flow rate.

Assuming that the settler is perfect and that no reaction occurs in it, a steady-state mass balance across it states that the mass of MLSS entering the settler must equal the mass leaving in the combined recycle and wastage streams:

$$(F + F_r)X_M = (F_w + F_r)X_{Mr} \tag{5.69}$$

where X_{Mr} is the MLSS concentration in the recycle stream. Recognizing that the solids concentration in the wastage flow, X_{Mw}, is equal to X_{Mr}, Eq. 5.69 may be rearranged and substituted into Eq. 5.1 to give:

$$\Theta_c = \frac{V}{F_w}\left(\frac{F_w + F_r}{F + F_r}\right) \tag{5.70}$$

This shows that the SRT of a system with a conventional configuration depends on both the influent and recycle flow rates in addition to the wastage flow rate and the bioreactor volume. Rearrangement of Eq. 5.70 gives:

$$F_w = \frac{V \cdot F_r}{(F + F_r)\Theta_c - V} \tag{5.71}$$

This shows that any time the recycle flow rate is changed, the wastage flow rate must be adjusted to maintain a constant SRT. In contrast, if the Garrett configuration were used, no adjustment of the wastage flow would be required. Furthermore, Eq. 5.71 also shows that the wastage flow from the conventional configuration must be adjusted any time the influent flow changes. Such a change is not required with the Garrett configuration, as shown by Eq. 5.2.

Once the impact of the recycle flow rate on the SRT is recognized, all of the system performance equations from Sections 5.1 and 5.2, which were written in terms of the SRT, may be used because the recycle flow rate has no impact on them at a fixed SRT. Its impact is indirect, through influencing the SRT if proper adjustment of the wastage rate is not done. In Chapter 10, we consider selection of the appropriate recycle flow rate for optimum settler performance.

5.4 PERFORMANCE OF A CSTR AS PREDICTED BY MODEL

The major value of the models in this chapter is as an aid to understanding how CSTRs behave under a variety of conditions. We saw during derivation of the equations that the most important operational variable is the SRT. Consequently, we will first use graphs generated with the parameter values in Table 5.2 to see how SRT influences bioreactor performance. Next, we will investigate the impact of adding inert and microbial solids to the influent to a CSTR. Finally, we will consider the impact of the values of the kinetic parameters and stoichiometric coefficients on bioreactor performance by considering situations representative of both high and low temperatures.

5.4.1 Effect of SRT

We see in Chapter 10 that the use of sedimentation as the means of biomass separation places both upper and lower limits on the SRT allowable in an operating bioreactor. However, to fully demonstrate the potential impact of SRT on bioreactor performance, the figures presented here were generated without regard to those limits. Consequently, most operating systems will not experience the broad range of conditions exhibited in the figures.

Table 5.2 Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used to Generate Figures in Section 5.4

Symbol	Units	Value
$\hat{\mu}_H$	hr^{-1}	0.50
K_S	mg/L as COD	50
Y_H	mg biomass COD formed/mg COD removed	0.60
b_H	hr^{-1}	0.0075
f_D	mg debris COD/mg biomass COD	0.20
F	L/hr	1.00
S_{SO}	mg/L as COD	500
X_{IO}	mg/L as COD	0[a]
$X_{H,IO}$	mg/L as COD	0[a]

[a]Unless specified otherwise.

Figure 5.3 shows the effect of SRT on the soluble substrate concentration. For the parameter values in Table 5.2, the minimum SRT is 2.24 hours, and at that SRT the effluent substrate concentration is equal to the influent concentration. This SRT is also called the washout point for the bioreactor because biomass can no longer exist in it. As the SRT is increased, however, growth can be established in the bioreactor and substantial substrate removal occurs even when the SRT is low. For example, an SRT of slightly more than four hours is all that is required to reduce

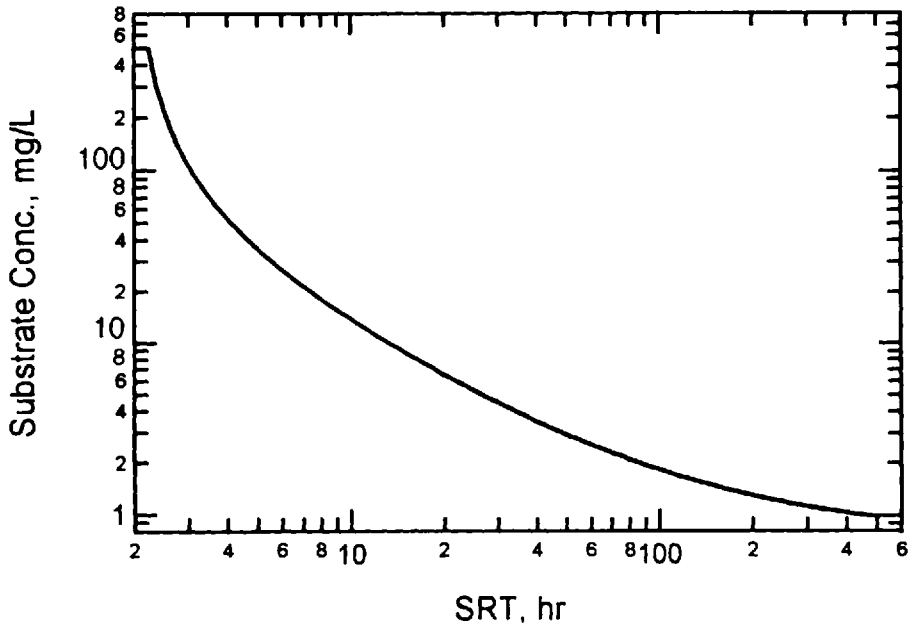


Figure 5.3 Effect of SRT on the concentration of soluble substrate (as COD) in a CSTR receiving a soluble substrate. Kinetic parameters and stoichiometric coefficients are listed in Table 5.2.

the substrate concentration from 500 to 50 mg/L as COD. This demonstrates an important characteristic of biological reactors: they are able to achieve substantial removal of soluble substrate at very short SRTs. However, incremental removal of substrate declines sharply as the SRT is increased, although such increases make the bioreactor more stable. For example, compare the differences in substrate concentration resulting from a 10% change in SRT at SRTs of 4, 40, and 400 hours. The minimum attainable substrate concentration (S_{Smin}) for the parameter values used to generate Figure 5.3 is 0.76 mg/L as COD and examination of the graph shows that the value is approached very slowly. When methods like COD and biochemical oxygen demand (BOD) are used to measure effluent substrate concentration on full scale systems, the SRT generally has little measurable effect beyond certain values. That is simply because the potential change is small compared to the error associated with the test method.

The dashed curve in Figure 5.4 shows the impact of SRT on the total mass of biomass (as COD) in the CSTR. It will be recalled from Eqs. 5.24 and 5.13 that the biomass concentration depends on the HRT whereas the substrate concentration is independent of it. Thus, the HRT must be considered to convert the mass values in Figure 5.4 into concentrations. However, since the influent flow rate was fixed for the simulations, as shown in Table 5.2, consideration of different HRTs is equivalent to consideration of different bioreactor volumes. Thus, the total biomass concentration in a CSTR with a volume of V liters being operated at a particular SRT can be obtained by dividing the mass associated with that SRT by V. For example, when the SRT is 100 hr, the bioreactor contains 20 g COD of total biomass (active plus

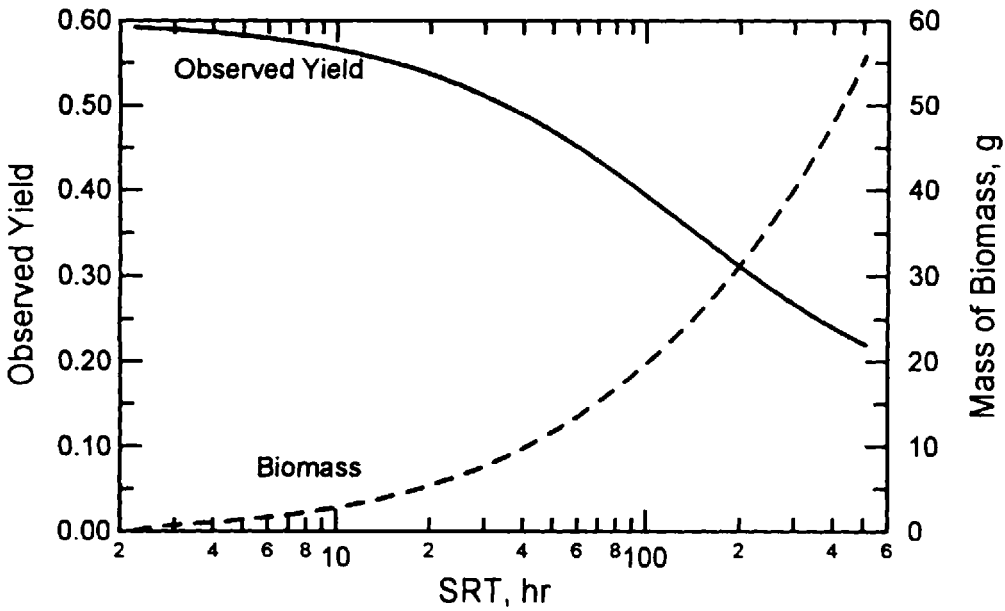


Figure 5.4 Effect of SRT on the observed yield and the total mass of biomass (as COD) in a CSTR receiving a soluble substrate. Kinetic parameters and stoichiometric coefficients are listed in Table 5.2.

debris). Thus, if the reactor volume were 2 L the concentration would be 10 g/L as COD, whereas if the volume were 4 L the concentration would be 5 g/L. The dashed curve is important because it demonstrates clearly that the mass, and thus the concentration, of biomass in a reactor of volume V increases as the SRT is increased. It is this increase in the mass of organisms that allows more substrate to be removed as the SRT is increased, even though the HRT is kept the same; more biomass can accomplish more in the same available time. A fixed mass of organisms is required to accomplish a given amount of substrate removal; only the concentration is influenced by the bioreactor volume.

The solid curve in Figure 5.4 is important because it illustrates the decrease in the observed yield that occurs as the SRT of a bioreactor is increased. This is because of the increased importance of decay as the SRT is increased, as indicated in Eq. 5.28. Only when the SRT is very short and the biomass is growing very rapidly will most substrate utilization go for growth, allowing the observed yield to approach the true growth yield, which was 0.60 for this case. For all other situations a significant amount of energy must be expended for maintenance and other purposes associated with decay, thereby lowering the observed yield.

Figure 5.4 presents the total mass of biomass, but we know that biomass decay will decrease the active fraction through the build up of biomass debris as the SRT is increased. Figure 5.5 shows that effect. At low SRT, when the specific growth rate is high, the impact of decay will be small so that little debris will be generated, making the active fraction large. As the SRT is increased, however, the build up of biomass debris in the bioreactor becomes significant and the active fraction drops, until at high SRTs only a small percentage of the biomass is actually contributing

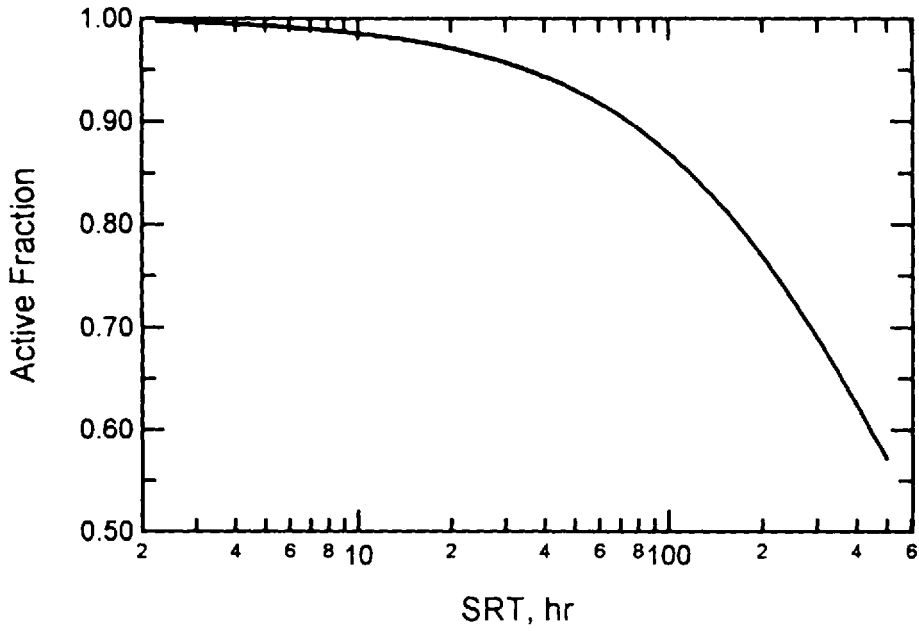


Figure 5.5 Effect of SRT on the active fraction of the biomass in a CSTR receiving a soluble substrate. Kinetic parameters and stoichiometric coefficients are listed in Table 5.2.

to substrate removal. Nevertheless, the amount of active biomass continues to increase as the SRT is increased, as can be seen by multiplying the active fraction from Figure 5.5 times the mass of biomass from Figure 5.4. As a consequence, increases in SRT are generally worthwhile, although a point of diminishing return will be reached. One price associated with a high SRT is the cost of moving inactive biomass continually around the system, yet little can be done about it because it is impossible to separate active biomass from debris. In fact, it is possible that accumulated debris contributes to improved settling properties at increased SRT.

A benefit associated with increased SRTs is that less excess biomass must be disposed of because more of it is oxidized through decay, maintenance energy needs, etc. as discussed above for Y_{obs}. This effect is illustrated in Figure 5.6. When the SRT is small, even though Y_{obs} is high, substrate removal is incomplete, as shown in Figure 5.3, and little excess biomass is synthesized. However, as the SRT is increased beyond the minimum SRT, more excess biomass is generated because of increased growth and substrate removal, all with a relatively high observed yield. As the SRT is increased further, the effluent substrate concentration becomes small relative to the influent concentration so that the term $S_{SO} - S_S$ becomes essentially constant. This occurs at an SRT of about 10 hr for the parameter values used to generate the graphs. Beyond that point, further increases in SRT increase the importance of decay, causing Y_{obs} and the net production of biomass to decline as shown by the decreasing mass which must be wasted. Those same events determine the shape of the oxygen consumption curve. At very short SRTs, substrate removal is incomplete and decay is of little importance, with the result that most of the electrons

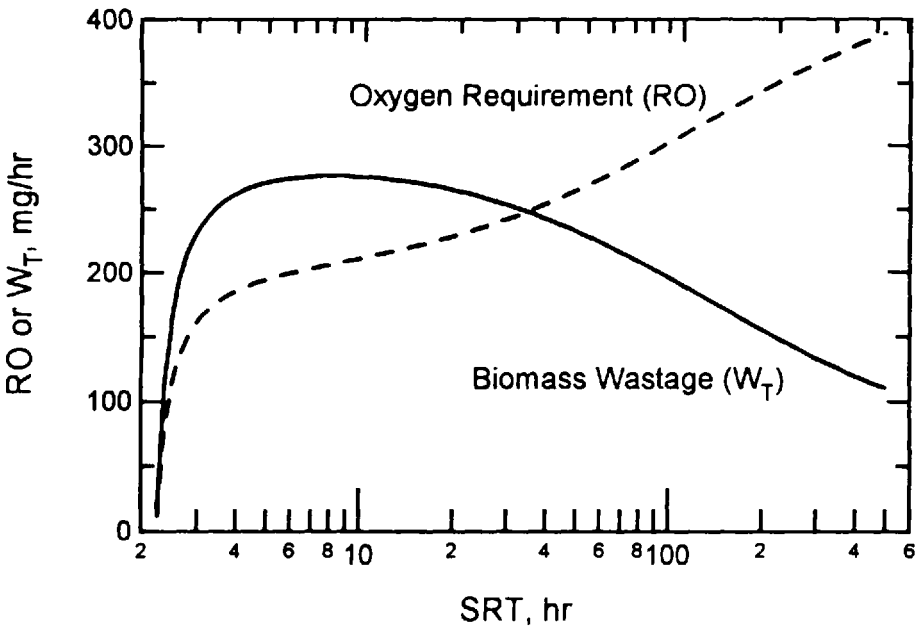


Figure 5.6 Effect of SRT on the biomass wastage rate (as COD) and oxygen requirement in a CSTR receiving a soluble substrate. Kinetic parameters and stoichiometric coefficients are listed in Table 5.2.

available in the influent substrate are either associated with the effluent substrate or the biomass formed. Thus, relatively little oxygen is required. As the SRT is increased slightly to the point where substrate removal is essentially complete (SRT $\approx$ 10 hr) but decay is not yet important, all of the oxygen use is associated with substrate removal and biomass growth, i.e., energy for synthesis. Further increases in the SRT result in more decay, and almost all of the increased oxygen requirement as the SRT is increased past 10 hr is associated with that. In other words, the decreased mass of excess biomass associated with longer SRTs is at the expense of an increased oxygen requirement. This suggests that the choice of SRT is often governed by the relative costs of supplying oxygen versus disposing of excess sludge.

Although Figures 5.3–5.6 were developed with parameter values representative of aerobic growth of heterotrophic bacteria, it is important to recognize that the shapes of the curves are representative of microbial growth in general, including aerobic growth of autotrophs and anoxic growth of heterotrophs. All that is necessary is an appropriate change in electron donor or acceptor and adjustment of the parameter values.

5.4.2 Effects of Influent Solids

In Section 5.2.2 it was seen that the impact of inert solids in the influent to a CSTR is to reduce the active fraction of the MLSS. This effect is illustrated in Figure 5.7. The solid curve is the same as the one in Figure 5.5 whereas the dashed one represents the case in which 100 mg/L as COD of inert solids are added to the influent.

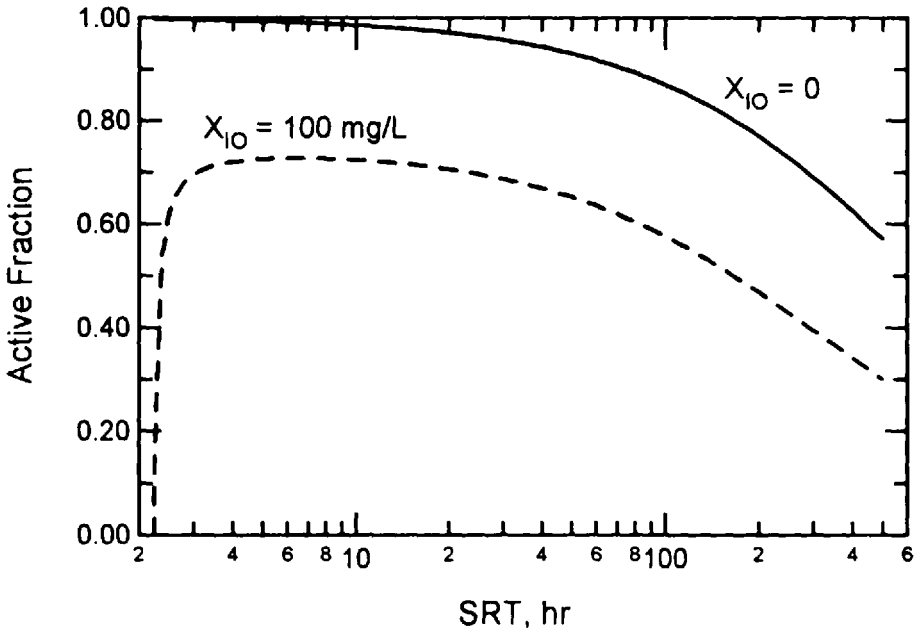


Figure 5.7 Effect of 100 mg/L (as COD) of influent inert organic solids on the active fraction of the biomass in a CSTR receiving a soluble substrate. Kinetic parameters and stoichiometric coefficients are listed in Table 5.2.

Comparison of the two curves reveals that only moderate amounts of inert solids in the influent to a CSTR can decrease the active fraction to less than 50%, especially at longer SRTs. As a result, final settlers and pumps for the recycle of biomass must be made larger to handle solids which contribute nothing to the process. Thus, it is generally more economic to reduce the concentration of inert solids prior to biological reactors.

As discussed in Section 5.2.3, one significant effect of having biomass in the influent to a CSTR is to prevent washout, thereby allowing substrate removal to occur at SRTs below the normal minimum. This is illustrated in Figure 5.8, where the impact on the soluble substrate concentration of having 50 mg/L as COD of active biomass in the influent is shown. The most dramatic effect is at SRTs near the minimum. Instead of having a discontinuity at the point of washout like the curve for the bioreactor without biomass in the influent, the concentration in the bioreactor receiving biomass slowly approaches the influent concentration as the SRT is made smaller and smaller. Under those conditions the microorganisms are growing and removing substrate at a very rapid specific rate, but the residence time in the bioreactor is too short to allow more complete removal to occur. Of course, higher concentrations of biomass in the influent will allow more substrate to be removed at short SRTs. The main importance of the effect illustrated in Figure 5.8 is as an explanation of why washout does not occur in circumstances where it might be expected. This is especially important during laboratory studies in which investigators attempt to measure $\hat{\mu}$ by observing the SRT at washout. Contamination of feed

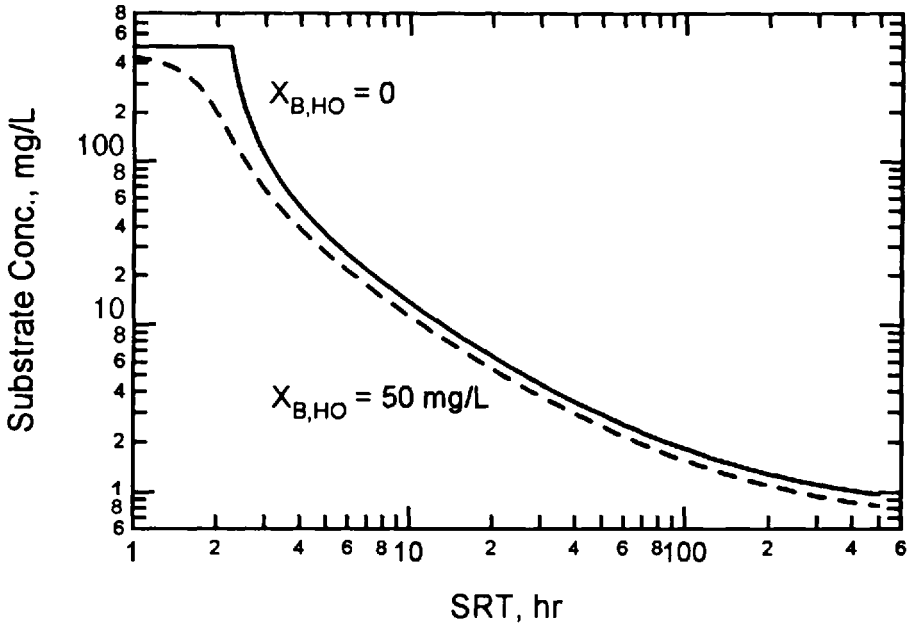


Figure 5.8 Effect of 50 mg/L (as COD) of active biomass in the influent to a CSTR on the soluble substrate concentration (as COD) in the reactor. Kinetic parameters and stoichiometric coefficients are listed in Table 5.2.

lines, thereby introducing biomass with the feed, can prevent the expected response and lead to error in the determination.

Another effect of influent biomass discussed previously is to reduce S_{Smin}, the minimum attainable substrate concentration from a single CSTR. It was seen in Eqs. 5.56 and 5.57 that the degree of reduction depends on the magnitude of the influent biomass concentration relative to the influent substrate concentration and this effect is illustrated in Figure 5.9. For the parameter values given in Table 5.2, S_{Smin} in the absence of influent biomass is 0.76 mg/L as COD, yet the presence of influent biomass can decrease that value significantly, as shown in the figure. This fact may be useful as engineers seek to reduce the concentrations of specific pollutants to very low levels. For example, although industrial wastewater treatment systems generally receive influent from several production areas, one may be the primary source of a targeted pollutant. If that waste stream were pretreated in a small bioreactor without biomass recycle prior to discharge to the main bioreactor, it would do two things: (1) provide a source of bacteria capable of degrading the targeted pollutant coming from the other production areas, and (2) reduce the concentration of the targeted pollutant in the influent to the main bioreactor. The combined effect of these two contributions would be to make the influent biomass to substrate concentration ratio large for the targeted pollutant, thereby allowing the main bioreactor to achieve a lower effluent substrate concentration than would be possible otherwise.

Re-examination of Figures 5.4 and 5.5 reveals that a CSTR with an SRT of 200 hr and an HRT of 10 hr would have a total biomass concentration of 3100

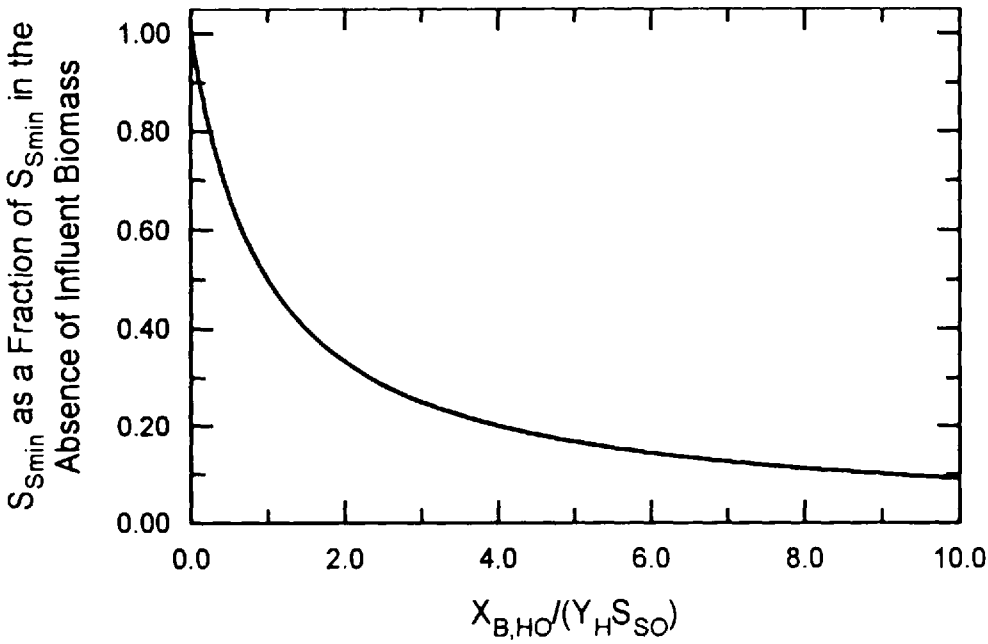


Figure 5.9 Effect of the influent biomass concentration relative to the influent substrate concentration on the minimum substrate concentration attainable in a CSTR. S_{Smin} is expressed as a fraction of the S_{Smin} value attainable in a similar bioreactor receiving no influent biomass.

mg/L with an active fraction of 0.76 if it were treating a wastewater with the characteristics in Table 5.2. What would be the fate of the excess biomass from that bioreactor if it were sent for treatment to another CSTR in which the ratio of the SRT to the HRT ratio remains fixed at 10? In other words, the SRT is increased by increasing the HRT proportionally. Since the concentration of soluble substrate in the waste biomass stream is negligible, Eqs. 5.64–5.68 describe the performance of the CSTR receiving the waste biomass and the results of their use are shown in Figure 5.10. There it can be seen that because of the buildup of debris in the bioreactor, the total biomass concentration will not go to zero as the SRT is increased, but will approach a limit, although the active biomass will become quite small. Furthermore, it can be seen that there is a point of diminishing return with regard to further increases in SRT because the active biomass declines rapidly at first, but then more slowly as the SRT is increased further. This is characteristic of the first order expression chosen to depict decay. It should be remembered that the model used assumes that debris is totally inert, whereas it will undergo some destruction given sufficient time, as discussed in Sections 2.4.2 and 3.3.1. Thus, it should be recognized that the residual stable biomass concentration will probably be less than that depicted by the model. Just as with the CSTR whose performance was depicted in Figure 5.6, in an aerobic process the destruction of biomass occurs at the expense of oxygen. Thus, the oxygen requirement is the mirror image of the total biomass curve. For simplicities sake, the influent flow rate to the bioreactor was taken as 1.0 L/hr, making the mass input rate of biomass equal to 3100 mg/hr as COD. Thus, it can be seen that about 50% of the oxygen demand of the influent biomass must ultimately be satisfied at longer SRTs. In other words, the final residual solids are highly stabilized.

5.4.3 Effects of Kinetic Parameters

The SRT of a CSTR is the primary control variable available to a designer or operator; however, it is not the only factor affecting the performance of such a bioreactor. Examination of the equations for the performance of a CSTR reveals that the values of each of the kinetic parameters and stoichiometric coefficients will influence them as well. The primary effects of $\hat{\mu}_H$ and K_S are on the substrate concentration. A higher value of $\hat{\mu}_H$ and a lower value of K_S allow the biomass to grow faster at a given substrate concentration, thereby giving a lower reactor substrate concentration for any given value of the SRT. The Monod parameters also exert a strong effect on the minimum SRT, so that organisms with high $\hat{\mu}_H$ and low K_S values can grow in CSTRs with short SRTs. The effect of the Monod parameters on biomass concentration is strongest at short SRTs where the effect on the substrate concentration is strongest. They have almost no effect at longer SRT values, however. In contrast to the Monod parameters, the primary effect of the decay coefficient is on the biomass concentration and the oxygen requirement at longer SRTs. A high decay coefficient means that the bioreactor will be more efficient in oxidizing the substrate to carbon dioxide; consequently, the biomass concentration will be low and the oxygen requirement high. This effect will be especially pronounced at long SRTs. Changes in the true growth yield will also primarily affect the biomass concentration and the oxygen requirement. High yields will result in more biomass, but the culture

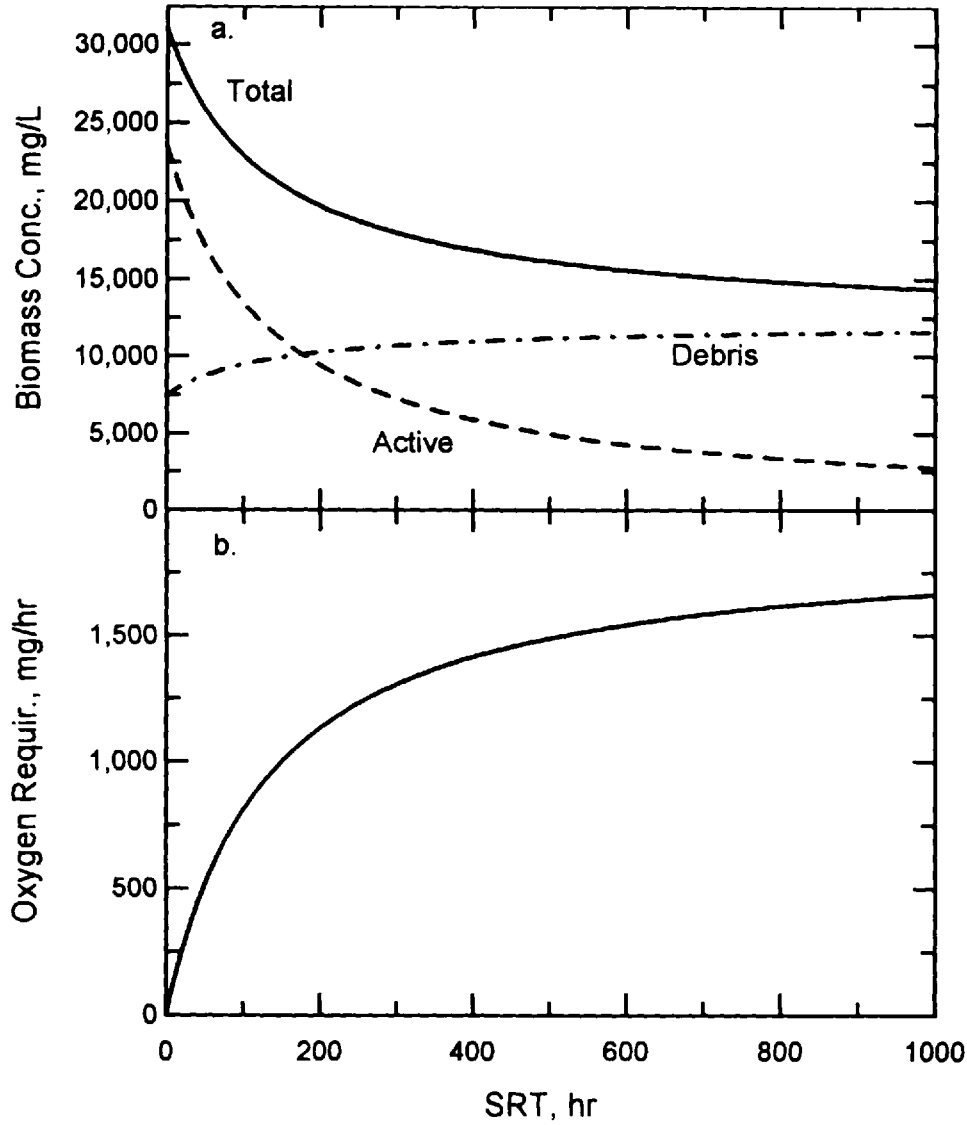


Figure 5.10 Effect of SRT on the performance of a CSTR receiving 1.0 L/hr of feed containing only 3100 mg/L as COD of biomass with an active fraction of 0.76. $\Theta_x/\tau = 10$. Kinetic parameters and stoichiometric coefficients are listed in Table 5.2.

will require less oxygen because more of the electrons in the substrate will be retained in the biomass synthesized.

Seldom does a situation occur in which only one parameter changes. Usually all will change. For example, as discussed in Section 3.9, temperature can affect all of them and the response of the system will depend on how the changes interact. An example of this is presented in Figure 5.11, which shows the substrate concentration, biomass wastage and oxygen requirement for the situations listed in Table

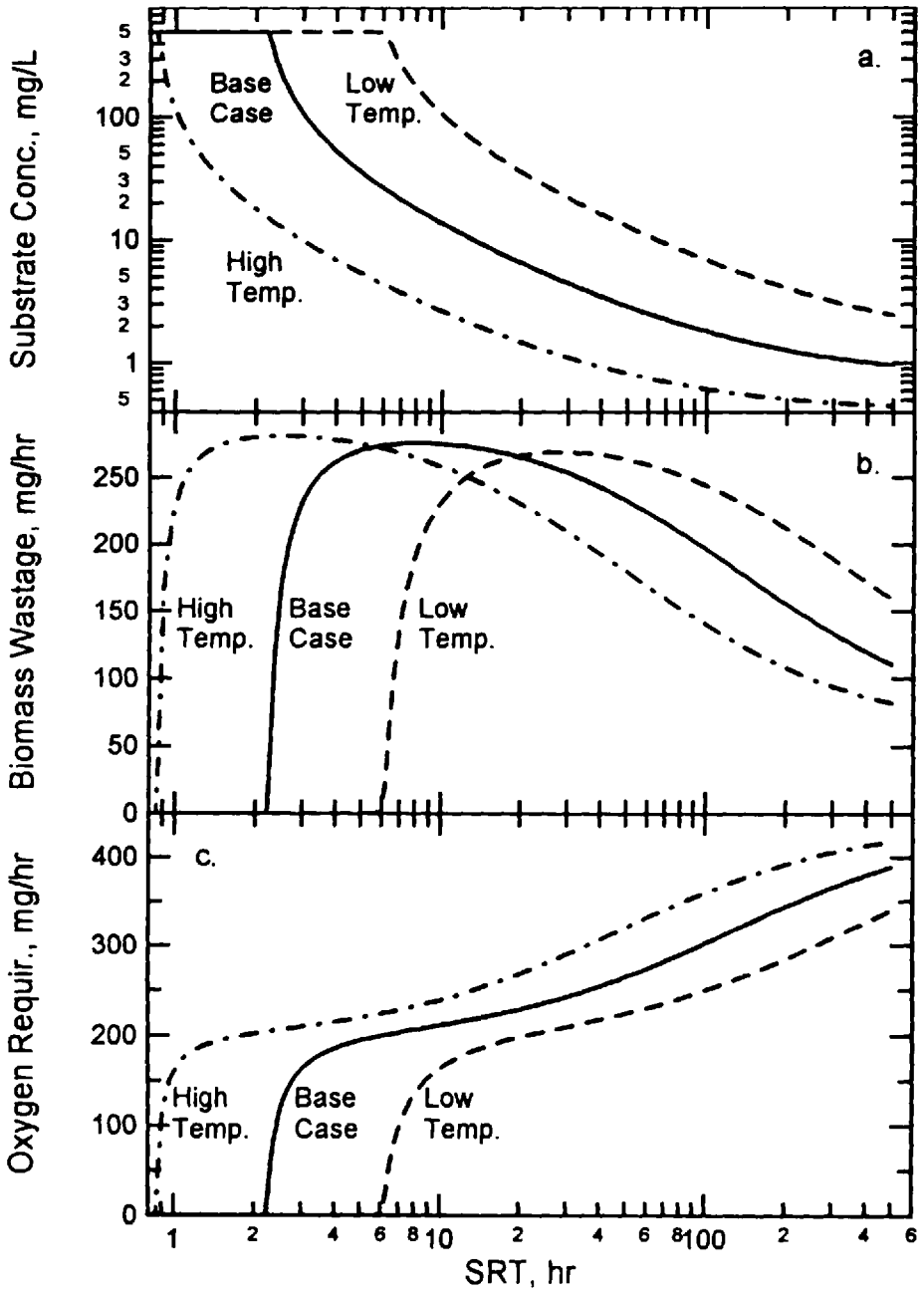


Figure 5.11 Effect of SRT on the performance of a CSTR at three different temperatures. Kinetic parameters and stoichiometric coefficients are listed in Table 5.3.

Table 5.3 Kinetic Parameters and Stoichiometric Coefficients Used to Generate Figure 5.11

	Value[a]	
Symbol	Low temperature	High temperature
$\hat{\mu}_H$[b]	0.20	1.25
K_S	100	25
Y_H	0.6	0.6
b_H[c]	0.0028	0.020
f_D	0.20	0.20

[a]The values for the base case and the units are given in Table 5.2.
[b]$\theta = 1.094$.
[c]$\theta = 1.104$.

5.3. The curves with dashes represent a low temperature situation, whereas those with dashes and dots represent a high temperature. The solid curves are the same as those presented earlier and are given for comparison. Both the low and high temperature cases represent a 10°C change from the base case, using temperature coefficients representative of the particular parameters. The true growth yield and the debris fraction were assumed to be independent of temperature. As would be anticipated from the previous discussion, the higher temperature results in more substrate removal at a given SRT whereas the low temperature results in less. Likewise, the high temperature results in a higher oxygen requirement, whereas the low temperature results in a lower one. Generally, the opposite is true of the biomass wastage although the curves overlap because of the impact of the temperature on the minimum SRT. In spite of more substrate removal, there is less net production of biomass in the bioreactor with the higher temperature. The curves in Figure 5.11 emphasize the point that one must learn to think in terms of the group of parameter values that characterize a particular substrate and culture, rather than thinking of individual coefficients.

5.5 KEY POINTS

1. Two retention times are important to the performance of CSTRs: the hydraulic residence time (HRT) and the solids retention time (SRT). The former represents the average length of time a fluid element stays in the bioreactor whereas the latter is the average length of time a solid particle stays there. The SRT can never be less than the HRT, but can be greater if the bioreactor contains a separator that removes particulate material from the effluent and returns it to the bioreactor.
2. The matrix format is a convenient way to present the stoichiometry and kinetics for multiple parallel reactions acting on several components.
3. The specific growth rate of the biomass in a CSTR is controlled by the SRT of the bioreactor, but is independent of its HRT. Since there is a functional relationship between the specific growth rate of biomass and

the concentration of substrate surrounding that biomass, control of the SRT allows control of the substrate concentration in a CSTR.

4. The concentration of biomass in a CSTR depends on the HRT, the SRT, and the amount of substrate removed. For a fixed SRT, the product of the HRT and the biomass concentration is a constant because the removal of a fixed mass of substrate generates a fixed mass of microorganisms.
5. The observed yield and active fraction of the biomass in a CSTR both decline as the SRT is increased because of the increased importance of death and decay at longer SRTs.
6. The oxygen requirement in an aerobic CSTR comes from a COD balance across the bioreactor, i.e., it must equal the total COD in minus the total COD out, including the COD of the biomass and the biomass debris. Consequently, any factor that decreases the observed yield, such as an increase in the SRT, will increase the oxygen requirement.
7. The SRT is preferable to the process loading factor (F/M ratio) as a design and control parameter for CSTRs because the latter requires knowledge of the active fraction before it can be related to the specific growth rate of the biomass whereas the former does not.
8. Both soluble nonbiodegradable COD and inert solids will be unaffected by biomass in a CSTR. The concentration of the former in the bioreactor will be the same as its concentration in the influent whereas the concentration of the latter will be greater by the factor Θ_x/τ because it becomes enmeshed in the MLSS and is lost from the system only through solids wastage.
9. The presence of active biomass in the influent to a CSTR decreases the specific growth rate of the biomass in the bioreactor, thereby reducing the effluent substrate concentration and increasing the biomass concentration associated with any given SRT. It also decreases the minimum substrate concentration that can be attained in the bioreactor.
10. When the influent to a CSTR contains biomass, but no soluble substrate, simplified equations can be derived for describing its performance.
11. Biodegradation of particulate substrates requires their solubilization by hydrolysis reactions. Such reactions are not included in the models in this chapter, and those models cannot be used to predict the performance of CSTRs receiving such substrates.
12. Two methods are used to waste biomass from a CSTR. The Garrett configuration wastes biomass directly from the bioreactor. The conventional configuration wastes biomass from the recycle flow returning biomass from the settler to the bioreactor. The Garrett configuration is easier to operate because it is not necessary to adjust the wastage flow rate each time a change is made in the recycle flow rate.
13. As the SRT is increased past the point of washout, the soluble substrate concentration in a CSTR declines rapidly, but as the SRT is increased further, it has less and less impact on soluble substrate removal.
14. When the SRT is small, the excess biomass production rate is small because of incomplete substrate removal. When the SRT is large, the excess biomass production rate is small because of the importance of biomass decay. It will reach a maximum at intermediate values of the

SRT. The oxygen requirement increases as the SRT is increased because more and more of the substrate and the biomass formed from it are oxidized.

15. Changes in $\hat{\mu}_H$ and K_S are reflected primarily in the substrate concentration whereas changes in Y_H and b_H have their largest effects on the biomass concentration.

5.6 STUDY QUESTIONS

1. A CSTR with a volume of 1000 L receives a flow of 100 L/hr. Ninety percent of the effluent exits through a biomass separator which removes all particulate material and returns it to the bioreactor. The remainder exits directly from the bioreactor. What is the HRT? What is the SRT?
2. Using the traditional approach to decay, write the matrix representing the stoichiometry and kinetics for aerobic growth of heterotrophic biomass on a soluble, noninhibitory substrate. Then use the information in the matrix to write the reaction rate term for active biomass and explain how you did it.
3. Derive Eq. 5.13 showing that the effluent substrate concentration from a CSTR is a function of only the SRT and is independent of the influent substrate concentration. Then explain what happens in the bioreactor to allow the effluent substrate concentration to be independent of the influent concentration.
4. Using derivations as needed, explain why the product of the biomass concentration and the HRT is a constant for a fixed SRT and influent concentration. Then explain why the steady-state performance of a CSTR equipped with a biomass separator is independent of its HRT.
5. The two major costs associated with the treatment of wastewaters in aerobic CSTRs are from the disposal of the excess biomass produced and the provision of ample oxygen. Describe how the SRT influences each of these and use that information as the basis of a discussion of the factors that must be considered by an engineer in choosing the design SRT for a CSTR.
6. A feed containing a soluble substrate with a biodegradable COD of 1000 mg/L is flowing at a rate of 100 L/hr into a 1000 L aerobic CSTR which contains a mixed community of microorganisms. The kinetic parameters and stoichiometric coefficients characterizing the culture are given in Table SQ5.1.
   a. Determine the following for an SRT of 100 hr.
      (1) The concentration of soluble biodegradable COD in the effluent.
      (2) The total biomass concentration in the bioreactor.
      (3) The active fraction of the biomass.
      (4) The g/hr of oxygen which must be supplied to the bioreactor.
      (5) The g/hr of excess biomass that must be disposed of.
      (6) The mg/hr of nitrogen that will be used for synthesis of new biomass.
      (7) The process loading factor.

Table SQ5.1 Kinetic Parameters and Stoichiometric Coefficients for a Mixed Microbial Community Growing on a Mixture of Organic Compounds

Symbol	Units	Value
$\hat{\mu}_H$	hr^{-1}	0.40
K_S	mg/L as COD	120
Y_H	mg biomass COD/mg substrate COD	0.54
b_H	hr^{-1}	0.004
f_D	mg debris COD/mg biomass COD	0.20

b. Would it be possible to achieve an effluent biodegradable COD concentration of 1.5 mg/L with this substrate and culture? Why?
c. What SRT would be required to reduce the soluble substrate concentration to 20 mg/L as COD? What concentration of active biomass would be present in the bioreactor?
d. If 150 mg/L as COD of inert organic solids are added to the influent, what fraction of the MLSS will be inert solids for the condition in part a? What will be the active fraction of the MLSS?

7. Explain through derivation why the concentration of inert solids in a CSTR is greater than the concentration in the influent by the factor Θ_x/τ.
8. Explain why the SRT is preferable to the process loading factor as a means of controlling a CSTR when the influent contains either inert solids or biomass.
9. Rework Study Question 6a, but include 110 mg/L as COD of biomass with an active fraction of 80 percent in the influent.
10. A CSTR is being used to degrade biomass. A stream containing biomass at a concentration of 800 mg/L as COD is flowing at a rate of 2.0 L/hr into a CSTR with a volume of 48 L which is being operated with an SRT of 240 hr. The active fraction of the biomass is 0.72, the decay coefficient, b, is 0.007 hr^{-1}, and the fraction of biomass contributing to biomass debris, f_D, is 0.20.
    a. What is the biomass concentration in the bioreactor (as COD)?
    b. What is its active fraction?
    c. What is the oxygen consumption rate in mg/hr?
    d. How many mg/hr (as COD) of biomass go to ultimate disposal?
11. Explain why the model presented in this chapter is not adequate for describing the performance of a CSTR receiving biodegradable particulate substrate.
12. Describe both the Garrett and conventional configurations for wasting biomass from CSTRs. Then explain (using derivations as needed) why operation of the Garrett configuration is simpler.
13. Using sketches as appropriate, describe the effects of SRT on the performance of a CSTR receiving a soluble substrate and explain conceptually why those effects occur.

14. Using Figures 5.4 and 5.5, generate a graph of active biomass as a function of SRT and discuss the implications of the curve to the selection of the SRT for a bioreactor.
15. Describe the major impacts of the presence of active biomass in the influent to a CSTR on the removal of soluble substrate. How can those effects be used to advantage in the treatment of wastewaters containing specific pollutants which must be removed to very low levels?
16. Using sketches as appropriate, describe the effects of SRT on the performance of a CSTR receiving only biomass and explain conceptually why those effects occur.

REFERENCES

1. Daigger, G. T. and C. P. L. Grady Jr., A model for the bio-oxidation process based on product formation concepts. *Water Research* **11**:1049–1057, 1977.
2. Garrett, M. T., Hydraulic control of activated sludge growth rate. *Sewage and Industrial Wastes* **30**:253–261, 1958.
3. Grady, C. P. L. Jr., L. J. Harlow, and R. R. Riesing, Effects of growth rate and influent substrate concentration on effluent quality from chemostats containing bacteria in pure and mixed culture. *Biotechnology and Bioengineering* **14**:391–410, 1972.
4. Henze, M., C. P. L. Grady Jr., W. Gujer, G. v. R. Marais, and T. Matsuo, Activated sludge model No. 1. *IAWPRC Scientific and Technical Reports*, No. 1, 1987.
5. Henze, M., C. P. L. Grady Jr., W. Gujer, G. v. R. Marais, and T. Matsuo, A general model for single-sludge wastewater treatment systems. *Water Research* **21**:505–515, 1987.
6. Irvine, R. L. and J. D. Bryers, Stoichiometry and kinetics of waste treatment. In *Comprehensive Biotechnology, Vol 4, The Principles, Applications and Regulations of Biotechnology in Industry, Agriculture and Medicine*, C. W. Robinson and J. A. Howell, eds. Pergamon Press: New York, NY, pp. 757–772, 1985.
7. Mason, C. A., J. D. Bryers and G. Hamer, Activity, death and lysis during microbial growth in a chemostat. *Chemical Engineering Communications* **45**:163–176, 1986.
8. McKinney, R. E. and R. J. Ooten, Concepts of complete mixing activated sludge *Transactions of the 19th Annual Conference on Sanitary Engineering*, University of Kansas, 32–59, 1969.
9. Rittmann, B. E., W. Bae, E. Namkung, and C.-J. Lu, A critical evaluation of microbial product formation in biological processes. *Water Science and Technology* **19**(7):517–528, 1987.
10. Stewart, M. J., Activated sludge process variations. The complete spectrum. *Water and Sewage Works Journal, Reference Number* **111**:241–262, 1964.

6

Multiple Microbial Activities in a Single Continuous Stirred Tank Reactor

In Chapter 5, we investigated the growth of aerobic heterotrophic bacteria in a single continuous stirred tank reactor (CSTR) receiving a soluble substrate. Through development of a simple model we saw that the SRT is an important determinant of bioreactor performance because it is related to the specific growth rate of biomass at steady-state. We also saw that there is a minimum SRT below which biomass growth cannot occur, as well as a minimum substrate concentration that can be achieved no matter how large the SRT. Finally, we saw how stoichiometry can be applied to determine the amount of electron acceptor required and the amount of excess biomass produced. All of these characteristics are of fundamental importance and apply to all types of biomass, both heterotrophic and autotrophic, in all types of environments, whether aerobic, anoxic, or anaerobic. Thus, even though the concepts in Chapter 5 are developed in the context of aerobic growth of heterotrophs, they are broadly applicable.

In spite of the broad utility of the concepts, the model developed in Chapter 5 has two characteristics that restrict its applicability in many wastewater treatment situations. One is that it is limited to soluble, readily biodegradable substrates, whereas most wastewaters contain particulate contaminants and soluble constituents of large molecular weight that must be reduced in size before they can be taken into the bacteria for biodegradation. If a model is to accurately depict the response of bioreactors receiving such wastewaters, it must include hydrolysis reactions. The other is that the biomass is assumed to be in a constant biochemical environment with no limitation by the electron acceptor. In many systems, however, limitations or alterations in the supply of electron acceptor cause shifts between aerobic and anoxic conditions, with short periods of anaerobiosis as well, and during these shifts the concentration of the electron acceptor may be limiting. Therefore, it would be desirable for a model to handle such situations.

In order to encourage practicing engineers to use modeling more extensively during the analysis of alternative wastewater treatment systems, in 1983 the International Association on Water Quality (IAWQ) [formerly the International Association on Water Pollution Research and Control (IAWPRC)] appointed a task group to review models for suspended growth cultures and to produce one capable of depicting the performance of wastewater treatment systems receiving both soluble and particulate substrates in which organic substrate removal, nitrification, and denitrification were all occurring. In other words, they were to consider most of the pro-

cesses discussed in Section 2.4. They completed their task in 1986 and submitted a report to IAWQ which was published in 1987,[17,18] outlining the major features of activated sludge model (ASM) No. 1. The task group was influenced by the published work of many researchers, but that of Marais and colleagues at the University of Cape Town in South Africa had a major impact on their thinking. A summary of much of the South African work can be found elsewhere.[12] Because ASM No. 1 is the result of the deliberations of several researchers with diverse opinions, and because it is capable of mimicking the performance of pilot[12] and full[3] scale systems, it will be adopted herein for investigating more fully the performance of suspended growth bioreactors. In this chapter, ASM No. 1 will be used to illustrate the impact in a single CSTR of the processes and events not covered in Chapter 5 and in Chapter 7 it will be used to investigate the performance of multiple bioreactor systems.

Because of the success of ASM No. 1, the IAWQ task group on mathematical modeling was reconstituted and asked to produce a consensus model capable of mimicking the performance of systems capable of performing organic substrate removal, nitrification, denitrification, and phosphorus removal. This was a complicated task because of the complexity of biological phosphorus removal and the evolving nature of our understanding of it. Nevertheless, they were successful, releasing their report in 1995,[19] calling the new model ASM No. 2. Use will be made of the model in Chapter 7, but it will not be explained in the same detail as ASM No. 1 because of the large number of components and processes involved. However, the major rate expressions associated with phosphorus removal in the model were presented in Section 3.7.

Modeling is now used extensively in biological wastewater treatment, in large part because of the success of ASM No. 1. Similar concepts have been applied to develop descriptive models for anaerobic wastewater treatment processes.[15,51] Space does not permit their investigation here, but the reader is encouraged to consult the primary literature concerning them.

6.1 INTERNATIONAL ASSOCIATION ON WATER QUALITY ACTIVATED SLUDGE MODELS

International Association on Water Quality ASM No. 1 is presented in matrix format in Table 6.1, where it can be seen to incorporate 8 processes and 13 components. Examination of the table reveals the utility of the matrix format, because application of the principles discussed in Section 5.1.2 allows immediate identification of the fate of each component and construction of the overall reaction rate term for it. It should be noted that components 1–8 are expressed in chemical oxygen demand (COD) units, whereas components 9–12 are given as nitrogen. The alkalinity is in molar units. These units are given in Table 6.2 along with the definition of each component.

6.1.1 Components in ASM No. 1

Components 1–5 and component 12 are all particulate. X_I is inert particulate organic material. The fact that there are no entries listed under it in Table 6.1 shows that it is neither generated nor destroyed in a biochemical reactor. However, if it is present

Table 6.1 Process Kinetics and Stoichiometry for Multiple Events in Suspended Growth Cultures as Presented by IAWQ Task Group on Mathematical Modeling[16,17]

Component[a] → i	1	2	3	4	5	6	7	8	9	10	11	12	13	
j Process ↓	X_I	X_S	$X_{B,H}$	$X_{B,A}$	X_D	S_I	S_S	S_O[b]	S_{NO}	S_{NH}	S_{ND}	X_{ND}	S_{ALK}	Process rate, r_j, ML^{-3}T^{-1}
1 Aerobic growth of heterotrophs			1				$-\frac{1}{Y_H}$	$\frac{1-Y_H}{Y_H}$		$-i_{NXB}$			$-\frac{i_{NXB}}{14}$	$\hat{\mu}_H\left(\frac{S_S}{K_S+S_S}\right)\left(\frac{S_O}{K_{OH}+S_O}\right)X_{B,H}$
2 Anoxic growth of heterotrophs			1				$-\frac{1}{Y_H}$		$-\frac{1-Y_H}{2.86\,Y_H}$	$-i_{NXB}$			$\frac{1-Y_H}{14(2.86\,Y_H)}-\frac{i_{NXB}}{14}$	$\hat{\mu}_H\left(\frac{S_S}{K_S+S_S}\right)\left(\frac{K_{OH}}{K_{OH}+S_O}\right)\cdot\left(\frac{S_{NO}}{K_{NO}+S_{NO}}\right)\eta_g X_{B,H}$
3 Aerobic growth of autotrophs				1				$\frac{4.57-Y_A}{Y_A}$	$\frac{1}{Y_A}$	$-i_{NXB}-\frac{1}{Y_A}$			$-\frac{i_{NXB}}{14}-\frac{1}{7Y_A}$	$\hat{\mu}_A\left(\frac{S_{NH}}{K_{NH}+S_{NH}}\right)\left(\frac{S_O}{K_{OA}+S_O}\right)X_{B,A}$
4 Death and lysis of heterotrophs		$1-f'_D$	−1		f'_D							$i_{NXB}-f'_D i_{NXD}$		$b_{L,H}X_{B,H}$
5 Death and lysis of autotrophs		$1-f'_D$		−1	f'_D							$i_{NXB}-f'_D i_{NXD}$		$b_{L,A}X_{B,A}$
6 Ammonification of soluble organic nitrogen										1	−1		$\frac{1}{14}$	$k_a S_{ND}X_{B,H}$
7 "Hydrolysis" of particulate organics		−1					1							$k_h\frac{X_S/X_{B,H}}{K_X+(X_S/X_{B,H})}\left[\left(\frac{S_O}{K_{OH}+S_O}\right)+\eta_h\left(\frac{K_{OH}}{K_{OH}+S_O}\right)\left(\frac{S_{NO}}{K_{NO}+S_{NO}}\right)\right]X_{B,H}$
8 "Hydrolysis" of particulate organic nitrogen											1	1		$r_h(X_{ND}/X_S)$
Observed conversion rates, ML^{-3}T^{-1}								$r_S=\sum_{j=1}^{n}\Psi_{Sj}r_j$						

[a]All organic compounds (1–7) and oxygen (8) are expressed as COD; all nitrogenous components (9–12) are expressed as nitrogen.
[b]Coefficients must be multiplied by −1 to express as oxygen.

Table 6.2 Definitions of Component Symbols in Table 6.1

Component number	Component symbol	Definition
1	X_I	Inert particulate organic matter, mg/L as COD
2	X_S	Slowly biodegradable substrate, mg/L as COD
3	$X_{B,H}$	Active heterotrophic biomass, mg/L as COD
4	$X_{B,A}$	Active autotrophic biomass, mg/L as COD
5	X_P	Debris from biomass death and lysis, mg/L as COD
6	S_I	Inert soluble organic matter, mg/L as COD
7	S_S	Readily biodegradable substrate, mg/L as COD
8	S_O	Oxygen, mg/L as COD
9	S_{NO}	Nitrate nitrogen, mg/L as N
10	S_{NH}	Ammonia nitrogen, mg/L as N
11	S_{ND}	Soluble biodegradable organic nitrogen, mg/L as N
12	X_{ND}	Particulate biodegradable organic nitrogen, mg/L as N
13	S_{ALK}	Alkalinity molar units

in the influent, it will accumulate with the degree being determined by the ratio of the SRT to the HRT (see Section 5.5.2). Component 2, X_S, is slowly biodegradable substrate. Although it is treated as a particulate constituent, its concentration in the influent to the bioreactor must be determined experimentally (see Chapter 8). It is destroyed by hydrolysis reactions, but is generated during biomass decay, which is modeled by the lysis:regrowth concept (see Section 3.3.2). Active heterotrophic biomass is depicted by $X_{B,H}$ whereas active autotrophic biomass (nitrifying bacteria) is given as $X_{B,A}$. Both are generated by growth on their respective substrates and both are lost by decay, leading to slowly biodegradable substrate and biomass debris, X_P. The latter is inert, and behaves in a manner similar to X_I. Component 12, X_{ND}, is particulate biodegradable organic nitrogen. It is formed by decay reactions since the slowly biodegradable substrate arising from biomass decay contains proteins and other nitrogen-containing organic compounds of high molecular weight. It is destroyed by hydrolysis.

Components 6–11 are all soluble, and with the exception of S_I, which is inert, are the constituents upon which the biomass acts. The presence of S_I in the matrix is simply to remind us that wastewaters contain nonbiodegradable soluble COD which passes through the bioreactor unaffected by biological activity (see Section 5.2.1). S_S is readily biodegradable substrate, which is removed by growth of heterotrophic biomass under aerobic or anoxic conditions and is generated by hydrolysis of slowly biodegradable organic matter. Its concentration in the wastewater entering a bioreactor must be determined experimentally, and the procedures for doing so are discussed in Chapter 8. Component 8 is oxygen, which is removed by aerobic growth of heterotrophic and autotrophic bacteria. The stoichiometric term for oxygen associated with heterotroph growth is the same as that in Table 5.1, but the term associated with autotroph growth contains the factor 4.57. That factor is required because ammonia is the substrate for autotrophic nitrifying bacteria and its concentration in the matrix (S_{NH}, component 10) is expressed as nitrogen, whereas oxygen is expressed as COD. Furthermore, Y_A has units of mg of biomass COD formed per mg

of nitrogen converted. Since the stoichiometric expression for oxygen in process 3 (autotrophic growth) comes from a COD-based stoichiometric equation, a factor must be included for the COD equivalents of ammonia-N in order to have consistent units. As discussed previously, the oxidation state of nitrogen is changed from −III to +V as nitrifiers form nitrate from ammonia. The amount of oxygen required to accept the electrons removed during this oxidation is 4.57 g O_2/g N, as indicated in Table 3.1. Thus, that factor must be included. Unlike the model in Table 5.1, which used the traditional approach to decay, no oxygen utilization is associated directly with biomass loss in Table 6.1 because it is modeled with the lysis:regrowth approach. Rather, the oxygen utilization associated with biomass loss occurs because of the use of readily biodegradable substrate generated by hydrolysis of the slowly biodegradable substrate formed by death and lysis. Component 9, S_{NO}, is nitrate-N. It is formed by aerobic growth of autotrophic bacteria and is lost as it serves as the electron acceptor for anoxic growth of heterotrophic bacteria. In the latter role, the oxidation state of the nitrogen is changed from +V to zero and the factor 2.86 appearing in the stoichiometric coefficient represents the oxygen equivalence of this change in units of g COD/g N, as shown in Table 3.1. Examination of column 10 shows that ammonia-N is involved in several reactions. Since ammonia is the preferred form of nitrogen for biomass growth, the term $-i_{N\,XB}$ is included in rows 1–3 to represent the amount of nitrogen incorporated into new biomass. No provision is made in this model for reduction of nitrate-N to ammonia-N for incorporation into biomass in the event insufficient ammonia is present. This restriction should be recognized. Other models[11,33] allow use of nitrate-N for biomass synthesis. The second stoichiometric coefficient in column 10 for aerobic growth of autotrophic bacteria represents the use of ammonia as a substrate and is analogous to the coefficients used for readily biodegradable substrate (column 7) removal by heterotrophic biomass in rows 1 and 2. Ammonia is formed by ammonification of soluble organic nitrogen, S_{NS}, which is the last nitrogen based soluble constituent. It, in turn, is formed by hydrolysis of particulate organic nitrogen.

In Section 3.2.10 the sensitivity of autotrophic biomass to low pH is discussed. Furthermore, in Section 3.2.5 it is seen that alkalinity is destroyed during their growth. If the wastewater contains insufficient alkalinity, the growth of autotrophic biomass will cease because a needed nutrient (carbon) is missing and because the pH will drop, inhibiting their activity. Thus, the destruction of alkalinity by autotrophic bacterial growth is an important event that must be considered by the engineer. Another factor influencing alkalinity is denitrification, which produces it, and in properly configured systems, this production can offset somewhat its destruction. The coefficients in column 13 account for the changes in alkalinity, S_{ALK}, associated with nitrogen conversions in the bioreactor. Although the IAWQ task group did not attempt to model the effects of those changes on pH, they noted that when the alkalinity falls below 50 mg/L as $CaCO_3$, the pH becomes unstable and can fall well below 6,[35] thereby hindering nitrification.

6.1.2 Reaction Rate Expressions in ASM No. 1

The growth of heterotrophic bacteria with associated use of substrate and electron acceptor is given by processes 1 and 2 in Table 6.1 for the situation in which ammonia serves as the nitrogen source for synthesis of new cell material. Process 1 is

aerobic growth and its rate is given by Eq. 3.35 with substitution of Eq. 3.46 for μ, the specific growth rate coefficient. In this case, reactant 1 is readily biodegradable substrate and reactant 2 is dissolved oxygen. The main purpose of the dissolved oxygen term is to turn off aerobic growth as the dissolved oxygen concentration becomes low, and to allow anoxic growth to begin if nitrate is present, as suggested by the rate term for process 2. The specific growth rate coefficient in that rate term is of the same form as Eq. 3.48, with $K_{O,H}$ serving as the inhibition coefficient K_{IO}. By using $K_{O,H}$ in that capacity, the oxygen terms in the rate expressions for processes 1 and 2 compliment each other, with one approaching zero as the other approaches one (their sum always equals one). It should be noted that both heterotrophic rate expressions go to zero under totally anaerobic conditions, i.e., in the absence of both oxygen and nitrate. Given long term acclimation to anaerobic conditions, fermentative reactions would allow growth of facultatively anaerobic bacteria, but such acclimation is not likely to occur in the systems for which the model was developed. Nevertheless, this limitation should be recognized and the model should not be used to simulate bioreactors in which fully anaerobic conditions develop. Comparison of the rate term for process 2 to Eq. 3.48 reveals the presence of an additional parameter, η_g. It is a correction factor for growth under anoxic conditions. As seen in Section 3.2.10, the $\hat{\mu}_H$ and K_S values for cultures grown on the same substrate under totally aerobic and totally anoxic conditions are very similar, whereas the yield is lower under anoxic conditions.[24] This suggests that an appropriate way to model two separate biomasses grown in different environments would be to use the same values of the kinetic parameters under the two conditions, but to use different yield values. A major purpose of the model in Table 6.1, however, is to model a biomass that is alternated between aerobic and anoxic conditions. In that case, although the entire biomass will be capable of aerobic growth, only a portion of it may be able to grow under anoxic conditions. The purpose of η_g is to correct for this condition. Because only a portion of the biomass may be capable of denitrification, η_g takes on values less than one, with those values depending somewhat on the system configuration.[16] Because of the empirical nature of η_g, it also corrects for differences in the yield values under the two environmental conditions, and thus, ASM No. 1 uses a common yield value for aerobic and anoxic conditions.

Process 3 in Table 6.1 is aerobic growth of autotrophic bacteria, which is modeled with Eq. 3.46 in which reactant 1 is ammonia-N and reactant 2 is dissolved oxygen. Two important things should be noted about the way this process is modeled. The first is that nitrification is considered to occur in a single step, with nitrate-N arising directly from ammonia-N. This is a simplification, because, as seen in Section 2.3.1, nitrification is a two step process, with nitrite as an intermediate. As discussed in Section 3.2.10, the kinetic parameters for *Nitrosomonas* and *Nitrobacter* are similar. Consequently, under balanced growth conditions nitrite is used as fast as it is formed so that its concentration is usually very low and of little importance. Therefore, to reduce the number of equations and to simplify the model, ASM No. 1 uses a one-step approach. It should be noted, however, that nitrite can accumulate in suspended growth cultures, particularly during startup or following severe temperature changes when the two bacterial populations are not balanced.[1,30] Thus, the model is really only appropriate for bioreactors at steady-state or for those receiving dynamic loads no more severe than the diurnal flows and concentrations normally entering domestic wastewater treatment systems. Other models consider both steps

in nitrification, although they are more complicated.[33] The second important characteristic of the process rate expression for nitrification is that no consideration is given to substrate and product inhibition, which are known to occur at high nitrogen concentrations, as discussed in Section 3.2.10. These factors were not considered because of a lack of adequate kinetic relationships, as discussed earlier, and because they do not normally occur at the nitrogen levels commonly found in domestic wastewater. Consequently, ASM No. 1 should not be applied to simulate the treatment of wastewaters containing nitrogen concentrations greatly in excess of those levels.

Loss of heterotrophic biomass, process 4, is modeled by the lysis:regrowth approach discussed in Section 3.3.2. A primary reason for adopting this approach is that no use of electron acceptor is directly associated with it, thereby making it easier to express the effects of alternative electron acceptors in the overall model. The loss of heterotrophic biomass is thought to continue at the same rate, regardless of the electron acceptor available,[12] and is modeled by Eq. 3.63. Similarly, the formation of biomass debris, slowly biodegradable substrate, and particulate biodegradable organic nitrogen are modeled by Eqs. 3.64–3.66, respectively. The nature of the electron acceptor will influence the rates of utilization of these constituents, however, as reflected in the other process rate expressions.

Process 5, loss of autotrophic biomass, is also modeled by the lysis:regrowth approach, although the amount of autotroph regrowth is not really significant, as discussed previously. Rather, heterotrophs grow on the organic substrate resulting from death and lysis of the autotrophic bacteria. As a consequence, the magnitude of the loss coefficient for autotrophs is the same as that for the traditional decay approach.

As nitrogen containing organic compounds undergo biodegradation, the nitrogen in them is released as ammonia, as discussed in Section 3.6. This release is reflected in process 6, which is modeled with Eqs. 3.79 and 3.80. These expressions are approximate because of a lack of information about them, as discussed in Section 3.6. They should be satisfactory, however, within the constraints already established for the model.

An important contribution of ASM No. 1 is consideration of the fate of particulate and other slowly biodegradable substrate, as reflected in process 7, which is hydrolysis. Although the fate of such material in suspended growth cultures is important, relatively little research has been done from which a rate expression can be developed (see Section 3.5). Nevertheless, based on the limited literature available, ASM No. 1 uses Eq. 3.77 as the basic rate expression. Comparison of it to the expression in Table 6.1, however, reveals that it was extended to include the effects of the electron acceptor. First, it will be noted that another correction factor, η_h, is included to reflect a retardation of hydrolysis under anoxic conditions. Like η_g, this correction factor is empirical and the rationale for its use is the same. Second, the rate of hydrolysis is assumed to go to zero in the total absence of oxygen or nitrate. Although hydrolysis is known to occur in anaerobic bioreactors (see Section 2.3.2), adaptation of facultative bacteria is required and evidently does not occur when a predominately aerobic or anoxic culture is subjected to short periods of anaerobiosis, because hydrolysis stops under such conditions.[34] Because death and lysis are thought to continue at the same rate regardless of the nature of the electron acceptor, there will be an accumulation of slowly biodegradable substrate when suspended growth

cultures are subjected to short periods without either oxygen or nitrate. In spite of the lack of certainty associated with the rate expression for process 7, the patterns of oxygen and nitrate-N utilization predicted by ASM No. 1 have been found to mimic well the performance of both pilot-[12,13] and full-scale[3] suspended growth systems with a number of configurations. Thus, although additional research is needed on this important topic, use of the model depicted in Table 6.1 should be satisfactory for the purposes intended herein.

The final process in Table 6.1 is conversion of particulate, biodegradable organic nitrogen, X_{NS}, into soluble, biodegradable organic nitrogen, S_{NS}. This rate is assumed to be proportional to the rate of hydrolysis of slowly biodegradable organic matter, as is modeled with Eq. 3.78.

Two events discussed in Chapters 2 and 3 are not included in Table 6.1: (1) soluble microbial product formation, and (2) phosphorus uptake and release. The impact of soluble microbial product formation is minor, and acts primarily to raise the concentration of soluble organic matter in the effluent from a bioreactor, as discussed previously. It was excluded for the same reason it was excluded from the simple model in Chapter 5. Phosphorus uptake and release will occur only when anaerobic zones are included in systems to allow a selective advantage for phosphate accumulating organisms (PAOs), as discussed previously. As seen above, however, some of the rate expressions in Table 6.1 have questionable validity under anaerobic conditions. For this reason, and because a model for the growth of only PAOs requires a matrix larger than the one in Table 6.1,[11,19,33,36,37] this process was not included here. IAWQ ASM No. 2 utilizes an expanded matrix to incorporate phosphorus removal by PAOs.[19] We use it in Chapter 7 to see how reactor conditions affect phosphorus removal.

6.1.3 Representative Parameter Values

The model depicted in Table 6.1 contains a large number of kinetic and stoichiometric parameters which must be evaluated for use in simulations. Techniques for conducting those evaluations are discussed in Chapter 8. Although the model should be calibrated for each situation under study, it is acceptable to use typical parameter values to demonstrate fundamental principles concerning suspended growth cultures, provided the reader recognizes that the conclusions are general and not directly applicable to any specific situation. Typical parameter values for domestic sewage at neutral pH and 20°C were compiled for ASM No. 1.[17,18] Consideration of those values, as well as the values given in Chapter 3, has resulted in the list given in Table 6.3. They will be used here and in Chapter 7 to demonstrate several things about suspended growth cultures that could not be demonstrated with the simple model in Chapter 5.

6.1.4 ASM No. 2

Activated Sludge Model No. 2 incorporates all of the events included in ASM No. 1, plus biological phosphorus removal. The latter is very complex[11,33,36,37] and a large number of components must be included to model it adequately, as seen in Section 2.4.6 and Section 3.7. Consequently, ASM No. 2 is much larger than ASM No. 1

Table 6.3 Typical Parameter Values at Neutral pH and 20°C for Domestic Wastewater

Symbol	Units	Value
	Stoichiometric coefficients	
Y_H	mg biomass COD formed/mg COD removed	0.60
f'_D	mg debris COD/mg biomass COD	0.08
$i_{N\,XB}$	mg N/mg COD in active biomass	0.086
$i_{N\,XD}$	mg N/mg COD in biomass debris	0.06
Y_A	mg biomass COD formed/mg N oxidized	0.24
	Kinetic parameters	
$\hat{\mu}_H$	hr^{-1}	0.25
K_S	mg/L as COD	20
$K_{O,H}$	mg/L as O_2	0.10
K_{NO}	mg/L as N	0.20
$b_{L,H}$	hr^{-1}	0.017
η_g	dimensionless	0.8
η_h	dimensionless	0.4
k_a	L/(mg biomass COD·hr)	0.0067
k_h	mg COD/(mg biomass COD·hr)	0.092
K_X	mg COD/mg biomass COD	0.15
$\hat{\mu}_A$	hr^{-1}	0.032
K_{NH}	mg/L as N	1.0
$K_{O,A}$	mg/L as O_2	0.75
$b_{L,A}$	hr^{-1}	0.004

and considerably more complex, including 19 components and 19 process rate equations that require 22 stoichiometric coefficients and 42 kinetic parameters.[19] Because of its size, ASM No. 2 will not be described in detail herein. However, because we use it in Chapter 7, some of its major characteristics will be presented.[19]

Some processes that were explicitly modeled in ASM No. 1 were simplified in ASM No. 2 in order to minimize its size. For example, processes 6 and 8 in Table 6.1, ammonification of soluble organic nitrogen and hydrolysis of particulate organic nitrogen, were eliminated. Their functions were made implicit by assuming that they occurred in stoichiometric proportion to soluble substrate removal and hydrolysis of slowly biodegradable organic matter. This accomplished the same thing as ASM No. 1, but with fewer process rate expressions. Organic phosphorus conversion to soluble phosphate was handled in a similar manner.

The events occurring under anaerobic conditions are quite different in the two models. ASM No. 1 assumed that growth and hydrolysis stopped under anaerobic conditions, although microbial death and lysis continued. This was adequate for the processes ASM No. 1 depicts, but is entirely inadequate for biological phosphorus removal. Consequently, ASM No. 2 includes fermentation, uptake of acetate for formation of PHB and other polyhydroxyalkanoic acids (PHAs), and release of soluble phosphate from hydrolysis of polyphosphate. The inclusion of fermentation required the partitioning of readily biodegradable substrate into two components, readily fermentable substrate and fermentation products, represented by acetate. Ac-

etate is produced from readily fermentable substrate under anaerobic conditions and is taken up by the PAOs, as depicted by Eq. 3.82, forming PHB, as given by Eq. 3.83. Under anoxic conditions, i.e., when nitrate-N is present as an electron acceptor, fermentation decreases and the common heterotrophic biomass competes with the PAOs for acetate.

The scope of activities of the common heterotrophic bacteria was expanded. Under anaerobic conditions, they ferment readily fermentable substrate, producing acetate, but they cannot grow. They can only grow under aerobic and anoxic conditions, and can use both readily fermentable substrate and acetate for that purpose. Because heterotrophic growth cannot occur under anaerobic conditions, ASM No. 2 is not capable of modeling a totally anaerobic system. It can only mimic the performance of an anaerobic zone in a system with aerobic and anoxic zones.

Our knowledge of PAOs is still evolving. As a consequence, several simplifying assumptions were made in ASM No. 2 with respect to their growth. For example, they are assumed to grow only under aerobic conditions and can use only stored PHB as a growth substrate, as indicated in Eq. 3.85. They are assumed to be unable to use nitrate-N as an electron acceptor or to use any other electron donor, either stored in the cell or in the medium. These are reasonable assumptions, but exceptions to them are known to exist.[23,32] Thus, as we gain additional knowledge, it is very likely that ASM No. 2 will undergo revision. Nevertheless, even in its initial form it is a very powerful tool that allows engineers to explore the complex microbial events occurring in biological phosphorus removal systems.

6.1.5 Application of International Association on Water Quality Activated Sludge Models

Activated sludge models No. 1 and No. 2 are considerably more complex than the one used in Chapter 5 (Table 5.1). As a consequence, it is impossible to attain analytical solutions for the concentrations of the various constituents in a bioreactor, as was done in Chapter 5. Rather, matrix solutions and numerical techniques must be used, depending on the complexity of the system under study. Several organizations have developed computer codes for solving the simultaneous mass balance equations for the constituents in the models, allowing their application to a variety of bioreactor configurations. One such code is SSSP,[7] which was developed for implementation of ASM No. 1 on microcomputers. It is menu driven and may be used for both steady-state and dynamic simulations. It was used to perform the simulations for single CSTRs in this chapter and for multiple bioreactor systems in Chapter 7. Another is ASIM,[15] which implements both models, as well as several others. It was used for the some of the simulations in Chapter 7. Table 6.4 lists several computer codes that are available for using both IAWQ activated sludge models. In addition, Dold[10] has developed a code that extends ASM No. 1 to include phosphorus removal, but it differs somewhat from ASM No. 2.

6.2 EFFECT OF PARTICULATE SUBSTRATE

A major limitation of the model presented in Chapter 5 is that it does not consider the biodegradation of particulate organic material, which is an important class of

Table 6.4 Computer Codes Implementing IAWQ Activated Sludge Models

Code name	Features	Contact information
SSSP	Implements Model No. 1	C. P. Leslie Grady Jr., Environmental Systems Engineering, Rich Environmental Research Lab, Clemson University, Clemson, SC 29634-0919 USA
EFOR	Implements Model No. 1 plus a clarifier model	Jan Peterson, I. Kruger AS, Gladsaxevej 363, DK-2860 Soborg, Denmark
ASIM	A flexible modeling tool that implements both Model No. 1 and No. 2, as well as several others	Willi Gujer, EAWAG, Swiss Federal Institute for Environ. Science and Technology, CH-8600 Dübendorf, Switzerland
GPS-X	A general purpose simulator that implements both Model No. 1 and No. 2 plus other unit operations	Hydromantis, Inc., 1685 Main St. West, Suite 302, Hamilton, Ontario L8S 1G5 Canada
SBRSIM	Implements Model No. 1 for a sequencing batch reactor	Jürgen Oles, Technical University Hamburg-Harburg, Eissendorfer Strasse 42, 2100 Hamburg 90, Germany

organic substrate in many wastewaters. Thus a suitable application of the model in Table 6.1 would be to see how the nature of the substrate influences the performance of a single CSTR, such as that depicted in Figure 5.1. To do this, two situations were considered, one in which all influent organic matter was soluble, and the other in which it was all particulate. The total concentration was the same in both cases, 500 mg/L as COD, as was the flow rate, 1000 m^3/day (1.0 m^3 = 1000 L), giving a total COD mass input rate of 500 kg/day. The solids retention time (SRT)/hydraulic residence time (HRT) ratio was held constant at 20 while the SRT was varied. In other words, the reactor volume was increased in proportion to the increase in SRT. This was done to make it easier to visualize the fate of particulate material, as well as the relative importance of growth and decay as SRT is changed. When the SRT/HRT ratio is held constant, if particulate material does not undergo reaction it has a constant concentration in the system, regardless of the SRT. Furthermore, as growth occurs on particulate substrate, the concentration of suspended matter (particulate substrate plus biomass) will decrease because the yield is less than one. In addition, growth associated with increased soluble substrate removal in the bioreactor is reflected by an increase in biomass concentration, whereas an increase in the importance of decay is reflected by a decrease. The biomass separator was assumed to be perfect so that it removed all undegraded particulate substrate from the effluent and returned it to the bioreactor. Thus, undegraded particulate substrate was removed only through the wastage flow. The parameter values used to describe the reactions are those given in Table 6.3, with the exception of $\hat{\mu}_A$, the maximum specific growth rate for autotrophic bacteria, which was set equal to zero to eliminate them from

Table 6.5 Process Kinetics and Stoichiometry Adopted for Considering the Impact of Particulate Substrate on the Performance of an Aerobic CSTR. Lysis:Regrowth Model for Biomass Loss

	Component[a]					
Process	X_S	$X_{B,H}$	X_D	S_S	S_O[b]	Process rate, r_j
Growth		1		$-\frac{1}{Y_H}$	$\left(\frac{1 - Y_H}{Y_H}\right)$	$\hat{\mu}_H \left(\frac{S_S}{K_S + S_S}\right)\left(\frac{S_O}{K_{O,H} + S_O}\right) K_{B,H}$
Death and lysis	$(1 - f'_D)$	-1	f'_D			$b_{L,H} \cdot X_{B,H}$
Hydrolysis	-1			1		$k_h \left(\frac{X_S/X_{B,H}}{K_X + (X_S/X_{B,H})}\right)\left(\frac{S_O}{K_{O,H} + S_O}\right) X_{B,H}$

[a]All components and coefficients are expressed as COD.
[b]Coefficient must be multiplied by -1 to express as oxygen.

consideration and limit the reactions to those associated with heterotrophs, as was done in Chapter 5. The influent contained sufficient ammonia-N for heterotrophic growth, but no nitrate-N, thereby eliminating nitrate-N, or reactions associated with it, from consideration as well. These selections simplified the model to include only processes 1, 4, and 7 and components 2, 3, 5, 7, and 8, as shown in Table 6.5. Finally, the dissolved oxygen concentration was held constant at 4.0 mg/L, which together with the $K_{O,H}$ value in Table 6.3, made the oxygen term in the rate expressions approach 1.0.

6.2.1 Steady-State Performance

Figure 6.1 shows the effect of SRT on the mixed liquor suspended solids (MLSS) concentration, the active fraction, and the oxygen requirement for bioreactors receiving the two types of substrate at a constant flow and concentration. The MLSS in the bioreactor receiving soluble substrate is composed of heterotrophic biomass and biomass debris, whereas the MLSS in the bioreactor receiving particulate substrate contains heterotrophic biomass, biomass debris, and unreacted particulate substrate, with the relative quantities depending on the SRT. The curves for the soluble feed are similar to those in Chapter 5 and can serve as a reference point with which to see the effect of particulate feed. The most obvious effect is that a longer SRT is required to get biomass growth on particulate substrate. As indicated by the oxygen requirement, washout occurs at an SRT of about 4.5 hr when the substrate is soluble, but at an SRT of about 26 hr when it is particulate. This reflects the fact that hydrolysis reactions are slow. At SRTs below 26 hr, nothing happens to the particulate substrate and it acts like inert material, giving a MLSS concentration equal to the influent concentration times the SRT/HRT ratio ($500 \times 20 = 10{,}000$) and an active fraction of zero, i.e., the MLSS is composed entirely of unreacted particulate substrate. As the SRT is increased past 26 hr, degradation of the particulate substrate begins, causing the MLSS concentration to drop and the active fraction and oxygen utilization to increase. In this region, the MLSS is composed of unreacted particulate substrate, active heterotrophic biomass, and some biomass debris. Eventually, at longer SRTs, bioreactor performance becomes independent of the feed type and is essentially the same for each. This occurs when both substrates are almost completely degraded so that system performance is governed primarily by biomass death and lysis, and the MLSS in each bioreactor is composed primarily of heterotrophic biomass and biomass debris.

Two significant points arise from Figure 6.1. The first is that long SRTs are required to achieve substantial degradation of particulate substrates. The second is that use of the process loading factor with a particulate substrate can be confusing. It will be recalled from Eq. 5.39 that the active fraction must be known before the process loading factor can be related to the specific growth rate of the biomass. Figure 6.1b, however, shows that the active fraction varies in a complex manner as the SRT is changed when the substrate is particulate. This is because the active fraction is the active biomass concentration divided by the MLSS concentration, which includes the undegraded particulate substrate. The SRT, on the other hand, is still related to the biomass specific growth rate by Eq. 5.12, and thus is more descriptive of process performance. Because most wastewaters contain some particulate

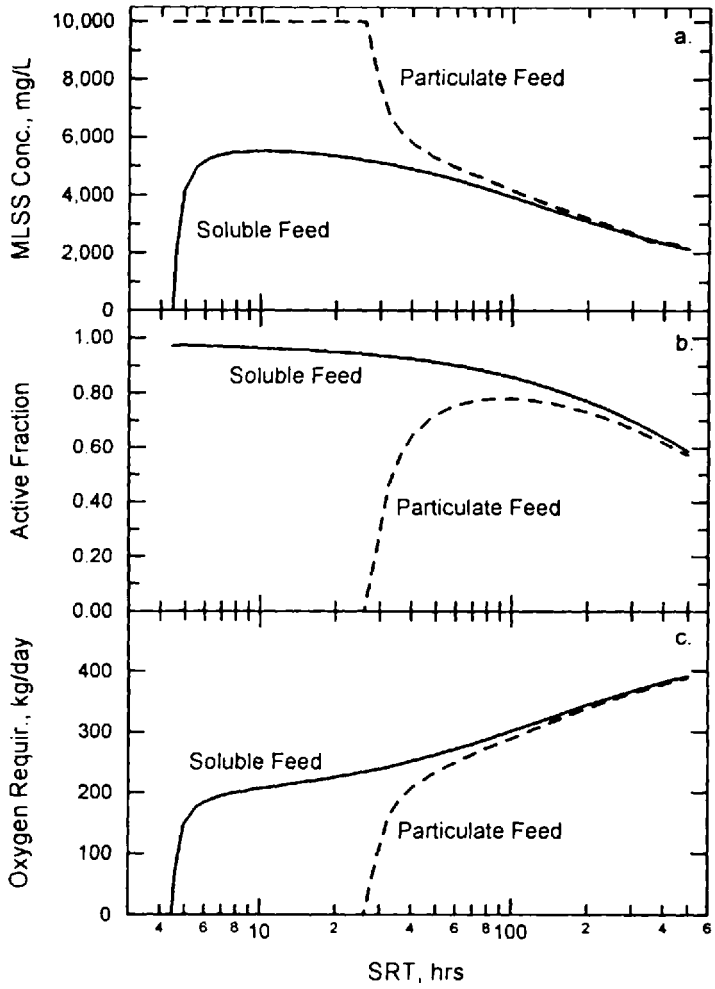


Figure 6.1 Effect of the nature of the organic matter in the feed on the performance of a CSTR with $\Theta_c/\tau = 20$. Influent biodegradable COD = 500 mg/L in each case. Flow = 1000 m^3/day. Particulate substrate was assumed to be removed by the biomass separator and retained in the reactor. Parameter values are listed in Table 6.3. $\hat{\mu}_A$ was set equal to zero.

substrate, SRT is preferable to process loading factor as a basic design and operational parameter for suspended growth bioreactors.

6.2.2 Dynamic Performance

So far we have only considered the steady-state performance of a CSTR, i.e., the performance that results when a bioreactor receives a constant influent flow at a constant concentration. Most wastewaters are subject to time dependent variability, however, and thus it would be beneficial to investigate the impact of the nature of the substrate under those conditions.

Because of variations in human activities, municipal wastewater treatment systems experience diurnal variations in the flow and concentration of the wastewater entering them. Figure 6.2 shows typical variations experienced over a 24 hr period, beginning at midnight as time 0. The patterns correspond to those observed at a large municipal plant in South Africa over a period of one week,[13] but the values have been normalized to a daily average flow of 100 m^3/day and a flow-weighted average concentration of 100 mg/L. The patterns are also typical of those experienced in the United States,[9] and will be adopted herein for demonstration purposes. As with kinetic and stoichiometric parameters, however, the necessity for determining the actual variations associated with a given wastewater cannot be overemphasized.

To determine the effect of the type of substrate on the dynamic response of a CSTR, the bioreactor in Figure 5.1 was subjected to the variations in flow and concentration shown in Figure 6.2. The flow values were adjusted to give a daily average flow of 1000 m^3/day and the biodegradable COD was adjusted to give a flow-weighted daily average concentration of 265 mg/L, a value commonly seen in United States domestic wastewater.[25] The SRT was set at 240 hr, a value sufficient to make the steady-state performance for the two substrate types essentially the same, as shown in Figure 6.1, while the HRT based on the daily average flow rate was set at 6 hr. As with the steady-state response, two situations were considered, one in which the organic matter was entirely soluble and one in which it was entirely particulate. All other constituents were assumed to be constant at concentrations sufficient to not limit the reactions. As before, the matrix in Table 6.5 described the system.

The response of the bioreactor is shown in Figure 6.3. Before considering the effect of the type of substrate on the bioreactor performance, we will examine only the soluble substrate case in order to understand why the bioreactor behaves as it does. Examination of the figure shows that the soluble substrate concentration has

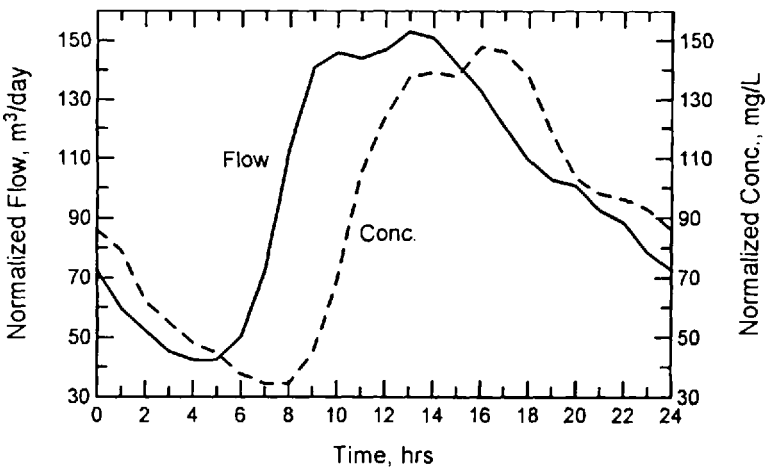


Figure 6.2 Typical diurnal patterns of wastewater flow and concentration for a community with little night time activity (after Dold and Marais[12]). The flow has been normalized to an average of 100 m^3/day. The concentration has been normalized to give a flow weighted average of 100 mg/L.

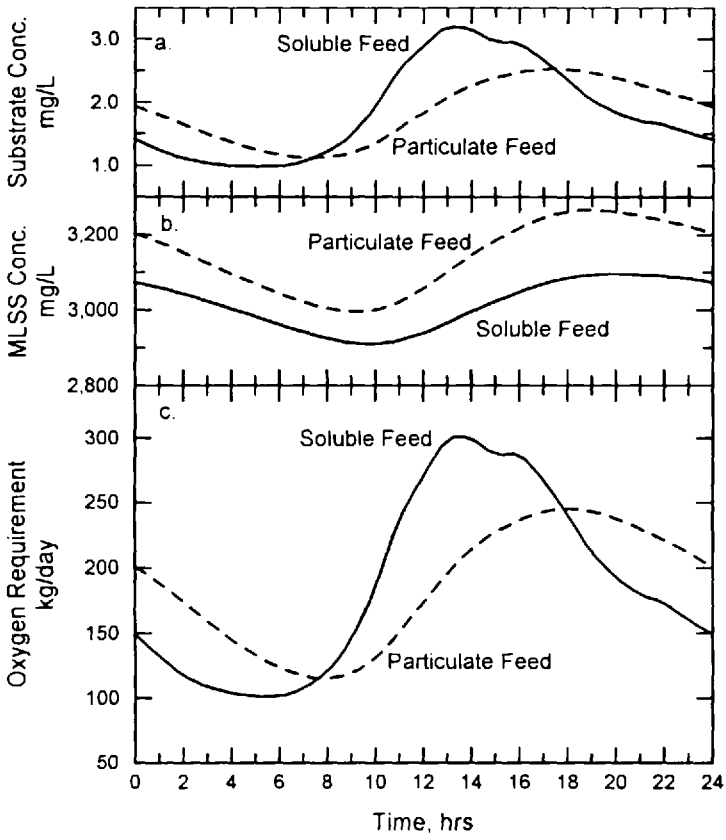


Figure 6.3 Effect of the nature of the organic matter in the feed on the response of a CSTR to the diurnal flow pattern in Figure 6.2. Flow weighted average influent biodegradable COD = 265 mg/L in each case. Daily average flow = 1000 m^3/day; SRT = 240 hrs; average HRT = 6 hrs. Parameter values are listed in Table 6.3. $\hat{\mu}_A$ was set equal to zero.

greater relative variations throughout the day than does the MLSS concentration. This is a direct result of the fact that the residence time of the MLSS in the bioreactor (the SRT) is much longer than the residence time of the soluble substrate (the HRT). As a result, variations in the MLSS concentration are dampened. In fact, the mass of MLSS in the bioreactor is almost constant throughout the day.

If we consider the mass of MLSS to be approximately constant throughout the day, we can then see why the soluble substrate concentration varies. Examination of Figure 6.2 reveals that the mass of substrate entering the bioreactor per unit time (flow × concentration) varies throughout the day. This means that the mass of substrate available to a unit mass of microorganisms also varies throughout the day. However, the rate at which a microorganism can remove substrate is controlled by the concentration of substrate surrounding it. Thus, as the mass flow rate of substrate into the bioreactor increases, the substrate concentration must rise to allow the microorganisms to remove substrate faster. Conversely, as the mass flow rate of substrate decreases, the microorganisms will drive the substrate concentration lower until

the rate of substrate removal is decreased to be consistent with the rate of input. Thus, the variation in substrate concentration is a direct consequence of the necessity for the microorganisms to vary their activity in response to the changing input rate of substrate. The variation in the oxygen requirement directly reflects that variation in activity.

The soluble substrate curves in Figure 6.3a also demonstrate an important point about the growth characteristics of the biomass in a CSTR receiving a time varying input; the specific growth rate is not constant over time. It will be recalled that the specific growth rate is controlled by the soluble substrate concentration, as expressed by the Monod equation (Eq. 3.36). Since the soluble substrate concentration is varying over time, so is the specific growth rate. This means that the bacteria are in a continually changing state. For the reactor configuration in Figure 5.1, the SRT is determined solely by the reactor volume and the wastage flow rate. Consequently, the SRT can be held constant, even though the specific growth rate of the bacteria is varying. In other words, the specific growth rate of the microorganisms in a CSTR that is not at steady-state is not fixed by the SRT. This can also be seen by performing a mass balance on heterotrophic biomass in the reactor. Such an exercise using the kinetics and stoichiometry in Table 6.5 reveals:

$$\mu_H = \frac{1}{\Theta_c} + b_{L,H} + \frac{1}{X_{B,H}} \frac{dX_{B,H}}{dt} \tag{6.1}$$

Since the heterotrophic biomass concentration varies over time in response to the changing input, so will the specific growth rate. It is important to recognize that the constant relationship between SRT and specific growth rate depicted in Eq. 5.12 is only valid for steady-state conditions.

Even though the specific growth rate is not determined solely by the SRT for a non–steady-state CSTR, the SRT is still a good indicator of average performance, with longer SRTs giving lower average substrate concentrations. Nevertheless, the average substrate concentration leaving a CSTR receiving a dynamic input will always be greater than the concentration leaving a steady-state CSTR. For the conditions imposed in Figure 6.3, the flow-weighted average output concentration for the soluble substrate case in 2.11 mg/L as opposed to 1.80 mg/L for a steady-state CSTR with the same SRT. The higher average substrate concentration results from the nonlinear nature of the Monod equation (Eq. 3.36) describing the relationship between substrate concentration and microbial activity. This is one reason why it is advantageous to practice equalization prior to a biochemical reactor.

Now consider the impact of the type of substrate on the dynamic behavior of the CSTR. Figure 6.3a shows the effluent soluble substrate concentrations resulting from the two feed types. When the influent feed is all particulate, soluble substrate arises from hydrolysis of the particulate substrate, which is a slow reaction. Thus, the response is dampened and the flow-weighted average concentration is lower (1.92 mg/L vs. 2.11 mg/L). Figure 6.3b shows that there will be little difference in the MLSS concentration or its variation for the two substrate types. The slightly higher concentration in the bioreactor receiving particulate substrate is because of a slight build up of that substrate in the system caused by the slow hydrolysis reactions, but the effect is small. Generally, one would not expect to be able to distinguish much difference in the amount of MLSS in the two systems.

The major impact of particulate substrate on the dynamic response of a CSTR is in the utilization of oxygen, as shown in Figure 6.3c. As might be expected from the previous discussion of the slow nature of hydrolysis, the impact of the presence of particulate substrate is to dampen the system response, thereby reducing the peak oxygen requirement. In addition, the need for hydrolysis to make substrate available causes a time lag in the occurrence of the maximum and minimum oxygen consumption rates. Examination of Figure 6.2 reveals that the minimum mass input rate occurs at about 6 hr and the maximum at about 13 hr, which correspond closely to the times of the minimum and maximum oxygen requirements in the bioreactor receiving soluble substrate, thereby demonstrating the rapidity with which biomass can respond to soluble substrate. In contrast, the need for hydrolysis of particulate substrate, in combination with the fact that its concentration does not vary rapidly because its residence time in the system is the SRT, delays the minimum response by about 3 hr and the maximum by about 5 hr. Since both systems use about the same amount of oxygen in a 24 hr period, this suggests that consideration must be given to the physical state of the substrate during design of the oxygen transfer system for a suspended growth bioreactor.

6.3 NITRIFICATION AND ITS IMPACTS

We see in Section 3.2.7 that the kinetics of growth of autotrophic nitrifying bacteria can be represented in the same manner as that of heterotrophic bacteria. Consequently, the general conclusions derived in Chapter 5 about biomass growth in CSTRs are equally applicable to them. We also see in Section 3.2.10 that the values of their kinetic coefficients are quite different from those for heterotrophs. This means that the specifics of their behavior in a given reactor environment will differ somewhat from that of the heterotrophs. In this section, we will investigate some of the characteristics of nitrifying bacteria that require special recognition and see how autotrophic and heterotrophic bacteria might influence one another when grown in the same bioreactor.

6.3.1 Special Characteristics of Nitrifying Bacteria

Comparison of the typical $\hat{\mu}$ values for heterotrophic and autotrophic bacteria in Table 6.3 shows that the value for autotrophs is almost an order of magnitude lower than that for heterotrophs, suggesting that the minimum SRT for nitrifying bacteria is almost an order of magnitude larger. As a consequence, they can be lost from bioreactors under conditions that allow heterotrophic bacteria to grow freely. This situation is aggravated by the fact that nitrifying bacteria are more sensitive to low temperatures and low dissolved oxygen concentrations, as will be discussed later in this section. Therefore, special consideration must be given to the choice of the SRT in systems containing autotrophic bacteria and it cannot be assumed that conditions suitable for the removal of soluble organic matter are suitable for the conversion of ammonia-N to nitrate-N.

Another characteristic of nitrifying bacteria is that their half-saturation coefficient is very low; the typical value given in Table 6.3 is 1.0 mg/L as N. It will be recalled that the half-saturation coefficient is the substrate concentration allowing

bacteria to grow at half of their maximal rate. This means that CSTRs containing autotrophic bacteria will have low ammonia-N concentrations even when the bacteria are growing relatively fast and that very low concentrations will result whenever the SRT is large enough to ensure stable growth. It also means that the ammonia-N concentration will rise rapidly as the SRT is decreased to the point of washout. As a consequence, nitrification has gained the reputation of being an all-or-none phenomenon. In other words, since the nitrogen concentration entering municipal wastewater treatment systems is on the order of 30 to 40 mg/L, the percent nitrification approaches 100 whenever the SRT is long enough to give stable growth and rapidly falls to zero as washout occurs. This is illustrated in Figure 6.4,[27] which includes data from the literature as well as the results of steady-state simulations with a model similar to the one employed herein.

There are some occasions in which the all-or-none phenomenon will not occur or will not be as drastic as that shown in Figure 6.4. One is when the bioreactor is subjected to diurnal loading, like that shown in Figure 6.2. It will be recalled from Section 6.2.2 that the flow-weighted average effluent substrate concentration from a CSTR receiving a diurnal input is not as low as the concentration from the same reactor receiving a constant input. The degree of difference between the two responses depends on the bioreactor SRT, as shown in Figure 6.5,[27] which was prepared by simulation with a model similar to the one presented herein. It also shows that complete nitrification can still be achieved in a CSTR receiving a diurnal input, but that longer SRTs are required. Another situation that can lead to incomplete nitrification is when the influent contains such a high ammonia-N concentration that both substrate and product inhibition can occur.[2,29] In that case the attainment of complete nitrification requires the use of multireactor systems in which denitrification can be used to reduce the product concentration, i.e., nitrite and nitrate. As pointed out earlier, substrate and product inhibition effects cannot yet be reliably modeled. Con-

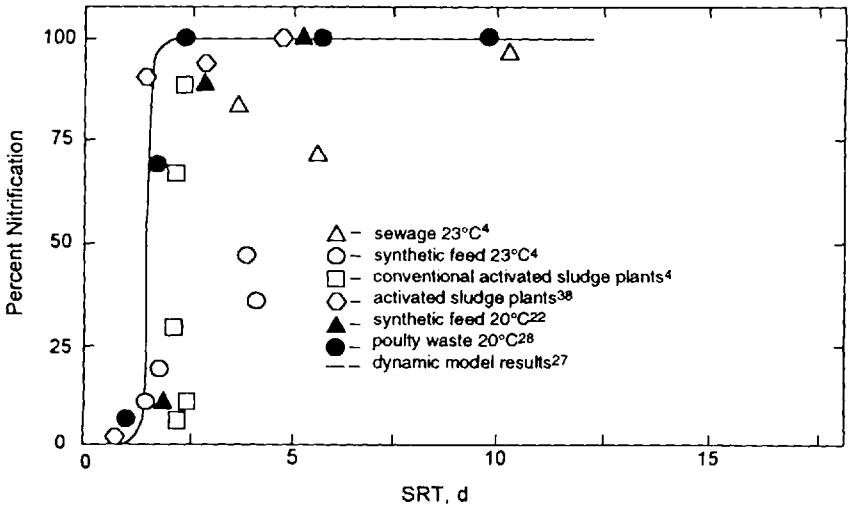


Figure 6.4 Effect of SRT on the steady state nitrification performance of a CSTR. The reference numbers refer to the sources of the data. (Adapted from Poduska and Andrews.[27])

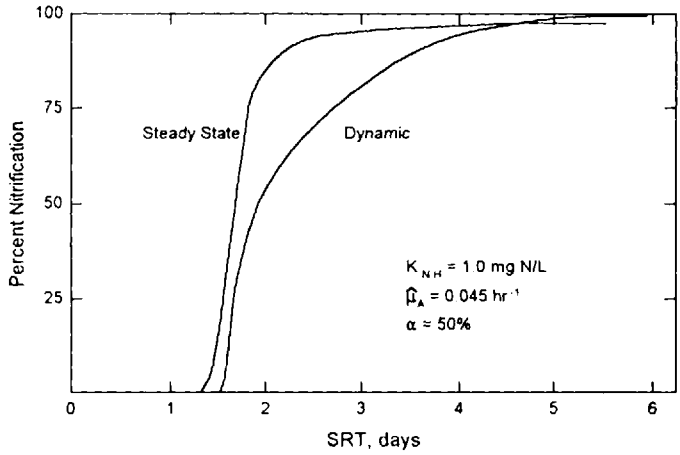


Figure 6.5 Comparison of steady state and dynamic performance of nitrification in a CSTR. (Adapted from Poduska and Andrews.[7])

sequently, pilot studies should always be run on wastes containing ammonia-N concentrations substantially higher than those found in domestic wastewater. A final circumstance in which a curve of a different shape may be obtained is when the wastewater contains chemicals inhibitory to the autotrophic bacteria. Such a curve is shown in Figure 6.6.[5] Inhibitory compounds can reduce $\hat{\mu}_A$ and increase K_{NH}, both of which reduce the percent nitrification achieved at a given SRT, making the curve

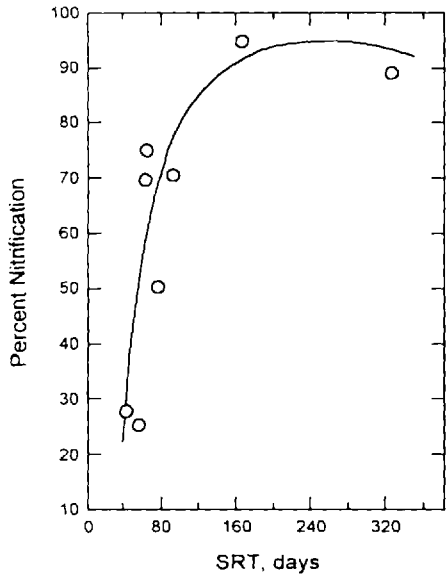


Figure 6.6 Effect of SRT on the performance of nitrification in a CSTR receiving a wastewater containing inhibitory compounds. (Adapted from Bridle et al.[5])

approach 100% more gradually. Both effects are illustrated in Figure 6.6, which was obtained with an industrial waste.[8]

Another important characteristic of nitrifying bacteria is their extreme sensitivity to the dissolved oxygen concentration, which necessitated the inclusion of a term for oxygen in the rate expression in Table 6.1. Examination of the values of K_O for autotrophs and heterotrophs in Table 6.3 shows that $K_{O,A}$ is much larger than $K_{O,H}$. This means that as the dissolved oxygen concentration is decreased, the term $S_O/(K_O + S_O)$ in the rate expression for autotrophs will become small more rapidly than the corresponding term in the rate expression for heterotrophs. Consequently, autotrophs will be affected by decreases in the oxygen concentration much more drastically than will heterotrophs. The importance of dissolved oxygen concentration to the growth of autotrophs is illustrated in Figure 6.7 where the minimum SRT required for their growth is plotted as a function of the oxygen concentration. The curve was generated with Eq. 5.16, with adjustment of $\hat{\mu}_A$ for the effects of oxygen as given by the rate expression in Table 6.1. The parameter values from Table 6.3 were used, as indicated in the figure. Examination of the figure shows that oxygen concentrations above 2.0 mg/L have little effect on the minimum SRT, and it is seldom necessary to maintain the concentration in excess of that value to get satisfactory nitrification. However, oxygen concentrations below 2.0 mg/L begin to have a strong effect and those below 0.5 mg/L have a drastic effect. A low dissolved oxygen concentration also diminishes the percent nitrification that can ultimately be achieved. Consequently, care in the specification of the oxygen transfer system is an important component of the design of a suspended growth bioreactor in which nitrification is to occur.

Nitrifying bacteria are also very sensitive to temperature, as reflected by the temperature coefficients in Section 3.9.2. Because $\hat{\mu}_A$ is small even at 20°C, low temperatures require bioreactors to have very long SRTs in order for nitrifying bacteria to grow. This is illustrated in Figure 6.8 which was generated with Eq. 5.16

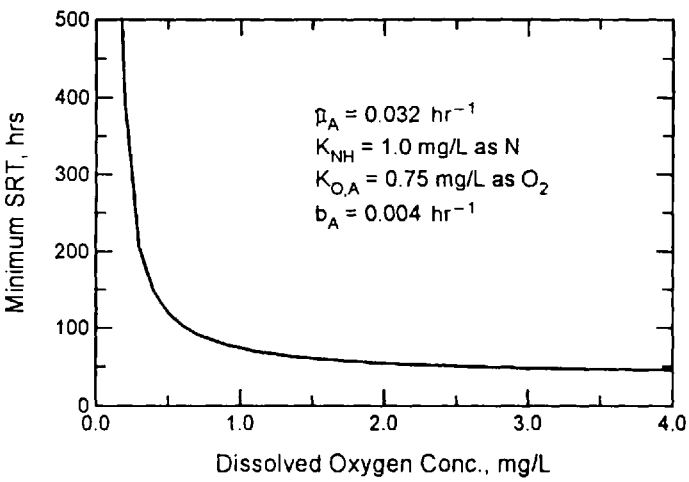


Figure 6.7 Effect of dissolved oxygen concentration on the minimum SRT required for nitrification in a CSTR receiving an ammonia-N concentration of 30 mg/L. Parameter values are for a temperature of 20°C.

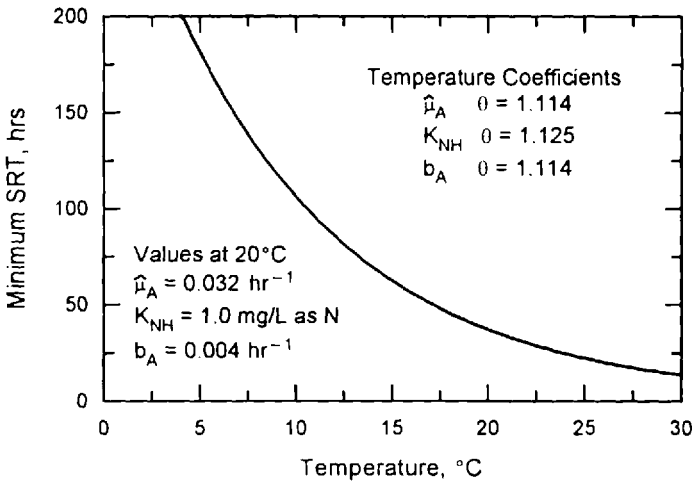


Figure 6.8 Effect of temperature on the minimum SRT required for nitrification in a CSTR receiving an ammonia-N concentration of 30 mg/L. The dissolved oxygen concentration is not limiting.

using kinetic parameters corrected for temperature with Eq. 3.95 and the coefficients shown in the figure. The dissolved oxygen concentration was assumed to be high enough to have no effect. Examination of the figure reveals that very large increases in the SRT are required to compensate for drops in temperature. This suggests that the choice of SRT for a nitrifying bioreactor must be made for the lowest temperature expected.

The model in Table 6.1 includes alkalinity as a component because of the sensitivity of nitrifying bacteria to low pH and the effect that the oxidation of ammonia-N has on alkalinity, and hence on pH. Equations 3.28 and 3.29 showed that 8.64 g of HCO_3^- alkalinity are destroyed for each g of nitrogen oxidized from ammonia-N to nitrate-N, or 7.08 g of alkalinity (as $CaCO_3$) per g of N. Domestic wastewaters often contain between 30 and 40 mg/L of nitrogen that can be oxidized during nitrification. This means that almost 300 mg/L of alkalinity (as $CaCO_3$) can be destroyed. Since many wastewaters do not contain this much alkalinity, it is apparent why consideration must be given to pH control in nitrifying bioreactors. Equation 3.50 may be used to determine the effect of pH on $\hat{\mu}_A$, with its resulting impact on the minimum SRT required to achieve nitrification. Such an activity is left as an exercise for the reader.

6.3.2 Interactions Between Heterotrophs and Autotrophs

From the discussion of the preceding section, it is apparent that there are several interactions between heterotrophic and autotrophic bacteria growing together in a CSTR. The most important concerns the dissolved oxygen concentration, which is why it is explicitly included in the model in Table 6.1. Because heterotrophic bacteria have larger $\hat{\mu}$ and smaller K_O values than autotrophs, they are capable of surviving

in bioreactors with SRTs and oxygen concentrations that would cause autotrophic bacteria to wash out. In other words, the heterotrophs control the dissolved oxygen concentration if the supply rate is insufficient to meet the needs of both groups. As a result, organic substrate removal can occur unimpeded under conditions that will not allow nitrification. Looked at another way, it is possible for a designer to choose an SRT and dissolved oxygen concentration that will allow only organic substrate removal or a combination that will allow both nitrification and organic substrate removal, but it is impossible to choose conditions that will allow nitrification without oxidation of any organic matter present. Since the designer can choose the conditions to be imposed on the bioreactor, he/she has control over the events that will occur in it, consistent with the kinetic characteristics of the organisms involved.

Simultaneous growth of heterotrophic and autotrophic bacteria need not be detrimental to the nitrifiers; it can be beneficial. As discussed previously, nitrifying bacteria are sensitive to inhibition by some organic compounds. If those compounds are biodegradable, it is possible to choose the SRT so that their concentrations in the bioreactor are too low to inhibit the nitrifiers, thereby allowing them to grow along with the heterotrophs destroying the inhibitors.[21] It is not necessary to perform nitrification in a separate bioreactor following destruction of the organic matter.

As stated before, domestic wastewater in the United States often contains about 265 mg/L of biodegradable COD and around 40 mg/L of reduced nitrogen.[25] Comparison of the Y_H and Y_A values in Table 6.3 in light of these concentrations reveals that the amount of autotrophic biomass formed in a bioreactor receiving such a feed will be small relative to the amount of heterotrophic biomass formed. Furthermore, when consideration is given to the fact that the MLSS in a suspended growth bioreactor contains inert particulate organic matter as well as biomass debris, it is likely that the contribution of the autotrophic bacteria to the MLSS concentration will be very small. That this is indeed the case is illustrated in Figure 6.9a. The curves in this figure were generated with the model in Table 6.1 using as input a feed with the characteristics listed in Table 6.6, which are considered to be representative of a domestic wastewater in the United States following primary sedimentation.[7] The curve without nitrification was obtained by setting $\hat{\mu}_A$ equal to zero during the simulation.

Unlike their effect on the MLSS concentration, however, nitrifying bacteria have a major impact on the amount of oxygen required in a CSTR receiving a feed like domestic wastewater. This can be seen by considering the stoichiometry of microbial growth as reflected by the coefficients in Table 6.1. For the true growth yields in Table 6.3, the oxygen utilization associated with heterotrophic growth (exclusive of biomass death and lysis) is $1 - 0.60$, or 0.40 g O_2/g COD removed, whereas that associated with autotrophic growth is $4.57 - 0.24$, or 4.33 g O_2/g N oxidized. Death and lysis makes the utilization greater for the heterotrophs, perhaps around 0.65 g O_2/g COD removed for a typical SRT, but have little impact on the autotrophic oxygen utilization because the value of the true growth yield (0.24) is small relative to 4.57. Given the concentrations of the two substrates in the influent, these values suggest that the autotrophs require around 173 g of oxygen (40×4.33) for each liter of influent whereas the heterotrophs require around 172 g (265×0.65). Thus, even though the concentration of reduced nitrogen entering the bioreactor is much less than the concentration of biodegradable COD entering, the large change in oxidation state of nitrogen associated with the production of nitrate, in combina-

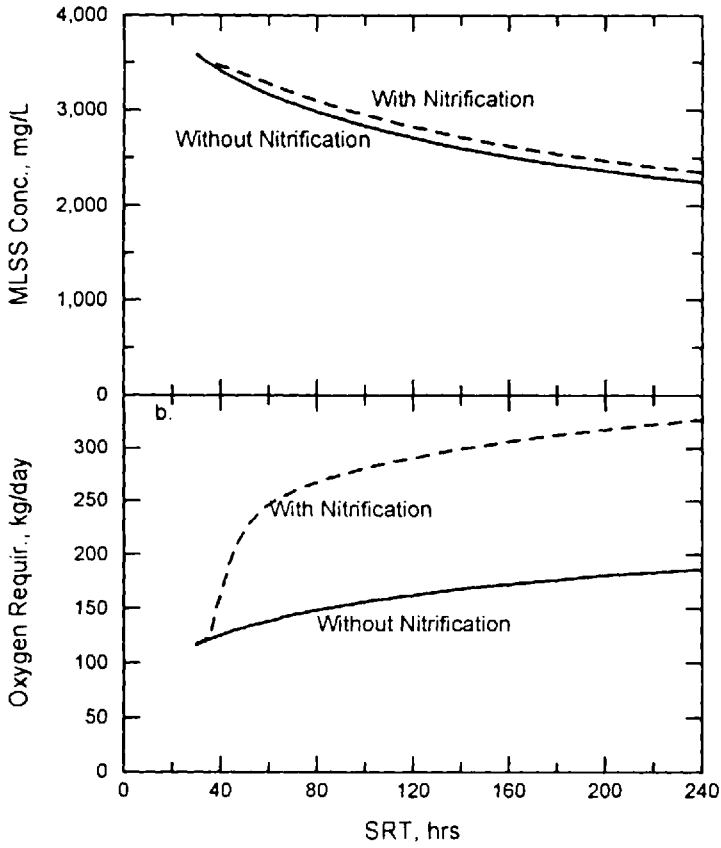


Figure 6.9 Effects of SRT and nitrification on the MLSS concentration and oxygen requirement in a CSTR with $\Theta_x/\tau = 20$. Parameter values are given in Table 6.3 and the influent conditions are given in Table 6.6. The value of $\hat{\mu}_A$ was set equal to zero to eliminate nitrification in one case. The dissolved oxygen concentration was fixed at 4.0 mg/L.

Table 6.6 Characteristics Considered to Be Representative of a United States Domestic Wastewater Following Primary Sedimentation[a]

Component	Concentration
Inert particulate organic matter	35.0 mg/L as COD
Slowly biodegradable substrate	150.0 mg/L as COD
Readily biodegradable substrate	115.0 mg/L as COD
Oxygen	0.0 mg/L as O_2
Soluble nitrate nitrogen	0.0 mg/L as N
Soluble ammonia nitrogen	25.0 mg/L as N
Soluble biodegradable organic nitrogen	6.5 mg/L as N
Particulate biodegradable organic nitrogen	8.5 mg/L as N
Alkalinity	5.0 mM/L

tion with the differences in biomass yield, means that similar amounts of oxygen are required by the two groups of microbes. This can be seen clearly in Figure 6.9b. Failure to consider this effect during bioreactor design will lead to an inability to transfer sufficient oxygen, with resulting impairment of nitrification performance, as discussed previously.

Because of the important effect of nitrification on the oxygen requirement in a bioreactor, there are circumstances in which it would be advantageous to be able to calculate it explicitly. This can be done by using the simple model in Chapter 5 with minor modification. For the simple decay model, an equation analogous to Eq. 5.33 can be used by recognizing that the factor 1.0 must be replaced with 4.57, as discussed in Section 6.1.1:

$$RO_A = F(S_{NHO} - S_{NH})\left[4.57 - \frac{(1 + f_D \cdot b_A \cdot \Theta_c)Y_A}{1 + b_A \cdot \Theta_c}\right] \tag{6.2}$$

where S_{NHO} represents the ammonia-N concentration in the feed that is available to the nitrifiers, i.e., the amount remaining after consideration of the amount used by the heterotrophs for biomass synthesis.

6.3.3 Effects of Nitrification in Bioreactors Receiving Only Biomass

In Section 5.4.2 and Figure 5.10, we investigate the performance of a CSTR receiving only biomass in the feed. Because the model in Chapter 5 considered only the growth of heterotrophic biomass, the discussion was limited to the fate of only heterotrophic bacteria under such circumstances. However, if the SRT of the bioreactor is sufficiently long, autotrophic bacteria will also grow, with an important impact on system performance. As expected from the preceding discussion, the major impacts of autotrophs on a bioreactor receiving only biomass in its feed are on the oxygen requirement and the pH. If f_D is 0.20, the destruction of 100 g of biomass COD will lead to the release of 20 g of biomass debris COD and the consumption of 80 g of oxygen. As reflected by $i_{N\,XB}$ and $i_{N\,XD}$, the nitrogen contents of biomass and biomass debris are around 0.086 and 0.06 g N/g biomass COD, respectively. Consequently, destruction of 100 g of biomass will also lead to the release of 7.4 g of ammonia-N (100[0.086−(0.20·0.06)]), which can act as substrate for autotrophic bacteria. If the oxygen requirement associated with oxidation of the ammonia-N is 4.33 g O_2/g N, growth of the autotrophs will increase the reactor's oxygen requirement by 40%. This can have a substantial impact on the amount of oxygen that must be supplied. Furthermore, oxidation of the nitrogen released by destruction of the biomass will result in the destruction of 52.4 g of alkalinity (as $CaCO_3$) for each 100 g of biomass COD destroyed. Since it is not unusual for the feedstream entering such a reactor to contain 4000 mg/L of biomass COD and for 50% of it to be destroyed, one can see that the occurrence of nitrification in the reactor will destroy substantial quantities of alkalinity. Consequently, it is often necessary to control pH in such systems. Failure to do so will create an unstable environment in which the pH oscillates, thereby hindering system performance.

6.4 DENITRIFICATION AND ITS IMPACTS

In the context of biochemical operations, anoxic growth of heterotrophic bacteria is simply an alternative mode of growth in response to the absence of oxygen and the presence of nitrate-N as the terminal electron acceptor. Because the resulting reduction of nitrate-N to N_2 removes nitrogen from the wastewater undergoing treatment, the process is also referred to as denitrification. Generally, it occurs when the appropriate conditions are purposely established in bioreactors, but as the rate expression in Table 6.1 suggests, it will occur any time the oxygen concentration is low, the nitrate concentration is high, and organic matter is present as an electron donor. This means that the potential for its occurrence exists in all biochemical reactors. It should be included as a possible reaction in any complete system model.

6.4.1 Characteristics of Denitrification

In Section 2.4.1, we see that four enzymes are involved in the reduction of nitrate-N to N_2: nitrate reductase, nitrite reductase, nitric oxide reductase, and nitrous oxide reductase. Nevertheless, just as nitrification was treated as a single step process in ASM No. 1, so is denitrification. This simplification will not seriously restrict the results of simulations and is necessary because insufficient information is available about the factors influencing each of the individual reaction rates to allow all of them to be included in a system model, although preliminary attempts have been made to model them in simple systems.[6] One modification of ASM No. 1 allows for direct use of nitrite during denitrification,[33] but the rate is not linked to the nitrite concentration, which could lead to erroneous results.

Denitrification is usually practiced to remove nitrogen from a wastewater, but that is not the only reason for including it in a treatment system.[13] In the previous section we saw that the inclusion of nitrification in a biochemical reactor requires that substantial amounts of oxygen be supplied. The electron accepting potential of that oxygen is not entirely lost, however, because much of it still resides in the nitrate produced. For example, 4.57 g O_2 are needed to convert a g of ammonia-N to nitrate-N through nitrification, but each g of nitrate-N has the electron accepting capacity of 2.86 g O_2 as shown in Table 3.1. Thus, if there were ways to use that electron accepting capacity in the oxidation of some of the organic matter in the wastewater, almost 63% of the energy expended in converting the ammonia-N to nitrate-N could be recovered. The development of treatment systems in which this can be done has been a major focus of environmental engineering during the past decade. We also see in the previous section that nitrification results in the destruction of substantial quantities of alkalinity, 7.08 g (as $CaCO_3$) per g of ammonia-N oxidized. Denitrification, on the other hand, destroys hydrogen ions and produces carbon dioxide, as shown in Eq. 3.22. This has the net effect of increasing alkalinity. As can be seen from the stoichiometric coefficient for alkalinity production in Table 6.1, approximately $(1-Y_H)/14$ moles of alkalinity will be produced for each gram of nitrate-N reduced to N_2. For a typical Y_H value, this amounts to approximately 3.5 g of alkalinity (as $CaCO_3$) per g of nitrate-N. Thus, about half of the alkalinity lost through the oxidation of ammonia-N to nitrate-N can be recovered during the subsequent reduction of nitrate-N to N_2. Through appropriate configuration or operation

of the bioreactor it is possible to make use of this to reduce the amount of chemical that must be purchased for pH control.[13] We will see later how this can be done.

6.4.2 Factors Affecting Denitrification

Even though denitrification is usually practiced as part of a treatment system within which the nitrate is formed by nitrification, it is easier to understand the unique characteristics of denitrification if it is considered in isolation. Such a consideration is not artificial because some industrial wastewaters contain high concentrations of nitrate and biological treatment is one option open to the engineer for its removal. In this section we will consider the growth of biomass in a CSTR receiving an influent containing nitrate. We will assume that the influent also contains sufficient ammonia-N to meet the biosynthetic needs of the biomass so that the only role of the nitrate-N is as the terminal electron acceptor. This means that the stoichiometry depicted in Table 6.1 is applicable.

Comparison of process 1 in Table 6.1, aerobic growth of heterotrophs, with process 2, anoxic growth of heterotrophs, reveals that the COD-based stoichiometry is the same. This was also seen in Chapter 3 when Eq. 3.24 was compared to Eq. 3.8. Thus, for the utilization of a given amount of organic matter in a CSTR, the amount of electron acceptor used (on a COD basis) will be the same for both growth conditions. This suggests that the oxygen requirement curves that we have looked at in preceding sections can also be expressed as nitrate-N requirement curves, as long as the nitrate-N is expressed on a COD or O_2 equivalent basis through use of the factor 2.86 g O_2/g N (or −2.86 g COD/g N) as shown in Table 3.1. This, in turn, tells us that the amount of nitrate-N that will be removed by growth on a given amount of influent substrate will depend on the SRT of the bioreactor. The longer the SRT, the greater the fraction of the electrons in the substrate that will go to the acceptor, and the greater the amount of nitrate-N that will be reduced. Thus, one factor affecting the removal of nitrate-N through denitrification is the SRT of the reactor in which the biomass is growing.

When considering denitrification it is necessary to reverse your thinking. During aerobic growth of heterotrophs, the goal is removal of the organic substrate and the electron acceptor is supplied in excess. However, during anoxic growth of heterotrophs, the goal is the removal of the electron acceptor, and sufficient organic substrate, i.e., electron donor, must be supplied to accomplish this. Consequently, engineers focus on the $\Delta S/\Delta N$ ratio, the amount of substrate COD that must be supplied to remove a given amount of nitrate-N, rather than on the electron acceptor requirement as was done for aerobic bioreactors. Consideration of the factors discussed in the preceding paragraph suggest that $\Delta S/\Delta N$ varies with SRT in a manner opposite to the way in which the oxygen requirement varies with the SRT in an aerobic bioreactor. Returning to Chapter 5, in which the traditional approach to decay was used, Eq. 5.33 allowed calculation of the amount of oxygen (electron acceptor) used:

$$RO = F(S_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H}{1 + b_H \cdot \Theta_c}\right] \tag{5.33}$$

Since RO represents the amount of electron acceptor used, and since each g of

nitrate-N can accept as many electrons as 2.86 g of oxygen, the equation can be rewritten as:

$$2.86 \cdot F(S_{NO_3} - S_{NO}) = F(S_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H}{1 + b_H \cdot \Theta_c}\right] \quad (6.3)$$

Recognition that $F(S_{NO_3} - S_{NO})$ is ΔN and that $F(S_{SO} - S_S)$ is ΔS, gives the following expression for $\Delta S/\Delta N$ using the traditional approach to decay:

$$\frac{\Delta S}{\Delta N} = \frac{2.86(1 + b_H \cdot \Theta_c)}{1 + (b_H \cdot \Theta_c) - Y_H(1 + f_D \cdot b_H \cdot \Theta_c)} \quad (6.4)$$

Because of the cycling of carbon in the lysis:regrowth approach to decay, it is not possible to derive an analogous expression using that approach, but the same principles apply and thus Eq. 6.4 is important for showing that $\Delta S/\Delta N$ varies with SRT in a manner opposite to that of the oxygen requirement. In other words, the amount of electron donor required to remove a given amount of nitrate-N will decrease as the SRT is increased. This happens because the increased importance of decay at longer SRTs allows a greater fraction of the electrons available in the donor to go to the acceptor rather than into biomass. Furthermore, it can be seen that the magnitude of this effect depends on the values of the kinetic and stoichiometric coefficients describing the system. The effect of SRT is illustrated in Figure 6.10[14] with data collected from a CSTR receiving methanol as the electron donor. Although ΔS was expressed as the mass of methanol rather than as the mass of COD, it is clear that the observations confirm the theory.

The calculation of the $\Delta S/\Delta N$ ratio above is based on the assumption that nitrate-N is the only available electron acceptor. In practice, however, it is difficult to totally eliminate the entrance of oxygen into a bioreactor. Since oxygen is the preferred electron acceptor, any entering the bioreactor will increase the amount of

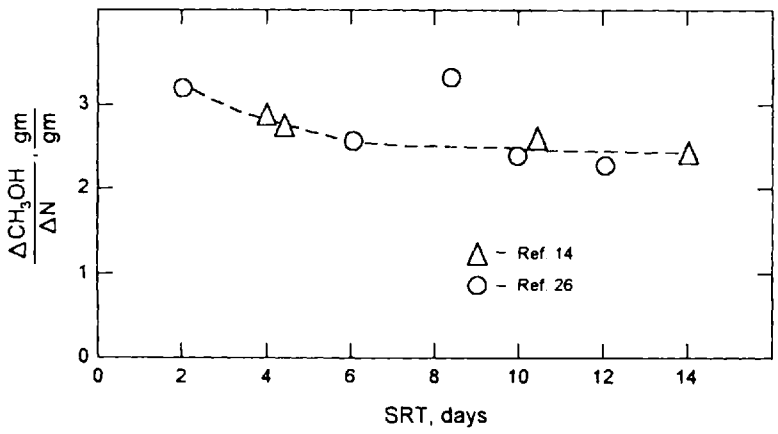


Figure 6.10 Effect of SRT on the $\Delta S/\Delta N$ ratio. The reference numbers refer to the sources of the data. (From D. J. Engberg and E. D. Schroeder, Kinetics and stoichiometry of bacterial denitrification as a function of cell residence time. *Water Research* **9**:1051–1054, 1975. Copyright © Elsevier Science Ltd.; reprinted with permission.)

electron donor that must be added to reduce a given amount of nitrate-N. In other words, it will increase the $\Delta S/\Delta N$ ratio.

The impact of the input of oxygen into a denitrification reactor can best be seen through use of the model in Table 6.1 because it considers both aerobic and anoxic growth of the heterotrophic bacteria and the effect of dissolved oxygen on each. To illustrate this point, a wastewater like that in Table 6.6 was assumed to be entering a CSTR operated at an SRT of 240 hr. The kinetic and stoichiometric coefficients describing the biomass in the bioreactor were assumed to have the values given in Table 6.3, except for $\hat{\mu}_A$ which was set equal to zero to ensure only heterotrophic reactions. Two situations were considered. In one, the influent nitrate-N concentration was set equal to 50 mg/L, which gave a mass input rate equivalent to 143 kg/day of oxygen. Given the values of the stoichiometric coefficients in Table 6.3, the amount of biodegradable COD entering the bioreactor (265 kg/day) was in excess of that needed to meet the required $\Delta S/\Delta N$ ratio. This case is called the excess COD case. In the other, the influent nitrate-N concentration was set equal to 60 mg/L, giving a mass input rate equivalent to 172 kg/day of oxygen, which was more than could be completely removed by the available COD. This is called the limiting COD case. Simulations were then conducted in which the rate of oxygen transfer into the bioreactor was set at various values and the results are shown in Figure 6.11. For the excess COD case, the input of a significant amount of oxygen could be tolerated without having an effect on the effluent nitrate-N concentration because the oxygen simply acted to allow more removal of COD. Eventually, however, a point was reached at which nitrate-N removal deteriorated because the total input of electron acceptor, i.e., nitrate-N plus oxygen, exceeded the amount of electrons available from the donor. For the limiting COD case, the entrance of even a small amount of oxygen caused the nitrate-N concentration to increase because any oxygen entering the bioreactor accepted electrons that otherwise would have gone for nitrate-N

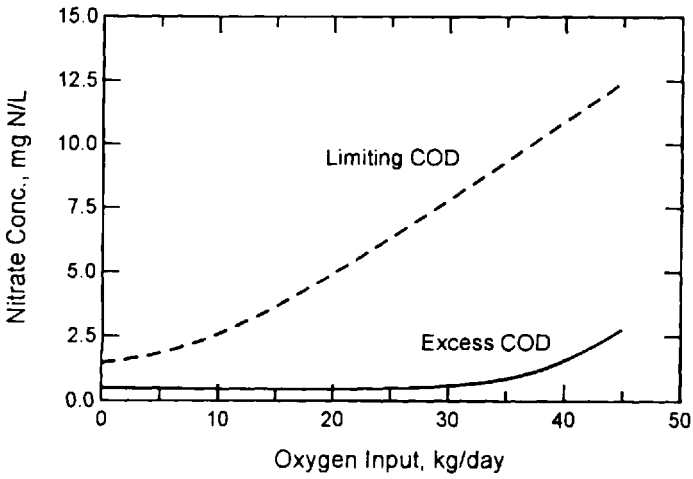


Figure 6.11 Effect of oxygen input rate on denitrification in a CSTR operated at an SRT of 240 hrs. Parameter values are given in Table 6.3 and the influent conditions are given in Table 6.6. For the limiting COD case the influent nitrate-N concentration was 60 mg/L whereas in the excess COD case it was 50 mg/L. The influent flow was 1000 m^3/day.

reduction. Although the magnitudes of the values given in Figure 6.11 are specific for the reactor conditions and parameter values used in the simulations, the results show clearly the importance of controlling the entrance of oxygen into a bioreactor in which denitrification is occurring.

The $\Delta S/\Delta N$ ratio is calculated from process stoichiometry, and simply tells us how much of one reactant will be removed in proportion to another. If the mass input rate of biodegradable COD into a denitrifying bioreactor is greatly in excess of that needed to remove the nitrate-N present, then the effluent biodegradable COD will be high (because there is insufficient acceptor to which to transfer electrons) and the nitrate-N concentration will be low and rate controlling. In other words, the term $S_S/(K_S + S_S)$ in the rate expression for process 2 in Table 6.1 will approach 1.0 and the term $S_{NO}/(K_{NO} + S_{NO})$ will be small. Conversely, if the input rate of biodegradable COD is less than that needed to remove the nitrate-N, then the effluent nitrate-N concentration will be high (because there is insufficient donor to provide the needed electrons) and the biodegradable COD concentration will be low and rate controlling. In other words, $S_{NO}/(K_{NO} + S_{NO})$ will approach 1.0 and $S_S/(K_S + S_S)$ will be small. The only way for both concentrations to be low and for both to simultaneously influence the rates of activity in the bioreactor is for the influent concentrations to closely match the required $\Delta S/\Delta N$ ratio. Given the influence of the entrance of small amounts of oxygen, as discussed above, and the variability associated with the values of the stoichiometric and kinetic coefficients in mixed microbial communities such as those used in wastewater treatment, this is difficult to achieve. For example, Figure 6.12[39] presents results from a study in which the relative amounts of biodegradable COD and nitrate-N (expressed as the carbon:nitrogen ratio) in the influent to a CSTR were varied. There it can be clearly seen that there was only a small range of influent ratios over which the effluent concentrations of both constituents were low. To overcome this problem, it is common practice to add the

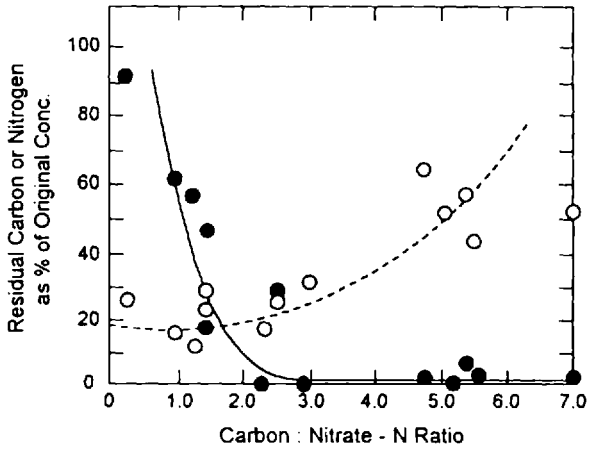


Figure 6.12 Effect of S_{SO}/S_{NOO} (expressed as C/N ratio) on the removal of carbon (○) and nitrogen (●) in a CSTR operated under anoxic conditions. (From K. Wuhrmann, Discussion of 'Factors affecting biological denitrification of wastewater' by R. N. Dawson and K. L. Murphy. *Advances in Water Pollution Research, Jerusalem, 1972*, 681–682, 1973. Reproduced by permission of Dr. K. L. Mechsner.)

organic substrate in slight excess of the amount required to remove the nitrate-N, and then to pass the effluent from the anoxic bioreactor through a small aerobic bioreactor in which any residual electron donor can be removed with oxygen as the electron acceptor.

6.5 MULTIPLE EVENTS

The purpose for development of the model in Table 6.1 was to allow engineers to simulate biochemical reactors in which all of the listed processes are occurring. Thus, it would be instructive to use it to investigate such a situation in a single CSTR. It is apparent, by now, however, that the conditions required for anoxic growth of heterotrophs and aerobic growth of autotrophs are mutually exclusive, since both are controlled by the dissolved oxygen concentration, but in the opposite manner. Consequently, if a CSTR is receiving a constant input and is operating at steady-state with a constant dissolved oxygen concentration, it is impossible for significant amounts of both nitrification and denitrification to occur. However, observations of treatment systems receiving diurnal variations in flow and concentration suggested that when the input was low, resulting in a high dissolved oxygen concentration, nitrification occurred, but when the input was high, driving the dissolved oxygen to very low concentrations, nitrification ceased and denitrification began, destroying part of the nitrate-N formed during the aerobic period. This made the daily average effluent total nitrogen concentration lower than in a system receiving adequate oxygen, suggesting that it might be possible to purposefully design a system in which the dissolved oxygen concentration varied sufficiently to allow both reactions to occur. In addition to reducing the amount of nitrogen discharged, this would reduce the amount of alkalinity destroyed and the amount of oxygen required, as discussed previously. Let us now consider each situation for a typical domestic wastewater.

6.5.1 Effects of Diurnal Variations in Loading

Figure 6.13 shows the simulated performance of a CSTR containing a biomass with the kinetic and stoichiometric coefficients in Table 6.3 and receiving an input with the variations shown in Figure 6.2. The average daily flow rate was 1000 m^3/day and the reactor volume was 500 m^3, giving an average HRT of 12 hr. The flow-weighted average concentrations of the various influent components were as shown in Table 6.6. The influent alkalinity, however, was assumed to have a constant concentration because it is influenced primarily by the characteristics of the carriage water. Two situations were considered. In one, the dissolved oxygen concentration was held constant at 2.0 mg/L throughout the entire 24 hr period. Given the value of $K_{O,H}$ in Table 6.3, this effectively eliminated denitrification. In the other, the mass transfer rate for oxygen was sufficient to maintain a dissolved oxygen concentration of 2.0 mg/L if the bioreactor received the daily average flow and concentrations. As shown in Figure 6.13a, however, because of the variable input, this resulted in excessive oxygen concentrations during periods of low loading, but inadequate concentrations during high loading. Under those conditions, significant denitrification occurred, as shown in Figure 6.13c. Comparison of the solid and dashed curves in parts b and d of the figure shows that while the lack of oxygen had only a minor

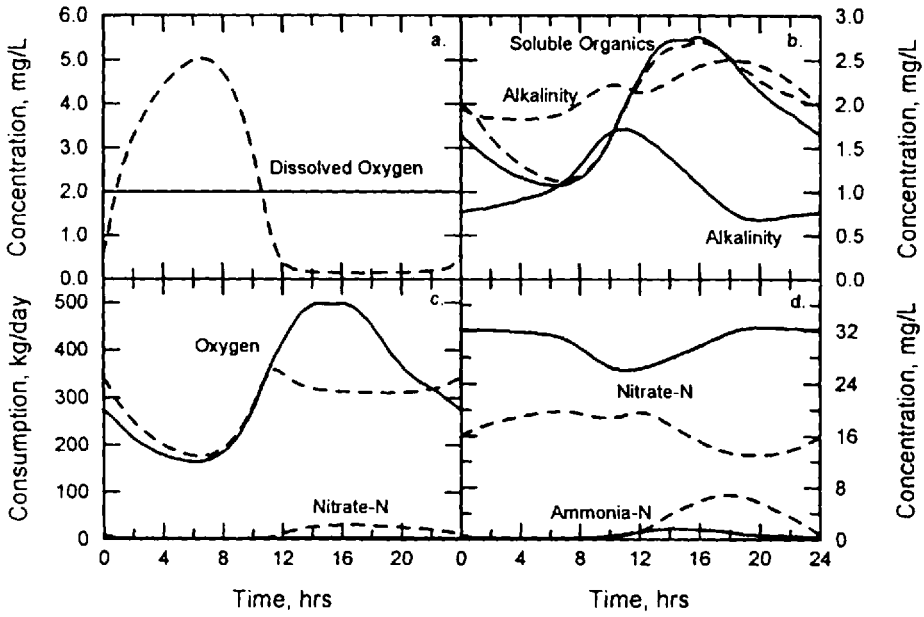


Figure 6.13 Response of a CSTR to a diurnal input. The input patterns are like those in Figure 6.2, but the flow weighted average concentrations are those given in Table 6.6. The daily average flow rate was 1000 m^3/day. The solid curves are for a reactor in which the dissolved oxygen concentration was held constant at 2.0 mg/L. The dashed curves are for a reactor in which the mass transfer coefficient for oxygen was held constant at 3.83 hr^{-1}. The average HRT = 12 hrs and the SRT = 240 hrs. The parameter values are given in Table 6.3.

effect on the concentration of soluble organic matter in the effluent, it significantly lowered the nitrate-N concentration and raised the ammonia-N concentration. The raising of the ammonia-N concentration was due to retardation of nitrification, but the decrease in the nitrate-N concentration was caused by a combination of less production from nitrification and more consumption by denitrification. Nevertheless, it is apparent that less nitrogen was discharged from the system that had limited oxygen transfer capacity. Furthermore, less alkalinity was destroyed.

While it is not desirable to design a supposedly aerobic system with inadequate oxygen transfer capacity, the results from the simulation show clearly that all events can occur in a single biomass provided that the SRT of the system is sufficiently long to allow the nitrifying bacteria to grow during the aerobic period. This suggests that it should be possible to design a system in a way that maximizes nitrogen removal by controlling the periods with and without oxygenation. Such a system would have minimal power input as well as minimal alkalinity destruction.

6.5.2 Intermittent Aeration

Batchelor[5] was among the first to use simulation to investigate the possibility of using intermittent aeration to achieve both nitrification and denitrification in a single CSTR receiving a constant influent. His study was conducted with a model that was conceptually similar to the one in Table 6.1, although it differed somewhat with

respect to both the kinetics and stoichiometry employed. In addition, the values of the kinetic and stoichiometric parameters were slightly different from those in Table 6.3. Nevertheless, the results of his simulations illustrate some important concepts regarding the major variables influencing the performance of such systems.

The situation considered by Batchelor[5] was of a wastewater containing 200 mg/L of biodegradable COD and 30 mg/L of ammonia-N, entering a CSTR with an SRT of 240 hr and an HRT of 4.3 hr. The bioreactor was operated with a cycle time (the time between initiations of aeration) of 0.5 hr while the aeration fraction (the fraction of time that the bioreactor was aerobic) was varied. In each case, the effluent nitrogen concentrations from the bioreactor achieved a stable oscillation and the average concentrations of ammonia-N, nitrate-N, and total nitrogen were calculated. Figure 6.14 shows clearly that at low aeration fractions, the average ammonia-N concentration rises whereas at high aeration fractions the average nitrate-N concentration rises. Furthermore, an optimum exists at which the discharge of nitrogen is minimized. This result is consistent with our previous discussions. At aeration fractions in excess of the optimum, nitrification is complete, but the anoxic period is insufficient to allow much reduction of nitrate-N. Conversely, at aeration fractions less than the optimum, the aerobic period becomes insufficient for growth of the nitrifying bacteria. If the aeration fraction is made so low that the product of the aeration fraction and the SRT is below the minimum SRT for the nitrifiers, they will wash out and no nitrogen removal will be achieved, other than that associated with incorporation of nitrogen into the heterotrophic biomass via synthesis. For the kinetic parameters used by Batchelor,[5] the minimum SRT for nitrification was 52.6 hr. Since the system SRT was 240 hr, this suggests that total process failure would occur at an aeration fraction of 0.22. The fraction of the SRT that is aerobic is called the aerobic SRT and it is now recognized as an important parameter in the design of systems in which both nitrification and denitrification are occurring.

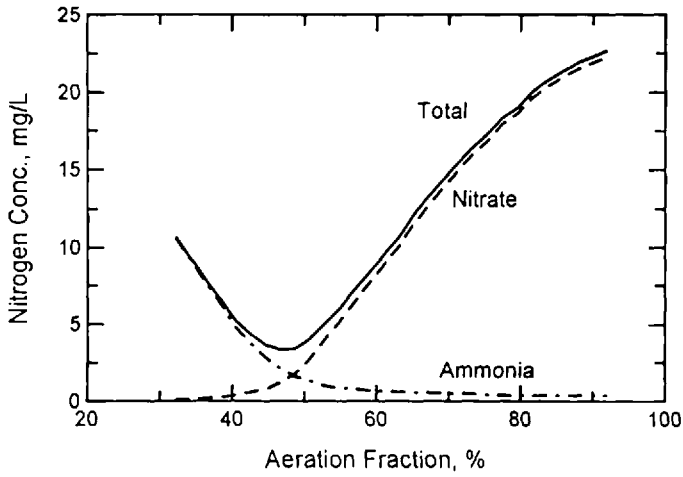


Figure 6.14 Effect of aeration fraction on the concentrations of ammonia-N, nitrate-N, and total-N in a CSTR operated with intermittent aeration at a cycle time of 0.5 hr. The HRT = 4.3 hrs and the SRT = 240 hrs. The influent biodegradable COD = 200 mg/L and the influent nitrogen concentration = 30 mg/L. (Adapted from Batchelor.[5])

The optimal aeration fraction in Figure 6.14 is 47%, but since the minimum allowable aeration fraction depends on the system SRT and the degree of denitrification also depends on the system SRT, we might expect the optimum aeration fraction to depend on the SRT as well. That this is the case is shown by the dark circles in Figure 6.15. The vertical bars in the figure indicate the range of aeration fractions at each SRT which results in total nitrogen concentrations within 1.0 mg/L of the minimum. That range is seen to broaden as the SRT is increased, suggesting that longer SRTs provide more latitude in operation of the bioreactor to achieve optimal nitrogen removal.

Both Figure 6.14 and Figure 6.15 were prepared from simulations conducted with a cycle time of 0.5 hr. However, we might also expect the cycle time to influence the degree of nitrogen removal. During aerobic periods, the ammonia concentration will fall and the nitrate-N concentration will rise as nitrification occurs. Conversely, during anoxic periods, denitrification will cause the nitrate-N concentration to fall while the continued influx of ammonia in the absence of nitrification will cause the ammonia concentration to rise. The longer the cycle time, the greater the amplitude of the cycles because a longer time will be available for each reaction. Since the effluent concentration is characterized by the daily average concentration, longer cycle times result in higher average nitrogen concentrations. On the other hand, as the cycle time is shortened, a point will be reached at which it is difficult for the biomass to shift rapidly enough from aerobic to anoxic metabolism, and vice versa. Consequently, there is also an optimum cycle time associated with each SRT and aeration fraction. At this time our understanding of the metabolic controls acting on the synthesis and activity of denitrifying enzymes is not sufficient to allow the optimum cycle time to be selected by simulation. Nevertheless, experimental studies have shown that cycle times on the order of 20 to 45 min work well.[20]

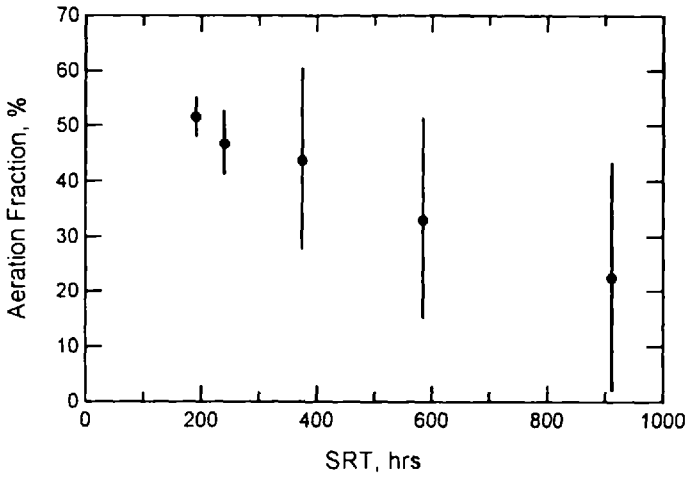


Figure 6.15 Effect of SRT on the optimal aeration fraction (dots) and the range of near optimal aeration fractions (lines) in a CSTR operated with intermittent aeration. The influent biodegradable COD = 200 mg/L and the influent nitrogen concentration = 30 mg/L. The dissolved oxygen concentration was fixed at 2.0 mg/L when the aeration was on and 0.0 mg/L when it was off. (Adapted from Batchelor.[5])

6.5.3 Closure

The important point to gain from the preceding is that engineers can exert control over the environment in biochemical reactors, thereby allowing processes to occur in a single system that would not otherwise occur together. This suggests that the engineer has considerable latitude in system design. The complexity of the interactions, however, makes it impossible to intuitively predict the outcome of all possible systems that the engineer might conceive. This, in turn suggests why it is necessary to work with models like those in Table 6.1. Through their application, engineers can explore large numbers of possible bioreactor systems to see how system layout and environment influence the outcome of the possible reactions.

It is apparent that the number of options available during design of a single CSTR is very limited. Thus, most biochemical operations employ reactors with spatial gradients in them, usually through use of reactors with large length to width ratios, but also through use of compartmentalized reactors. Since both can be modeled as tanks in series systems, in the next chapter we will apply ASM No. 1 to study the performance of a number of such systems.

6.6 KEY POINTS

1. International Association on Water Quality activated sludge model No. 1 incorporates eight processes acting on thirteen components. The processes are aerobic and anoxic growth of heterotrophic biomass, aerobic growth of autotrophic bacteria, decay of both heterotrophic and autotrophic biomass, ammonification of soluble organic nitrogen, and hydrolysis of both particulate organic substrate and particulate organic nitrogen.
2. The thirteen components incorporated into ASM No. 1 include six particulate and seven soluble ones. The particulate ones are inert organic matter, slowly biodegradable substrate, heterotrophic biomass, autotrophic biomass, biomass debris, and organic nitrogen. The soluble ones are inert organic matter, readily biodegradable substrate, dissolved oxygen, nitrate-N, ammonia-N, organic nitrogen, and alkalinity.
3. In ASM No. 1, biomass growth is expressed by a double Monod equation in which both the electron donor and acceptor are considered, biomass loss is modeled by the lysis:regrowth approach, ammonification is considered to be first order with respect to the soluble organic nitrogen concentration, and hydrolysis is modeled by a surface mediated reaction term in which the rate is controlled by the slowly biodegradable substrate to biomass ratio and the electron acceptor concentration.
4. Activated sludge model No. 2 includes biological phosphorus removal in addition to carbon oxidation, nitrification, and denitrification. Because biological phosphorus removal is a complicated process, ASM No. 2 is considerably more complex than ASM No. 1.
5. Because hydrolysis is a slow reaction, long SRTs are required to obtain substantial degradation of particulate substrate.
6. The major impact of the presence of particulate substrate in the feed to a CSTR receiving a dynamic input is to dampen the variability in the

oxygen requirement, thereby decreasing the peak requirement and delaying the occurrence of the maximum and minimum values.

7. Because the half-saturation coefficient for nitrifying biomass is very small relative to the concentration of ammonia-N in most wastewaters, nitrification behaves in an all-or-none manner in which nitrification is either almost complete or washout occurs. Furthermore, the washout SRT is very sensitive to the dissolved oxygen concentration and the temperature.
8. Because of the kinetics and stoichiometry of nitrification, the growth of autotrophic bacteria in a CSTR in which heterotrophs are growing can have a major impact on the amount of oxygen required while having little effect on the MLSS concentration.
9. The release of ammonia-N as the result of biomass destruction in a CSTR receiving a feed containing only biomass provides a significant amount of substrate for nitrifying bacteria. Growth of those bacteria will have a substantial impact on the oxygen requirement in the bioreactor as well as on the amount of alkalinity destroyed.
10. Denitrification can have several benefits: it can remove nitrogen by converting nitrate-N to nitrogen gas; it can recover approximately 63% of the energy expended during nitrification by using the resulting nitrate-N as a terminal electron acceptor for organic substrate removal; and it can recover about half of the alkalinity destroyed during nitrification.
11. The $\Delta S/\Delta N$ ratio is an important characteristic of denitrifying bioreactors because it determines how much electron donor must be provided to convert a given amount of nitrate-N to nitrogen gas. However, because the ratio is influenced by factors such as the SRT and the presence of oxygen, it is not generally possible to operate a CSTR so that the effluent concentrations of both the electron donor and nitrate-N are low. Consequently, the electron donor is usually provided in excess and any residual is removed in a small aerobic bioreactor with oxygen as the electron acceptor.
12. If a CSTR with an SRT sufficient for nitrification receives influent that follows typical diurnal patterns of flow and concentration, it is possible for denitrification to occur during periods of peak loading if the input rate of oxygen is insufficient to maintain significant concentrations of dissolved oxygen.
13. By subjecting a CSTR to intermittent aeration, it is possible to have both nitrification and denitrification occur in a single bioreactor receiving a constant input. Under those circumstances, the fraction of time that the system is aerobic is an important determinant of system performance, and there will be an optimum fraction that minimizes the effluent nitrogen concentration.

6.7 STUDY QUESTIONS

1. What are the thirteen components and the eight processes considered in IAWQ ASM No. 1? Construct a matrix that indicates which processes act on which components, using a plus sign to indicate when the concentra-

tion of a component is increased by the process and a negative sign to indicate when it is decreased.

2. What are the eight processes considered in ASM No. 1 and how is each modeled?
3. Simplify the matrix representing ASM No. 1 for the following situations:
   a. A totally aerobic bioreactor receiving only soluble constituents in the feed.
   b. A totally aerobic bioreactor receiving only particulate constituents in the feed.
   c. A totally anoxic bioreactor receiving particulate organic constituents and soluble nitrate and ammonia nitrogen.
   d. A totally aerobic bioreactor receiving ammonia-N as the only electron donor.
4. Draw a sketch comparing the steady-state performance of CSTRs receiving soluble and particulate substrates, and use it to contrast the usefulness of the process loading factor and SRT as design and operational parameters for bioreactors receiving particulate substrate.
5. Draw a sketch contrasting the response to typical diurnal loading patterns of the oxygen requirement in CSTRs receiving soluble and particulate substrates? Why do the two reactors behave differently?
6. What is meant by the statement that nitrification behaves in an all-or-none fashion? Why does it do so? What are the impacts of temperature and dissolved oxygen concentration on nitrification?
7. What are the effects of nitrification in a CSTR receiving an influent with the characteristics of typical domestic wastewater? Why do they occur?
8. Using a computer code for IAWQ ASM No. 1 and typical temperature coefficients for the growth of heterotrophic and autotrophic biomass, investigate the effects of temperature (10° to 35°C) and SRT (48 to 360 hr) on the effluent concentrations of ammonia-N and readily biodegradable substrate from a CSTR receiving an influent with the characteristics given in Table 6.6. Use your results to discuss what is likely to happen to the performance of such a bioreactor throughout the year as the temperature changes and how SRT can be used to influence that performance.
9. What are the effects of nitrification in a CSTR receiving an influent containing heterotrophic biomass as its major constituent? What are the implications of those effects to the design and operation of such a reactor?
10. What is meant by the term $\Delta S/\Delta N$ ratio, how is it used in characterizing the performance of an anoxic CSTR, and why does it decline as the SRT of the bioreactor is increased?
11. Why is it difficult to operate an anoxic CSTR in a way that will ensure that the effluent concentrations of both the organic substrate and nitrate-N are low?
12. When a single CSTR receiving influent at a constant rate is subjected to intermittent aeration, the fraction of time that the system is aerobic will affect the effluent nitrogen concentration. Draw a sketch showing the effect of aeration fraction on the effluent nitrogen concentrations from such a system. Why does it look as it does?

13. For the situation described in Study Question 12, why does the optimal aeration fraction decrease as the SRT of the bioreactor is increased? Why does the range of near optimal aeration fractions increase?

REFERENCES

1. Alleman, J. E., Elevated nitrite occurrence in biological wastewater treatment systems. *Water Science and Technology* **17**(2/3):409–419, 1985.
2. Anthonisen, A. C., R. C. Loehr, T. B. S. Prakasam and E. G. Srinath, Inhibition of nitrification by ammonia and nitrous acid. *Journal, Water Pollution Control Federation* **48**:835–852, 1976.
3. Baillod, C. R., Oxygen utilization in activated sludge plants: simulation and model calibration. U. S. EPA Report No. EPA/600/S2-88/065, 1989.
4. Balakrishnan, S. and W. W. Eckenfelder, Nitrogen relationships in biological treatment processes. I. Nitrification in the activated sludge process. *Water Research* **3**:73–81, 1969.
5. Batchelor, B., Simulation of single-sludge nitrogen removal. *Journal of Environmental Engineering* **109**:1–16, 1983.
6. Betlach, M. R. and J. M. Tiedje, Kinetic explanation for accumulation of nitrite, nitric oxide and nitrous oxide during bacterial denitrification. *Applied and Environmental Microbiology* **42**:1074–1084, 1981.
7. Bidstrup, S. M. and C. P. L. Grady Jr., SSSP—Simulation of single-sludge processes. *Journal, Water Pollution Control Federation* **60**:351–361, 1988.
8. Bridle, T. R., D. C. Climenhage, and A. Stelzig, Start-up of a full scale nitrification-denitrification treatment plant for industrial waste. *Proceedings of the 31st Industrial Waste Conference, 1976, Purdue University*, Ann Arbor Science, Ann Arbor, MI, 807–815, 1977.
9. Corbitt, R. A., ed., *Standard Handbook of Environmental Engineering*. McGraw-Hill, New York, NY, 1990.
10. Dold, P. L., *BioWin*. EnviroSim Associates, Oakville, Ontario, Canada, 1991.
11. Dold, P. L., Incorporation of biological excess phosphorus removal in a general activated sludge model. Paper, Department of Civil Engineering and Engineering Mechanics, McMasters University, Hamilton, Ontario, Canada, 1991.
12. Dold, P. L. and G. v. R. Marais, Evaluation of the general activated sludge model proposed by the IAWPRC task group. *Water Science and Technology* **18**(6):63–89, 1986.
13. Dold, P. L. and G. v. R. Marais, Benefits of including unaerated zones in nitrifying activated sludge plants. *Water Science and Technology* **19**(1/2):195–207, 1987.
14. Engberg, D. J. and E. D. Schroeder, Kinetics and stoichiometry of bacterial denitrification as a function of cell residence time. *Water Research*, **9**:1051–1054, 1975.
15. Gujer, W., ASIM: Activated Sludge SIMulation Program, Version 3.0. Swiss Federal Institute for Environmental Science and Technology, Dübendorf, Switzerland, 1994.
16. Henze, M., Nitrate versus oxygen utilization rates in wastewater and activated sludge systems. *Water Science and Technology* **18**(6):115–122, 1986.
17. Henze, M., C. P. L. Grady, Jr., W. Gujer, G. v. R. Marais, and T. Matsuo, Activated sludge model No. 1. *IAWPRC Scientific and Technical Reports*, No. 1, 1987.
18. Henze, M., C. P. L. Grady Jr., W. Gujer, G. v. R. Marais, and T. Matsuo, A general model for single-sludge wastewater treatment systems. *Water Research* **21**:505–515, 1987.
19. Henze, M., W. Gujer, T. Mino, T. Matsuo, M. C. Wentzel, and G. v. R. Marais, Activated sludge model No. 2. *IAWQ Scientific and Technical Reports*, No. 3, 1995.

20. Inomae, K., H. Araki, K. Koga, Y. Awaya, T. Kusuda, and Y. Matsuo, Nitrogen removal in an oxidation ditch with intermittent aeration. *Water Science and Technology* **19**(1/2): 209–218, 1987.
21. Joel, A. R. and C. P. L. Grady Jr., Inhibition of nitrification—Effects of aniline following biodegradation. *Journal, Water Pollution Control Federation* **49**:778–788, 1977.
22. Johnson, W. K. and G. J. Schroepfer, Nitrogen removal by nitrification and denitrification. *Journal, Water Pollution Control Federation* **36**:1015–1036, 1964.
23. Kuba, T., G. J. F. Smolders, M. C. M. van Loosdrecht, and J. J. Heijnen, Biological phosphorus removal from wastewater by anaerobic-anoxic sequencing batch reactor. *Water Science and Technology* **27**(5/6):241–252, 1993.
24. McClintock, S. A., J. H. Sherrard, J. T. Novak and C. W. Randall, Nitrate versus oxygen respiration in the activated sludge process. *Journal, Water Pollution Control Federation* **60**:342–350, 1988.
25. Metcalf and Eddy, Inc., *Wastewater Engineering: Treatment, Disposal, and Reuse*. Third Edition McGraw-Hill, New York, NY 1991.
26. Moore, S. F. and E. D. Schroeder, An investigation of the effect of residence time on anaerobic bacterial denitrification. *Water Research* **4**:685–694, 1970.
27. Poduska, R. A. and J. F. Andrews, Dynamics of nitrification in the activated sludge process. *Proceedings of the 29th Industrial Waste Conference*. Purdue University Engineering Extension Series No. 145, 1005–1025, 1974.
28. Prakasam, T. B. S. and R. C. Loehr, Microbial nitrification and denitrification in concentrated wastes. *Water Research* **6**:859–869, 1972.
29. Prakasam, T. B. S., Y. D. Joo, E. G. Srinath, and R. C. Loehr, Nitrogen removal from a concentrated waste by nitrification and denitrification. *Proceedings of the 29th Industrial Waste Conference*. Purdue University Engineering Extension Series No. 145, 497–509, 1974.
30. Randall, C. W. and D. Buth, Nitrite build-up in activated sludge resulting from temperature effects. *Journal, Water Pollution Control Federation* **56**:1039–1044, 1984.
31. Sam-Soon, P.A.L.N.S., M. C. Wentzel, P. L. Dold, R. E. Loewenthal and G. v. R. Marais, Mathematical modelling of upflow anaerobic sludge bed (UASB) systems treating carbohydrate waste waters. *Water SA* **17**:91–106, 1991.
32. Satoh, H., T. Mino, and T. Matsuo, Uptake of organic substrates and accumulation of polyhydroxyalkanoates linked with glycolysis of intracellular carbohydrates under anaerobic conditions in the biological excess phosphate removal process. *Water Science and Technology* **26**(5/6):933–942, 1992.
33. Sen, D. and C. W. Randall, General activated sludge model for biological nitrogen and excess phosphorus removal. In *Design and Retrofit of Wastewater Treatment Plants for Biological Nutrient Removal*, C. W. Randall, J. L. Barnard and H. D. Stensel, eds., Technomic Publishing, Lancaster, PA, 311–333, 1992.
34. van Haandel, A. C., G. A. Ekama and G. v. R. Marais, The activated sludge process. 3. Single sludge denitrification. *Water Research* **15**:1135–1152, 1981.
35. Water Research Commission, *Theory, Design and Operation of Nutrient Removal Activated Sludge Processes*, Water Research Commission, Pretoria, South Africa.
36. Wentzel, M. C., P. L. Dold, G. A. Ekama, and G. v. R. Marais, Enhanced polyphosphate organism culture in activated sludge systems. Part III: kinetic model. *Water SA* **15**:89–102, 1989.
37. Wentzel, M. C., G. A. Ekama and G. v. R. Marais, Kinetics of nitrification denitrification biological excess phosphorus removal systems—A review. *Water Science and Technology* **23**(4-6):555–565, 1991.
38. Wuhrmann, K., Nitrogen removal in sewage treatment processes. Proceedings. *International Association of Theoretical and Applied Limnology*, **XV**, 580–596, 1964.
39. Wuhrmann, K., Discussion of 'Factors affecting biological denitrification of wastewater' by R. N. Dawson and K. L. Murphy. In *Advances in Water Pollution Research, Jerusalem, 1972*, S. H. Jenkins, ed., Pergamon Press, Oxford, England, 681–682, 1973.

This page intentionally left blank

7

Multiple Microbial Activities in Complex Systems

Chapters 5 and 6 introduce us to the response of microbial cultures in single continuous stirred tank reactors (CSTRs) and demonstrated the importance of solids retention time (SRT) in determining that response. Although such reactors have found extensive use in research and in the treatment of some industrial wastewaters, more complex reactor systems are commonly used in practice and it is important to consider them. Consequently, in this chapter we will use International Association on Water Quality (IAWQ) activated sludge model (ASM) No. 1 to investigate the responses of complex bioreactor systems when they contain biomass reacting in accordance with the processes discussed in Chapter 6 and listed in Table 6.1. Since the events occurring in these various systems are all the same, the responses will demonstrate how bioreactor configuration affects system performance, thereby providing a foundation from which engineering decisions can be made about the appropriate system configuration for a given objective. In addition, we will introduce the concepts of biological phosphorus removal (BPR) through use of ASM No. 2.

7.1 MODELING COMPLEX SYSTEMS

Table 1.2 shows that several activated sludge types use CSTRs in series or plug flow reactors with dispersion. Actually, for purposes of modeling system response, the latter can be represented adequately as CSTRs in series,[13] and only that configuration need be considered. Thus, this chapter will focus on the theoretical performance of a number of systems containing CSTRs linked in various ways.

7.1.1 Representing Complex Systems

Conventional activated sludge (CAS) systems use long rectangular bioreactors with influent and biomass recycle introduced together at one end. Even though such bioreactors are often called plug–flow reactors (PFRs), the mixing patterns in them are equivalent to three to five CSTRs in series.[14] High purity oxygen activated sludge (HPOAS), on the other hand, actually uses a series of bioreactors in order to achieve efficient utilization of the oxygen supplied, with four or five in series being common configurations. Because five CSTRs in series can be considered representative of both of these important activated sludge variations, that configuration will be investigated first.

Step feed activated sludge (SFAS) is an important variation on the CAS process in which influent is introduced at several points along a rectangular aeration basin while biomass recycle is introduced only at the beginning.[7] It was originally developed as a means for distributing the oxygen demand more evenly throughout the basin, but is now used primarily for its effect on the concentration of MLSS entering the final settler.[11,22] Since the reactor is the same as in CAS, it is also possible to represent this configuration as five CSTRs in series with all biomass recycle to the first bioreactor, but with the influent distributed evenly among the bioreactors. This configuration will be studied in Section 7.3 to see how the simple act of changing the feed distribution influences system performance.

Table 1.2 indicates that contact stabilization activated sludge[24] (CSAS) also uses CSTRs in series. While it is true that the system uses two CSTRs, they are only in series with respect to the biomass recycle flow, not the main wastewater flow. Rather, the main wastewater flow passes through only one bioreactor (the contact tank) while the biomass recycle flow passes through another bioreactor (the stabilization tank) before entering the contact tank. Thus, this system can be represented as two CSTRs in series with all biomass recycle entering the first bioreactor and all influent entering the second. This unique configuration responds in a way that is quite distinct from either of the first two configurations, and it is considered in Section 7.4.

One observation from Chapter 6 is that aerobic and anoxic growth of heterotrophic biomass cannot both occur to significant degrees in a single bioreactor maintained under constant environmental conditions because the two processes have mutually exclusive requirements. We see in Section 6.5.2, however, that partial denitrification can be achieved by imposing time variant changes in the dissolved oxygen concentration. While such a scheme can be used in small systems, it is not practically feasible for large ones because of problems associated with manipulation of large air flows. An alternative approach is to use two bioreactors in series, but with one bioreactor maintained under anoxic conditions and the other under aerobic conditions so that the wastewater and associated biomass encounter both environments during their passage through the system. One such system is the modified Ludzack–Ettinger[13] (MLE) process, in which the anoxic bioreactor is first. It is often used to treat domestic wastewaters, which are devoid of nitrate but rich in organic and ammonia nitrogen. Because nitrate is only formed under aerobic conditions, a recirculation stream returns mixed liquor from the second bioreactor to the first, thereby providing nitrate which the heterotrophic biomass can use as terminal electron acceptor during growth on the organic matter in the wastewater. Systems of this type—which are called single-sludge systems because they allow carbon oxidation, nitrification, and denitrification to occur in a single biomass that is passed through different biochemical environments—have become quite important in domestic wastewater treatment for biological nutrient removal (BNR). Thus, it is important that we investigate their performance through simulation.

One characteristic of the MLE process is that the effluent contains appreciable quantities of nitrate-N since nitrification occurs in the last bioreactor in the chain. One way to overcome this is to add a third bioreactor that is maintained under anoxic conditions, thereby allowing additional denitrification to occur. Unfortunately, discharge of an actively denitrifying biomass to a final settler results in biomass settling problems because the evolving nitrogen gas attaches to the biomass flocs and makes

them rise, rather than settle. One way to eliminate this is to aerate the biomass for a short period prior to settling, thereby stopping denitrification. The Bardenpho[1] process is one of several BNR systems that use these concepts. It can be represented as four tanks in series with all influent and biomass recycle entering the first and with mixed liquor recirculation (MLR) from the second to the first. As suggested above, the first and third bioreactors are anoxic whereas the second and fourth are aerobic. It is studied in Section 7.6.

As shown in Table 1.2, BPR processes also employ CSTRs in series to create the unique environment required for selection of phosphate accumulating organisms (PAOs). The simplest of those processes uses two bioreactors in series, with all influent wastewater and all biomass recycle entering the first bioreactor[18]. As will be recalled from Section 2.4.6, PAOs only develop their unique ability to store large quantities of phosphate when they are subjected to alternating anaerobic and aerobic conditions. Because of this, and because of the necessity to capitalize on the readily biodegradable substrate in the influent wastewater, the first bioreactor in a two-stage BPR system is anaerobic and the second is aerobic. Such systems are becoming increasingly important in domestic wastewater treatment. Consequently, it is important that we also investigate their performance through simulation.

The ability to represent several important biochemical operations as systems made of tanks in series simplifies the task of modeling their performance. All that is required is that a mass balance equation like Eq. 4.2 be written for each component in each bioreactor, including the settler, giving a set of equations that must be solved simultaneously. If the influent to the system has time invariant characteristics, the derivative terms in the equations may be set to zero, reducing the differential equations to algebraic ones. Otherwise, they remain as differential equations. In either case, the presence of biomass recycle and MLR requires numerical techniques to be used for their solution. As mentioned in Section 6.1.4, several computer codes are available for solving the equations that result from application of ASM No. 1 to a single CSTR with biomass recycle. They are also applicable to CSTRs in series. For example, the code SSSP,[3] which was used in Chapter 6, allows simulation of systems containing up to nine CSTRs with any distribution of feed or biomass recycle among them and with MLR from any bioreactor to any other bioreactor upstream of it. Because of its ability to simulate all of the systems discussed above that do not involve BPR, it was used to investigate their performance. Simulation of BPR processes is more complex than simulation of the other processes because more reactions and components are involved. Henze et al.[9] incorporated the concepts of Wentzel et al.[26,27] and others to expand ASM No. 1 to include biological phosphorus removal, thereby developing ASM No. 2. Gujer[8] then implemented the model in a computer code named ASIM, which was used to simulate the performance of a two-stage BPR process.

Examination of Table 1.2 reveals that biological nutrient removal can also be accomplished in sequencing batch reactors (SBRs) and they are finding increasing use for that purpose.[10,18] The utility of SBRs stems from their characteristic of varying temporally in a manner that is analogous to the spatial variations in a perfect plug–flow reactor. As a consequence, by changing the aeration pattern over time, carbon oxidation, nitrification, denitrification, and phosphorus removal can be accomplished in the same way as in the tanks-in-series systems discussed above. Therefore, it is important that we consider the performance of SBRs and their similarity to the other

systems. Oles and Wilderer[15] have developed a computer code, SBRSIM, for simulation of SBR performance. It is based on ASM No. 1. In Section 7.8 we briefly review the operating strategies for SBRs and examine the performance of a system that is analogous to the MLE process.

7.1.2 Significance of SRT

The solids retention time, SRT, is used as the fundamental variable in Chapters 5 and 6 because it is functionally related to the steady-state specific growth rate of the biomass in a CSTR. No such relationship holds for a chain of CSTRs, however. This is because the SRT is defined with respect to the entire system, while the biomass in each bioreactor has a unique specific growth rate in response to the concentration of the particular nutrient limiting growth in that bioreactor. Although one might question the significance of the SRT in such systems, it is still a variable of fundamental importance because it represents a net average specific growth rate for the system. This follows from the fact that SRT is defined as the mass of biomass contained in the system divided by the mass wasted per unit time (Eq. 5.1), whereas specific growth rate is defined as the mass of biomass formed per unit time divided by the biomass present (Eq. 3.35). Since at steady-state, the mass of biomass wasted must equal the net mass formed, i.e., the growth minus the loss due to lysis, death, and decay, it follows that the SRT is inversely proportional to the net average specific growth rate of the entire system. Consequently, it determines important system characteristics, such as the electron acceptor requirement and the excess biomass production rate. It also influences strongly the concentrations of various components in the effluent, although they will also be influenced by the system configuration, as we will see.

If SRT is to continue to be a variable of fundamental importance, it is necessary to have a means of controlling it in practice and of setting it during simulations. This is best done through modification of the Garrett[6] configuration discussed in Section 5.3.1. As before, biomass should be wasted directly from the bioreactors, rather than from the biomass recycle flow, but in this case it should be from each bioreactor in proportion to its volume. The efficacy of this procedure can be seen in the following. Consider a system containing N bioreactors of volume V_i, containing MLSS at concentration X_{Mi}, with wastage from each at flow rate F_{wi}. In that case, the SRT is defined as:

$$\Theta_c = \frac{\Sigma V_i \cdot X_{Mi}}{\Sigma F_{wi} \cdot X_{Mi}} \tag{7.1}$$

If wastage is from each bioreactor in proportion to its volume, then:

$$F_{wi} = F_W(V_i/V_T) \tag{7.2}$$

where F_w is the total wastage flow rate and V_T is the total system volume. Substitution of Eq. 7.2 for F_{wi} in Eq. 7.1 and rearrangement gives:

$$\Theta_c = \frac{V_T \Sigma V_i \cdot X_{Mi}}{F_W \Sigma V_i \cdot X_{Mi}} \tag{7.3}$$

or

$$\Theta_c = \frac{V_T}{F_W} \tag{7.4}$$

which is the same as Eq. 5.3. Thus, if wastage is from each bioreactor in proportion to its volume, the SRT will be controlled solely by the total bioreactor volume and the total wastage flow rate, regardless of the MLSS concentration in each bioreactor. Because of its simplicity, this technique will be used in all simulations in this chapter, even though the wastage streams will not be shown in the schematic diagrams to simplify them.

7.1.3 Importance of Process Loading Factor

In Section 5.1.5 we argue that for a single CSTR at steady-state the SRT is a better control parameter than the process loading factor or F/M ratio (U). This does not mean that the process loading factor is unimportant, however. Rather, for a tanks-in-series system it is quite significant because the value in each bioreactor is proportional to the specific growth rate of the biomass in that bioreactor. This can be seen by reviewing Eq. 5.38 in Section 5.1.5. For the general situation under consideration here, it can be rewritten as

$$\mu_{Hi} = \frac{Y_H F(S_{SO} - S_{Si})}{V_i \cdot f_A \cdot X_{Mi}} \tag{7.5}$$

If for any given bioreactor, the assumption can be made that $S_{Si} << S_{SO}$, then

$$\mu_{Hi} = \frac{Y_H \cdot U_i}{f_A} \tag{7.6}$$

The active fraction, f_A, changes little from tank to tank and can be considered to be constant for a given system, as can Y_H. Consequently, the value of the process loading factor in each bioreactor (U_i) tells us how the state of the microbial community is changing as it moves from bioreactor to bioreactor in the chain. This is important information that cannot be gained from the overall system SRT. In Chapter 10, we see how control of the process loading factor in individual bioreactors can be used to influence the competition between members of the microbial community to produce activated sludges with desired characteristics.

In SBRs, influent is added for only a portion of the cycle, called the fill period. Equation 7.5 can be used to define an "instantaneous" process loading factor for an SBR by letting F be the volumetric flow rate during the fill period and by ignoring S_{Si}, which is time dependent. Short fill periods require a large F for a given delivered volume, thereby giving a large instantaneous process loading factor. Conversely, long fill periods require a small F, giving low instantaneous process loading factors. While only an approximation, the calculated instantaneous process loading factor in an SBR can be used in a manner analogous to the individual reactor process loading factors in a tanks-in-series system.

7.2 CONVENTIONAL AND HIGH PURITY OXYGEN ACTIVATED SLUDGE

7.2.1 Description

Figure 7.1 presents a schematic diagram of the system used to simulate conventional and high purity oxygen activated sludge systems. All influent and all biomass recycle enters the first bioreactor and passes from bioreactor to bioreactor down the chain. For the purposes of this chapter, the bioreactors were considered to be of equal volume, but different residence time distributions can be attained by using bioreactors of different size.[12] The influent flow rate used in simulations was 1000 m^3/day and the volume of each bioreactor was 50 m^3, giving a total system volume of 250 m^3 and a system hydraulic residence time (HRT) of 6 hr, which is a value commonly used in practice. The biomass recycle flow rate was fixed at 500 m^3/day, unless otherwise specified, giving a recycle ratio, $\alpha(F_r/F)$, of 0.5, which is also commonly used in practice. All bioreactors are aerobic and the dissolved oxygen concentration was controlled at 2.0 mg/L in each, thereby eliminating denitrification. This concentration was chosen because it is easily and economically achieved in practice and allows unhindered nitrification, as shown in Figure 6.7.

To allow direct comparison to the results in Chapter 6, the characteristics of the influent were assumed to be those listed in Table 6.6 and the values of the kinetic and stoichiometric coefficients of the biomass growing on that influent were assumed to be those in Table 6.3.

7.2.2 Effect of SRT on Steady State Performance

The effect of SRT on the steady-state performance of the tanks-in-series reactor system is given by the solid curves in Figure 7.2, which show the concentrations in the last bioreactor of the chain, and represent the concentrations entering the final settler. For comparison, the dashed curves show the performance of a single CSTR with a volume equal to the total system volume of the chain, i.e., the systems have the same HRT. Because the influent to the systems contains particulate organic matter, simulations were only performed at SRTs in excess of one day.

The most striking thing about the curves in Figure 7.2 is that although the biomass and MLSS concentrations in the two systems are essentially the same, the tanks-in-series system achieves slightly better effluent quality, with lower concentra-

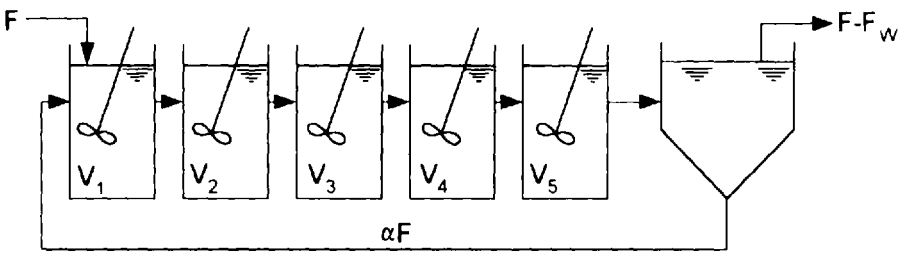


Figure 7.1 Schematic diagram of five CSTRs in series with all influent and all biomass recycle to the first reactor. Although not shown, solids wastage is directly from all reactors. This configuration simulates CAS and HPOAS systems.

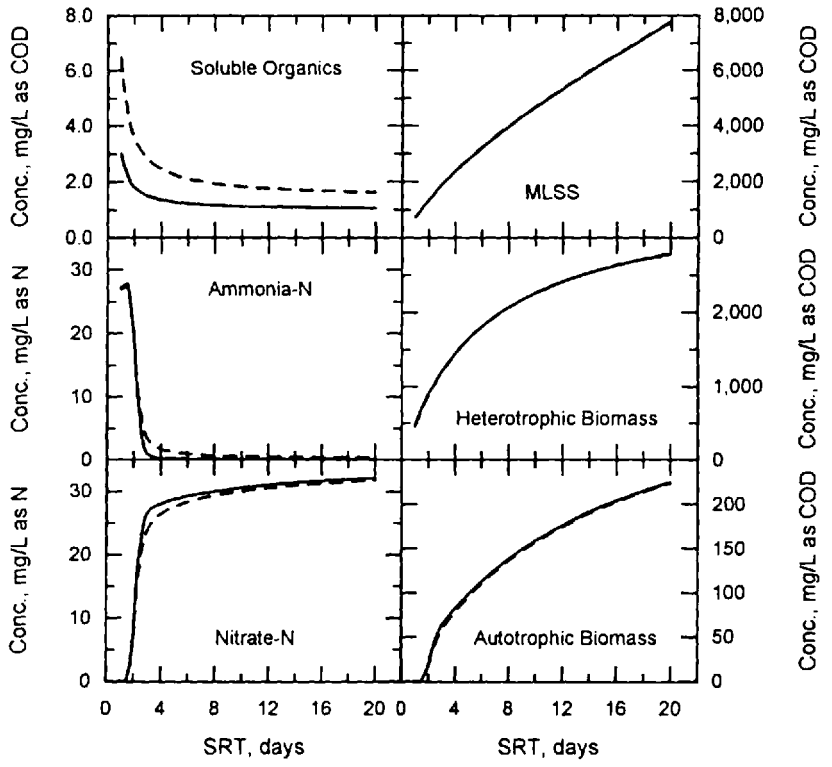


Figure 7.2 Effect of SRT on the steady-state concentrations of various constituents in the last reactor of the CAS or HPOAS system depicted in Figure 7.1. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Influent flow = 1000 m^3/day. Influent concentrations are given in Table 6.6, biomass recycle flow = 500 m^3/day; volume of each reactor = 50 m^3. Parameters are listed in Table 6.3. The dissolved oxygen concentration was held constant at 2.0 mg/L.

tions of soluble organic substrate and ammonia-N and slightly higher concentrations of nitrate-N. This is because microbial growth and substrate utilization behave in a first order manner when the substrate concentration is low, and first order reactions achieve a higher extent of reaction in chains of reactors than in single CSTRs.[12] It should be noted, however, that in contrast to previous simulation studies,[5] the effect of bioreactor configuration on soluble organic substrate concentration is not large. This is because ASM No. 1 includes the generation of soluble organic substrate due to biomass death and lysis, which was not considered in previous studies. This small effect of bioreactor configuration on organic substrate removal is consistent with experimental observation.[23] It will also be noted that the effect of bioreactor configuration on nitrification is larger than the effect on soluble substrate removal. This follows from the fact that most nitrogen released as a result of heterotrophic biomass death and lysis will be used for heterotrophic biomass regrowth, with only a small fraction being available as the substrate for nitrifier growth. It is doubtful that such a large effect would be seen in practice, however, because a single CSTR will offer

more protection to the nitrifiers against high concentrations of inhibitory organic matter.

Since microbial death and lysis is a first order reaction, one might expect the tanks-in-series system to exhibit more biomass loss than the single CSTR, thereby resulting in lower concentrations of heterotrophic and autotrophic biomass. The fact that this does not occur may be surprising at first, but in reality, there is a logical reason. The residence time of the biomass in the system is equal to the SRT and is independent of the residence time of the fluid within the system, i.e. the HRT. Because the biomass recycle flow rate is of the same magnitude as the influent flow rate, the biomass is recycled around the system many times within one SRT, causing it to approach a completely mixed condition even though the fluid passes through a cascade. As long as the SRT is large with respect to the system HRT, the tendency of the biomass recycle to make the biomass well mixed will overpower the cascade effect of the fluid in the chain in determining the extent of reaction of the biomass by death and lysis. Thus, the biomass concentration in the chain will be very similar to the concentration in a CSTR with equal SRT. The tanks-in-series system will only experience more biomass loss when the HRT approaches the SRT so that the biomass experiences a cascade effect during its residence time.

Figure 7.3 shows the effect of SRT on the total system oxygen requirement and the solids wastage rate. Again, for comparison, the response of a single CSTR is shown as dotted curves. The lack of difference between the two systems at longer SRTs is consistent with the arguments in the preceding paragraph concerning the homogeneous nature of the MLSS caused by biomass recycle. The slight difference between the oxygen requirements in the two types of bioreactors at lower SRTs is due primarily to the greater degree of nitrification in the tanks-in-series system. The similarity in total oxygen requirement and solids wastage rate means that these characteristics can be estimated with the simple equations for a single CSTR in Chapter

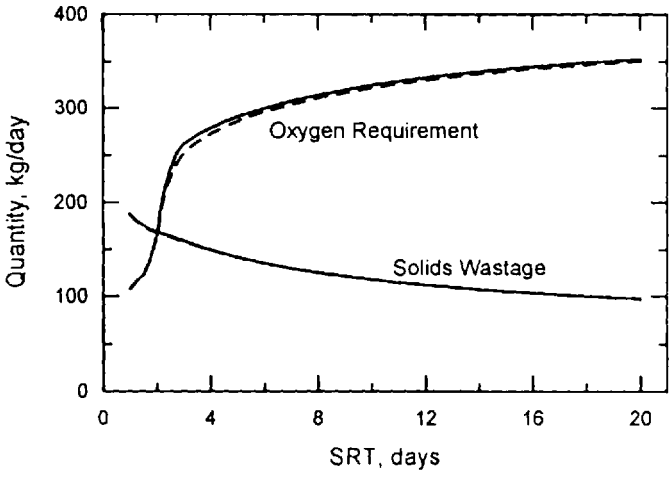


Figure 7.3 Effect of SRT on the total steady-state oxygen requirement and solids wastage rate for the CAS or HPOAS system depicted in Figure 7.1 operating under the conditions listed in Figure 7.2. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3.

5, even when a tanks-in-series reactor is to be used. Furthermore, these similarities tell us that the mass of MLSS in the two systems will be similar at a fixed SRT. Consequently, the simple equations for a CSTR can also be used to estimate the $X_M \cdot \tau$ product. This ability to use the simple equations is very helpful during process design, as seen in Chapter 10.

7.2.3 Dynamic Performance

Figures 7.2 and 7.3 present only the steady-state responses of the chain of CSTRs, but, as seen in Chapter 6, biochemical operations are often subjected to diurnal changes in loading. Thus, it is important to consider how bioreactor configuration influences dynamic response. Because 10 days is a commonly used value for the SRT in activated sludge systems, the dynamic response of the system in Figure 7.1 was studied at that SRT by imposing the diurnal loading pattern shown in Figure 6.2. The average flow rate to the system was 1000 m^3/day, the same as in the steady-state simulations, and the flow-weighted average concentrations of the various constituents were the same as those in Table 6.6. Because the alkalinity of a wastewater is determined primarily by the characteristics of the carriage water, rather than by the waste constituents, the alkalinity was assumed to be constant at the value in Table 6.6.

The output response of the tanks-in-series system to repeated application of the loading pattern in Figure 6.2 is shown in Figure 7.4 by the solid curves. For comparison, the response of a single CSTR with the same SRT and average HRT is shown by the dashed curves. Only the soluble constituents are shown because the particulate constituents are relatively constant due to the length of the SRT relative to the HRT. Examination of the curves shows that the variability in the effluent concentrations of the two reactants, organic substrate and ammonia-N, was much less for the chain of CSTRs than for the single CSTR. This is a direct result of the hydraulic differences between the two systems. Any change in the influent to a single CSTR is seen instantly in the effluent because the bioreactor and the effluent have the same concentration. The chain experiences time delays, however, as the fluid moves from tank to tank. This gives a greater opportunity for degradation or transformation of the substrates. The effect is particularly significant for ammonia because of the small maximum specific growth rate of nitrifying bacteria.

An important assumption in all of the simulations was that the dissolved oxygen concentration was constant at 2.0 mg/L, a situation that can be achieved in practice through DO control. We see in Figure 6.13, however, that imposition of the typical diurnal loading pattern causes considerable variation in the oxygen requirement within a single CSTR and that failure to meet that requirement causes a deterioration in system performance. Thus, it is important to consider how the diurnal loading influences the oxygen requirement in each bioreactor of the chain so that appropriate provisions can be made for delivering the needed amount. This is shown in Figure 7.5. In this figure, the vertical lines represent the range in the oxygen consumption rate experienced in each bioreactor as the loading varies over a 24-hr period. As expected, because the first bioreactor receives the variable waste load directly, it experiences the greatest variability, exhibiting a four-fold range in oxygen consumption rate. The second bioreactor receives only the substrate that is not removed by the first, and the magnitude of the oxygen requirement and the variability associated

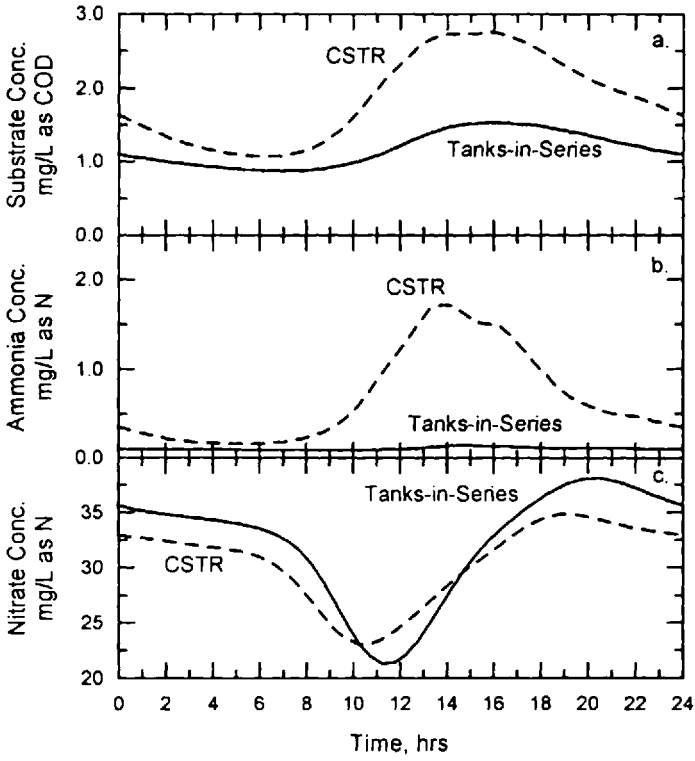


Figure 7.4 The time dependent response from the CAS or HPOAS system depicted in Figure 7.1 when subjected to the diurnal loading patterns shown in Figure 6.2. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Average influent flow = 1000 m^3/day. Average influent concentrations are given in Table 6.6. Biomass recycle flow = 500 m^3/day; volume of each reactor = 50 m^3; SRT = 10 days. Parameters are listed in Table 6.3. The dissolved oxygen concentration was held constant at 2.0 mg/L.

with it are lower, and so on down the chain. This means that any system that uses a tanks-in-series type reactor must be designed to handle much different oxygen requirements at different points in the system. The dots on the bars represent the steady-state oxygen requirement in each bioreactor caused by the entrance of a constant flow and concentration of wastewater to the system. They clearly show two things: (1) a system designed only on the basis of the steady-state requirement would be inadequate during periods of peak loading, and (2) equalization of influent flow and concentration allows use of a smaller oxygen transfer system. The figure also shows that the peak to average ratio changes from bioreactor to bioreactor, becoming particularly large in bioreactor three. This is primarily due to nitrification, which occurs further downstream than organic substrate removal during periods of high loading. This is because nitrifying bacteria have a much smaller maximum specific growth rate than heterotrophic bacteria. As a consequence, the nitrifiers cannot increase their metabolism as much in response to the increased input rate of substrate, allowing a larger fraction of the ammonia-N to pass through to the downstream

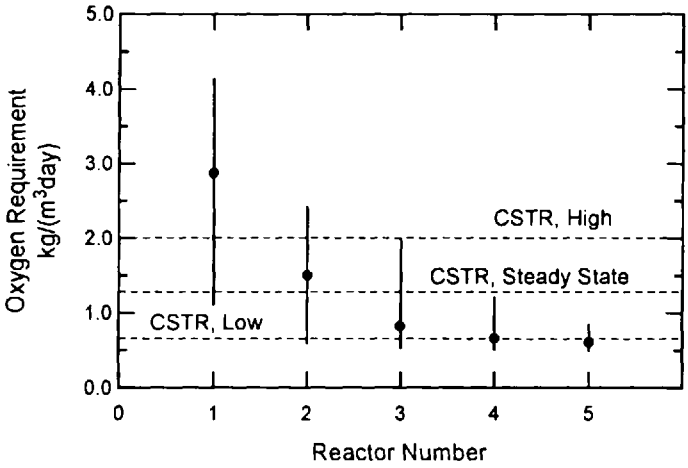


Figure 7.5 The variability of the oxygen requirement in each reactor for the CAS or HPOAS system described in Figure 7.4. The vertical lines represent the range observed over a 24 hr period and the dots represent the requirement when the influent flow and concentrations are constant at the flow-weighted average values (steady-state). For comparison, the dashed lines show the extreme diurnal requirements, as well as the steady-state requirement, for a single CSTR with a volume of 250 m^3.

reactors during periods of high loading. For comparison purposes, the dashed lines in Figure 7.5 show the high, low, and steady-state oxygen requirements in a single CSTR. They show clearly that the first tanks in the chain require much higher oxygen input rates and experience a greater variation in requirement due to the lower degree of dampening associated with the small bioreactor. One reason for use of single CSTRs is to take advantage of the smaller variability in oxygen requirement.

7.2.4 Variations Within the System

The dots in Figure 7.5 demonstrate that each tank in a chain of CSTRs has a different reaction rate associated with it even when the system is operating at steady-state. This suggests that further investigation of that aspect of system performance would be worthwhile. Thus, changes in the steady-state concentrations of three soluble constituents and the MLSS are shown in Figure 7.6 for three different SRTs.

The first thing to note in Figure 7.6 is that the MLSS concentration has little tank-to-tank variability. The same is true for the heterotrophic and autotrophic biomass concentrations, although they are not shown. This lack of change is because the SRT is large relative to the HRT, causing the biomass recycle to make the mixed liquor homogeneous, as discussed previously. A decline in the MLSS concentration as the solids move down the chain would only occur when the HRT approaches the SRT.

In contrast to the behavior of the particulate material, there are significant decreases in the concentrations of soluble organic substrate and ammonia-N as the fluid moves down the chain. All substrate is added to the first tank where it is mixed with biomass that has been returned from the settler. The presence there of high

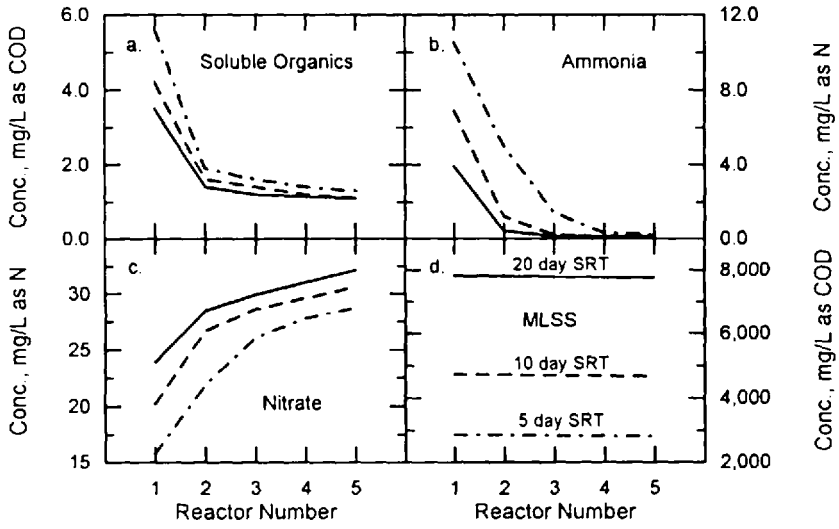


Figure 7.6 Effect of SRT on the steady-state concentrations of various constituents in each reactor of the CAS or HPOAS system described in Figure 7.2.

concentrations of soluble organic matter and ammonia-N allows both the heterotrophic and autotrophic biomass to attain relatively high specific growth and substrate removal rates, which, when combined with the high biomass concentration, gives rapid removal rates. This causes considerable reduction in soluble organic and ammonia-N concentrations in the first bioreactor relative to the feed. Since the mass flow rates of soluble organic substrate and ammonia-N into the second bioreactor are much less than the rates into the first, the specific growth and substrate removal rates are reduced, but the high biomass concentration allows maintenance of an overall rate that is sufficiently large to cause more substrate to be removed. In other words, the process loading factor is very high in the first bioreactor, but lower in the later ones. This means that the activity of the biomass varies as it moves from tank to tank.

In spite of the similarity in mechanism associated with the removal of soluble organic matter and ammonia-N, comparison of parts a and b of Figure 7.6 shows that there are significant differences in their concentration profiles, particularly at low SRT. These differences result in part from the low maximum specific growth rate associated with autotrophic biomass. Because of it, the percentage of the influent ammonia-N removed in the first tank is smaller than the percent of the soluble organic matter removed. This causes a greater percentage of the ammonia-N to move to the second tank, etc., thereby causing nitrification to occur further down the bioreactor chain than organic substrate removal. In addition, because the half-saturation coefficient for autotrophs is lower than the coefficient for heterotrophs, the autotrophs tend to grow near their maximal rate in the first few bioreactors, particularly at lower SRTs where their mass in the system is small. Consequently, tank-to-tank removal of ammonia-N is more nearly linear than is removal of organic substrate. In the final tanks, the ammonia-N concentration approaches zero, making the specific removal rate quite low, but the presence of a high autotrophic biomass concentration keeps

the overall rate high enough to scavenge any remaining ammonia from the system. The heterotrophic biomass respond in a similar manner, but the organic substrate concentration does not approach zero because substrate is continually being resupplied by biomass death and lysis. Thus, an equilibrium is achieved in which the use of organic substrate just balances its release, causing a pseudo steady-state concentration to be maintained.

The tank-to-tank variability in the activity of the biomass is reflected best by the oxygen requirement, which is shown in Figure 7.7. There it can be seen that as in Figure 7.5, more oxygen is used in the first tanks because more substrate is removed in them. Furthermore, oxygen consumption in the later tanks reaches a relatively constant value reflective of endogenous metabolism and decay of the biomass.

The effect of SRT on the tank-to-tank performance is shown in Figures 7.6 and 7.7 by the three different curves. Because the system with the greatest SRT contains more biomass, organic substrate removal and nitrification occur more rapidly in it, causing lower concentrations to be attained in early tanks than in systems with lower SRT. In addition, at lower SRT much less removal of ammonia-N occurs in the early tanks, for reasons discussed above. Comparison of parts b and c of Figure 7.6 shows that even though about the same ammonia-N concentration is ultimately achieved at all three SRTs, more nitrate-N is produced at higher SRTs. This is due to the greater death and lysis that occurs at longer SRTs, releasing more nitrogen, which acts as substrate for the nitrifying bacteria, producing more nitrate-N. SRT has a fairly complicated effect on the pattern of oxygen consumption down the chain because it is influenced by organic substrate removal, nitrification, and biomass decay, all of which depend on the SRT. The longer the SRT, the greater the organic substrate removal in the first tank, and the greater the oxygen requirement associated with that activity. Likewise, the longer the SRT, the greater the importance of biomass decay in the later tanks, also causing greater oxygen consumption. In intermediate tanks, however, differing degrees of nitrification occur at different SRTs,

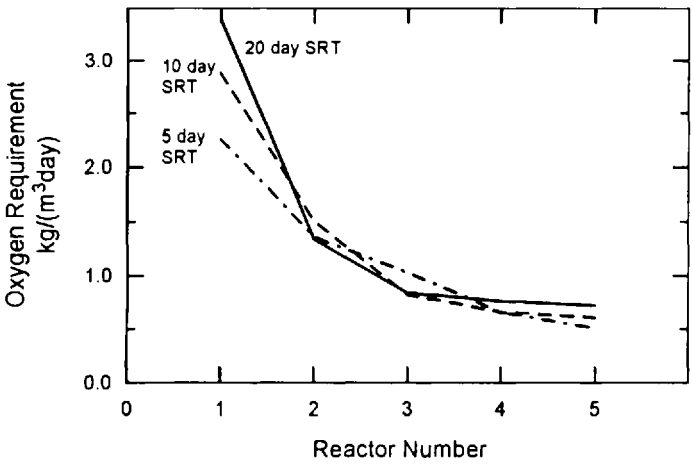


Figure 7.7 Effect of SRT on the steady-state oxygen requirement in each reactor of the CAS or HPOAS system described in Figure 7.2.

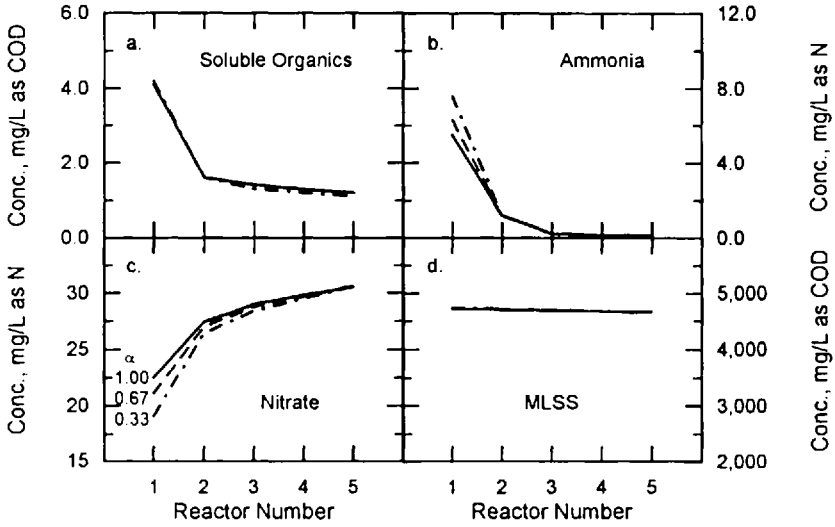


Figure 7.8 Effect of the biomass recycle ratio (α) on the steady-state concentrations of various constituents in each reactor of the CAS or HPOAS system described in Figure 7.2. SRT = 10 days.

depending on the amount that has occurred in the early tanks. This makes it difficult to generalize about the effect of SRT on the amount of oxygen required in such tanks. Each situation will depend on the character of the wastewater. Consequently, any system employing this configuration must be built to ensure operational flexibility.

Figures 7.8 and 7.9 show that as with a single CSTR, the rate of biomass recycle around a simple chain of CSTRs has relatively little effect on system per-

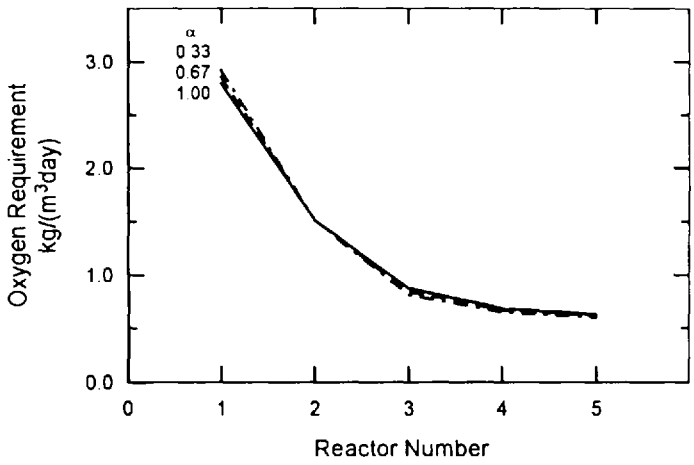


Figure 7.9 Effect of the biomass recycle ratio (α) on the steady-state oxygen requirement in each reactor of the CAS or HPOAS system described in Figure 7.2. SRT = 10 days.

formance. Theoretically, higher recycle flows around a chain of reactors act to make the system more homogeneous.[12] For example, in the extreme, an infinite recycle ratio would make the system behave like a CSTR. In practice, however, biomass recycle ratios in excess of 1.0 are seldom used, and thus the simulations presented in Figures 7.8 and 7.9 were limited to that value. Although the profiles do show slightly more homogeneity at higher recycle ratios, the effects are not significant, demonstrating that biomass recycle cannot be used as an operational tool for influencing substrate removal or oxygen uptake in a simple chain of CSTRs.

7.3 STEP FEED ACTIVATED SLUDGE

7.3.1 Description

Figure 7.10 presents the schematic diagram for the configuration used to simulate an SFAS system. As in Figure 7.1, five equal sized CSTRs in series were used, with all biomass recycle to the first bioreactor, but in this case the feed was distributed evenly among the bioreactors. All other characteristics of the system, including the flow rates, feed concentrations, etc., were the same as those used to simulate the performance of the tanks-in-series system.

7.3.2 Effect of SRT on Steady-State Performance

The effect of SRT on the steady-state performance of the system in Figure 7.10 is shown by the solid curves in Figure 7.11. For comparison, the performance of a single CSTR with volume equal to the total volume of the SFAS system is shown by the dashed curves. Examination of parts a, b, and c of the figure reveals that the effluent concentrations of soluble constituents from the SFAS system are slightly higher than those from the single CSTR. The differences are relatively small, however, and would probably be difficult to distinguish in practice. Further evidence for the similarity in performance between the SFAS system and a single CSTR is provided in Figure 7.12, where it is seen that the oxygen requirements and solids wastage rates are almost identical in the two systems. This suggests that the mass of MLSS in the SFAS system is approximately the same as that in a single CSTR with the same SRT.

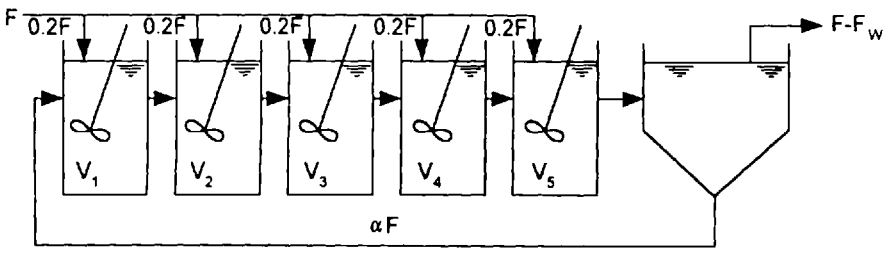


Figure 7.10 Schematic diagram of five CSTRs in series with all biomass recycle to the first reactor and the influent distributed evenly among the reactors. Although not shown, solids wastage comes directly from all reactors. This configuration simulates step feed activated sludge (SFAS).

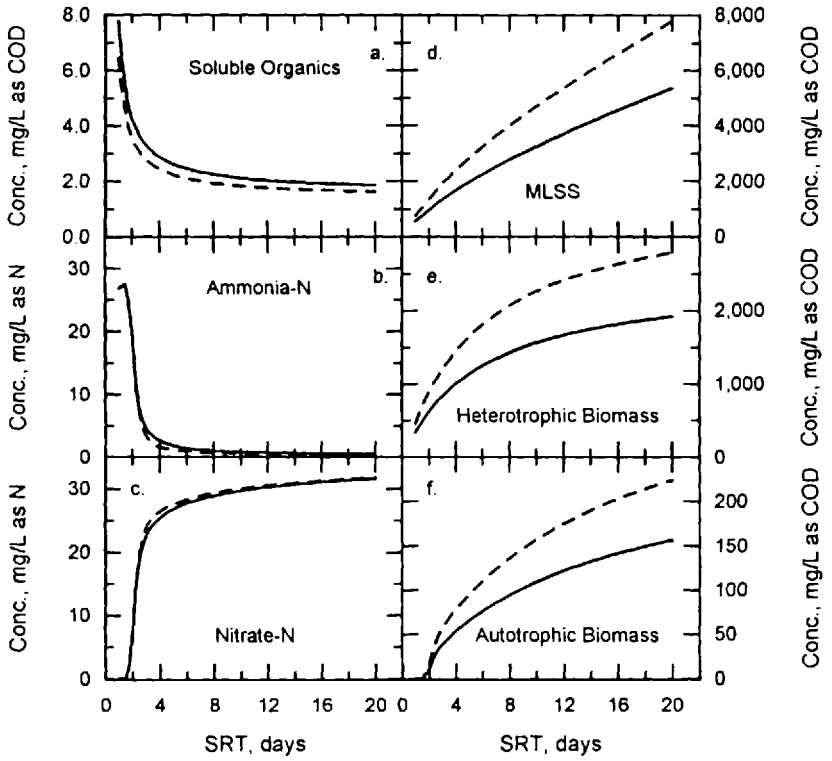


Figure 7.11 Effect of SRT on the steady-state concentrations of various constituents in the last reactor of the SFAS system depicted in Figure 7.10. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Influent flow = 1000 m^3/day. Influent concentrations are given in Table 6.6. Biomass recycle flow = 500 m^3/day; volume of each reactor = 50 m^3. Parameters are listed in Table 6.3. The dissolved oxygen concentration was held constant at 2.0 mg/L.

The failure of the SFAS system to perform as well as a simple CSTR (and by extension, as well as a simple chain) is a direct result of the feed distribution. If the effect of the wastage flow rate, F_w, is assumed to be small, a mass balance on MLSS about the final settler in a system with biomass recycle reveals that:

$$X_{M_r} \approx \frac{1 + \alpha}{\alpha} X_M \tag{7.7}$$

where X_M is the MLSS concentration entering the settler from the last bioreactor, X_{M_r} is the concentration in the biomass recycle flow, and α is the recycle ratio (F_r/F). For the operational conditions described in the legend of Figure 7.11, α has a value 0.5, making the MLSS concentration in the biomass recycle stream three times the concentration in the last bioreactor of the chain. Because the influent flow is distributed evenly among the bioreactors, only one-fifth of it enters the first bioreactor, which provides little dilution of the biomass recycle flow and makes the MLSS concentration in the first bioreactor very high. Furthermore, as the MLSS moves from bioreactor to bioreactor, more influent flow enters, diluting it. This establishes

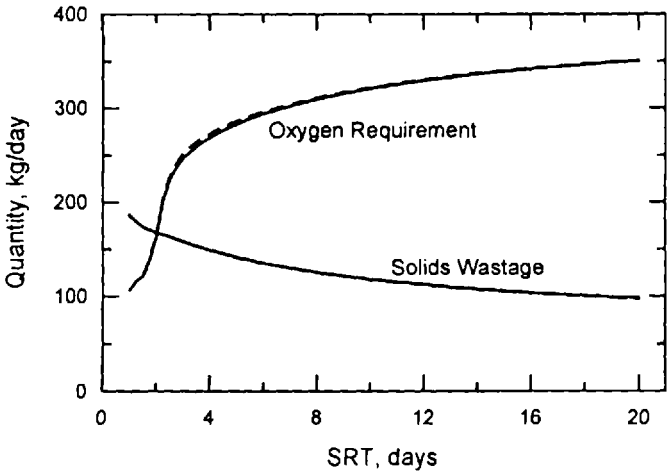


Figure 7.12 Effect of SRT on the total steady-state oxygen requirement and solids wastage rate for the SFAS system depicted in Figure 7.10 operating under the conditions listed in Figure 7.11. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3.

a MLSS concentration gradient through the system. It will be recalled from Figure 7.12, however, that for a given SRT, the SFAS system has the same mass wastage rate of biomass as a single CSTR, suggesting that both systems contain the same mass of MLSS. Because of this, and because the volume of the single CSTR is equal to the total volume in the SFAS chain, the MLSS concentration in the single CSTR must equal the average concentration in the SFAS system. Consequently, since the SFAS system contains a MLSS concentration gradient, the MLSS concentration in its last tank must be less than the concentration in the single CSTR. That this is true is shown in Figure 7.11. Now, since the influent flow rate into the last bioreactor of the chain is one-fifth the flow rate into the single CSTR, and since the volume of the last bioreactor is one-fifth the volume of the single CSTR, the mass flow rate of substrate per unit volume into the last bioreactor of the chain is the same as that into the single CSTR. However, since the biomass concentration is less, there is less biomass per unit of substrate added, i.e., the process loading factor is higher, allowing less substrate to be removed, as observed. Thus, the very configuration of the SFAS system prevents it from performing as well as a single CSTR or a simple chain. This raises the question as to why such a system would be used. One reason can be seen by examining the dynamic performance of the system.

7.3.3 Dynamic Performance

The dynamic performance of the SFAS system when subjected to the same diurnal loading pattern as the tanks-in-series system is shown in Figure 7.13. Again, the response of a single CSTR is shown for comparison. As would be expected from the entrance of feed directly into the last bioreactor with its low biomass concentration, the dynamic performance of the SFAS system is worse than that of the CSTR. Furthermore, although the difference is relatively small for soluble organic substrate,

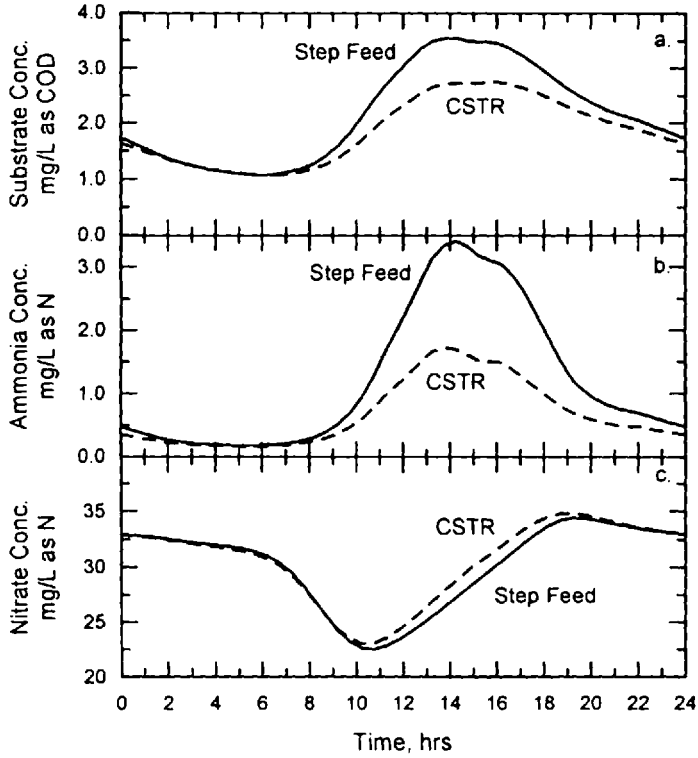


Figure 7.13 The time dependent response from the SFAS system depicted in Figure 7.10 when subjected to the diurnal loading patterns shown in Figure 6.2. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Average influent flow = 1000 m^3/day. Average influent concentrations are given in Table 6.6. Biomass recycle flow = 500 m^3/day; volume of each reactor = 50 m^3; SRT = 10 days. Parameters are listed in Table 6.3. The dissolved oxygen concentration was held constant at 2.0 mg/L.

it is substantial for nitrification, with the maximum effluent ammonia-N concentration being almost twice as high as that from the CSTR. This is a direct result of the low maximum specific growth rate of nitrifying bacteria, which prevents them from responding rapidly enough to the changing input rate of nitrogen into the last bioreactor. In addition, comparison of the effluent ammonia-N concentration to that from the simple chain in Figure 7.4 reveals that a SFAS system subjected to diurnal loading produces an effluent that is much worse than the chain. Thus, the dynamic performance of the SFAS system does not justify its use.

A major reason for employing a SFAS system, however, is revealed by Figure 7.14, where it can be seen that the distribution of the load along the length of the chain gives an oxygen consumption pattern that is much closer to that of a single CSTR than was the simple chain. This makes it much easier to provide the needed oxygen. Consequently, if a tanks-in-series system is not performing properly because of an inability to supply sufficient oxygen to the first tanks in the system, redistribution of a portion of the influent to tanks further down the chain may be able to

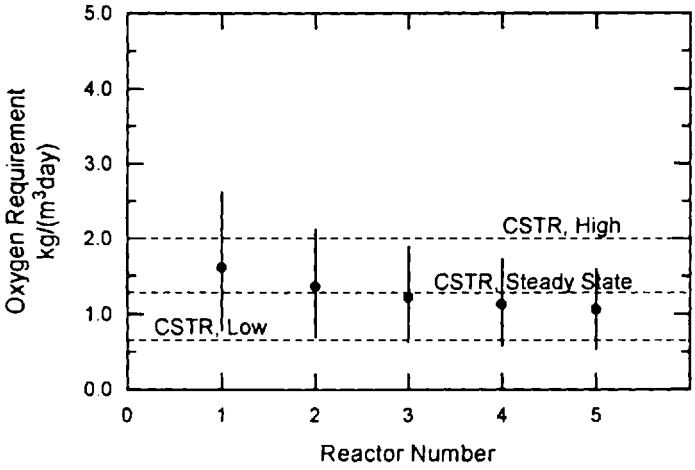


Figure 7.14 The variability of the oxygen requirement in each reactor for the SFAS system described in Figure 7.13. The vertical lines represent the range observed over a 24 hr period and the dots represent the requirement when the influent flow and concentrations are constant at the flow-weighted average values (steady-state). For comparison, the dashed lines show the extreme diurnal requirements, as well as the steady-state requirement, for a single CSTR with a volume of 250 m^3.

alleviate the problem and produce an acceptable effluent, even though it may not be as good as that from a properly designed tanks-in-series system.

These simulations have only examined the case of distributing the feed evenly along the chain. Obviously, many possible distribution patterns exist and each will have a different impact on system performance. A study of the impact of alternative patterns will be left as an exercise for the reader.

7.3.4 Variations Within the System

Tank-to-tank variations in concentration within an SFAS system are shown in Figure 7.15 for steady-state operation. Part d of the figure shows the gradient in MLSS concentration discussed earlier, which is steeper at longer SRTs. This gradient is the other major reason for using the SFAS system.[11] Occasionally the settling characteristics of activated sludge will deteriorate so that the mass loading rate of solids entering the final settler must be reduced to prevent the settler from failing. Since the flow rate of wastewater being treated cannot be reduced, this requires reduction of the MLSS concentration. One way to do this is to reduce the SRT, but this may negatively impact other aspects of process performance, such as causing loss of nitrification. If the system were operating in the tanks-in-series mode, however, and contained the appropriate piping, switching to the SFAS mode would establish the MLSS concentration gradient, thereby reducing the concentration of MLSS entering the final settler while maintaining the same system SRT. Then, after the biomass settling characteristics had returned to normal, operation could be returned to the tanks-in-series mode.

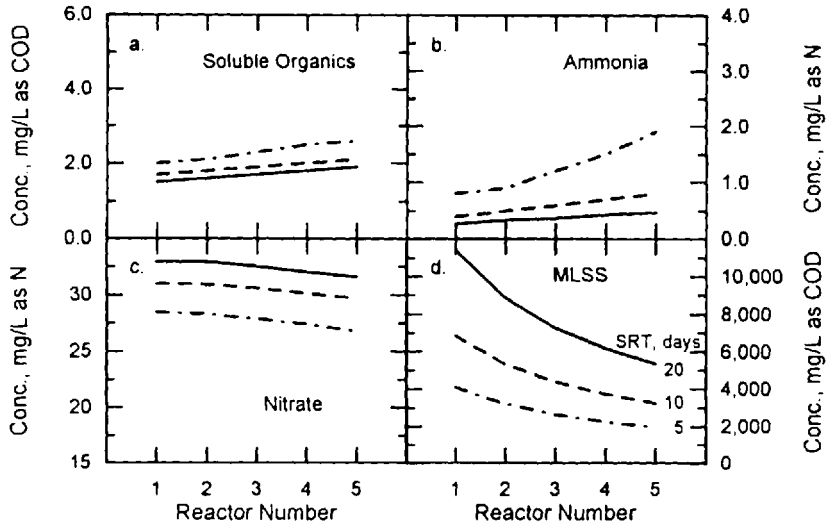


Figure 7.15 Effect of SRT on the steady-state concentrations of various constituents in each reactor of the SFAS system described in Figure 7.11.

The concentration profiles of soluble organic substrate and ammonia-N are the opposite of the MLSS profiles. This follows from the fact that less biomass is available in the later bioreactors to act on the substrate and ammonia entering them, thereby causing the concentrations to rise. In other words, the process loading factor increases from bioreactor to bioreactor down the chain. The tank to tank changes in substrate and ammonia-N concentrations are less at longer SRTs, however, because the larger mass of biomass in the system at longer SRTs allows more complete reaction. This is particularly evident for nitrification, which is a slower reaction than organic substrate removal.

The steady-state profiles in oxygen requirement through the system are shown in Figure 7.16. They respond more systematically to changes in SRT than the profiles through the tanks-in-series system, shown in Figure 7.7, primarily because the soluble organic substrate and ammonia-N profiles are less severe. The major impact of an increase in SRT is to increase the oxygen requirement by increasing the importance of biomass decay.

The effect of the biomass recycle ratio on the steady-state concentration profiles is shown in Figure 7.17, while the effect on the oxygen requirement is illustrated in Figure 7.18. Examination of those figures shows that changes in the recycle ratio have a greater effect on the effluent quality of this system than on that of the simple chain (Figure 7.8) or the single CSTR (upon which the recycle ratio has no effect). This is due entirely to the hydraulics of the system. All of the systems whose responses are shown in Figures 7.17 and 7.18 have the same SRT. Thus, they all contain the same mass of MLSS. Furthermore, they all have the same reactor volumes, and they all have the same average MLSS concentrations. However, as shown by Eq. 7.7, as the recycle ratio is increased, the difference between the concentration of MLSS entering the settler and the concentration leaving in the recycle flow decreases. This means that the MLSS concentration gradient along the bioreactor chain

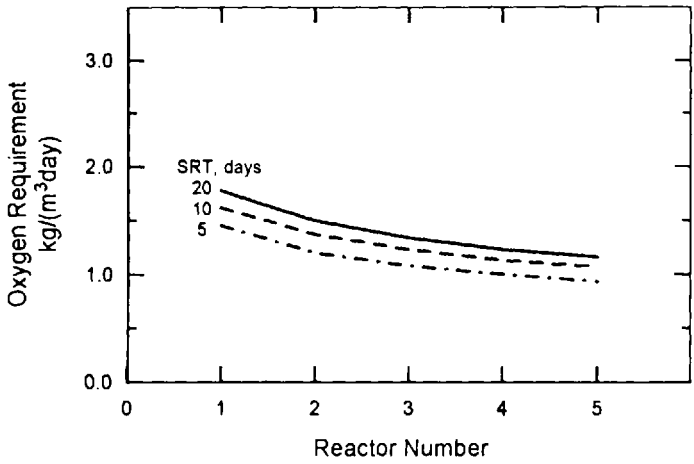


Figure 7.16 Effect of SRT on the steady-state oxygen requirement in each reactor of the SFAS system described in Figure 7.11.

is reduced, as shown in Figure 7.17d. Furthermore, since the recycle ratio doesn't influence the average MLSS concentration, the MLSS concentration in the first bioreactors must decrease with increasing recycle while the MLSS concentration in the last ones must increase. This decreases the process loading factor in the last bioreactors, allowing more substrate and ammonia-N to be removed, thereby making their effluent concentrations lower. Finally, since an increase in the recycle ratio

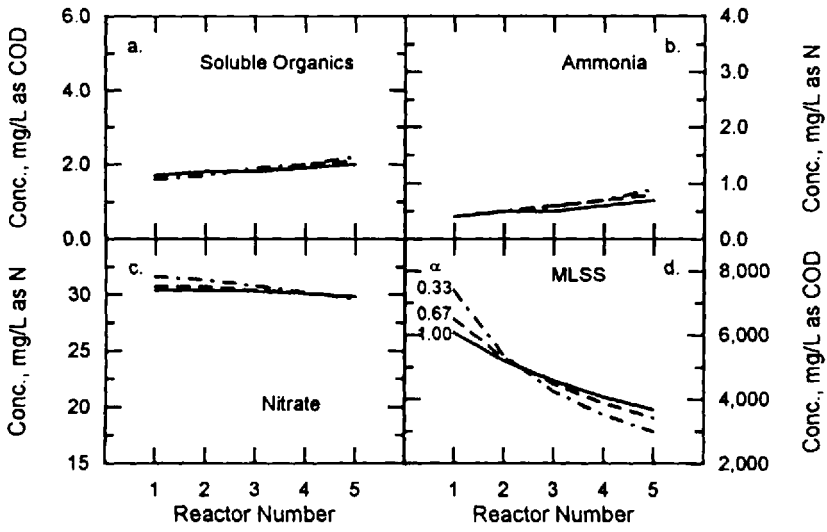


Figure 7.17 Effect of the biomass recycle ratio (α) on the steady-state concentrations of various constituents in each reactor of the SFAS system described in Figure 7.11. SRT = 10 days.

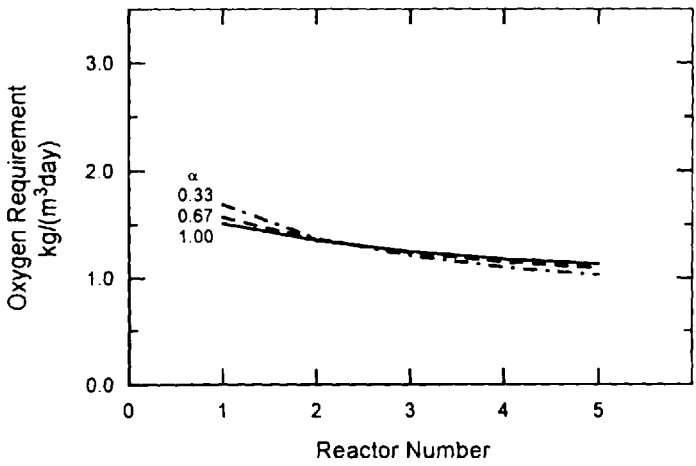


Figure 7.18 Effect of the biomass recycle ratio (α) on the steady-state oxygen requirement in each reactor of the SFAS system described in Figure 7.11. SRT = 10 days.

makes the system more homogeneous, it will make the utilization of oxygen more uniform, as shown in Figure 7.18.

Although the effects associated with changes in the recycle ratio are relatively small for this system configuration, they demonstrate clearly that the performance of suspended growth biological reactors is not always independent of the recycle ratio. In fact, it has an even stronger effect on the next reactor configuration.

7.4 CONTACT STABILIZATION ACTIVATED SLUDGE

7.4.1 Description

As described previously, in the contact stabilization version of the activated sludge process, the biomass recycle flow enters a bioreactor (stabilization basin) where it undergoes aeration and reaction prior to its return to the bioreactor receiving the influent wastewater (contact tank). This system can be simulated as a two bioreactor chain, as shown in Figure 7.19, in which all biomass recycle enters the first bioreactor and all influent enters the second. To maintain consistency with previous simulations, a total bioreactor volume of 250 m^3 was assumed, split evenly between the two bioreactors. The influent flows and concentrations were the same as those used in previous simulations.

7.4.2 Effect of SRT on Steady State Performance

The effect of the SRT on the steady-state performance of a CSAS system is shown by the solid curves in Figure 7.20, while the dashed curves present the performance of a single CSTR for comparison. Examination of the figure reveals that the effluent concentrations of soluble organic matter and ammonia-N are both higher than that of the CSTR, showing that the contact stabilization system does not perform as well.

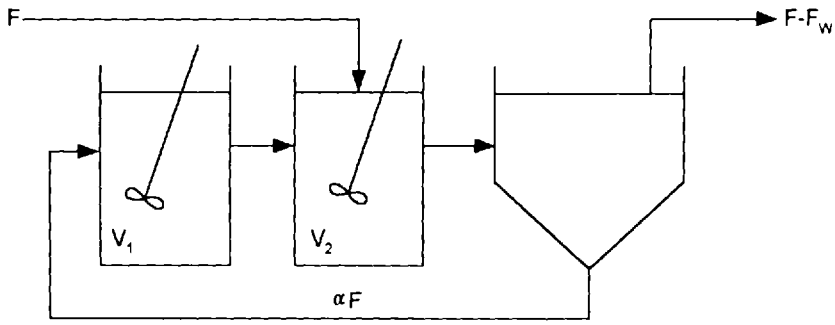


Figure 7.19 Schematic diagram of two CSTRs in series with all influent to the second (contact) reactor and all biomass recycle to the first (stabilization) reactor. Although not shown, solids wastage is directly from both reactors. This configuration simulates contact stabilization activated sludge.

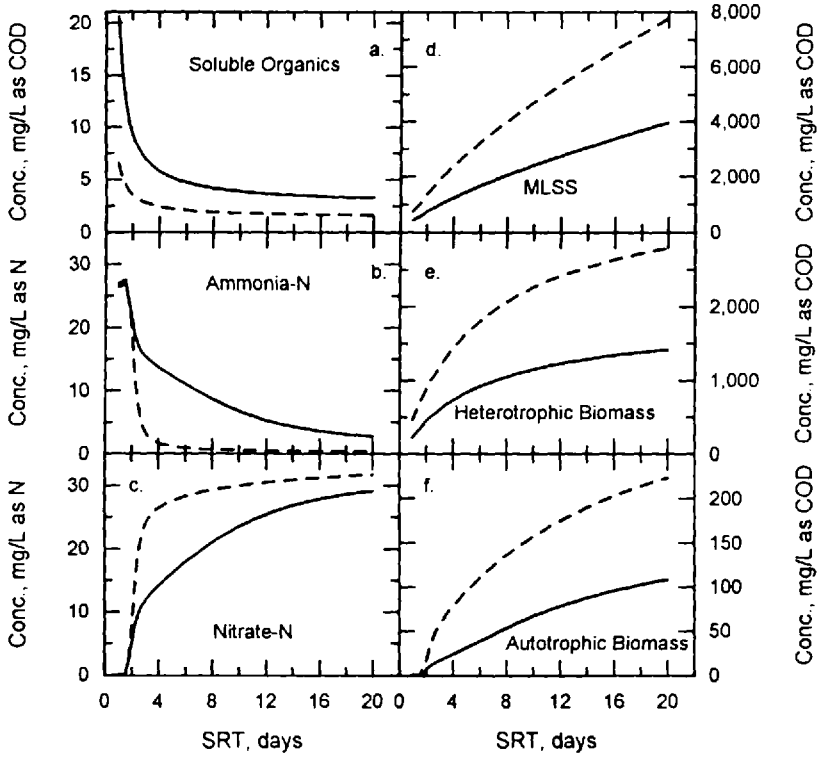


Figure 7.20 Effect of SRT on the steady-state concentrations of various constituents in the contact (second) reactor of the CSAS system depicted in Figure 7.19. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Influent flow = 1000 m^3/day. Influent concentrations are given in Table 6.6. Biomass recycle flow = 500 m^3/day; volume of each reactor = 125 m^3. Parameters are listed in Table 6.3. The dissolved oxygen concentration was held constant at 2.0 mg/L.

The reason becomes apparent when we consider the distribution of biomass within the system, as we did for the SFAS system. Examination of Figure 7.21 reveals that in spite of the slightly poorer performance, the CSAS system has almost the same excess biomass production as a single CSTR. As was argued for the SFAS system, this suggests that the CSAS system contains approximately the same mass of MLSS as the CSTR. However, the only flow entering the first bioreactor of the CSAS system is biomass recycle flow, which contains a much higher concentration of MLSS than the flow entering the settler. Since the CSAS system contains the same mass of MLSS as the CSTR, but the first bioreactor in it contains MLSS at a very high concentration, the MLSS concentration in the second bioreactor must be less than the concentration in the single CSTR, as shown in Figure 7.20. Furthermore, since the second bioreactor in the CSAS system receives the same influent wastewater flow rate as the single CSTR, but has only half the volume of the CSTR and contains a lower concentration of MLSS, the CSAS system cannot perform as well as the single CSTR. In other words, the process loading factor for the contact tank is much higher than it is for the CSTR, so less substrate will be removed. However, as we will see later, performance of the CSAS process can be changed by altering both the relative volumes of the two bioreactors and the recycle flow rate. This means that the degree of difference in performance between the two systems depends on the configuration chosen for the CSAS system. In addition, wastewater characteristics influence system performance. For example, colloidal and particulate organic matter are entrapped in the MLSS in the contact tank and undergo more complete biodegradation in the stabilization tank. This suggests that the CSAS system is better suited for wastewaters containing a higher fraction of their organic content in the colloidal form than in the soluble form. These and other factors influencing system performance will be discussed later when we consider the choice of system configuration

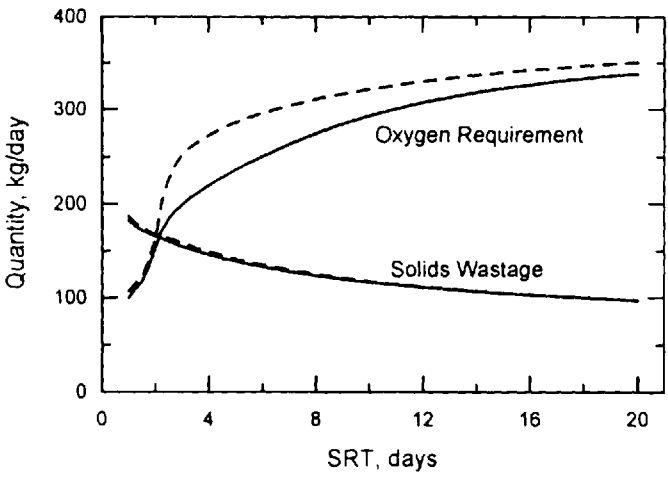


Figure 7.21 Effect of SRT on the total steady-state oxygen requirement and solids wastage rate for the CSAS system depicted in Figure 7.19 operating under the conditions listed in Figure 7.20. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3.

for activated sludge systems. For now, however, we will concentrate on understanding system performance for the standard wastewater being used in all of these simulations.

Consider first Figure 7.20a where the removal of soluble organic matter is considered. The most striking thing about the performance of the CSAS system is how closely it parallels the single CSTR. The shapes of the curves are the same; only the magnitudes are different. This follows from the fact that the mass of heterotrophic biomass in the two systems is similar, which is caused by the nature of the organic substrate in the influent. Recall from Table 6.6 that the influent contains more particulate than soluble organic substrate. This particulate substrate is entrapped in the MLSS, making it available for microbial attack in both the contact and stabilization tanks. Because the SRT is the same in the CSAS system and the single CSTR, and because particulate substrate is attacked in both tanks, the opportunity for degradation of the particulate substrate is essentially the same in the two systems. That degradation results in the growth of heterotrophic biomass, which can then attack the soluble organic matter in the contact tank, resulting in even more biomass. Since more than half of the organic substrate is particulate and will be removed totally in the contact tank, and a substantial portion of the soluble organic substrate is removed at the SRTs studied, there is relatively little difference in the mass of heterotrophic biomass formed in the two systems, and thus they perform in a similar manner. The concentration of soluble substrate in the effluent from the CSAS system is higher simply because it is removed only in the contact tank, which has a smaller volume and a lower biomass concentration than the single CSTR, giving the contact tank a higher process loading factor.

In contrast, the shape of the ammonia-N curve for the CSAS system is quite different from that of the single CSTR. This difference is also reflected by the shapes of the autotrophic biomass curves for the two systems and is primarily due to the fact that most of the influent nitrogen is present in the soluble form, but also in part to the low maximum specific growth rate of autotrophic nitrifying bacteria. Examination of Table 6.6 reveals that the wastewater contains some particulate organic nitrogen, which will be converted to ammonia-N as the particulate organic matter undergoes biodegradation. Because the particulate organic nitrogen is entrapped in the MLSS, it is present throughout the entire system and its biodegradation provides ammonia-N that is available to the nitrifying bacteria in both tanks. In addition, that portion of the mass inflow of soluble nitrogen that is recycled through the stabilization tank is also available to the nitrifiers in both bioreactors. For the recycle ratio used (0.5), this is about one third of the influent soluble nitrogen. The remainder is only available to the nitrifiers in the contact tank. When the SRT is short, but long enough to prevent washout of the nitrifiers, they will grow using the nitrogen that is available throughout the entire system and provide the basis for ammonia-N oxidation in the contact tank. However, the quantity of nitrifiers formed will be limited primarily by the amount of nitrogen that is available in the stabilization tank, because the residence time of the biomass is greater in it and there is greater opportunity for degradation of the particulate organic nitrogen, thereby making ammonia-N available to the nitrifiers. Nitrifiers grown in the stabilization tank will then pass to the contact tank where they can oxidize a portion of the ammonia-N entering from the feed. This limitation forms the break in the ammonia-N curve at short SRTs. As the SRT is increased, further reductions in the ammonia-N concentration will be due to its

greater utilization in the contact tank. Because of the low half-saturation coefficient associated with autotrophic biomass growth, the autotrophs will be growing at their maximal rate in the contact tank and thus the mass of ammonia-N removed will be governed by the mass of autotrophs present. As long as excess ammonia-N is available, the mass of autotrophs will increase almost linearly with the SRT, which means that the ammonia-N concentration will decrease almost linearly, as shown. Only when the ammonia-N concentration drops sufficiently to cause the specific growth rate of the autotrophs in the contact tank to be governed by that concentration does the curve depart from linearity.

Figure 7.21 shows that significantly less oxygen is used in the CSAS system than in the single CSTR even though the solids wastage rates are very similar. This difference is due to the differences in the amount of nitrification. It will be recalled from the discussion in Section 6.3 that nitrification has a major impact on oxygen utilization, but almost no impact on biomass production.

7.4.3 Dynamic Performance

Because the influent enters only the contact tank, because the concentration of biomass in that tank is low, and because its volume is half that of the single CSTR, we might expect the dynamic performance of this bioreactor system to be worse than any we have encountered so far, and that is the case, as shown in Figure 7.22. The nitrification performance of the system is particularly poor. Reexamination of Figure 7.20 shows that at an SRT of 10 days, the value used for the dynamic simulation, steady-state nitrification is incomplete, with an effluent ammonia-N concentration of about 7 mg/L. This means that even at steady-state, the nitrifying bacteria are growing near their maximal rate in the contact tank. Consequently, when the diurnal load is applied, no excess nitrification capacity exists to oxidize the additional ammonia-N that enters during peak loading periods, causing most of it to pass through to the effluent. Relatively complete nitrification only occurs when the influent mass flow rate of ammonia is sufficiently low for the mass of nitrifiers in the system to handle it.

The diurnal oxygen requirements in each of the bioreactors of the CSAS system are shown in Figure 7.23, along with the requirement in a single CSTR. The surprising thing about the curves is that the oxygen utilization rate in the stabilization tank is almost as dynamic as the utilization rate in the contact tank. Because the influent only flows through the contact tank and the stabilization tank receives a constant flow rate, one might expect the stabilization tank to show a less severe response. There are two reasons why it does not. One is that over half of the organic loading to the CSAS system is due to particulate organic matter, which is degraded in both bioreactors. Since its input varies in a diurnal manner, so does its degradation. The other is the transport of ammonia-N into the stabilization tank through the biomass recycle flow. Since the ammonia-N concentration gets quite high in the contact tank, an appreciable quantity enters the stabilization tank where the longer HRT and higher nitrifier mass allow its oxidation. The time lag associated with the transport of these materials causes a shift in the times at which the maximum and minimum uptake occur in the two tanks, however.

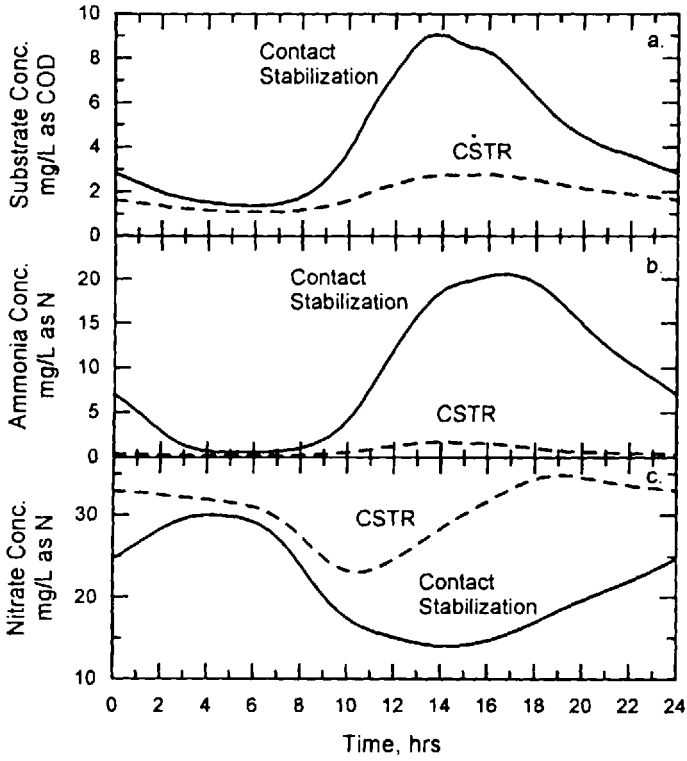


Figure 7.22 The time dependent response of the effluent from the CSAS system depicted in Figure 7.19 when subjected to the diurnal loading patterns shown in Figure 6.2. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Average influent flow = 1000 m^3/day. Average influent concentrations are given in Table 6.6. Biomass recycle flow = 500 m^3/day; volume of each reactor = 125 m^3; SRT = 10 days. Parameters are listed in Table 6.3. The dissolved oxygen concentration was held constant at 2.0 mg/L.

7.4.4 Effects of System Configuration

We saw earlier that the recycle ratio influences the performance of the SFAS system because of its effect on the distribution of biomass in the system. Thus, we would expect the recycle ratio to also affect the performance of the CSAS system, which it does, as shown in Figure 7.24. In this figure, which was generated for an SRT of 10 days, the dashed lines represent the concentrations in the contact (second) reactor, and thus represent the concentrations entering the settler, whereas the solid lines represent the concentrations in the stabilization (first) reactor.

Consider first the concentrations of heterotrophic biomass. The mass of heterotrophic biomass in the system is essentially independent of the recycle ratio between 0.1 and 1.0 because organic substrate removal is almost complete (relative to the influent) for all of those values. Thus, the differences in the concentrations in the two bioreactors shown in Figure 7.24e reflect primarily the effect of the recycle ratio on the concentrations of biomass entering the settler and leaving in the biomass

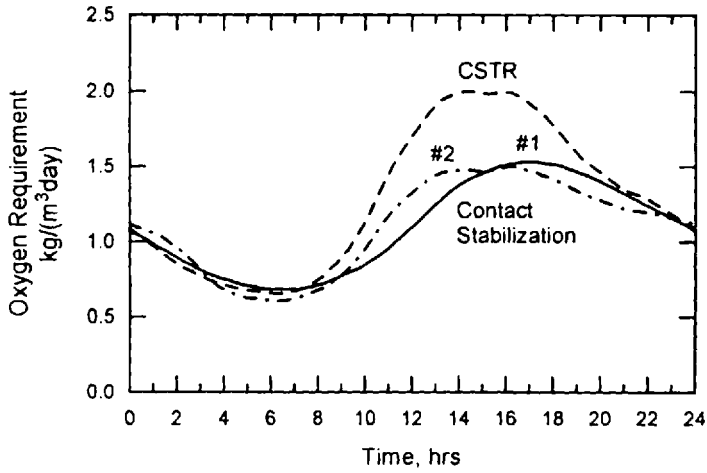


Figure 7.23 The time dependent variability in the oxygen requirement in each reactor of the CSAS system described in Figure 7.22. The solid curve represents the stabilization (first) reactor and the dashed-dot curve the contact (second) reactor. For comparison, the dashed curve shows the requirement in a single CSTR with a volume of 250 m^3.

recycle stream as given by Eq. 7.7. Because the microbial mass is fixed, an increase in the recycle ratio simply shifts heterotrophic biomass from the stabilization tank to the contact tank. An increase in the mass of heterotrophic biomass in contact with the wastewater allows more soluble organic constituents to be removed, thereby improving system performance, as shown in Figure 7.24a. As might be expected, almost all soluble organic matter is gone from the stabilization tank regardless of the recycle ratio and the residual simply reflects a balance between its utilization and its production through biomass death and lysis.

The response of the autotrophic biomass is very different from that of the heterotrophic biomass and reflects the fact that the mass of autotrophic bacteria increases as the recycle ratio is increased. When the recycle ratio is small, only a small percentage of the biomass is in the contact tank. As a consequence, only a small fraction of the ammonia-N is oxidized. Furthermore, because the recycle flow rate is small, only a small portion of the ammonia-N in the contact tank effluent is transported to the stabilization tank for oxidation. Thus, only a small percentage of the total mass of nitrogen passing through the system is oxidized and only a small mass of autotrophic biomass is formed. An increase in the recycle ratio has two effects, however. First, it shifts more of the biomass from the stabilization tank to the contact tank, allowing more ammonia-N oxidation in the contact tank, thereby forming more autotrophic biomass. Second, it transports a greater fraction of the unreacted ammonia-N into the stabilization tank, allowing formation of even more autotrophic biomass. By the time the recycle ratio is around 0.8, the majority of the nitrogen flowing into the system is being oxidized, giving a relatively constant mass of nitrifiers, so that further changes in the recycle ratio simply act to redistribute them in the same manner as the heterotrophic biomass.

One other point about nitrification needs clarification, and that concerns the concentration of nitrate-N in the stabilization tank. It can be seen in Figure 7.24c

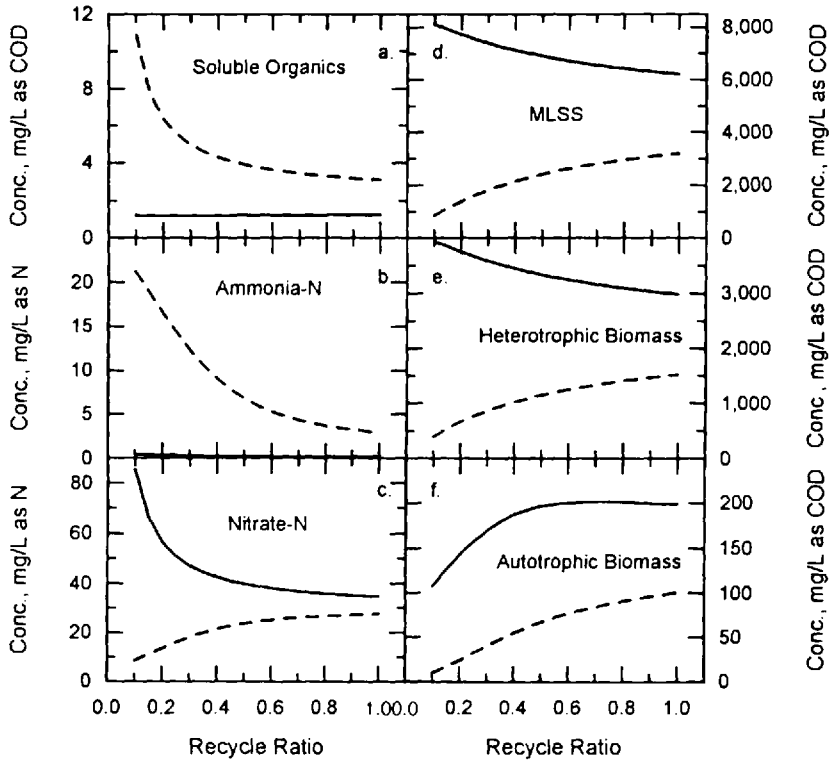


Figure 7.24 Effect of the recycle ratio (α) on the steady-state concentrations of various constituents in each reactor of the CSAS system described in Figure 7.20. SRT = 10 days. The solid curves represent the stabilization (first) reactor and the dashed curves the contact (second) reactor.

that the nitrate-N concentration is very high when the recycle ratio is very low. However, even though that concentration is high, it represents only a small fraction of the nitrogen flowing through the system. Furthermore, most of that nitrate comes from ammonia-N released as a result of biomass decay and particulate substrate degradation in the stabilization tank. Because the MLSS concentration is high there, the concentration of nitrogen released will be high. As stated above, however, it does not represent a large mass, as reflected by the low nitrate-N concentration in the final system effluent, which is the same as the concentration in the contact tank.

Figure 7.25 shows the effect of the recycle ratio on the oxygen utilization in the two tanks. Since an increase in the recycle ratio shifts biomass from the stabilization tank to the contact tank, it shifts the oxygen requirement in a like manner. In addition, however, an increase in the recycle ratio also allows more nitrification in both tanks, although more of the nitrification increase will occur in the contact tank. Thus, there is a greater increase in the oxygen requirement in the contact tank than there is a decrease in the stabilization tank.

Another way to shift the relative amounts of biomass in the two tanks is to change their sizes. Thus, simulations were done in which the total bioreactor volume was held constant but the relative sizes of the two tanks were changed. The SRT

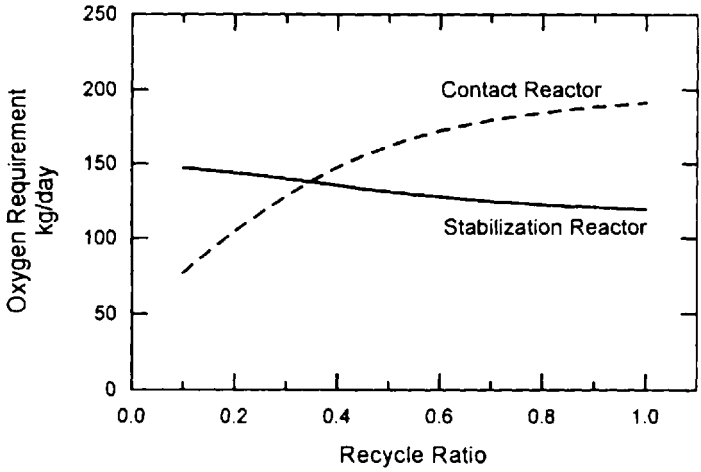


Figure 7.25 Effect of the recycle ratio (α) on the steady-state oxygen requirement in each reactor of the CSAS system described in Figure 7.20. SRT = 10 days. The solid curves represent the stabilization (first) reactor and the dashed curves the contact (second) reactor.

was held constant at 10 days and all other conditions were the same as for the simulations shown in Figure 7.20. The results are shown in Figure 7.26 where they are plotted as a function of the fraction of the total system volume in the contact tank (bioreactor 2). When all of the volume is in that tank the system is just a single CSTR. Thus, the responses shown for that fraction are the same as those shown earlier for a single CSTR.

Consider first the concentrations of MLSS and heterotrophic biomass shown in parts d and e of Figure 7.26. For an SRT of 10 days, the mass of each in the system is relatively constant over the range of relative tank sizes considered. This suggests that the changes in the concentrations of each shown in the figure are due primarily to their distributions within the system. Consequently, the concentration curves must satisfy two criteria. First, the sum of the masses in each tank, i.e., volume times concentration, must be relatively constant. Second, the concentration in the stabilization tank must be related to the concentration in the contact tank in approximately the same way that the concentration in the biomass recycle flow is related to the concentration entering the settler, i.e., through Eq. 7.7. In other words, as more volume is shifted to the contact tank, a smaller percentage of the system biomass is at the concentration of the biomass recycle and thus the concentration in the system must increase in order to contain the same mass in the same total volume. Those changes in biomass concentration, in turn, impact on the concentrations of soluble constituents leaving the contact tank.

As might be expected from the previous discussion, the concentrations of soluble organic matter and ammonia-N leaving the system decrease as the fraction of the system volume in the contact tank increases. There are two reasons for this: (1) the fraction of the biomass in contact with the wastewater increases, as discussed above, and (2) the residence time of the wastewater in contact with the biomass increases. Both of these act to reduce the process loading factor in the contact tank, which acts to reduce the substrate concentration in it. It will be noted in Figure 7.26,

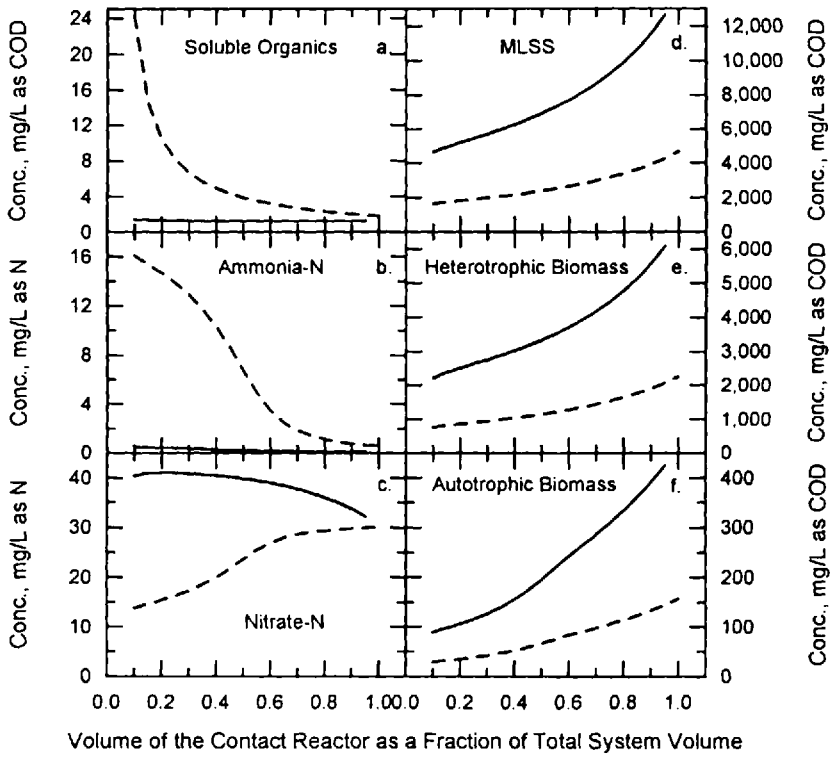


Figure 7.26 Effect of the relative volumes of the two reactors on the steady-state concentrations of various constituents in each reactor of the CSAS system described in Figure 7.20. The total system volume was constant at 250 m^3. SRT = 10 days. The solid curves represent the stabilization (first) reactor and the dashed curves the contact (second) reactor.

however, that the responses of soluble organic substrate and ammonia-N are quite different. This follows from the fact that heterotrophic biomass arises from the destruction of both soluble and particulate organic matter, whereas autotrophic biomass comes primarily from destruction of soluble material, as discussed previously for changes in SRT. As a consequence, the fraction of autotrophs in the biomass increases as the fraction of the system volume in the contact tank increases, as reflected by the differences between the heterotrophic and autotrophic biomass curves.

Figure 7.27 shows the oxygen requirements in the two bioreactors as their relative sizes are changed. As would be anticipated from the previous discussion, oxygen demand is shifted from the stabilization tank to the contact tank as volume is shifted to the contact tank. The fact that the increase in the oxygen requirement in the contact tank exceeds the decrease in the requirement in the stabilization tank reflects the increased amount of nitrification that occurs as the relative volumes are changed.

The CSAS process is the most complex studied so far, since effluent quality can be affected by SRT, recycle ratio, and relative reactor volumes. This suggests that its design is more complex than that of simpler systems. Although application of optimization techniques is the best way to arrive at a sound design, in Chapter

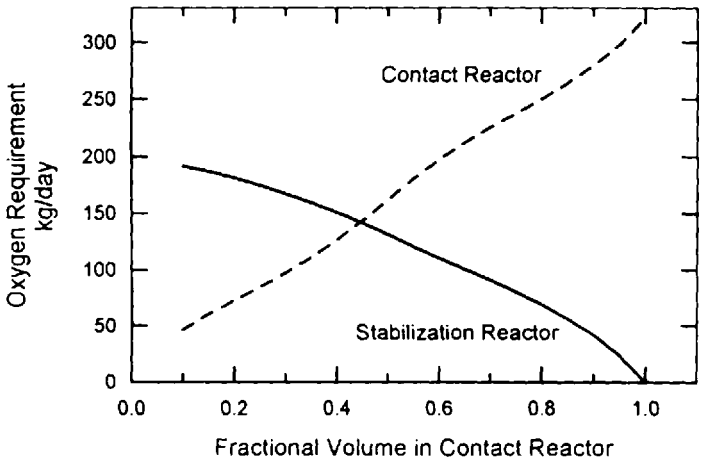


Figure 7.27 Effect of the relative volumes of the two reactors on the steady-state oxygen requirement in each reactor of the CSAS system described in Figure 7.20. The total system volume was constant at 250 m^3. SRT = 10 days. The solid curves represent the stabilization (first) reactor and the dashed curves the contact (second) reactor.

10 we will review some general rules that can be used by the engineer designing such a system.

7.5 MODIFIED LUDZACK–ETTINGER PROCESS

7.5.1 Description

All of the systems considered so far in this chapter have been totally aerobic. As a consequence, no significant denitrification occurs, and thus no nitrogen is removed; it is simply transformed from ammonia-N to nitrate-N. If one wished to add a denitrification reactor to any of the systems studied previously, it would be necessary to add an electron donor because by the time the SRT is sufficiently long to allow oxidation of the ammonia-N to nitrate-N, all of the soluble organic substrate has been degraded. As a consequence, the only source of organic matter to serve as the electron donor for denitrification is through hydrolysis of particulate substrate entrapped in the MLSS or released through biomass death and lysis. Hydrolysis is a slow reaction, however, and very large anoxic bioreactors would be needed to achieve even partial denitrification in this manner. Adding organic matter, such as methanol, to the bioreactor would overcome this problem, but adds to the system's operating costs.

Ludzack and Ettinger[13] reasoned that it should be possible to use the readily biodegradable substrate in the wastewater itself as an electron donor to achieve partial denitrification if there was a way to bring the nitrate formed in the reactor system back to a point where the substrate was available. They achieved this by separating the bioreactor into two compartments, with the first receiving no aeration, and by pumping a stream of mixed liquor from the actively nitrifying aerobic zone back to the anoxic zone to carry nitrate to it. This system is referred to as the modified

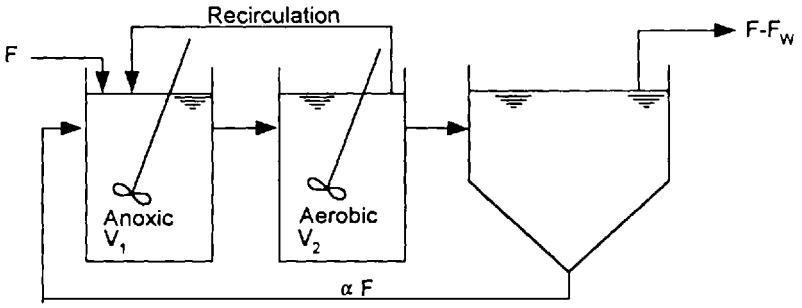


Figure 7.28 Schematic diagram of two CSTRs in series with all influent and all biomass recycle to the first reactor, in which the first reactor is anoxic and receives MLR flow from the second, which is aerobic. Although not shown, solids wastage is directly from both reactors. This configuration simulates the MLE process.

Ludzack–Ettinger system to differentiate it from some of their earlier work. Although many other systems have since been developed within which both nitrification and denitrification occur with greater efficiency,[19,25] the MLE system represents one of the simplest within which both processes take place. Thus, it provides a convenient system with which to investigate through modeling the effects of various system variables on both processes.

The system chosen to simulate the MLE process is shown schematically in Figure 7.28. The bioreactors are each 125 m^3, as in the last system, but all influent flow and biomass recycle enter the first one, which is anoxic with a dissolved oxygen concentration of zero. The second bioreactor is maintained at an oxygen concentration of 2.0 mg/L, as has been done in all previous simulations. To provide nitrate for anoxic growth of the heterotrophic bacteria in the first bioreactor, mixed liquor from bioreactor 2 is pumped to bioreactor 1 at a rate of 2000 m^3/day, or twice the influent flow rate. To distinguish this stream from the biomass recycle stream, which has a higher MLSS concentration and is at a lower flow rate, it will be called the mixed liquor recirculation (MLR) stream.

7.5.2 Effect of SRT on Steady State Performance

The effects of SRT on the concentrations in the second bioreactor of the MLE system are shown by the solid curves in Figure 7.29. For comparison, the concentrations in a single CSTR with a volume of 250 m^3 are shown as the dashed curves. As expected, the most obvious difference between the two systems is the nitrate-N concentration, which is much lower in the MLE system due to the denitrification in the first bioreactor. There are some other differences that should be recognized, however.

Examination of the ammonia-N, nitrate-N, and autotrophic biomass curves shows that a longer system SRT is required for the onset of nitrification in the MLE system. This is because nitrifiers can grow in only half of the system volume, the second half where dissolved oxygen is present. As discussed in Section 6.5.2, the aerobic SRT is recognized as the variable of importance in determining the growth of autotrophic bacteria in systems containing anoxic zones. Since the mass of biomass in the aerobic zone is half the total mass in this MLE system, the aerobic SRT

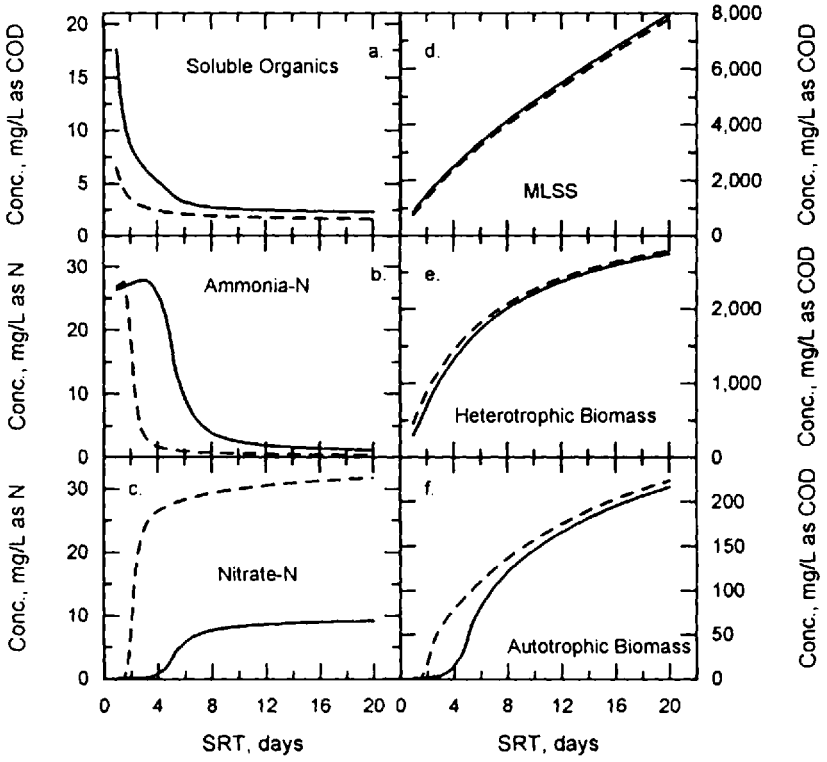


Figure 7.29 Effect of SRT on the steady-state concentrations of various constituents in the aerobic (last) reactor of the MLE system depicted in Figure 7.28. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Influent flow = 1000 m^3/day. Influent concentrations are given in Table 6.6. Biomass recycle flow = 500 m^3/day; mixed liquor recirculation flow = 2000 m^3/day; volume of each reactor = 125 m^3. Parameters are listed in Table 6.3. The dissolved oxygen concentration is zero in the anoxic (first) reactor and 2.0 mg/L in the aerobic (second) reactor.

is one half of the total SRT. Examination of the curves shows that nitrification begins in the single CSTR when the SRT is about 1.5 days. Thus, we would expect nitrification to begin in the MLE system when the SRT is around three days, which it does. Furthermore, nitrification is well established in the CSTR when the SRT has reached 8 days and in the MLE system when it has reached 16 days. Thus, aerobic SRT is a valid concept for this system as well. Careful examination of the ammonia-N and autotrophic biomass curves reveals, however, that the curves for the single CSTR would not be coincident with the curves for the MLE system if both are plotted against aerobic SRT. This is primarily because of biomass decay in the systems, which responds to total SRT, not just aerobic SRT. For a given aerobic SRT, the net specific growth rates of the autotrophs in the two systems are different because more decay occurs in the MLE system due to its greater total SRT. Furthermore, since decay releases ammonia-N from the biomass, the input rate of ammonia-N into the MLE system is greater at a given aerobic SRT because of the larger

amount of decay. These factors act together to cause the observed differences in the responses of the autotrophic biomass in the two systems.

The differences between the organic substrate curves are also related to the fact that only part of the system is aerobic. It will be recalled that a correction factor, η_g, is used in ASM No. 1 to express the fact that soluble substrate removal is slower under anoxic conditions. This retardation of the reaction rate in the anoxic zone acts to allow less soluble substrate removal in the MLE system. Furthermore, at shorter SRTs, incomplete nitrification means that insufficient electron acceptor is returned to the anoxic zone to allow biodegradation of all of the organic substrate. Thus, more substrate enters the aerobic zone, but because it is smaller than the single CSTR, less substrate can be removed.

One final difference that can be observed in Figure 7.29 is that the MLSS concentration is slightly higher in the MLE system even though the concentrations of heterotrophic and autotrophic biomass are lower. This reflects the presence of more slowly biodegradable substrate in the MLE system as a result of a lower rate of hydrolysis under anoxic conditions. This difference is also reflected in the solids wastage curves shown in Figure 7.30a.

Figure 7.30a also shows the oxygen requirements in the two systems. The most obvious difference between the two is that less oxygen is used in the MLE system. This is because of the use of nitrate-N as the terminal electron acceptor for the organic substrate that is oxidized in the first (anoxic) MLE reactor. If the nitrate utilization curve in Figure 7.30b is multiplied by 2.86, the oxygen equivalence factor for nitrate-N, and added to the oxygen requirement curve for the MLE system, the sum will be found to be similar to the oxygen requirement in the single CSTR at the longer SRTs where nitrification is fully established. The curves will not coincide at lower SRTs, however, because of differences in the extent of nitrification.

Another difference between the two systems, not shown in the figures, is that the MLE system will have a higher alkalinity. This is because denitrification increases alkalinity, as shown in Eq. 3.22 and discussed in Section 6.4.1.

7.5.3 Effects of System Configuration

Two things can be altered in the MLE system, the amount of mixed liquor recirculation flow returning nitrate to the anoxic bioreactor and the relative sizes of the anoxic and aerobic zones. Like the single CSTR, however, the total system HRT has a negligible effect on system performance.[28] Thus, it would be instructive to examine each of the variables that has an effect. In so doing, the system SRT will be fixed at 10 days, which is adequate to allow relatively complete nitrification in the system with equal sized bioreactors.

Figure 7.31 shows the effect of varying the MLR rate, expressed as a ratio of the influent flow. Two types of curves are shown. The solid ones represent the concentrations in the first (anoxic) bioreactor and the dashed ones the concentrations in the second (aerobic). As might be expected, because the recycle and MLR flows distribute the MLSS around the system many times within one SRT, the MLSS and biomass behave in a completely mixed fashion and their concentrations are essentially the same in the two bioreactors. Furthermore, the concentrations of soluble organic matter (SOM) and ammonia-N in the aerobic bioreactor are independent of the recirculation ratio because they are both controlled primarily by the aerobic SRT,

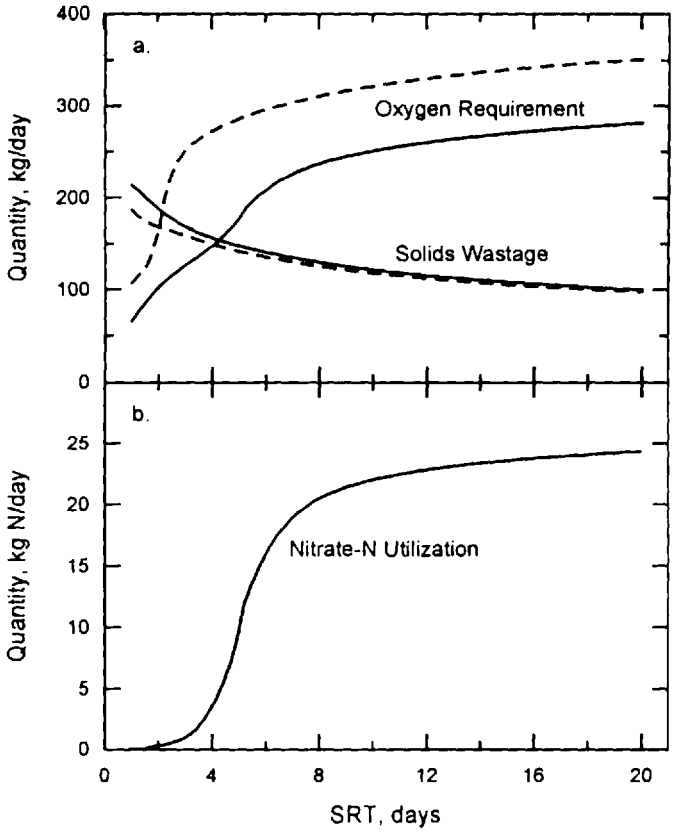


Figure 7.30 Effect of SRT on the total steady-state oxygen requirement, nitrate utilization rate, and solids wastage rate for the MLE system depicted in Figure 7.28 operating under the conditions listed in Figure 7.29. For a comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3.

which is constant. The only use of ammonia-N in the anoxic bioreactor is for biomass synthesis. Since that requirement is relatively small, the major factor influencing the ammonia-N concentration there is dilution by the recirculation flow, which has a uniformly low concentration. Consequently, the greater the recirculation ratio, the more dilution will occur, and the lower the concentration.

As far as the reactions in the system are concerned, the constituents most influenced by the recirculation are the soluble organic matter in the anoxic bioreactor and nitrate-N in both. When the recirculation flow is zero, nitrate can only enter the anoxic bioreactor through the biomass recycle flow (ratio = 0.5). Because the recycle flow rate is small, reaction in the anoxic bioreactor is limited by the availability of electron acceptor and thus the anoxic bioreactor contains appreciable soluble organic matter and almost no nitrate-N. Furthermore, the minimal flow of nitrate-N to the first bioreactor limits the amount of denitrification in the system, thereby making the nitrate concentration in the effluent, i.e., the second bioreactor, high. As the recirculation flow is increased, more nitrate-N is returned to the first bioreactor, providing

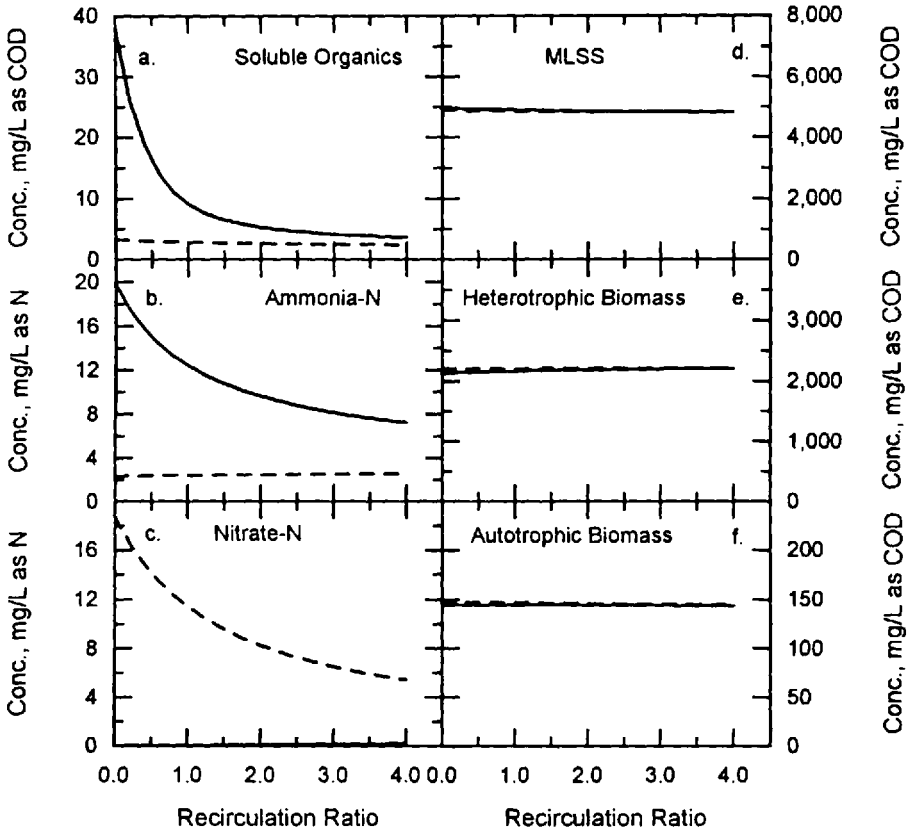


Figure 7.31 Effect of the mixed liquor recirculation ratio on the steady-state concentrations of various constituents in each reactor of the MLE system described in Figure 7.29. SRT = 10 days. The solid curves represent the anoxic (first) reactor and the dashed curves the aerobic (second) reactor.

more electron acceptor and allowing more anoxic heterotrophic growth, thereby reducing the concentration of soluble organic matter in that reactor. This increased denitrification also acts to diminish the amount of nitrate-N in the system effluent, as shown by the concentration in the second bioreactor. Up to a recirculation ratio of around 2.0, the availability of electron acceptor appears to limit the reactions in the anoxic bioreactor because appreciable soluble organic matter is present, but almost no nitrate-N. Above that point, however, the reactions are limited by both the electron donor and the electron acceptor because the concentrations of both are low. If the recirculation ratio were increased beyond 4.0, reaction in the anoxic bioreactor would become limited by the availability of electron donor and the concentration of nitrate-N in it would increase. This, in turn, would cause the concentration of nitrate-N in the aerobic bioreactor to reach a limit below which further reduction could not be achieved. That final limit is determined by the relative amounts of organic matter (electron donor) and nitrogen (potential electron acceptor) in the influent. If the concentration of organic matter in the influent were higher than in this example, the

lowest possible effluent nitrate-N concentration would be lower, whereas if the wastewater contained less organic matter relative to the amount of nitrogen present, the lowest attainable nitrate-N concentration would be higher. For this wastewater, which is similar to domestic sewage, it can be seen that recirculation ratios above 4.0 have little effect on the concentration of nitrate-N in the effluent. Thus, larger recirculation ratios are seldom used.

The mass utilization rates of nitrate and oxygen in the two bioreactors are shown in Figure 7.32. The nitrate utilization curve reflects the events just described, with the amount of nitrate used at zero recirculation being due to its return to the anoxic bioreactor by biomass recycle. As more recirculation is used, less oxygen is required because more of the electrons associated with the organic matter in the wastewater are being transferred to nitrate-N. As discussed previously, reduction of the amount of nitrate-N discharged is not the only reason a designer might decide to incorporate denitrification into a bioreactor system.[4] It may also be done to reduce power costs. In that case, the designer would choose the recirculation ratio that minimized total system power costs, including that for aeration and recirculation pumping.

Figure 7.33 shows the effects of varying the relative volumes of the two bioreactors, with the results plotted as a function of the volume in the aerobic bioreactor. Examination of Figure 7.33f shows that washout of the nitrifying bacteria began as the aerobic fraction of the system was reduced below 0.4. Since the system SRT is 10 days, this corresponds to an aerobic SRT of 4 days. Reexamination of Figure 7.29 reveals that aerobic SRTs in excess of 4 days are required to have maximal nitrification with the kinetic parameters used in these simulations. Since stable nitrification is a prerequisite for successful operation of the MLE system (or any other system in which denitrification is to be achieved at the expense of nitrate formed in it), the minimum fraction of the bioreactor volume allocated to the aerobic zone is

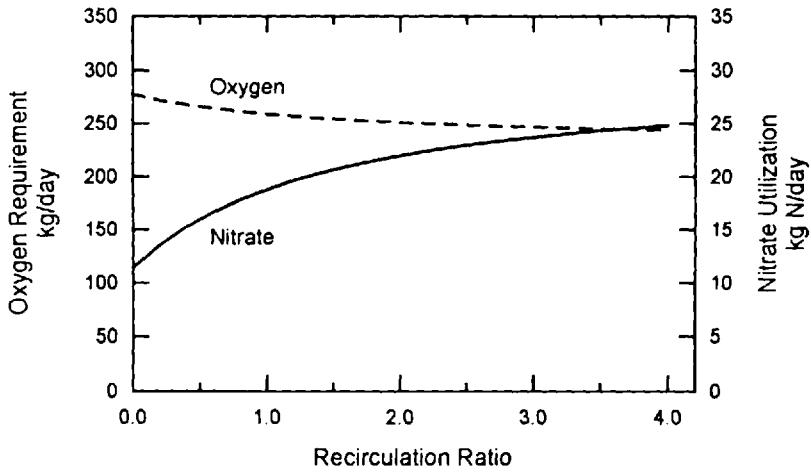


Figure 7.32 Effect of the mixed liquor recirculation ratio on the total steady-state oxygen requirement and nitrate utilization rate in the MLE system described in Figure 7.29. SRT = 10 days.

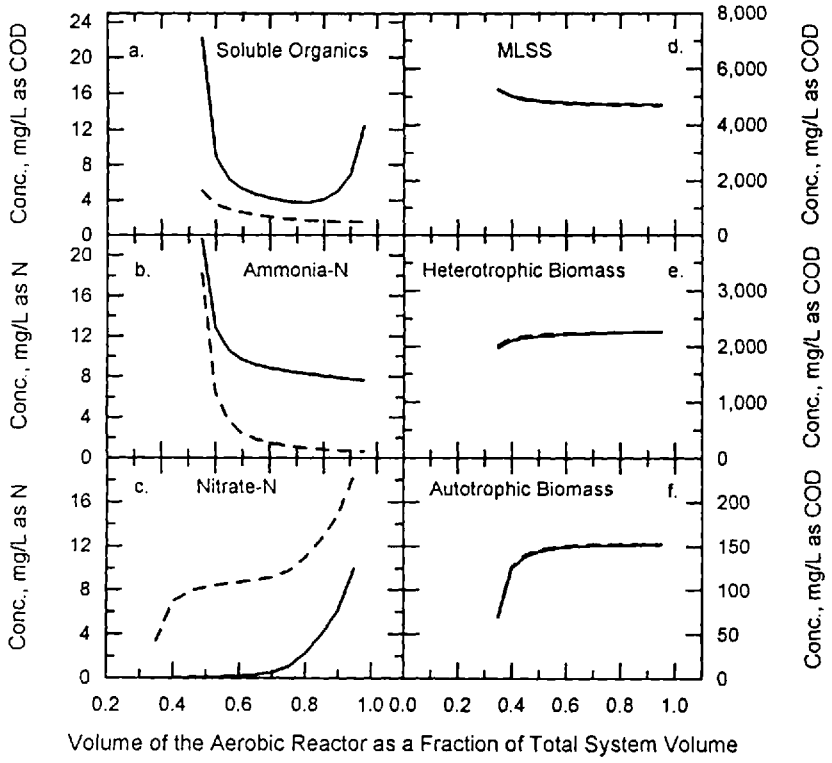


Figure 7.33 Effect of the relative volumes of the two reactors on the steady-state concentrations of various constituents in each reactor of the MLE system described in Figure 7.29. The total system volume was constant at 250 m^3. SRT = 10 days. The solid curves represent the anoxic (first) reactor and the dashed curves the aerobic (second) reactor.

determined by the minimum allowable aerobic SRT. It will be recalled that ASM No. 1 was not developed for simulating prolonged anaerobic periods. However, once washout of the autotrophic biomass has occurred, no nitrate will be present, which means that bioreactor 1 will be anaerobic rather than anoxic. Because the model is not valid under those conditions and because it was apparent that failure of nitrification was occurring at an aerobic fraction of 0.35, lower aerobic fractions were not investigated.

The rise of ammonia-N in both bioreactors and the drop in nitrate-N in the aerobic bioreactor as the aerobic fraction is reduced reflects the washout of the autotrophic biomass discussed above. Furthermore, the rapid rise in the concentration of soluble organic matter as the aerobic fraction is reduced below 0.4 also reflects this washout. When the amount of nitrate-N being returned to the first bioreactor is greatly reduced, the reactions in it became severely limited by the availability of electron acceptor, preventing complete utilization of the organic matter. This limitation is also reflected in the heterotrophic biomass and MLSS concentrations. The former drops due to curtailed growth in the anoxic bioreactor whereas the latter rises due to the accumulation of particulate organic matter.

As the aerobic fraction is increased above 0.7 the nitrate-N and soluble organic matter concentrations in the anoxic bioreactor begin to rise because the HRT is insufficient to allow complete reaction. This effect becomes especially severe at aerobic fractions above 0.8, and both soluble organic matter and nitrate-N leave the anoxic bioreactor unreacted.

Aerobic fractions between 0.5 and 0.7 produce effluents with about the same total nitrogen concentrations (ammonia-N plus nitrate-N), although a greater fraction of that nitrogen is in the form of nitrate-N when the aerobic fraction is larger. This suggests that designers of MLE systems have some latitude in the distribution of the system volume between the anoxic and aerobic zones. Furthermore, that latitude increases as the system SRT is increased, suggesting that systems with longer SRTs have greater operational flexibility.

Figure 7.34 shows the effects of the aerobic fraction of the system volume on the nitrate and oxygen utilization rates. The curves are consistent with previous explanations. Nitrate utilization is low at small aerobic fractions because little is being produced and is low at large aerobic fractions because insufficient time is available in the anoxic bioreactor for its utilization. Likewise, oxygen utilization is low at small aerobic fractions because little nitrification is occurring, increases as more nitrification is achieved, and finally increases again as denitrification is curtailed, requiring more of the electrons associated with organic matter in the influent to be transferred to oxygen as the terminal acceptor.

Finally, it was pointed out at the beginning of this section that system HRT has little impact on performance.[28] This point should be emphasized. The curves in Figures 6.33 and 6.34 change little when the system HRT is changed for a constant SRT. The important variable is the fraction of the HRT that is anoxic or aerobic, not the actual residence time in each zone.

7.6 BARDENPHO PROCESS

7.6.1 Description

As discussed in Section 7.1.1, a disadvantage of the MLE reactor configuration is that the effluent will always contain appreciable quantities of nitrate-N because nitrification occurs in the last bioreactor and the MLR is from that bioreactor. The Bardenpho[1] process overcomes this by adding an anoxic bioreactor after the aerobic one in which denitrification can occur by biomass decay and the utilization of slowly biodegradable substrate. In addition, to prevent biomass settling problems associated with denitrification in the final settler, a small aerobic bioreactor is usually used as the final zone.[1] This reactor configuration is illustrated in Figure 7.35. Selection of the best combinations of bioreactor sizes is a complex question that requires the use of optimization techniques[28] and it will be discussed in Chapter 11. For the purpose of this discussion, however, the total bioreactor volume was kept at 250 m^3, the value used in all of the simulations in this chapter, and the aerobic fraction was kept at 50%, the value used in Figures 7.29–7.32. The sizes of bioreactors one through four were selected as 50, 100, 75, and 25 m^3, respectively. The MLR ratio from bioreactor two to bioreactor one was maintained at 2.0, the value used in Figures 7.29, 7.30, 7.33, and 7.34.

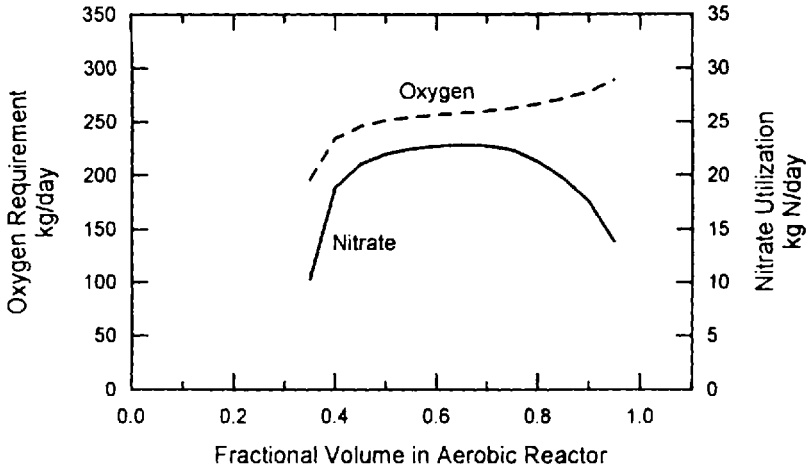


Figure 7.34 Effect of the relative volumes of the two reactors on the total steady-state oxygen requirement and nitrate utilization rate in the MLE system described in Figure 7.29. The total system volume was constant at 250 m^3. SRT = 10 days.

7.6.2 Effect of SRT on Steady-State Performance

The effect of SRT on the performance of the Bardenpho system is shown by the solid curves in Figures 7.36 and 7.37. For comparison, the performance of the MLE system from Figures 7.29 and 7.30 is shown by the dashed curves. Comparison of the curves shows that both systems contain similar quantities of biomass, but that the Bardenpho system achieves an effluent with less ammonia-N and less nitrate-N than the MLE system by achieving more denitrification. This is done even though the aerobic and anoxic fractions are the same as in the MLE system. Even lower effluent concentrations could be attained by proper selection of the aerobic fraction,

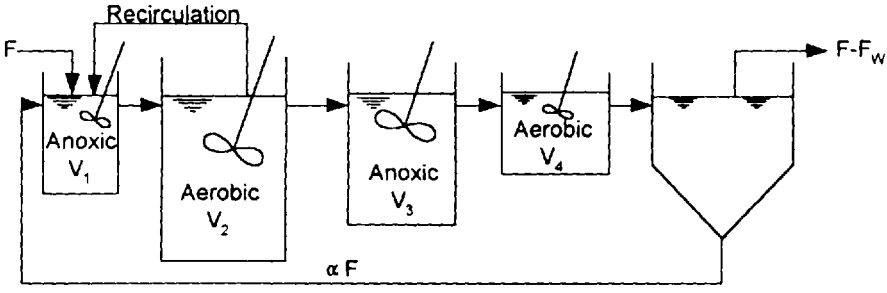


Figure 7.35 Schematic diagram of four CSTRs in series with all influent and all biomass recycle to the first reactor, in which the first and third reactors are anoxic and the second and fourth are aerobic. The first reactor receives mixed liquor recirculation flow from the second. Although not shown, solids wastage is directly from all reactors. This configuration simulates the Bardenpho process.

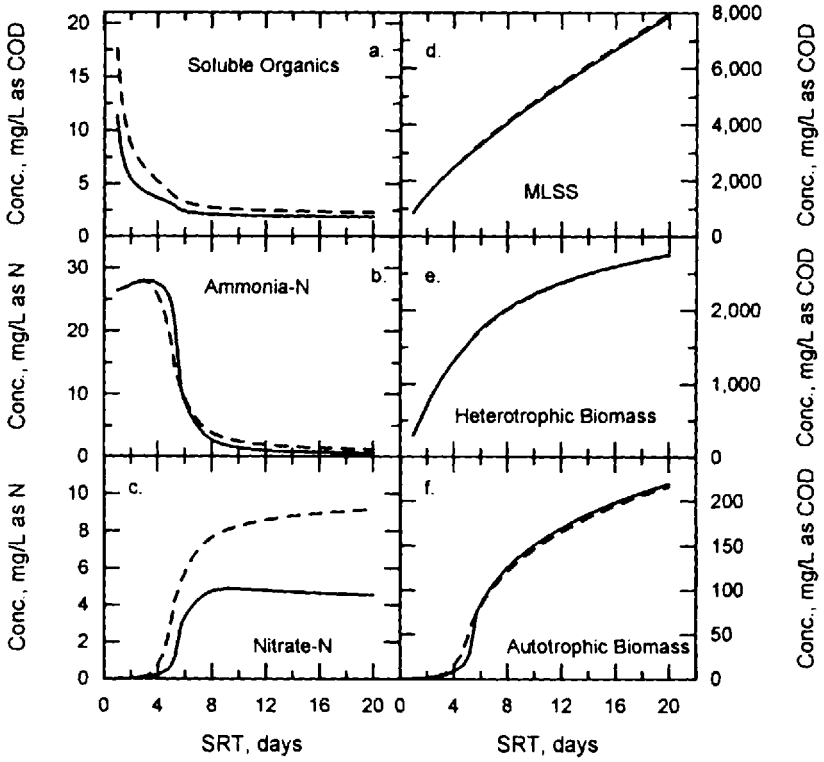


Figure 7.36 Effect of SRT on the steady-state concentration of various constituents in the last reactor of the Bardenpho system depicted in Figure 7.35. For comparison, the dashed curves represent the performance of the MLE system described in Figure 7.29. Influent flow = 1000 m^3/day. Influent concentrations are given in Table 6.6. Biomass recycle flow = 500 m^3/day; mixed liquor recirculation flow = 2000 m^3/day, reactor volumes: V_1 = 50 m^3; V_2 = 100 m^3; V_3 = 75 m^3; V_4 = 25 m^3. Parameters are listed in Table 6.3. The dissolved oxygen concentration was zero in the first and third (anoxic) reactors and 2.0 mg/L in the second and fourth (aerobic) reactors.

volume distribution among the various bioreactors, and recirculation ratio. However, it was not the intent here to minimize effluent nitrogen concentrations; rather, the point was to show the effect of adding additional anoxic and aerobic bioreactors.

A significant thing to note in Figure 7.36 is that the growth pattern of the autotrophic bacteria is different in the two bioreactor systems, even though the aerobic SRTs are the same. This emphasizes the point made earlier that although the concept of aerobic SRT is important to understanding the fate of nitrifying bacteria in systems containing anoxic zones, it is not the sole factor influencing their growth. System configuration is also important. This suggests that pilot scale studies coupled with system simulation are required to arrive at sound designs for such complex systems. In Chapter 11, we see how the results from such studies have been combined with full scale plant experience to allow development of design guidelines for biological nutrient removal systems.

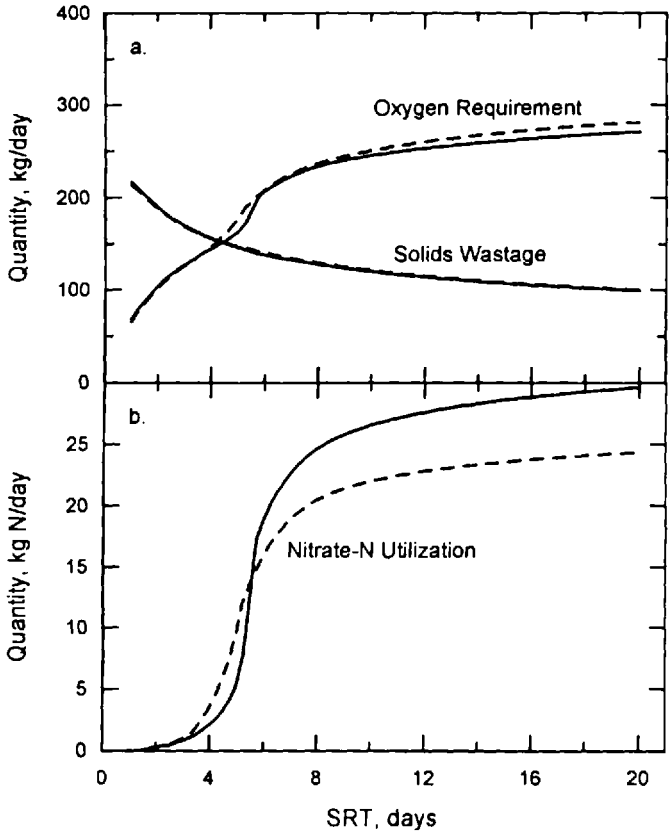


Figure 7.37 Effect of SRT on the total steady-state oxygen requirement, nitrate utilization rate, and solids wastage rate for the Bardenpho system depicted in Figure 7.35 operating under the conditions listed in Figure 7.36. For comparison, the dashed curves represent the performance of the MLE system described in Figure 7.29.

7.7 BIOLOGICAL PHOSPHORUS REMOVAL PROCESS

7.7.1 Description

As discussed in Section 2.4.6, certain bacteria, known collectively as phosphate accumulating organisms (PAOs), have the interesting characteristic of concentrating phosphate in Poly-P granules when they are cycled between aerobic and anaerobic conditions. The Poly-P acts as an energy reserve that allows the bacteria to rapidly take up acetate under anaerobic conditions, storing it as PHB and other polyhydroxyalkanoic acids (PHAs). Soluble phosphate is released in the process. The PHAs, in turn, provide energy for growth under aerobic conditions. They also allow soluble phosphate to be taken up and stored as Poly-P. The difference in energetics between aerobic and anaerobic metabolism is such that more phosphate can be taken up than was released, providing a mechanism for concentrating phosphate within the biomass, allowing it to be removed via solids wastage.

The history of the development of biological phosphorus removal processes is one of the most fascinating ones in environmental engineering, beginning with ob-

servations of unexplained phosphorus removal in full scale CAS systems, arguments over the reasons for that removal, simultaneous development of processes by various groups, and conflicts over patent claims and infringements. Unfortunately, space does not permit a review of that history here, but the reader is urged to consult Randall[17] and Stensel[21] for part of the story.

The simplest process flow sheet for biological phosphorus removal incorporates two bioreactors in series, with the first being anaerobic and the second aerobic, as shown in Figure 7.38. This process was first presented in the open literature by Barnard[3] who termed it the Phoredox process to indicate that phosphorus removal will occur when a sufficiently low redox potential is achieved through use of an anaerobic zone. He reasoned that the anaerobic zone should be placed first in the process train to take advantage of the electrons available in the raw wastewater, just as is done in the MLE and Bardenpho processes of denitrification.[2] This same configuration was patented by Air Products and Chemicals, Inc. of Allentown, PA under the trademark Anaerobic/Oxic,® or A/O process.[20] The major difference between the Phoredox and A/O processes is that in the latter the anaerobic and aerobic zones are divided into a number of equally sized completely mixed compartments.[21] For use in this section, we will consider only a single compartment for each. However, a large number of process flow sheets are available for biological phosphorus removal, both alone and in concert with nitrogen removal. They are discussed in Chapter 11.

The simple process flow sheet shown in Figure 7.38 provides an opportunity to observe some very interesting interactions among the various types of bacteria in wastewater treatment systems. In the MLE and Bardenpho processes we observed interactions between heterotrophs and autotrophs. Introduction of the anaerobic zone allows the specialized PAOs to interact with both of those groups. To introduce those interactions and the effects that they have on design of BPR systems, simulations were performed in ASIM[5] using ASM No. 2. As discussed in Section 6.1.4, ASM No. 2 is very complex because it seeks to incorporate a number of complicated processes that are not yet fully understood. Nevertheless, it is sufficiently conceptually accurate to illustrate several important points, and it is for that purpose that it is used herein. The limitations inherent in the assumptions of the model will be pointed out as needed as the simulated performance of a Phoredox system is discussed.

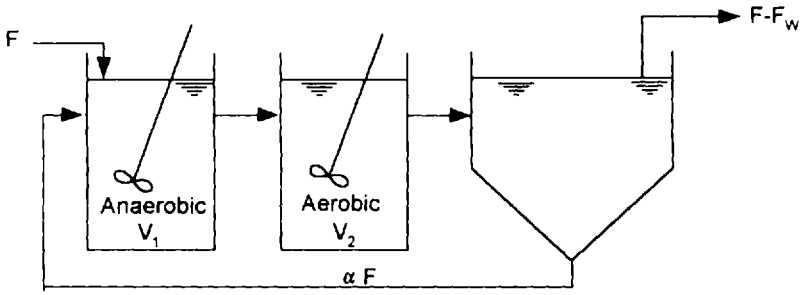


Figure 7.38 Schematic diagram of two CSTRs in series with all influent and all biomass recycle to the first reactor, which is anaerobic. Although not shown, solids wastage is from the second reactor, which is aerobic. The configuration simulates the Phoredox process.

Table 7.1 Wastewater Characteristics Used to Simulate the Performance of the Phoredox Process[5]

Component	Concentration
Inert particulate organic matter	25.0 mg/L as COD
Slowly biodegradable substrate	125.0 mg/L as COD
Readily (fermentable) biodegradable substrate	30.0 mg/L as COD
Volatile acids (acetate)	20.0 mg/L as COD
Oxygen	0.0 mg/L as O_2
Soluble nitrate nitrogen	0.0 mg/L as N
Soluble ammonia nitrogen	16.0 mg/L as N
Soluble biodegradable organic nitrogen	0.9 mg/L as N
Particulate biodegradable organic nitrogen	5.0 mg/L as N
Soluble phosphate phosphorus	3.6 mg/L as P
Soluble biodegradable organic phosphorus	0.3 mg/L as P
Particulate biodegradable organic phosphorus	1.2 mg/L as P
Alkalinity	5.0 mM/L

The system chosen to represent the Phoredox process, shown in Figure 7.38, contains two bioreactors in series with the first being anaerobic and the second aerobic. As done previously the system has a total volume of 250 m^3, receives 1000 m^3/d of wastewater flow, and has a biomass recycle rate of 500 m^3/d from the clarifier to the first bioreactor. In this case, however, 20% of the total system volume is allocated to the first bioreactor. The first bioreactor is assumed to receive no dissolved oxygen whereas the second receives sufficient oxygen to maintain the dissolved oxygen concentration at 2.0 mg/L. The characteristics of the wastewater entering the system are given in Table 7.1. The major difference between them and the characteristics used in the previous simulations is that the readily biodegradable substrate has been divided into two components, acetate and readily fermentable substrate. Acetate is found in many wastewaters, particularly if the sewers are septic, and plays a major role in the metabolism of the PAOs, as discussed above. The concentrations of the various constituents differ somewhat from the concentrations used in the previous simulations. They were chosen because they were listed in ASM No. 2[9] and ASIM.[8] With the exception of abiotic phosphate precipitation, which was "turned off" to demonstrate the effect of biological phosphorus removal alone, the default parameter values in ASIM[8] were used. They are very similar to those in ASM No. 2[9] but are not listed here because the entire model has not been presented, as discussed in Section 6.1.4. Interested readers should consult the original sources.

7.7.2 Effect of SRT on Steady-State Performance

The effects of SRT on the concentrations of various constituents in the second bioreactor (and the effluent) of the Phoredox system are shown by the solid curves in Figure 7.39. For comparison, the concentrations in a single CSTR with a volume of 250 m^3 receiving the same wastewater are shown as the dashed curves. Quantitative differences between those dashed curves and others presented earlier are because of

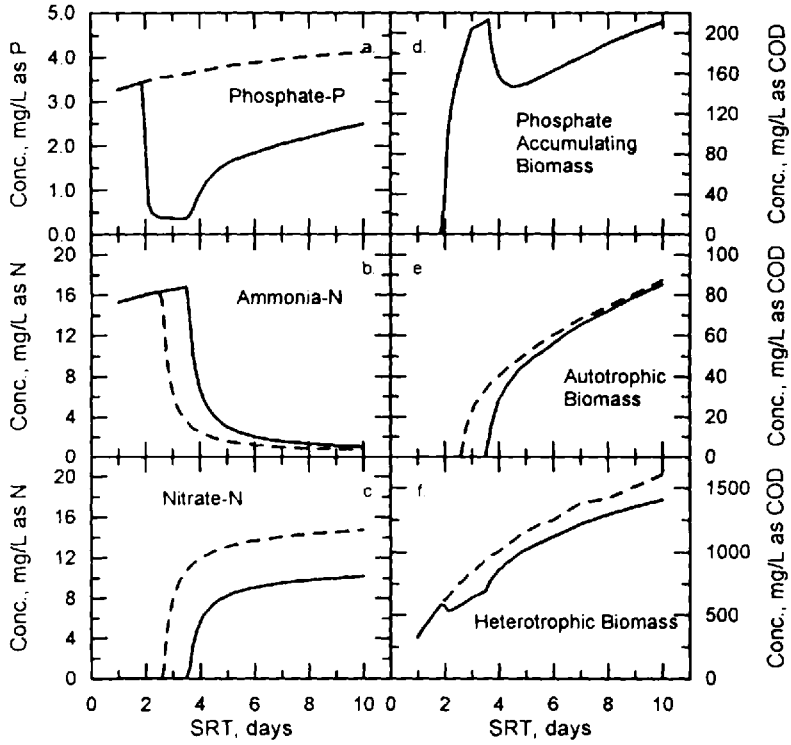


Figure 7.39 Effect of SRT on the steady-state concentrations of various constituents in the aerobic (last) reactor of the Phoredox system depicted in Figure 7.38. For comparison, the dashed curves represent the performance of a single CSTR with a volume of 250 m^3. Influent flow = 1000 m^3/day. Influent concentrations are given in Table 7.1. Biomass recycle flow = 500 m^3/day; volume of the anaerobic (first) reactor = 50 m^3, volume of the aerobic (second) reactor = 200 m^3. Parameters are the default parameters in ASIM.[3] The dissolved oxygen concentration is zero in the anaerobic reactor and 2.0 mg/L in the aerobic reactor.

differences in the wastewater characteristics and the kinetic and stoichiometric coefficients in the two activated sludge models. They are qualitatively similar, however.

The responses of the soluble phosphate and the PAOs are shown in panels a and d of Figure 7.39. No PAOs grow in the single CSTR because the proper environment is not provided, and the only phosphorus removal is that associated with its role as a macronutrient for biomass growth. Examination of Figure 7.39a reveals that the soluble phosphate concentration in the single CSTR increases as the SRT is increased. There are two reasons for this. First, the observed yield of biomass decreases as the SRT is increased, thereby decreasing the amount of nutrients required. Second, soluble phosphate is released from slowly biodegradable substrate as it is hydrolyzed. Hydrolysis is greater at longer SRTs, allowing more phosphate to be released. The net effect of these events is to make the amount of phosphate released exceed the amount incorporated into biomass. Thus, the concentration increases. Furthermore, at short SRTs, no phosphorus is removed in the Phoredox system because the SRT is below the minimum required for growth of PAOs. The minimum

SRT relates only to the aerobic SRT because growth of PAOs occurs only under aerobic conditions. The presence of the anaerobic zone is necessary for their growth, however, because they grow at the expense of stored PHAs, which are only formed under anaerobic conditions as acetate is taken up and stored at the expense of Poly-P. Once the minimum SRT is exceeded, a significant population of PAOs is established in the Phoredox system and phosphate uptake by them in the aerobic bioreactor is able to reduce the phosphate to very low levels.

Examination of Figure 7.39f reveals that there is a reduction in the amount of heterotrophic biomass relative to the single CSTR over the SRT range where good phosphorus removal is occurring. This is because of competition between the PAOs and the common heterotrophs for substrate. In the anaerobic bioreactor of the Phoredox system, the PAOs are able to store acetate. The common heterotrophs, on the other hand, cannot store or use acetate in the anaerobic zone; nor can they grow as long as oxygen or nitrate-N is unavailable as an electron acceptor. They can only produce acetate by fermentation of readily fermentable substrates. Consequently, the amount of substrate available to the common heterotrophs in the aerobic zone, where growth occurs, has been reduced by the activity of the PAOs in the anaerobic zone, thereby reducing the quantity of heterotrophs that can be formed.

Examination of Figure 7.39a reveals that excellent phosphorus removal occurs as long as the SRT is between 2.2 and 3.6 days, but once the SRT exceeds 3.6 days, phosphorus removal deteriorates sharply. This is because of nitrification and denitrification. Once the aerobic SRT is sufficiently long, autotrophic biomass can grow, converting ammonia-N to nitrate-N. The nitrate-N, in turn, is returned to the anaerobic zone via the biomass recycle, providing an electron acceptor for the heterotrophs in that zone. That allows them to compete with the PAOs for acetate and readily fermentable substrate in the anaerobic zone. It also reduces the amount of fermentation that occurs, thereby reducing the quantity of acetate produced. The impact of both of these mechanisms is to reduce the mass of PAOs that can be grown, as shown in Figure 7.39d, which reduces the amount of soluble phosphate that can be taken up in the system. As a consequence of these effects, Phoredox and A/O processes tend to be operated with short SRTs. Furthermore, because it is often desirable to achieve carbon oxidation, nitrification, denitrification, and phosphorus removal all in a single system, several process flow sheets have been devised to overcome these effects. They are discussed in Chapter 11.

In summary, the important point to draw from Figure 7.39 is that there is a limited range of SRTs over which a simple Phoredox (or A/O) system will work properly. If the SRT is too short, the PAOs will wash out. If it is too long, nitrifying bacteria will be able to grow, providing an inorganic electron acceptor to the anaerobic zone which allows the common heterotrophs to out-compete the PAOs for substrate, thereby reducing phosphate removal.

7.7.3 Effects of System Configuration

The role of the anaerobic bioreactor in the Phoredox process is two fold. First and foremost, it provides the selective advantage that allows PAOs to grow in the system. It is there that they take up acetate, forming the PHAs that will serve as their energy source for growth in the aerobic bioreactor. Second, it provides a regime wherein fermentation may occur, providing acetate in excess of that available in the influent.

As important as the anaerobic tank is, the entire system cannot be anaerobic because aerobic conditions are required for growth of both PAOs and heterotrophic bacteria. Thus, there is an optimal balance between the sizes of the two zones. In order to illustrate that balance, simulations were performed in which the total system volume was held constant while the relative sizes of the two zones were varied. The system SRT was held constant at 4 days. The results of the simulations are shown in Figures 7.40 and 7.41. The former shows the responses of the soluble constituents while the latter presents the particulate ones. The solid curves represent the anaerobic (first) bioreactor and the dashed curves the aerobic (second) bioreactor. The dashed curves, therefore, also represent the system effluent with respect to the soluble constituents.

Examination of Figure 7.40a reveals that the best phosphorus removal occurs when the volume of the anaerobic zone is between 25% and 48% of the total system volume. It should be emphasized that this range is unique to this SRT, wastewater characteristics, etc. It will be different for other situations. The important point is that there is indeed an optimal combination that maximizes the concentration of PAOs (Figure 7.41a), thereby allowing maximum phosphate-P removal. Panels a and d of Figure 7.40 also reveal that over that range, the concentration of phosphate-P reaches a maximum in the anaerobic zone while the concentration of acetate reaches

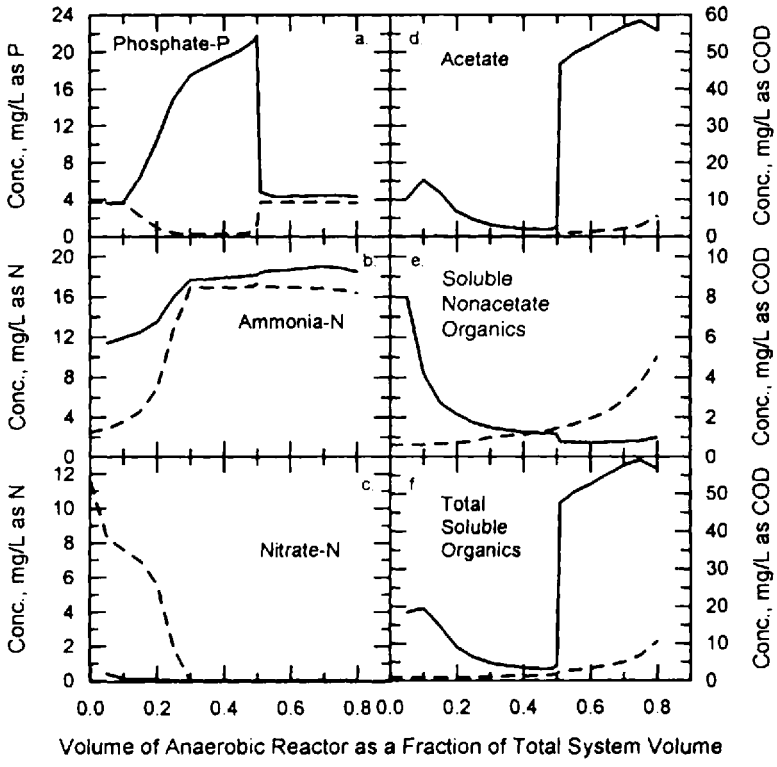


Figure 7.40 Effect of the relative volumes of the two reactors on the steady-state concentrations of various soluble constituents in each reactor of the Phoredox system described in Figure 7.38. The total system volume was constant at 250 m^3. SRT = 4 days. The solid curves represent the anaerobic (first) reactor and the dashed curves the aerobic (second) reactor.

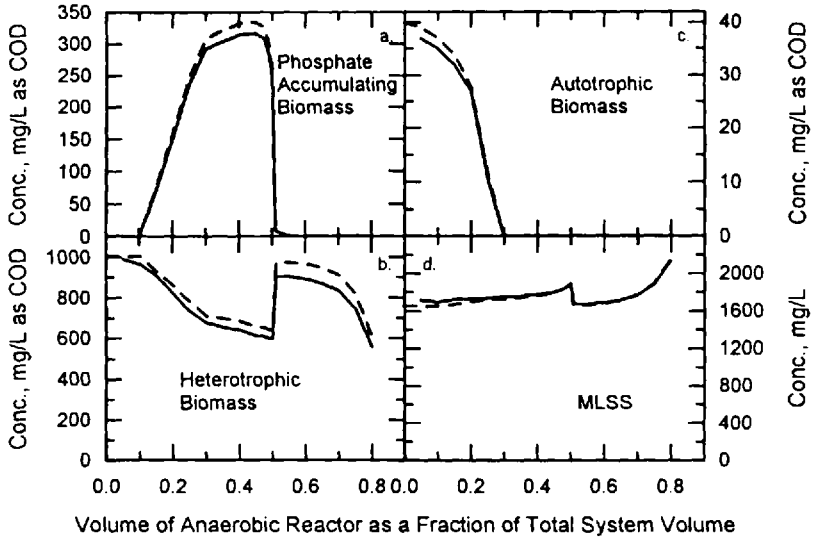


Figure 7.41 Effect of the relative volumes of the two reactors on the steady-state concentrations of various particulate constituents in each reactor of the Phoredox system described in Figure 7.38. The total system volume was constant at 250 m^3. SRT = 4 days. The solid curves represent the anaerobic (first) reactor and the dashed curves the aerobic (second) reactor.

a minimum. This is because of the mechanism of phosphate removal. Recall that in the anaerobic zone the PAOs hydrolyze stored Poly-P to gain energy for the uptake and storage of acetate as PHAs, releasing orthophosphate in the process (see Section 2.4.6). Over the optimal volume split range, maximum utilization has been made of the available acetate in the anaerobic tank, leading to maximum release of soluble phosphate. That phosphate is then taken up again in the aerobic bioreactor as the PAOs grow by using the PHAs as substrate and store Poly-P.

The upper end of the optimal range is controlled by the minimum aerobic SRT for growth of the PAOs in the aerobic zone. Once the anaerobic zone becomes so large that the aerobic SRT is below the minimum required for growth of the PAOs, they wash out and the system fails from the perspective of phosphorus removal. This can be seen clearly in Figures 7.40a and 7.41a. The population of heterotrophic bacteria increases (Figure 7.41b) because they no longer have to compete with the PAOs for acetate. Ultimately, as the anaerobic zone gets even larger, the entire system begins to act like a totally anaerobic system. The model was not intended for simulating such systems, so anaerobic fractions greater than 0.8 are not shown.

The lower end of the optimal range is a little more complicated. If the anaerobic zone is very small, there is insufficient time for acetate production by fermentation and its uptake by the PAOs. Thus, the PAOs can't grow at all in the system (Figure 7.41a). In addition, when the anaerobic zone is very small, the aerobic SRT is sufficiently long to allow autotrophic nitrifying bacteria to grow, thereby providing nitrate-N as an electron acceptor in the anaerobic zone through the biomass recycle. As a consequence, the anaerobic zone actually behaves as an anoxic zone, such as in the MLE process. Moderately sized anaerobic zones allow for some growth of

PAOs, but their full potential cannot be reached because of the return of nitrate-N to the anaerobic zone, which reduces the amount of fermentation that can occur and allows the heterotrophic bacteria to out-compete the PAOs for acetate. However, as the anaerobic zone is made larger, there is more opportunity for fermentation, with a gain in PAOs at the expense of heterotrophs (Figure 7.41a and 7.41b). Furthermore, as the size of the anaerobic zone is increased, the aerobic SRT is decreased, thereby reducing the population of autotrophic bacteria (Figure 7.41c) and the concentration of nitrate-N being returned to the first bioreactor (Figure 7.40c). This allows better growth of the PAOs. Ultimately, the point is reached at which the aerobic SRT is smaller than the minimum SRT for the autotrophs and they wash out (Figure 7.41c), allowing the first tank to operate in a truly anaerobic mode, which maximizes phosphorus removal.

In summary, the important point to gain from these simulations is that there is an optimal split between the anaerobic and aerobic zones of the Phoredox (or A/O) system. Optimal phosphorus removal occurs when the aerobic SRT is small enough to exclude growth of autotrophic nitrifying bacteria, yet large enough to allow the PAOs to grow, and the time in the anaerobic zone is sufficiently large to allow efficient fermentation and uptake of the resulting acetate.

7.8 SEQUENCING BATCH REACTOR

7.8.1 Description

All of the process variations we have considered so far are continuous processes. As a consequence, the environments required to achieve a variety of objectives must be encountered spatially as the wastewater and biomass move from tank to tank within the system. Because each tank has a fixed volume, the relative amount of process time spent under each environmental condition is fixed for a given influent flow rate. Alteration of those times requires alteration of the sizes of the various tanks, something that may or may not be achieved easily. It is possible, however, to accomplish the same results in a batch reactor by altering the environment temporally. In this situation, if the relative times devoted to the particular environments are not attaining the desired result, they can be changed easily by reprogramming the controllers that turn the pumps and blowers on and off. This flexibility is the major advantage associated with batch bioreactors.

The term sequencing batch reactor (SBR) stems from the sequence of steps that the reactor goes through as it receives wastewater, treats it, and discharges it, since all steps are accomplished in a single tank. A typical sequence is illustrated in Figure 7.42. The cycle starts with the fill period in which the wastewater enters the bioreactor. The length of the fill period is chosen by the designer and depends upon a variety of factors, including the nature of the facility and the treatment objectives. The main effect of the fill period, however, is to determine the hydraulic characteristics of the bioreactor. If the fill period is short, the process will be characterized by a high instantaneous process loading factor, thereby making the system analogous to a continuous system with a tanks-in-series configuration. In that case, the biomass will be exposed initially to high concentrations of organic matter and other wastewater constituents, but the concentrations will drop over time. Conversely, if the fill period is long, the instantaneous process loading factor will be small and the system

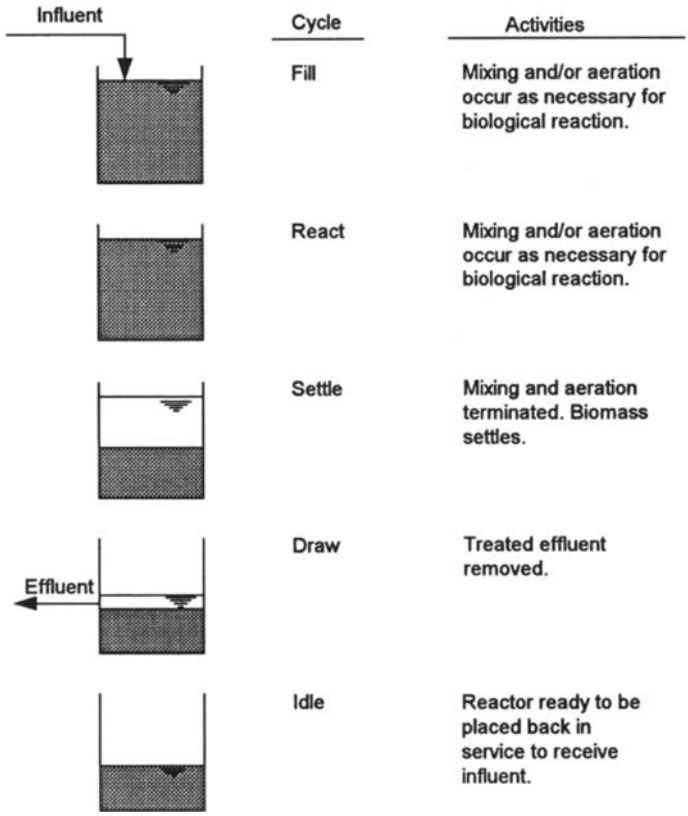


Figure 7.42 The sequence of events in a sequencing batch reactor (SBR).

will be similar to a completely mixed continuous flow system in its performance. This means that the biomass will experience only low and relatively constant concentrations of the wastewater constituents.

The fill period is followed by the react period in which the biomass is allowed to act upon the wastewater constituents. Actually, reactions, i.e., biomass growth and substrate utilization, also occur during the fill period, so the fill period should really be thought of as "fill plus react," with react continuing after fill has ended. Since a certain total react period will be required to achieve the process objectives, if the fill period is short, the separate react period will be long, whereas if the fill period is long, the separate react period will be short to nonexistent. The two periods are usually separately specified, however, because of the impact that each has on the performance of the system.

The environmental conditions established during the fill and react periods will determine which events occur. For example, if the fill and react periods are aerobic throughout, the events will be limited to carbon oxidation and nitrification. Consequently, performance of the SBR will lie somewhere between that of conventional and completely mixed activated sludge, depending on the length of the fill period. The elimination of aeration while maintaining mixing will allow denitrification to occur if nitrate is present. If the nitrate is generated by nitrification during the react

period and is present in the liquid retained in the bioreactor at the end of the cycle, then imposition of a mixed but unaerated interval during the early portion of the fill and react periods will make the SBR behave like a continuous flow MLE system. Furthermore, inclusion of another mixed but unaerated interval later in the react period can make the SBR act like the Bardenpho process. On the other hand, if the SBR is operated with a short SRT so that no nitrate is produced, mixing without aeration during fill and react can lead to the selection of PAOs, allowing the SBR to behave like the Phoredox or A/O continuous systems. It is apparent from these few examples that SBRs can be designed and operated to mimic many different continuous flow processes. Readers wanting more information about this ability should consult the review by Irvine and Ketchum.[10]

After the react period has been completed, all mixing and aeration are stopped and the biomass is allowed to settle. Just as in continuous flow processes, this accomplishes two things. It produces a clear effluent suitable for discharge and it retains biomass for SRT control. Solids wastage may be accomplished at the end of the settle period, simulating the conventional wastage strategy for continuous flow systems, or it may be done at the end of the react period, simulating the Garrett[11] wastage strategy. Regardless of when wastage is done, however, after sufficient settling has occurred, treated effluent may be removed by decantation during the draw period. The amount of liquid and biomass retained in the bioreactor constitutes the biomass recycle for the next cycle. If a large volume is retained relative to the influent volume in order to provide nitrate for an initial denitrification period, then the retained volume is analogous to biomass recycle plus MLR in a continuous process.

Finally, an idle period is generally allowed in each cycle to provide flexibility. This is particularly important for a system with several SBRs because it allows their operation to be synchronized for maximum effectiveness. Mixing and aeration may or may not be used during the idle period, depending on the overall process objectives, and the length of the idle period may vary from cycle to cycle as system needs dictate. The beginning of the new fill period terminates the idle period and initiates a new cycle.

7.8.2 Analogy to Continuous Systems

In the previous section, we saw that an SBR is capable of mimicking the operation of many different types of continuous systems, depending upon how the cycle is conducted. As a consequence, the performance of an SBR is similar to the performance of the analogous continuous system, provided that the two have the same SRT. Just as with continuous systems, the SRT determines the amount of biomass in the SBR, thereby determining its overall average performance. Furthermore, as with the other systems described in this chapter, it is not possible to derive analytical expressions describing SBR performance. Because the biomass in an SBR is in a dynamic state, a set of dynamic differential equations must be written and solved to determine system performance. Oles and colleagues[15,16] have developed a computer code that does this for ASM No. 1 and it can be used to examine the performance of SBRs accomplishing carbon oxidation, nitrification, and denitrification. Use of the program requires specification of the SBR operating characteristics. This may be accomplished by analogy with continuous systems.

The hydraulic and solids retention times for continuous flow systems were defined by Eqs. 4.15 and 5.1:

$$\tau = \frac{V}{F} \tag{4.15}$$

$$\Theta_c = \frac{V \cdot X}{F_w \cdot X_w} \tag{5.1}$$

In the case of an SBR, the wastewater flow rate, F, and the solids wastage flow rate, F_w, must be defined in terms of the total volume processed and wasted per day, respectively. The retention times defined by the above equations must be considered to be nominal retention times when applied to an SBR because the reactor volume, V, serves two functions, reaction plus settling. If the SBR is properly designed, all reactions will have been completed by the end of the react period and little change will occur in the soluble constituents during the settle period. In other words, all biochemical events can be assumed to take place during only the fill and react periods, with nothing except liquid–solid separation occurring during the settle and decant periods. This is the same assumption that was made in writing the equations describing continuous reactor performance in that time spent in the settler was not considered. To make Eqs. 4.15 and 5.1 equivalently applicable to SBRs, we must define an effective HRT, τ_c, and an effective SRT, Θ_{cc}, by multiplying both equations by the fraction of a cycle devoted to fill plus react, ζ:

$$\tau_c = \frac{\zeta \cdot V}{F} \tag{7.8}$$

$$\Theta_{cc} = \frac{\zeta \cdot V \cdot X}{F_w \cdot X_w} \tag{7.9}$$

If wastage is accomplished at the end of the react period before settling is started, then X_w will be equal to X and Eq. 7.9 will simplify to:

$$\Theta_{cc} = \frac{\zeta \cdot V}{F_w} \tag{7.10}$$

An SBR and a continuous system will have similar MLSS concentrations as long as the effective HRT and SRT of the SBR are equal to the HRT and SRT of the continuous system. As a consequence, they will have similar overall process performance if the hydraulic regime and other characteristics are similar. It should be stressed, however, that this will only be true as long as the effective HRTs and SRTs for the two systems are comparable.

Having set the effective HRT and SRT for an SBR, the major task is to determine the number of cycles required per day, N_c, thereby fixing the volume of wastewater applied per cycle, F_c, since by definition:

$$F_c = \frac{F}{N_c} \tag{7.11}$$

Fixing the number of cycles per day also fixes the cycle time because it is just the reciprocal of N_c. It should be emphasized that one cannot simply choose N_c freely. Rather it is tied to a number of other decision variables. One important decision variable is the volume retained in the SBR, V_{br}, following the draw period for the

purpose of returning biomass for the next cycle. Let α represent that volume expressed as a fraction of the feed added per cycle:

$$\alpha = \frac{V_{br}}{F_c} \tag{7.12}$$

It is analogous to the biomass recycle ratio in a continuous flow system. Another important decision variable is the volume retained in the SBR, V_{nr}, following the draw period for the purpose of returning nitrate for use as an electron acceptor at the beginning of the next cycle. Let β represent that volume expressed as a fraction of the feed added per cycle:

$$\beta = \frac{V_{nr}}{F_c} \tag{7.13}$$

It is analogous to the mixed liquor recirculation ratio in a continuous flow system. Recognizing that the volume of the SBR, V, must accommodate the incoming flow per cycle plus the volume retained per cycle gives:

$$V = F_c(1 + \alpha + \beta) \tag{7.14}$$

Substitution of Eqs. 7.11 and 7.14 into Eq. 7.8 relates the number of cycles per day to all of the important decision variables:

$$N_c = \frac{\zeta(1 + \alpha + \beta)}{\tau_c} \tag{7.15}$$

Thus, it can be seen why N_c cannot be chosen freely. All of the decision variables are linked and the choice of one will affect the choice of another. Nevertheless, by proper choice of α, β, and τ_c it is possible to have an SBR that mimics closely the performance of a continuous flow system with an equal SRT.

7.8.3 Effects of Cycle Characteristics

To illustrate the similarity in performance of continuous flow and SBR systems, simulations were performed with SBRSIM[16] using the parameter values listed in Table 6.3 and the wastewater characteristics in Table 6.6. Because SBRSIM implements IAWQ ASM No. 1, the reactions included were the same as those in the simulations with SSSP presented previously. The situation simulated was analogous to the MLE system depicted in Figure 7.28, with one important difference. A ten minute fill period was used, which made the SBR system behave like a plug–flow continuous reactor. Thus, rather than having a CSTR for the anoxic and aerobic zones, each behaved in a plug–flow manner.

The first set of simulations was done to illustrate the effect of SRT on system performance, so the conditions employed were analogous to those described in the legend to Figure 7.29. This means that the effective HRT was six hours, α was 0.5, and β was 2.0. Furthermore, the first half of the combined fill plus react period was anoxic with no dissolved oxygen whereas the second half was aerobic with a dissolved oxygen concentration of 2.0 mg/L. As stated above, the fill period lasted for ten minutes and thus occurred under anoxic conditions.

Before examining the effects of SRT on the performance of the SBR, it would be instructive to examine the changes occurring within a cycle, thereby establishing

the basis for system performance. Figure 7.43 illustrates the changes in the concentrations of readily biodegradable substrate (soluble organics), ammonia-N, and nitrate-N during a fill and react period of 101 minutes when the effective SRT was ten days. The anoxic and aerobic periods each occupied 50% of the fill plus react time. As would be anticipated, the concentrations of soluble organics and ammonia-N rose during the fill period as wastewater was added to the volume of mixed liquor retained from the previous cycle. The concentration of nitrate-N, which came from the volume retained from the previous cycle, dropped as it served as electron acceptor for biodegradation of the readily biodegradable substrate. It should be noted that carbon oxidation occurred throughout the fill period and that the reaction rate was sufficiently high to limit the buildup of soluble organics. Upon completion of the fill period, the remainder of the soluble organics and nitrate-N were rapidly depleted. In this case, the mass of nitrate-N retained from the previous cycle was well balanced with the mass of readily biodegradable COD added so that all nitrate-N was removed while little soluble organic matter remained. Nitrification was the main event during the aerobic react period and the reaction behaved in an almost zero order manner over much of the time as evidenced by the almost linear ammonia-N and nitrate-N curves. The soluble organics rose slightly because of their production by hydrolysis reactions. The model predicts that their concentration does not reach zero because they are continually being produced while they are being degraded. It is clear that the length of the anoxic period was excessive. Complete denitrification was achieved

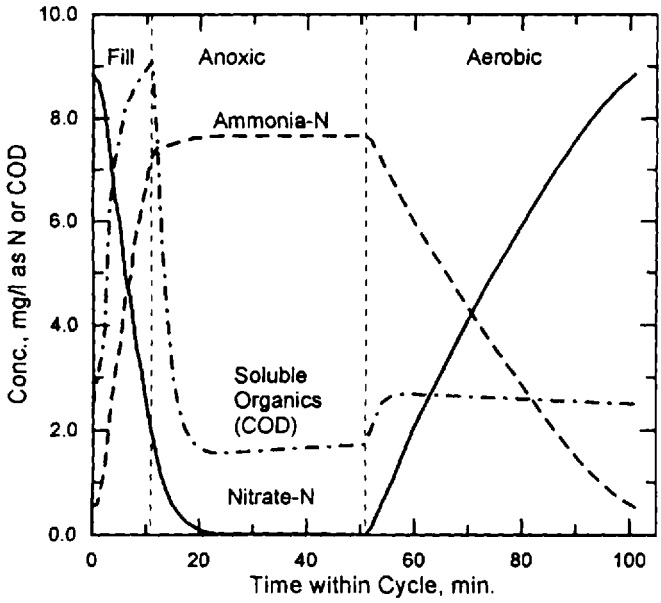


Figure 7.43 Performance of an SBR during a single cycle. The SBR was operated in a manner to mimic the performance of the MLE system shown in Figure 7.28. Effective HRT = 6 hr; effective SRT = 10 days; biomass recycle ratio = 0.5; mixed liquor recirculation ratio = 2.0; aerobic fraction = 50%. Influent flow = 1000 m^3/day. Influent concentrations are given in Table 6.6. Parameters are listed in Table 6.3. The dissolved oxygen concentration was zero during the anoxic period and 2.0 mg/L during the aerobic period.

within 25 minutes, after which little occurred due to lack of an electron acceptor. Furthermore, complete nitrification was not achieved during the aerobic period. Consequently, for this SRT, it would have been better to devote less time to anoxic conditions and more to aerobic conditions. The effects of the aerobic fraction of the fill plus react period will be addressed later.

The effect of SRT on the performance of this SBR system is shown by the solid lines in Figure 7.44. For comparison, the performance of an analogous MLE system is shown by the dashed lines. They are the same as the second bioreactor curves in Figure 7.29. The similarity in the performance of the two systems is apparent; the differences are due to the plug–flow nature of the SBR which allows it to achieve better removal of the soluble constituents. The particulate constituents are very similar, however, for the same reason that the particulate constituents were the same in the CSTR and tanks-in-series systems. This emphasizes the point that continuous flow and batch systems with the same effective SRT will contain the same amount of biomass, thereby making the biomass concentrations the same when the effective HRTs are the same. Consequently, the simple analytical model for a CSTR

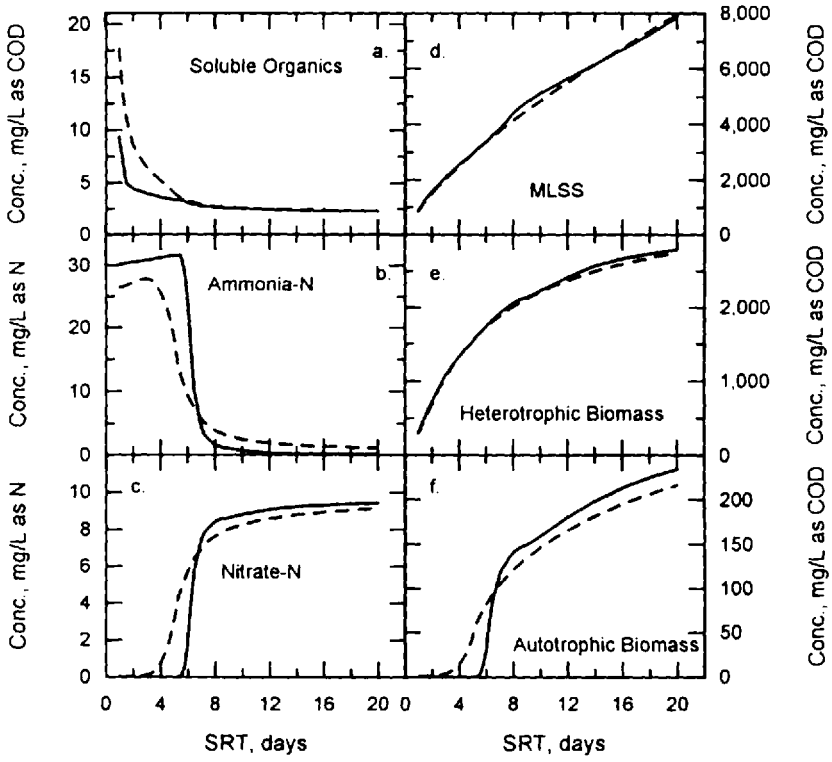


Figure 7.44 Effect of the effective SRT on the concentrations at the end of a cycle in an SBR operated in a manner to mimic the performance of the MLE system shown in Figure 7.28. For comparison, the dashed curves represent the concentrations in the aerobic (second) tank of that system. They are the same as the curves in Figure 7.29. With the exception of the SRT, the characteristics were the same as those listed in Figure 7.43.

in Chapter 5 can be used to help select the desired effective HRT for an SBR just as it can be used for the other bioreactor systems discussed in this chapter.

It was stated earlier that the volume retained in an SBR at the end of each cycle is analogous to the combined effects of biomass recycle and mixed liquor recirculation in a continuous flow system. Consequently, changing that volume should have an effect on system performance analogous to the effects of changing the mixed liquor recirculation ratio on the MLE system. The solid lines in Figure 7.45 show the effects of changing the retained mixed liquor recirculation volume expressed as a ratio of the influent volume added per cycle, β. For comparison, the effects on the analogous MLE system are shown as dashed lines. They are the same as the second bioreactor curves in Figure 7.30. The similarity between the two systems is obvious. More complete nitrification occurs in the SBR because of its plug–flow nature (due to the short fill period), resulting in slightly more autotrophic biomass and slightly higher nitrate-N concentrations. Otherwise, the effects are essentially the same.

The third variable that was examined when we investigated the MLE system was the fraction of the total system volume that was aerobic. The analogous variable

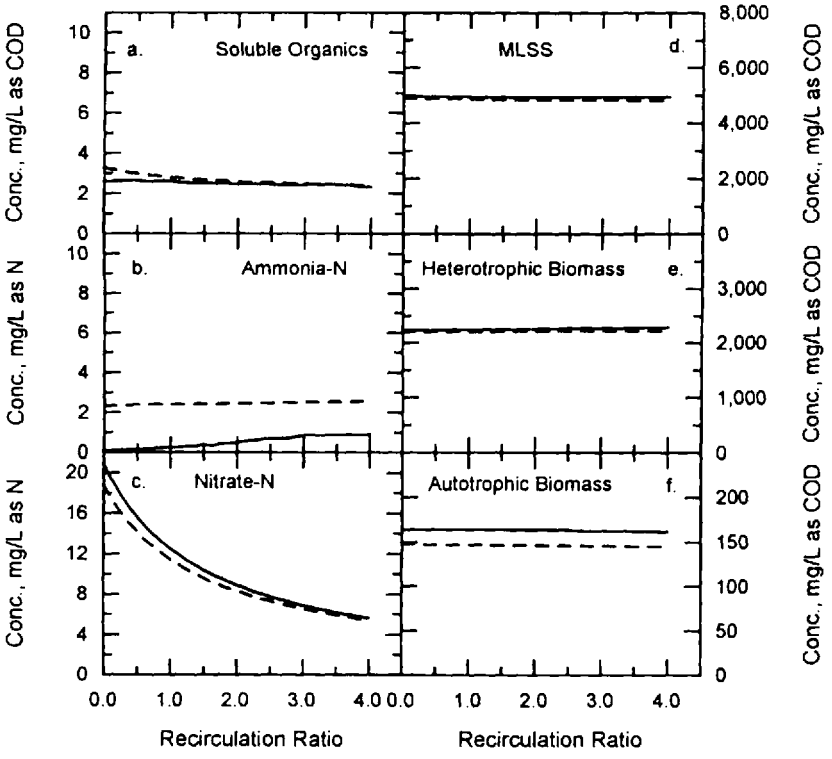


Figure 7.45 Effect of the mixed liquor recirculation ratio (β) on the concentrations at the end of a cycle in an SBR operated in a manner to mimic the performance of the MLE system shown in Figure 7.28. For comparison, the dashed curves represent the concentrations in the aerobic (second) tank of that system. They are the same as the curves in Figure 7.31. With the exception of the recirculation ratio, the characteristics were the same as those listed in Figure 7.43.

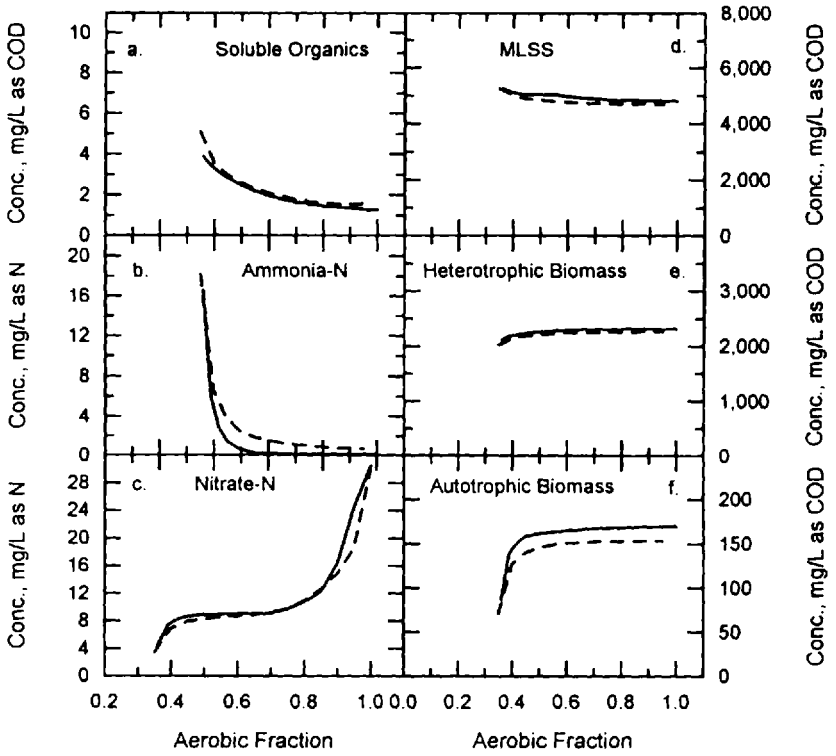


Figure 7.46 Effect of the fraction of the fill plus react period that is aerobic (aerobic fraction) on the concentrations at the end of a cycle in an SBR operated in a manner to mimic the performance of the MLE system shown in Figure 7.28. For comparison, the dashed curves represent the concentrations in the aerobic (second) tank of that system. They are the same as the curves in Figure 7.33. With the exception of the aerobic fraction, the characteristics were the same as those listed in Figure 7.43.

for the SBR is the fraction of the fill plus react period that is aerobic. Both can be expressed as the aerobic fraction. Figure 7.46 illustrates the effect of the aerobic fraction on the performance of both systems. Again, the similarity is striking, with the only differences being in the ammonia-N and autotrophic biomass concentrations, both of which are due to the plug–flow nature of this SBR. It will be recalled from Figure 7.43 that an anoxic period of around 30 minutes was adequate for complete denitrification at an SRT of 10 days, the value used in the simulations for Figure 7.46. This suggests that best system performance would occur with an aerobic fraction of 60–70%, which is indeed the case, as shown in Figure 7.46.

In summary, SBR systems can be operated in ways that mimic the performance of the continuous flow systems described in this chapter. As long as the effective HRT and effective SRT are the same, the two systems will contain similar biomass concentrations and have similar performance. Furthermore, variation of the length of the fill period will allow the performance of the SBR to approach any degree of mixing between complete mixing and plug–flow. As a consequence of their flexibility, SBRs are finding increased usage for wastewater treatment. Decisions related

to their design are addressed in Chapters 10 and 11 along with the continuous flow activated sludge and nutrient removal systems.

7.9 KEY POINTS

1. Several important variations of the activated sludge process can be simulated as systems containing CSTRs in series with various feed points and recycle streams. They are conventional activated sludge (CAS), step feed activated sludge (SFAS), contact stabilization activated sludge (CSAS), modified Ludzack–Ettinger (MLE) process, the Bardenpho process, and biological phosphorus removal (BPR) processes.
2. Biomass in each bioreactor of a multiple reactor system has a unique specific growth rate in response to the concentration of the growth limiting nutrient(s) in the bioreactor, whereas the SRT is defined with respect to the system as a whole. Thus, the SRT is not related to the specific growth rate in the same way that it is in a single bioreactor. Nevertheless, because the SRT is related to the net average specific growth rate of the entire system, it is still an important design and operational variable.
3. The process loading factor in each bioreactor of a chain of CSTRs is proportional to the specific growth rate of the bacteria in that bioreactor. Similarly, the instantaneous process loading factor achieved during the fill period for a sequencing batch reactor (SBR) is representative of the specific growth rate of the bacteria during that period.
4. At a given SRT, conventional or plug–flow activated sludge systems achieve slightly better removal or transformation of soluble constituents than a single CSTR, but exhibit almost identical biomass concentrations, oxygen utilization rates, and excess biomass production rates.
5. A CAS system exhibits less diurnal variability in its effluent organic substrate and ammonia-N concentrations than a CSTR operating at the same SRT, but experiences greater variability in its oxygen requirement.
6. Most soluble organic matter (SOM) is removed in the early part of a CAS system, whereas nitrification occurs over a greater part of the bioreactor system. The longer the SRT, the nearer the influent end of the bioreactor the reactions are completed.
7. Within the normal range found in practice, changes in the recycle ratio of biomass around a CAS system have little impact on system performance.
8. At a given SRT and system HRT, the MLSS concentration entering the settler from the last tank of an SFAS system with influent distributed evenly along the system is less than the concentration from a CAS system or from a single CSTR. System performance, however, is only slightly worse than that of a single CSTR.
9. An SFAS system with influent distributed evenly along the system exhibits more diurnal variability in its effluent organic substrate and ammonia-N concentrations than a CSTR operating at the same SRT, but experiences about the same variability in its oxygen requirement.

10. The MLSS in an SFAS system has a concentration gradient from the influent to the effluent end, with the influent end of the system having higher concentrations. As a consequence, soluble organic substrate and ammonia-N exhibit the opposite concentration gradient.
11. Increasing the recycle ratio around an SFAS system decreases the concentration gradient through the system and gives better system performance.
12. Although the removal of soluble organic substrate by a CSAS system is not as complete as the removal in a single CSTR, its response to changes in SRT is qualitatively similar. The response of ammonia-N to such changes is both qualitatively and quantitatively different, however.
13. Because the contact tank in a CSAS system has a lower MLSS concentration and a smaller volume than a single CSTR with the same SRT and system volume, the variations in the effluent concentrations of soluble organic substrate and ammonia-N in response to typical diurnal variations in loading are greater than those of the CSTR.
14. The biomass recycle ratio has a large impact on the performance of a CSAS system. Up to a point, the larger it is, the lower the concentrations of soluble organic substrate and ammonia-N in the effluent.
15. For a fixed total system volume, the larger the contact tank in a CSAS system, the more the system performs like a single CSTR of equal SRT and total system HRT.
16. When operated at an SRT sufficient to achieve a high degree of nitrification, the MLE process discharges less nitrate-N and uses less oxygen than a single CSTR because denitrification occurs in the first (anoxic) bioreactor.
17. Although the aerobic SRT is an important determinant of the degree of nitrification achieved in systems with unaerated zones, it is not the only one.
18. If the rate of recirculation of mixed liquor from the aerobic zone to the anoxic zone of an MLE system is insufficient to return enough electron acceptor for the amount of electron donor entering in the influent, the removal of nitrate-N by the system will not be as high as it could be.
19. For a fixed system volume and SRT, there is a region of optimum aerobic volume fraction that results in the minimum discharge of nitrogen from an MLE system.
20. By providing a second anoxic zone that achieves denitrification through use of slowly biodegradable substrate, the Bardenpho process can produce effluents with less total nitrogen than the MLE process can.
21. Selection of the sizes of the bioreactors in a Bardenpho system to achieve the minimum possible discharge of nitrogen requires use of optimization techniques.
22. There is a limited range of SRT over which a simple Phoredox (or A/O) system will work properly. If the SRT is too short, the PAOs will wash out. If it is too long, nitrifying bacteria will be able to grow, providing an inorganic electron acceptor to the anaerobic zone which allows the common heterotrophs to out-compete the PAOs for substrate, reducing the amount of phosphate removed.

23. There is an optimal split between the anaerobic and aerobic zones in the Phoredox (or A/O) system. Optimal phosphorus removal occurs when the aerobic SRT is small enough to exclude growth of autotrophic nitrifying bacteria, yet large enough to allow the PAOs to grow, and the time in the anaerobic zone is sufficiently large to allow efficient fermentation and uptake of the resulting acetate.
24. Sequencing batch reactors (SBRs) are batch reactors that can be used to treat wastewaters on a semicontinuous basis by going through a repeated sequence of steps consisting of fill, react, settle, decant, draw, and idle. They are very flexible in operation and can be used to mimic many complex continuous reactor systems by altering the length of the fill and react periods and by employing different environmental conditions (aerobic, anaerobic, anoxic) during those periods.
25. Continuous flow processes and SBRs have similar performance as long as they have the same effective SRT and HRT, as well as similar environmental conditions.

7.10 STUDY QUESTIONS

1. State the main characteristics of each of the following biochemical operations and draw sketches showing how each may be simulated: conventional activated sludge (CAS), step feed activated sludge (SFAS), contact stabilization activated sludge (CSAS), modified Ludzack–Ettinger (MLE) process, the Bardenpho process, and two-stage biological phosphorus removal (BPR).
2. Define the SRT for a system containing N bioreactors of volume V_i. What is the simplest way of controlling the SRT in such a system? Why?
3. Why does a CAS system remove or transform more soluble material than a CSTR with the same SRT, yet still have the same excess biomass production rate?
4. Draw a sketch comparing the variability in the oxygen requirement of a CAS system to that in a CSTR with the same SRT when both receive the same diurnal loading. What are the most important differences between the two systems? Why do they occur?
5. Draw a sketch depicting the tank to tank change in the concentrations of soluble organic matter and ammonia-N down a chain of bioreactors representative of CAS with an SRT sufficiently long to allow nitrification to occur. Why does the sketch look as it does? How will a change in the SRT influence the sketch? Why?
6. Why do changes in the recycle ratio of biomass around a CAS system have little impact on system performance?
7. Why is the MLSS concentration entering the settler from the last tank of an SFAS system less than the concentration from a CAS system or from a single CSTR with the same SRT and system HRT?
8. Why does an SFAS system exhibit more variability in its effluent organic substrate and ammonia-N concentrations in response to a diurnal load

than does a CSTR with equal SRT and system HRT? How do the variations in oxygen requirement compare? Why?

9. Why do the concentrations of soluble organic substrate and ammonia-N increase as the flow moves from tank to tank in an SFAS system?
10. Why does an increase in the recycle ratio around an SFAS system cause slightly better system performance?
11. Using a computer code for IAWQ ASM No. 1, investigate the impact of distributing the influent to the SFAS system described in Figure 7.11 so that it is equally divided among the first four bioreactors, with none entering the fifth. How does the steady-state performance of this system over a range of SRTs compare to the system in Figure 7.11? How does it compare to the CSTR? How does the dynamic response of the system to a diurnal load compare to that of the other two systems? Discuss the reasons for the differences in response.
12. Why is the response of ammonia-N to changes in the SRT of a CSAS system qualitatively different from the response of soluble organic matter?
13. Even though the stabilization tank in the CSAS system receives biomass recycle flow at a constant rate, it still experiences variability in its oxygen requirement when the system receives typical diurnal variations in loading. Why?
14. Why does the biomass recycle ratio have a large impact on the performance of a CSAS system?
15. For a fixed total system volume, as the volume of the contact tank is increased in a CSAS system, the concentration of ammonia-N changes in a manner that is qualitatively different from the manner in which the soluble organic substrate concentration changes. Why?
16. Using a computer code for ASM No. 1, investigate the effect of HRT on the performance of the CSAS system described in Figure 7.20. Maintain the SRT constant at a value of 12 days and vary the HRT over the range of 1.0 to 24 hr by changing the total system volume while keeping the two bioreactors of equal size. Explain your results.
17. Why does the MLE process fail when its operational conditions are such that nitrification cannot occur?
18. Although the aerobic SRT is an important determinant of the extent of nitrification achieved in a system with unaerated zones, equivalent degrees of nitrification will not be achieved in systems with and without unaerated zones, even though they may have the same aerobic SRTs. Why?
19. Draw a sketch showing the effect of mixed liquor recirculation ratio on the performance of the MLE process and explain why it has the effect that it does. Why does the impact of increased recirculation reach a point of diminishing return?
20. Why is there a region of optimum aerobic volume fraction that results in the minimum discharge of nitrogen from an MLE system of fixed total volume and SRT?
21. Using a computer code for ASM No. 1, investigate the effect of HRT on the performance of the MLE system described in Figure 7.29. Maintain the SRT constant at a value of 12 days and vary the HRT by changing

the total system volume while keeping the aerobic fraction constant at 50%. Explain your results.

22. Using a computer code for ASM No. 1, explore the interactions between mixed liquor recirculation ratio and aerobic volume fraction for the MLE system described in Figure 7.29. Maintain the SRT constant at a value of 12 days and the total volume constant at 250 m^3. Vary the recirculation ratio over the range of 3 to 5 and the aerobic volume fraction over the range of 0.4 to 0.8. Plot total nitrogen concentration in the effluent as a function of aerobic volume fraction with recirculation ratio as the parameter and determine the best combination for the system. Then explain your results.
23. A Bardenpho system can achieve an effluent with less total nitrogen in it than an MLE system can, even though the two systems have the same SRT, total system volume, and aerobic volume fraction. Why?
24. Using a computer code for ASM No. 1, investigate the effect on the effluent nitrogen concentration of changing the distribution of anoxic volume between bioreactors one and three in the Bardenpho system described in Figure 7.36. Use an SRT of 10 days. Why does the system respond as it does?
25. Explain why the occurrence of nitrification has a deleterious effect on biological phosphorus removal in the Phoredox process. Then suggest a process configuration that would achieve a high degree of biological phosphorus removal, complete nitrification, and partial denitrification, and explain why it would work.
26. Using a computer code for ASM No. 2, demonstrate the effects of the key process variables on the performance of the process you suggested in Study Question 25. Use the wastewater characteristics listed in Table 7.1 and the default kinetic parameters from the computer code.
27. Describe how you would set up an SBR system to be analogous to the four-tank Bardenpho system described in Figure 7.36.
28. Using a computer code for ASM No. 1, repeat Study Question 24 for the SBR system identified in Study Question 27.

REFERENCES

1. Barnard, J. L., Biological denitrification. *Water Pollution Control* **72**:705–717, 1973.
2. Barnard, J. L., A review of biological phosphorus removal in the activated sludge process. *Water SA* **2**:136–144, 1976.
3. Bidstrup, S. M. and C. P. L. Grady Jr., SSSP—simulation of single-sludge processes. *Journal, Water Pollution Control Federation* **60**:351–361, 1988.
4. Dold, P. L. and G. v. R. Marais, Benefits of including unaerated zones in nitrifying activated sludge plants. *Water Science and Technology* **19**(7):195–207, 1987.
5. Erickson, L. E. and L. T. Fan, Optimization of the hydraulic regime of activated sludge systems. *Journal, Water Pollution Control Federation* **40**:345–362, 1968.
6. Garrett, M. T., Hydraulic control of activated sludge growth rate. *Sewage and Industrial Wastes* **30**:253–261, 1958.
7. Gould, R. H., Operating experiences in New York City. *Sewage Works Journal* **14**:70–80, 1942.

8. Gujer, W., ASIM: *Activated Sludge SIMulation Program, Version 3.0*, Swiss Federal Institute for Environmental Science and Technology, Dübendorf, Switzerland, 1994.
9. Henze, M., W. Gujer, T. Mino, T. Matsuo, M. C. Wentzel, and G. v. R. Marais, Activated sludge model No. 2. *IAWQ Scientific and Technical Reports*, No. 3, 1995.
10. Irvine, R. L. and L. H. Ketchum Jr., Sequencing batch reactors for biological wastewater treatment. *CRC Critical Reviews in Environmental Control* **18**:255–294, 1989.
11. Keinath, T. M., Operational diagrams and control of secondary clarifiers. *Journal, Water Pollution Control Federation* **57**:770–776, 1985.
12. Levenspiel, O., *Chemical Reaction Engineering*, 2nd ed., Wiley, New York, 1972.
13. Ludzack, F. J. and M. B. Ettinger, Controlling operation to minimize activated sludge effluent nitrogen. *Journal, Water Pollution Control Federation* **34**:920–931, 1962.
14. Murphy, K. L. and B. I. Boyko, Longitudinal mixing in spiral flow aeration tanks. *Journal of the Sanitary Engineering Division, ASCE* **96**:211–221, 1970.
15. Oles, J. and P. A. Wilderer, Computer aided design of sequencing batch reactors based on the IAWPRC activated sludge model. *Water Science and Technology* **23**(4–6):1087–1095, 1991.
16. Oles, J., P. A. Wilderer, J. Schelling and J. Gritzner, *SBRSIM—Simulation of Reaction Kinetics in Sequencing Batch Reactors Based on the IAWPRC Activated Sludge Model*, Version 1.03, Technical University Hamburg-Harburg, Germany, 1992.
17. Randall, C. W., Introduction. In: *Design and Retrofit of Wastewater Treatment Plants for Biological Nutrient Removal*, C. W. Randall, J. L. Barnard and H. D. Stensel, eds., Technomic Publishing, Lancaster, Pennsylvania, pp. 1–24, 1992.
18. Randall, C. W., H. D. Stensel and J. L. Barnard, Design of activated sludge biological nutrient removal plants. In: *Design and Retrofit of Wastewater Treatment Plants for Biological Nutrient Removal*, C. W. Randall, J. L. Barnard and H. D. Stensel, eds., Technomic Publishing, Lancaster, Pennsylvania, pp. 97–183, 1992.
19. Sedlak, R. I. *Principles and Practice of Nutrient Removal from Municipal Wastewater*, The Soap and Detergent Association, New York, 1989.
20. Spector, M. L., U.S. Patent 4,162,153, July 24, 1979.
21. Stensel, H. D., Principles of biological phosphorus removal In: *Principles and Practice of Nutrient Removal from Municipal Wastewater*, R. I. Sedlak, ed., The Soap and Detergent Association, New York, pp. 123–143, 1989.
22. Thompson, D., D. T. Chapman, and K. L. Murphy, Step feed control to minimize solids loss during storm flows. *Research Journal, Water Pollution Control Federation* **61**: 1658–1665, 1989.
23. Toerber, E. D., W. L. Paulson and H. S. Smith, Comparison of completely mixed and plug flow biological systems. *Journal, Water Pollution Control Federation* **46**:1995–2014, 1974.
24. Ulrich, A. H. and M. W. Smith, The biosorption process of sewage and waste treatment. *Sewage and Industrial Wastes* **23**:1248–1253, 1951.
25. Water Research Commission, *Theory, Design and Operation of Nutrient Removal Activated Sludge Processes*, Water Research Commission, Pretoria, South Africa
26. Wentzel, M. C., P. L. Dold, G. A. Ekama and G. v. R. Marais, Enhanced polyphosphate organism culture in activated sludge systems. Part III: Kinetic model. *Water SA* **15**:89–102, 1989.
27. Wentzel, M. C., G. A. Ekama and G. v. R. Marais, Kinetics of nitrification denitrification biological excess phosphorus removal systems—a review. *Water Science and Technology* **23**(4–6):555–565, 1991.
28. Youker, B. A., Design of single-sludge systems: the modified Ludzack–Ettinger and Bardenpho processes. Master of Engineering Report, Clemson University, Clemson, South Carolina, 1988.

8

Techniques for Evaluating Kinetic and Stoichiometric Parameters

In the preceding chapters of Part II, we have examined models for characterizing the performance of ideal suspended growth bioreactors. Before those models can be used for design and evaluation of wastewater treatment systems, however, values must be available for the kinetic and stoichiometric parameters in them. Some of those values may be obtained from the literature or from experience with the particular wastewater to be treated. Generally, however, parameters must be evaluated experimentally during treatability studies. In this chapter, we will review procedures for evaluating the parameters in the models presented in Chapters 5–7.

8.1 TREATABILITY STUDIES

Treatability studies must be carefully designed. That is because biological reactors are subject to their own version of the Heisenberg uncertainty principle. Biological wastewater treatment systems all use mixed microbial communities, which involve complex interactions among their members. In other words, bioreactors contain their own ecosystems. As a consequence, the physical characteristics of the treatment system and the manner in which it is operated will influence the composition of the community that develops within it, as well as the physiological state of the microorganisms in that community. The nature of the community and the state of the microorganisms in it, in turn, will define the kinetics exhibited by it, which will determine its performance. So, when we go into the laboratory, set up a bench-scale bioreactor, seed it with biomass from a suitable source, and begin to feed it with a wastewater, we are, to a large degree, predetermining the outcome of the study, although we may have no idea before hand just what that outcome will be. This is not to suggest that treatability studies are a waste of time. Quite to the contrary. They are essential to successful design, particularly for unique wastewaters. This follows from the fact that the nature of the wastewater has the primary impact on community composition, with bioreactor configuration playing a secondary, but important, role.

Because of the impact of the nature of the bioreactor system on the composition of the microbial community that develops, it is necessary to approach the study of complex systems in a staged manner. Generally, initial studies are done with the simplest possible system. If the focus is only on oxidation of carbon and/or nitrogen, then the system will usually contain a single continuous stirred tank reactor (CSTR).

However, if denitrification is to be considered along with carbon and nitrogen oxidation, then a modified Ludzack–Ettinger (MLE) system consisting of two CSTRs in series must be used. Finally, consideration of biological phosphorus removal would require two to three CSTRs in series, depending upon whether nitrification will also occur. Regardless of the bioreactor configuration, if gravity sedimentation is to be used for biomass recycle in the full-scale system, then it should be employed during the treatability study as well. As seen in Chapter 10, bioflocculation is still not clearly understood, although certain bacteria are known to be more prone to growth in floc than others. If we are dependent upon such bacteria in the full-scale system, it is essential that they be present in the microbial community used to develop the kinetic and stoichiometric data from which that system will be designed.

The simplest system should be run at a number of SRTs while appropriate performance data are collected. Generally, it is desirable to operate at a minimum of four different solids retention time (SRT) values. Because of the importance of SRT in determining both the nature of the microorganisms present and their physiological state, the bioreactors should be run for an extended period of time before data collection begins. Usually, the time is expressed in terms of SRT, although what constitutes a sufficient number of SRTs is subject to debate. It is generally accepted that three SRTs are the minimum acceptable to achieve stability, and most treatability studies use that value because of the cost of running bioreactors. Sampling should continue long enough to determine whether the system is stable. This normally requires several SRTs, with many studies using two as a practical lower limit, although four would be preferable. Because of the necessity to operate at several SRTs for extended periods of time, most studies run several systems simultaneously.

During operation, it is important that tight control be maintained over SRT. This means that computation of the SRT (or the wastage rate, F_w, required to maintain a desired SRT) must consider the biomass lost in the effluent from the settler, X_{Me}:

$$\Theta_c = \frac{VX_M}{F_wX_{Mw} + (F - F_w)X_{Me}} \tag{8.1}$$

For simplicity, most bench-scale bioreactors are operated with the Garrett flow scheme with wastage directly from the bioreactor. It is generally not feasible to waste continually from a bench-scale bioreactor because of the small flows involved. However, wastage should be done a sufficient number of times per day to limit the amount of biomass removed each time to no more than 5% of the total biomass in the system.

After the data have been analyzed and initial estimates have been obtained for the parameter values, simulation studies should be run with an appropriate model to allow investigation of alternative bioreactor configurations. From those studies, the systems considered to be the most likely to meet the effluent criteria can be chosen and subjected to preliminary engineering analysis. One or more of those systems should then be chosen for further bench- or pilot-scale testing to verify the selection and tune the model parameters for the biomass that develops in the chosen system.

Throughout the entire staged approach to treatability testing, care should be taken to ensure that conditions are optimal for microbial growth. Bioreactors cannot be expected to perform satisfactorily if environmental conditions are inadequate. For example, no inorganic nutrients should be limiting. Guidelines are given in Section 3.8.2 to help in the determination of the proper amounts. In addition, sufficient oxygen should be provided to aerobic bioreactors so that it is not rate limiting and

the pH should be maintained near neutrality. Finally, the temperature should be maintained as constant as possible because the kinetic and stoichiometric parameters will be unique to the temperature. If large seasonal temperature fluctuations are expected in the final facility, then studies should be conducted at the extreme values to allow correction of the parameter values for temperature effects.

8.2 SIMPLE SOLUBLE SUBSTRATE MODEL WITH TRADITIONAL DECAY AS PRESENTED IN CHAPTER 5

This model should be applied only to wastewaters without significant quantities of particulate organic matter for which the primary focus is on carbon oxidation. It can also be applied to situations in which nitrification is an objective. The focus here will be on carbon oxidation, but the reader can extend the principles presented to nitrification, in which case the substrate would be ammonia-N rather than soluble, biodegradable organic matter. Generally, particulate organic matter is operationally defined as the material that will be retained on a 0.45 μm pore-size filter. Many colloidal sized particles will pass such a filter, and therefore, in a strict sense, "soluble" organic matter may not all be truly soluble. Nevertheless, for purposes of parameter estimation it is generally acceptable to apply the model of Chapter 5 to any wastewater in which the organic matter will all pass such a filter.

8.2.1 Data to Be Collected

For this application the test bioreactors should be simple CSTRs with biomass recycle. They should be operated at a number of SRTs and the following data should be collected during the steady-state period following stabilization:

S_{CO} = Soluble chemical oxygen demand (COD) in the influent (mg/L)

S_C = Soluble COD in the bioreactor (mg/L)

X_T = Total biomass COD in the bioreactor (mg/L)

X_{Tw} = Total biomass COD in the waste solids (mg/L) (This will be the same as X_T if the Garrett flow scheme is used.)

X_{Te} = Total biomass COD in the final effluent (mg/L)

f_A = Active fraction of biomass

V = Reactor volume under aeration (L)

F = Influent flow rate (L/h)

F_w = Waste solids flow rate (L/h)

Several points should be noted about the data to be collected. To be consistent with the models in Chapters 5–7, the biomass concentrations are expressed in COD units. They can also be measured in mass units, either as total suspended solids (TSS) or as volatile, i.e., organic, suspended solids (VSS). If that is done, the yield, Y_H, will have units of TSS or VSS formed per unit of substrate COD removed. In such a situation, in order to have consistent units, the yield must be multiplied by a conversion factor, i, when it is used in the computation of the oxygen requirement,

RO, by Eq. 5.33. The value of the conversion factor will depend upon the method of measuring the biomass concentration, as discussed in Sections 2.4.1 and 5.1.3. When biomass is measured as TSS, i is denoted as $i_{O\,XB,T}$ and is generally considered to have a value of 1.20 g COD/g TSS (as shown in Table 3.1) unless data are collected that suggest otherwise. Similarly, when biomass is measured as VSS, i is denoted as $i_{O\,XB,V}$ and is considered to have a value of 1.42 g COD/g VSS, as shown in Table 3.1. The active fraction of the biomass, f_A, is the most difficult data to collect during treatability studies. As a consequence, most studies do not try to measure it. In Section 8.3 we examine how to estimate parameters in the absence of such data. A number of techniques have been proposed for measuring the active fraction, but all are tedious and subject to error. Consequently, they are used mostly in a research setting. The most direct method is the slide culture technique of Postgate,[32] which involves plating bacteria on microscope slides and observing the fraction that divide. This works well with soluble substrates but becomes more difficult when particulate materials are present. An indirect method involves quantifying the amount of adenosine triphosphate (ATP) present per unit of biomass. It has been used successfully because the amount of ATP per viable cell is relatively independent of SRT and ATP is quickly lost from nonviable cells.[31,38] Another indirect method involves measurement of the amount of deoxyribonucleic acid (DNA) present per unit of biomass. Like ATP, it is relatively independent of the SRT[37] and is quickly degraded when cells die and lyse.[35]

The data collected during the treatability study will be used to estimate the values of $\hat{\mu}$, K_S, Y_H, b_H, and f_D. In the process of doing this we will also have to estimate the soluble inert COD, S_I. Because many of the equations describing the performance of a CSTR can be reduced to linear form, graphical procedures have commonly been used to estimate the parameters. Linear transformations usually change the structure of the error in a data set, and nonlinear parameter estimation techniques are preferable whenever possible. However, because an explanation of such techniques is beyond the scope of this book, the linear techniques will be described. Because some of the parameters appear in more than one equation it is necessary to determine them in a sequential manner when the linear techniques are employed. Regardless of the estimation technique employed, however, it is important to recognize that all parameters estimated from a data set are interrelated. Consequently, an error in the estimation of one will influence the estimated values of the others. This means that more emphasis should be placed on the parameter set as a whole than on any individual values within it.

8.2.2 Determination of Y_H and b_H

The first parameters to be estimated are the biomass yield, Y_H, and the traditional decay coefficient, b_H. As presented in Chapter 5, Y_H has units of mg biomass COD formed per mg of substrate COD used whereas b_H has units of hr^{-1}. Both can be obtained from a rearranged form of Eq. 5.20:

$$X_{B,H} \cdot \tau = \frac{\Theta_c \cdot Y_H(S_{SO} - S_S)}{1 + b_H \cdot \Theta_c} \tag{5.20}$$

Examination of Eq. 5.20 reveals that the units on Y_H must be consistent with the units of $X_{B,H}$, i.e., $X_{B,H}$ must be measured in COD units to give a Y_H value in COD

units. As discussed in the preceding section, it is often desirable to measure the biomass concentration in TSS or VSS units. In that case, Y_H will have similar units. It makes no difference which unit system is used provided that it is used consistently and that the appropriate COD conversion factor, $i_{O\,XB,T}$ or $i_{O\,XB,V}$, is used when COD balances are performed. During the experimental studies, measurements were made of the total biomass concentration, X_T, not the active biomass, $X_{B,H}$. Thus, use must be made of the active fraction to get $X_{B,H}$:

$$X_{B,H} = f_A \cdot X_T \tag{8.2}$$

Measurements were also made of the soluble COD in the feed, S_{CO}, and in the bioreactor, S_C, not the biodegradable COD, which is the substrate. The concentrations of biodegradable COD can be obtained from the measured soluble COD values by subtracting the inert soluble COD, S_I, which passes through the bioreactor:

$$S_{SO} = S_{CO} - S_I \tag{8.3}$$

$$S_S = S_C - S_I \tag{8.4}$$

However, for use in Eq. 5.20, knowledge of the inert soluble COD is not required because it cancels out:

$$S_{SO} - S_S = S_{CO} - S_C \tag{8.5}$$

Substitution of Eqs. 8.2 and 8.5 in Eq. 5.20 yields:

$$f_A \cdot X_T \cdot \tau = \frac{\Theta_c \cdot Y_H(S_{CO} - S_C)}{1 + b_H \cdot \Theta_c} \tag{8.6}$$

Data are collected on the effect of SRT (Θ_c) on f_A, X_T, and S_C, with τ and S_{CO} as controlled input values. Consequently, sufficient information is available for estimation of Y_H and b_H. The most suitable method for doing this depends on the structure of the errors in the data.[2] If the errors have constant variance, then a nonlinear least squares technique applied directly to Eq. 8.6 without transformation is the most appropriate method.[2] Such techniques are available in commercially available statistical software, such as SAS,[3,4] as well as in many spreadsheet and graphics programs. If nonlinear least squares estimation is inappropriate or if the analyst does not have access to appropriate software, then a linear least squares technique must be used. Linear least squares procedures are available on most engineering calculators as well as in almost all spreadsheet and graphics programs. Equation 8.6 can be linearized to give:

$$\frac{S_{CO} - S_C}{f_A \cdot X_T \cdot \tau} = \frac{b_H}{Y_H} + \frac{1}{Y_H} \cdot \frac{1}{\Theta_c} \tag{8.7}$$

A plot of $(S_{CO} - S_C)/(f_A \cdot X_T \cdot \tau)$ vs. $1/\Theta_c$ will give a straight line with a slope of $1/Y_H$ and an ordinate intercept of b_H/Y_H. This is illustrated in Figure 8.1.

If the study is conducted in CSTRs without biomass separators, then Θ_c and τ will be the same, and Eq. 5.21 must be used:

$$X_{B,H} = \frac{Y_H(S_{SO} - S_S)}{1 + b_H \cdot \tau} \tag{5.21}$$

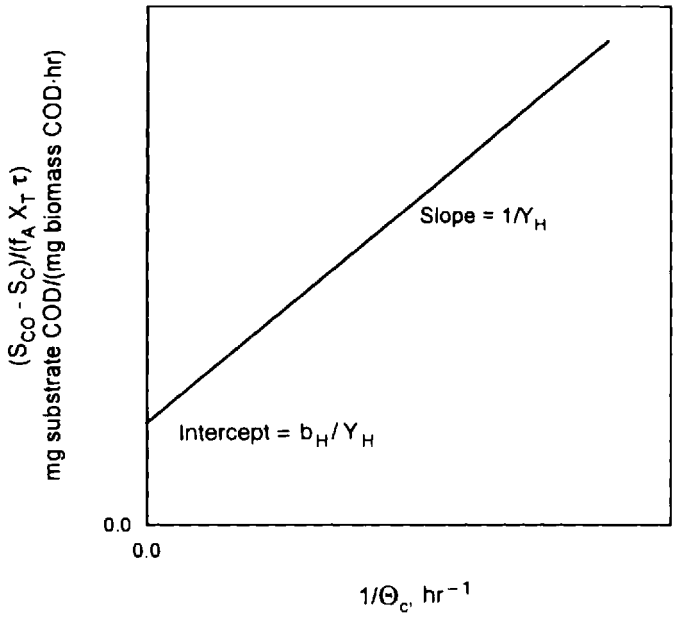


Figure 8.1 Plot of Eq. 8.7 for the determination of Y_H and b_H.

Making the same substitutions and linearizing gives:

$$\frac{S_{CO} - S_C}{f_A \cdot X_T} = \frac{1}{Y_H} + \frac{b_H}{Y_H} \cdot \tau \tag{8.8}$$

A plot of $(S_{CO} - S_C)/(f_A \cdot X_T)$ vs. τ will give a straight line with a slope of b_H/Y_H and an ordinate intercept of $1/Y_H$. This is illustrated in Figure 8.2.

This technique is better than alternative linearizations when the error is either normally or log-normally distributed and the coefficient of variation is less than 11%, or when the error is uniformly distributed, regardless of the coefficient of variation.[28]

8.2.3 Determination of f_D

The value of f_D can be determined from Eq. 5.26 using nonlinear least squares analysis with b_H as a fixed value obtained from the preceding analysis:

$$f_A = \frac{1}{(1 + f_D \cdot b_H \cdot \Theta_c)} \tag{5.26}$$

Alternatively, rearrangement gives:

$$\frac{1}{f_A} = 1 + f_D \cdot b_H \cdot \Theta_c \tag{8.9}$$

A plot of $1/f_A$ vs. Θ_c will give a straight line with a slope of $f_D \cdot b_H$. The ordinate intercept should pass through 1.0. Because b_H is known, f_D can be calculated.

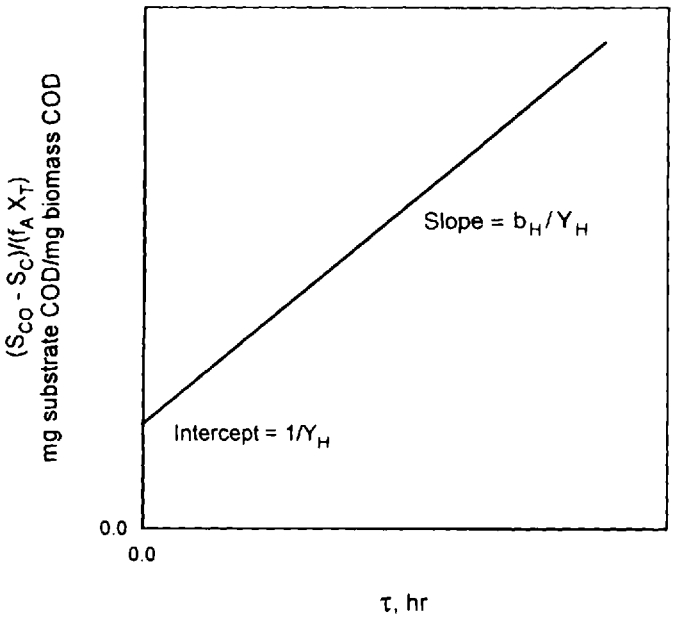


Figure 8.2 Plot of Eq. 8.8 for determination of Y_H and b_H.

8.2.4. Estimation of Inert Soluble Chemical Oxygen Demand, S_I

Before the kinetic parameters describing microbial growth and substrate utilization can be estimated, data must be available on the soluble substrate concentration, S_S. This requires knowledge of the inert soluble COD, S_I, as shown in Eq. 8.4. The easiest way to determine S_I is to remove an aliquot of the mixed liquor from one of the bioreactors operating at an SRT of 10 days or more, place it in a batch reactor, and aerate it. The soluble COD should be measured over time and when it reaches a stable residual value, that value can be considered to be equivalent to the concentration of inert soluble COD in the feed.[19]

Mamais et al.[24] have suggested that S_I can be estimated as the truly soluble COD remaining in the effluent from a bioreactor operated with a long SRT. The rationale for this technique is that at longer SRTs, the amount of soluble, readily biodegradable COD remaining in the effluent will be negligibly small. Consequently, essentially all soluble COD remaining will be nonbiodegradable. The truly soluble COD is obtained by flocculating an effluent sample from the bioreactor with the longest SRT with $ZnSO_4$ at pH 10.5 (forming $Zn(OH)_2$ floc) prior to filtration through a 0.45 μm membrane filter. The flocculation step effectively removes colloidal organic matter that might pass through the filter, leaving only S_I.

Both of the above techniques are approximations. As discussed in Section 3.4, bacteria produce soluble microbial products as they degrade organic matter. Consequently, part of the inert organic matter remaining will actually be of microbial origin. However, because the models employed herein do not explicitly account for soluble microbial products, it is acceptable to consider it as part of the inert soluble

COD as long as a constant influent soluble COD, S_{CO}, is used in the treatability study. Germirli et al.[12] have proposed a simple technique whereby the residual COD from the test described above may be partitioned into inert soluble COD from the influent and from microbial activity. It may be used in situations in which it is necessary to explicitly account for soluble microbial products.

8.2.5 Estimation of Monod Parameters, $\hat{\mu}_H$ and K_S.

Once S_I is known, S_S can be calculated with Eq. 8.4, thereby allowing $\hat{\mu}_H$ and K_S to be determined from Eq. 5.13 using the value of b_H determined previously:

$$S_S = \frac{K_S(1/\Theta_c + b_H)}{\hat{\mu}_H - (1/\Theta_c + b_H)} \tag{5.13}$$

As with Y_H and b_H, the best way to determine $\hat{\mu}_H$ and K_S is through use of a nonlinear least squares analysis. S_S is treated as the dependent variable, with $1/\Theta_c + b_H$ as the independent variable. If a nonlinear least squares routine is unavailable, or if the error structure is inappropriate for nonlinear techniques, Eq. 5.13 can be linearized to allow linear least squares analysis to be used. Three techniques are available.[7]

Hanes Linearization. If the plot is to be drawn by eye without use of a least squares analysis, the Hanes linearization is preferable:

$$\frac{S_S}{1/\Theta_c + b_H} = \frac{K_S}{\hat{\mu}_H} + \frac{S_S}{\hat{\mu}_H} \tag{8.10}$$

As illustrated in Figure 8.3, a plot of $S_S/(1/\Theta_c + b_H)$ vs. S_S will yield a straight line with a slope of $1/\hat{\mu}_H$ and an ordinate intercept equal to $K_S/\hat{\mu}_H$. The scales can be chosen to give good estimates of $\hat{\mu}_H$ and K_S, but the linear least squares technique cannot be used to find the line of best fit because both axes contain terms that are subject to error, i.e., S_S.

Hofstee Linearization. If it is desired to use the linear least squares technique to determine the line of best fit, the Hofstee linearization should be used:

$$\frac{1/\Theta_c + b_H}{S_S} = \frac{\hat{\mu}_H}{K_S} - \frac{1}{K_S}(1/\Theta_c + b_H) \tag{8.11}$$

This is illustrated in Figure 8.4. The use of the reciprocal of S_S may amplify the error in S_S and make it difficult to fit the line by eye. Furthermore, because the independent variable $(1/\Theta_c + b_H)$ appears in both axes there will be some degree of inevitable correlation.

Lineweaver–Burk Linearization. To eliminate the correlation in Eq. 8.11, a slightly rearranged form of the equation may be used:[28]

$$\frac{1}{S_S} = \frac{\hat{\mu}_H}{K_S}\left(\frac{1}{1/\Theta_c + b_H}\right) - \frac{1}{K_S} \tag{8.12}$$

This is illustrated in Figure 8.5. The Lineweaver–Burk method has been found to give a deceptively good fit, even with unreliable data points, and thus some authors have recommended against its use.[7] Ong,[28] however, has found that Eq. 8.12 works well for bioreactor data when the error is either normally or log-normally distributed.

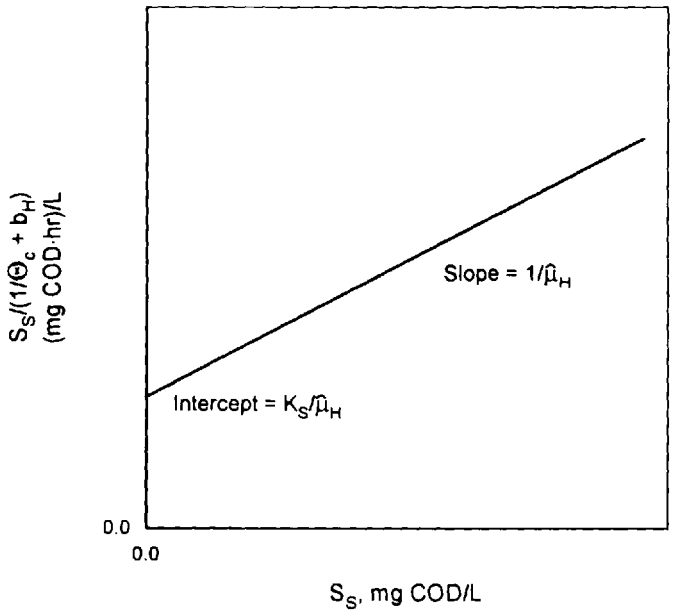


Figure 8.3 Hanes plot (Eq. 8.10) for determination of $\hat{\mu}_H$ and K_S.

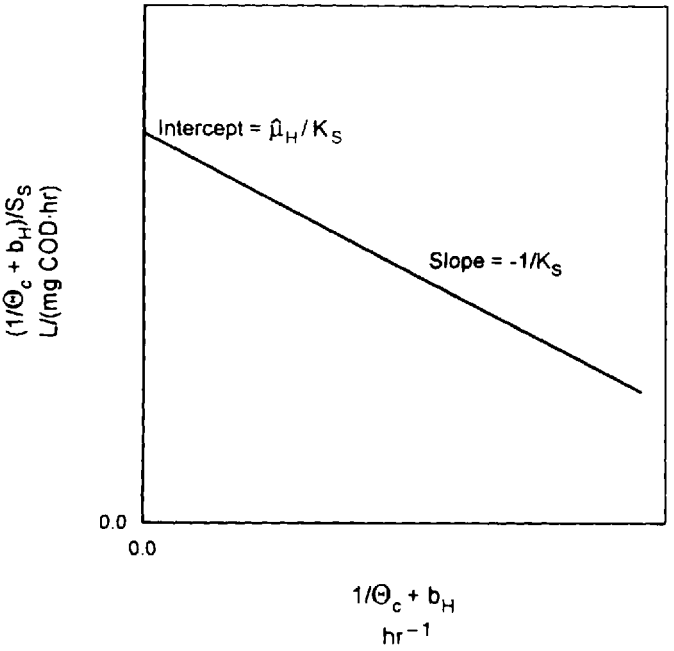


Figure 8.4 Hofstee plot (Eq. 8.11) for determination of $\hat{\mu}_H$ and K_S.

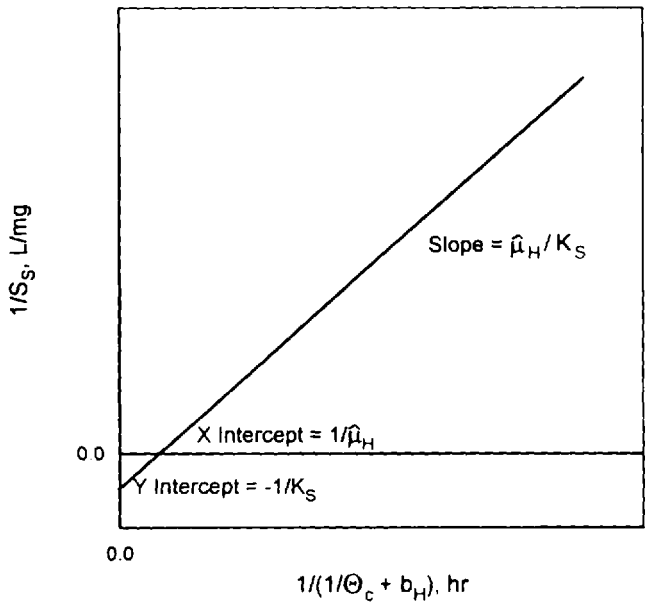


Figure 8.5 Lineweaver–Burk plot (Eq. 8.12) for determination of $\hat{\mu}_H$ and K_S.

8.2.6 Estimation of k_e

In situations in which the organic compounds in the wastewater are relatively easy to degrade, the soluble biodegradable COD in the effluent from the CSTRs is likely to be small relative to the half-saturation coefficient. This makes it difficult to obtain independent estimates of $\hat{\mu}_H$ and K_S. In that situation it is better to use the first order approximation of the Monod equation to express the biodegradation kinetics of the wastewater constituents. It was given in Eq. 5.41:

$$S_S = \frac{1}{Y_H \cdot k_e} (1/\Theta_c + b_H) \tag{5.41}$$

In that case, a plot of S_S versus $(1/\Theta_c + b_H)$ will yield a straight line that passes through the origin. Its slope will be $1/(Y_H \cdot k_e)$ and can be determined from linear least squares analysis. Because the value of Y_H is already known, the slope can be used to determine the value of k_e, the mean reaction rate coefficient.

Example 8.2.1

A treatability study was performed on an industrial wastewater using CSTRs with biomass recycle. The wastewater contained a complex mixture of soluble organic compounds with a total soluble COD of 350 mg/L. The wastewater contained no particulate matter. The studies were run in lab scale bioreactors that had a volume of 6.0 L. The flow rate was maintained at a constant rate of 1.0 L/hr and the SRT was maintained at the desired values by wasting excess biomass directly from the reactors. The wastage rate was corrected for the loss of biomass in the effluent. Batch studies with biomass removed from one of the reactors revealed

Table E8.1 Data Collected During a Treatability Study with a Soluble Industrial Wastewater

SRT hrs	Soluble COD mg COD/L	Total biomass mg COD/L	Active fraction
48	44.2	1,000	0.95
96	39.9	1,740	0.91
144	38.7	2,320	0.87
192	38.2	2,800	0.83
288	37.6	3,600	0.76
384	37.4	4,280	0.71

that the concentration of inert soluble COD was 35 mg/L. Using the data provided in Table E8.1, estimate the values of the parameters $\hat{\mu}_H$, K_S, Y_H, b_H, and f_D. Use the linearized forms of the equations to do this.

a. The first task is to determine Y_H and b_H. This is done by plotting $(S_{CO} - S_C)/(f_A \cdot X_T \cdot \tau)$ versus $1/\Theta_c$ as indicated in Eq. 8.7. The values of S_C, the soluble COD, f_A, the active fraction, and X_T, the total biomass concentration, are given in Table E8.1. Because the bioreactors all have a volume of 6 L and receive a flow rate of 1.0 L/hr, the HRT, τ, has a value of 6.0 hr. The resulting plot is shown in Figure E8.1. Application of linear least squares analysis to the data reveals that the slope is 2.00 and the ordinate intercept is 0.0119. The slope is equal to $1/Y_H$; consequently, Y_H has a value of 0.50. Because the biomass concentration was measured as COD, the units of Y_H

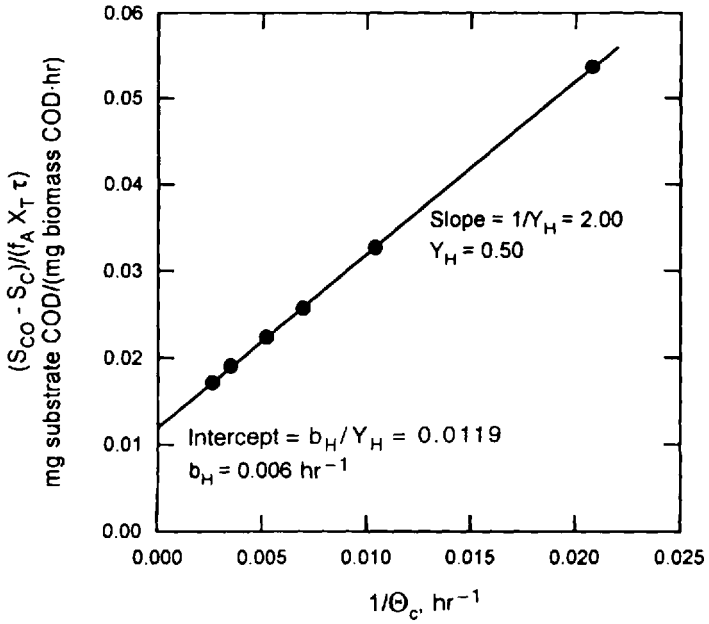


Figure E8.1 Determination of Y_H and b_H in Example 8.2.1.

are mg biomass COD formed/mg substrate COD removed. The intercept is equal to b_H/Y_H; consequently, b_H has a value of 0.006 hr^{-1}.

b. The second task is to estimate the value of f_D. This is done by plotting $1/f_A$ versus Θ_c as indicated by Eq. 8.9. The resulting plot is shown in Figure E8.2. Application of linear least squares analysis to the data reveals that the slope of the line is 0.00109 hr^{-1}. The slope is equal to $f_D \cdot b_H$. Since b_H has a value of 0.006 hr^{-1}, f_D has a value of 0.18. It is dimensionless.

c. The final task is to determine the values of $\hat{\mu}_H$ and K_S. Before this can be done, it is necessary to calculate the biodegradable COD concentration, S_S. This is done by subtracting the inert soluble COD, S_I from the soluble COD values given in Table E8.1, as indicated by Eq. 8.4. The Hanes linearization will be used because it does not require taking the reciprocal of S_S. Because the values of S_S are generally small, taking their reciprocal can greatly amplify the error in them, making it difficult to see the trends. The Hanes linearization requires plotting $S_S/(1/\Theta_c + b_H)$ versus S_S as indicated in Eq. 8.10. The plot is shown in Figure E8.3. It suggests that $\hat{\mu}_H$ has a value of 0.10 hr^{-1} and K_S a value of 25 mg COD/L. One thing that is evident from the plot is that small errors in the estimated values of S_S have a large effect on the estimates of the slope and the intercept. This is because the substrate concentration was low at all SRTs studied. While this suggests that it would be good to collect data at shorter SRTs, where the values of S_S would be larger, this would be difficult to do because bioflocculation would be poor (See Section 10.2.1). If the scatter in the data is so great that no confidence can be placed in the resulting parameter values, then it would be better to determine k_s, the mean reaction rate coefficient, which combines $\hat{\mu}_H$, K_S, and Y_H into a single first order coefficient. This is done by plotting S_S versus $(1/\Theta_c + b_H)$ as indicated by Eq. 5.41. Figure E8.4 shows the result of doing

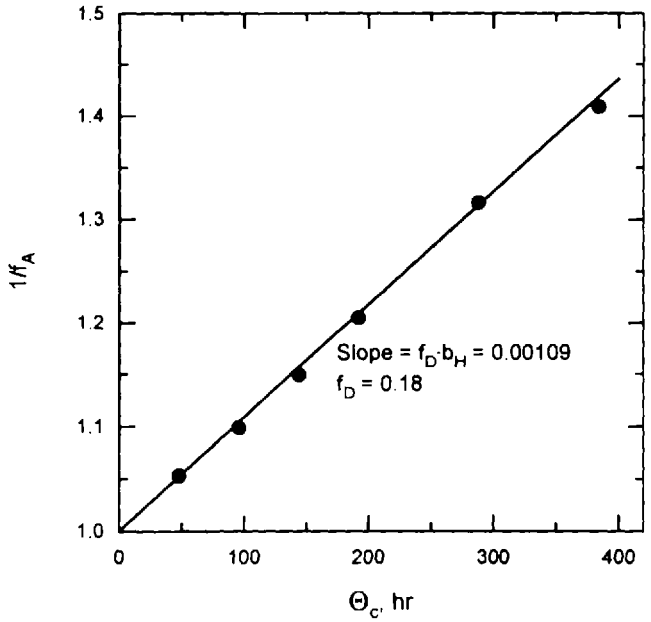


Figure E8.2 Determination of f_D in Example 8.2.1.

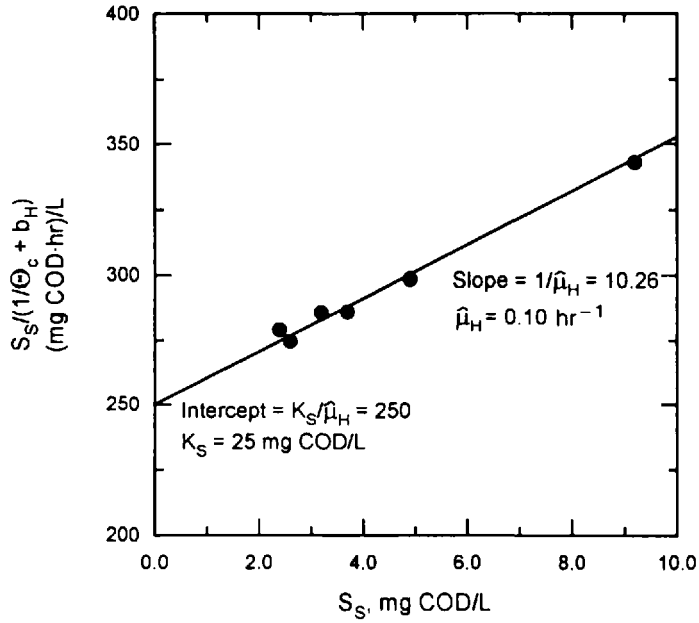


Figure E8.3 Determination of $\hat{\mu}_H$ and K_S in Example 8.2.1.

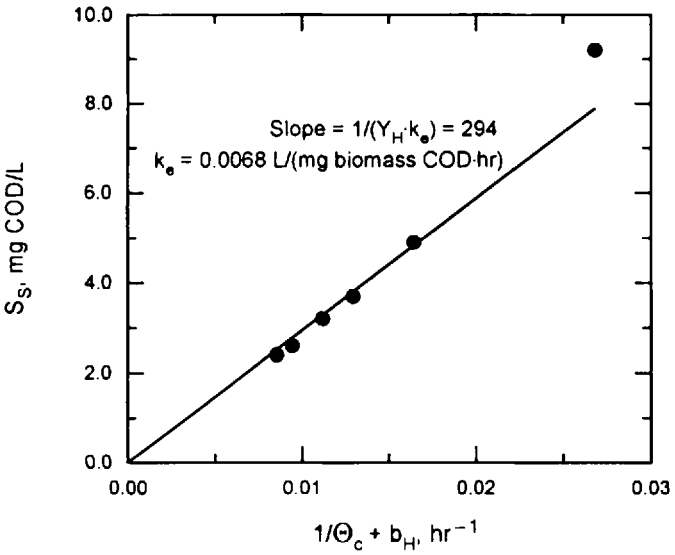


Figure E8.4 Determination of k_e in Example 8.2.1.

that in this case. Examination of the plot shows that only the lowest five points can be approximated as forming a straight line. For S_S values above 5 mg/L, S_S is not negligible with respect to K_S and thus the first order approximation of the Monod equation does not apply. However, for S_S values below 5 mg/L, the linear approximation is acceptable. As can be seen in the plot, k_1 has a value of 0.0068 L/(mg biomass COD · hr). This is slightly lower than the theoretical value 0.0080 L/(mg biomass COD · hr) which comes from its definition and the previously determined values of $\hat{\mu}_H$, K_S, and Y_H (See Section 5.1.6). The lower measured value is a direct result of the approximate nature of first order kinetics and the mean reaction rate coefficient.

8.3 SIMPLE SOLUBLE SUBSTRATE MODEL WITH TRADITIONAL DECAY IN THE ABSENCE OF DATA ON THE ACTIVE FRACTION

As discussed in Section 8.2.1, the active fraction of the biomass, f_A, is difficult to assess and thus data will generally not be collected on it during routine treatability studies for wastewater treatment. As a result, it will generally be impossible to determine b_H and Y_H by the techniques outlined in Section 8.2.2. Rather, an additional experiment must be conducted which will allow independent determination of b_H. Once b_H is known, the value of Y_H can be determined by using an assumed value for f_D of 0.20. The justification for assuming a value for f_D is that it is an inherent characteristic of biomass, and as such, should not change greatly from system to system.[10] The validity of this argument will be tested as more data are collected and analyzed using the techniques in the preceding section.

8.3.1 Data to Be Collected

With the exception of f_A, data will be collected from CSTRs exactly as described in Section 8.2.1. In addition, however, an experiment will be conducted for the direct determination of the decay coefficient for heterotrophic biomass, b_H. Biomass should be removed from one of the CSTRs and placed into a batch reactor where it will be aerated for several days. Generally, the longer the experiment is run, the more accurate the assessment of b_H will be. On a regular basis, biomass should be removed from the batch reactor and placed into a respirometer where the oxygen uptake rate (OUR) can be measured. Data on the decline in OUR over time will be used to assess b_H.[35] Historically, many investigators have used data on the change in TSS or VSS concentration over time for the determination of b_H, but the OUR technique has been shown to give more reproducible results provided proper precautions are taken.[35] The control of pH is one of those precautions. Since the bioreactor used for wastewater treatment will be maintained at neutral pH, it is important that the batch reactor used to provide the data for determining b_H also be maintained at neutral pH. The other major precaution concerns nitrification. If nitrification is well established in the CSTR from which the biomass is obtained, then continued nitrification will cause no problems in the experiment, provided that the pH is controlled and adequate oxygen is provided. This is because the ammonia released by biomass decay will be oxidized as it is released and no accumulation will occur. Consequently, the addi-

tional oxygen uptake associated with nitrification will be proportional to the oxygen uptake associated with heterotrophic decay and will not alter the change in the OUR over time.[25] It will only alter the magnitude of the OUR. If nitrification is not well established in the CSTR from which the biomass is obtained for the batch reactor, then the onset of nitrification in the batch reactor will distort the OUR data, making it unusable for the intended purpose. In that situation, nitrification should be inhibited during the OUR measurements by the addition of 20 mg/L of thiourea.[29]

8.3.2 Determination of b_H

The value of b_H can be obtained by two techniques from the data on the change in OUR over time. The rationale is as follows. For the traditional decay approach, the rate of oxygen utilization associated with decay was given by Eq. 3.58.

$$OUR = -r_{SO} = (1 - f_D)b_H \cdot X_{B,H} \quad (O_2 \text{ units}) \tag{3.58}$$

In a batch bioreactor that receives only biomass there will be no soluble substrate, so the only reaction will be decay. Performing a mass balance on active biomass for the batch reactor and substituting the appropriate terms from the matrix in Table 5.1 reveals:

$$\frac{dX_{B,H}}{dt} = -b_H \cdot X_{B,H} \tag{8.13}$$

which when integrated over time t tells us:

$$X_{B,H}|_t = X_{B,HO} \cdot e^{-b_H t} \tag{8.14}$$

Substitution of Eq. 8.14 into Eq. 3.58 reveals that the oxygen uptake rate in the batch bioreactor at any time t is given by:

$$OUR|_t = (1 - f_D)b_H \cdot X_{B,HO} \cdot e^{-b_H t} \tag{8.15}$$

Equation 8.15 can be used directly to estimate b_H through the application of nonlinear least squares techniques. Examination of the equation shows that there are three unknowns, b_H, $X_{B,HO}$, and f_D, but that f_D and $X_{B,HO}$ cannot be determined independently of each other. As discussed above, however, a value of 0.20 can generally be assumed for f_D, allowing $X_{B,HO}$ and b_H to be determined. It should be noted that $X_{B,HO}$ is specific to the biomass used in the batch test and has no utility as a parameter.

In the absence of nonlinear techniques, Eq. 8.15 may be transformed, allowing linear least squares techniques to be used. Taking the natural log of both sides reveals:

$$\ln(OUR)|_t = \ln[(1 - f_D)b_H \cdot X_{B,HO}] - b_H \cdot t \tag{8.16}$$

Consequently, an alternative technique is to plot ln(OUR) versus time and use linear least squares to obtain the slope, which is $-b_H$. This is illustrated in Figure 8.6. The ordinate intercept can be used to estimate $X_{B,HO}$ from an assumed value for f_D, but this is seldom done because $X_{B,HO}$ has limited utility, as noted above.

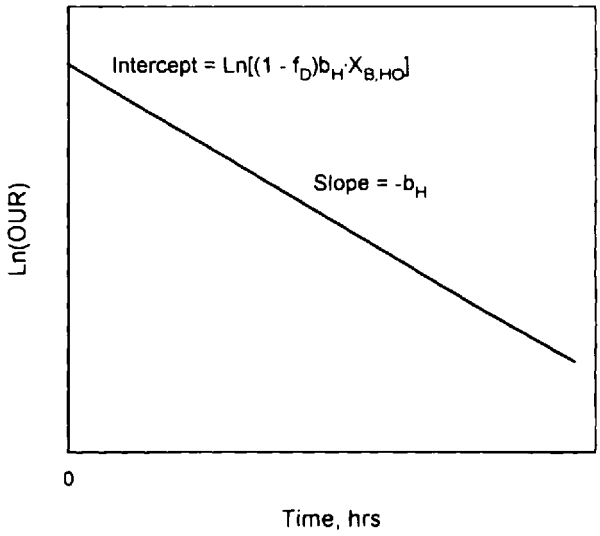


Figure 8.6 Plot of Eq. 8.16 to determine b_H.

8.3.3 Determination of Y_H

Once the value b_H has been estimated and a value has been assumed for f_D, it is possible to estimate Y_H from a rearranged form of Eq. 5.24:

$$\frac{X_I}{S_{SO} - S_S} = Y_H \left[\left(\frac{\Theta_c}{\tau} \right) \frac{(1 + f_D \cdot b_H \cdot \Theta_c)}{(1 + b_H \cdot \Theta_c)} \right] \tag{8.17}$$

Recall from Eq. 8.5 that $S_{SO} - S_S$ is equal to $S_{CO} - S_C$. Because data have been collected from the CSTRs on the effect of SRT on X_I and S_C, a plot may be made of $X_I/(S_{CO} - S_C)$ versus the bracketed term on the right side. It will have a slope equal to Y_H.

If the treatability study is conducted in CSTRs without biomass separators, then Θ_c and τ will be the same, requiring a rearranged form of Eq. 5.25 to be used:

$$\frac{X_I}{S_{SO} - S_S} = Y_H \left[\frac{(1 + f_D \cdot b_H \cdot \tau)}{(1 + b_H \cdot \tau)} \right] \tag{8.18}$$

Again, $S_{CO} - S_C$ may be used in place of $S_{SO} - S_S$.

8.3.4 Determination of S_I, $\hat{\mu}$, K_S, and k_e

Once f_D, b_H and Y_H are known, the estimation of the other parameters may proceed as described in Sections 8.2.4, 8.2.5, and 8.2.6.

Example 8.3.1

A treatability study was performed on an industrial wastewater using CSTRs with biomass recycle. The wastewater contained a complex mixture of soluble organic

Table E8.2 Data Collected During a Treatability Study with a Soluble Industrial Wastewater

SRT hrs	Soluble COD mg COD/L	Total biomass mg COD/L
48	29.6	1,420
96	27.6	2,425
144	27.0	3,214
192	26.8	3,879
288	26.5	5,004
384	26.4	5,981

compounds with a total soluble COD of 400 mg/L. The wastewater contained no particulate matter. The studies were run in lab scale bioreactors that had a volume of 6.0 L. The flow rate was maintained at a constant rate of 1.0 L/hr and the SRT was maintained at the desired values by wasting excess biomass directly from the reactors. The wastage rate was corrected for the loss of biomass in the effluent. Data collected from the bioreactors are shown in Table E8.2. In addition, biomass was removed from the bioreactor with an SRT of 96 hr, doubled in concentration by settling, and placed in a batch reactor to allow measurement of the OUR over time for estimation of b_H. The data from this test are shown in Table E8.3. Using the information provided, estimate the values of the parameters b_H and Y_H. Assume that f_D has a value of 0.20.

a. The first task is to determine b_H from the batch data given in Table E8.3. This is done by plotting the natural log of the OUR versus time. The resulting plot is shown in Figure E8.5. The value of b_H is equivalent to the slope of the line, which is 0.007 hr^{-1}.

b. Once b_H has been evaluated it is possible to estimate Y_H from Eq. 8.17, provided a value is available for f_D. No data are available on the active fraction of the biomass, so it is necessary to assume a value for f_D. The risk

Table E8.3 Data Collected During a Batch Aeration Test for the Determination of b_H

Time hrs	OUR mg O_2/(L·hr)
0	24.0
12	22.0
24	20.2
36	18.6
48	17.1
72	14.5
96	12.2
120	10.3
144	8.7
192	6.2
240	4.5

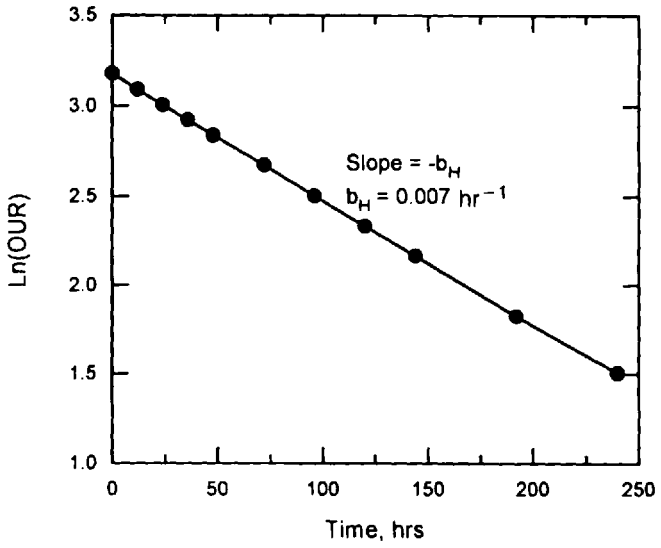


Figure E8.5 Determination of b_H in Example 8.3.1.

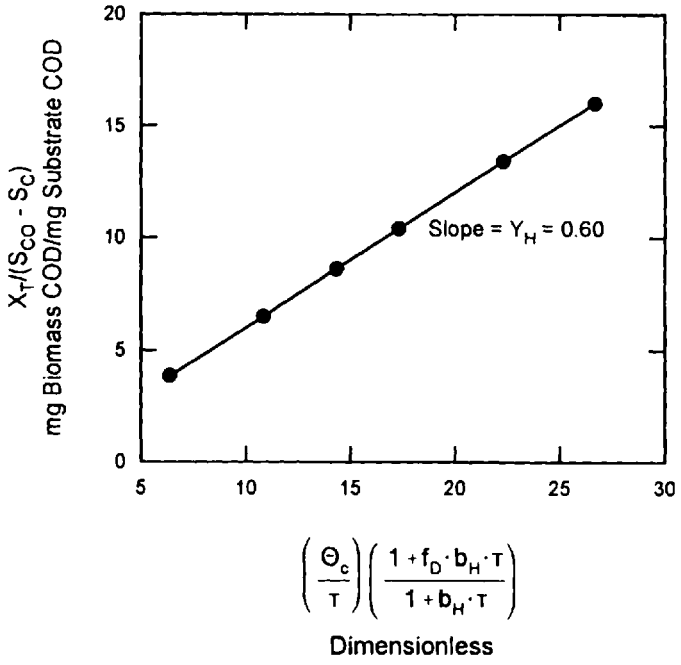


Figure E8.6 Determination of Y_H in Example 8.3.1.

in doing this is small because f_D is not thought to vary greatly from system to system. Figure E8.6 shows the plot of Eq. 8.17 when f_D is assumed to be 0.20. The slope is equal to the value of Y_H, which is 0.60 mg biomass COD/mg substrate COD.

8.4 USE OF BATCH REACTORS TO DETERMINE MONOD KINETIC PARAMETERS FOR SINGLE SUBSTRATES

The values of the Monod kinetic parameters $\hat{\mu}_H$ and K_S determined by the techniques discussed in Section 8.2.5 are for general organic matter as measured by COD. Frequently, however, it is desirable to know the kinetic parameters associated with the biodegradation of a single organic compound in a wastewater. Because of the relatively low specific growth rates at which suspended growth bioreactors operate, it is permissible to use parameter values determined during metabolism of the compound as the sole carbon and energy source without regard for metabolism of the other organic compounds present in the wastewater,[11] provided that none of the other compounds have strongly inhibitory effects. Such parameter values may be determined during batch experiments, but care must be exercised in the way in which those experiments are performed. This follows directly from the effect that growth conditions have on the physiological state of the biomass.

8.4.1 Intrinsic Versus Extant Kinetics

The physiological state of a microbial culture is determined by the manner in which it is grown. This follows from the fact that the physiological state is a measure of the macromolecular composition of the cells in the culture. If the culture is grown at a very high rate, the cells' protein synthesizing system will be fully developed for rapid growth and the cells will contain high levels of all enzymes. If the culture is grown at a slow rate, the cells will adjust the level of their protein synthesizing system to conserve resources. This means that they will contain lower enzyme levels. Consideration must be given to the physiological state of a culture and how it may change when batch experiments are designed for measuring kinetic parameter values. This is necessary because the parameter values measured in the experiment will depend upon the physiological state of the biomass.

To assist in considering the effects of physiological state, the following nomenclature system has been proposed.[17] Two extreme conditions represent the limits that the physiological state of a culture may attain during a kinetic experiment. At one extreme, if the physiological state is not allowed to change, the resulting kinetic parameter values are reflective of the conditions of the biomass in the bioreactor from which the microorganisms used in the kinetic test were obtained. Because those parameters reflect the conditions currently existing in the parent bioreactor, they have been called "extant." At the other extreme, if the physiological state of the culture is allowed to change during the test to the point that the cells' protein synthesizing system is fully developed and the bacteria have an enzyme system that allows them to grow at the fastest rate possible on the test substrate at the given temperature and pH, the resulting kinetic parameters are said to be "intrinsic." The name follows

from the fact that the parameter values are dependent only on the nature of the substrate and the types of bacteria in the culture. They are independent of the history of the culture, i.e., they are intrinsic. Parameter values obtained from experiments that allow the physiological state to be between these two extremes may be either "defined" or "undefined," depending upon the care with which the experiment is run and reported.

It is unclear which type of kinetic parameter set, intrinsic or extant, is of most utility to environmental engineers and scientists. It appears, however, that intrinsic parameters are most useful for comparing the biodegradability of organic compounds whereas extant parameters are most useful for predicting the performance of an operating bioreactor with respect to the removal of a given organic compound.[16] Consequently, we will briefly examine how to determine each type.

8.4.2 Intrinsic Kinetics

The key to determining intrinsic kinetic parameter values during a batch test is to provide sufficient substrate to allow the bacteria performing the biodegradation to fully develop their protein synthesizing and enzyme systems. This can usually be accomplished when the initial substrate to biomass ratio ($S_{SO}/X_{H,IO}$) is at least 20 when both concentrations are expressed as COD.[3] In addition, the initial substrate concentration should be greater than the expected value of K_S. Since intrinsic K_S values for individual substrates tend to be less than 10 mg/L as COD,[15] experiments in which S_{SO} is 20 mg/L as COD have proven to be successful.[3] Consequently, the initial biomass concentration should be on the order of 1 mg/L as COD. It should be noted that the initial biomass concentration applies only to that portion of the biomass that is active in biodegradation of the test compound. This will generally be only a fraction of the total biomass in a suspended growth system like activated sludge.

The substrate and biomass are placed into a batch reactor and the course of biodegradation is followed over time. Three types of data may be collected and used: (1) substrate disappearance, (2) biomass growth, or (3) oxygen consumption. This follows directly from the proportionality of the three rates during balanced growth as depicted in Eq. 3.34. The equivalency of the three data sets has also been shown experimentally.[6] Mass balance equations for substrate and biomass in a batch reactor must be written using the simplified model of Chapter 5 and solved simultaneously by appropriate numerical methods using assumed values for the parameters. If oxygen consumption data are to be used for parameter estimation, the resulting theoretical substrate and biomass curves can be converted into the equivalent oxygen consumption curve using a COD balance as depicted by Eq. 3.32. The theoretical data curve is then compared to the actual data curve and the parameter values are adjusted until the best agreement is achieved between the theoretical and actual curves. The parameter values associated with the best-fit curve are considered to be the best estimates of them. Because of the nature of the Monod (and Andrews) equation, a robust fitting routine is required to find the true best estimate of the parameters.[33] Details of a test procedure employing the collection of oxygen consumption data may be found in Brown et al.[3]

8.4.3 Extant Kinetics

The requirements for determining extant values of the Monod (or Andrews) kinetic parameters are just the opposite of the requirements for the determination of intrinsic values. This follows from the fact that extant parameter values reflect the conditions of the biomass in the bioreactor from which they were obtained. Consequently, the test conditions need to be such that few changes occur in the physiological state of the biomass or the composition of the microbial community during the test. This can be achieved by keeping the value of $S_{SO}/X_{B,HO}$ small during batch tests, with 0.02 being a typical value. Again, it should be emphasized that $X_{B,HO}$ should reflect only that portion of the biomass that is capable of degrading the test compound. Because the amount of substrate added is very small relative to the amount of biomass present, biomass growth may be neglected in any given test, allowing use of only the mass balance equation for substrate. This simplifies the computations somewhat but means that data can only be collected on substrate loss or oxygen consumption.

Two respirometrically based procedures are available for determining extant kinetic parameter values. One, developed by Lamb et al.[22] and refined by Chudoba and colleagues,[4,5] relies on substrate injections into a respirometer at a number of concentrations. The net oxygen uptake rate (OUR) in response to an injection is proportional to the substrate utilization rate (see Eq. 3.34), which in turn is proportional to the growth rate, with the yield being the proportionality factor (Eq. 3.44). Dividing the OUR by the concentration of biomass actively involved in the biodegradation of the substrate gives the specific oxygen uptake rate (SOUR), which can be used to calculate the specific growth rate associated with the concentration of substrate supplied by the injection:

$$\mu_H = \left(\frac{Y_H}{1 - Y_H}\right) (\text{SOUR}) \tag{8.19}$$

Making injections at several substrate concentrations provides a data set showing the effect of substrate concentration on specific growth rate. Consequently, the Monod (or Andrews) equation may be fit to the data set by using the techniques in Section 8.2.5. As discussed in that section, the best procedure for estimating $\hat{\mu}_H$ and K_S is by applying nonlinear parameter estimation to the Monod or Andrews equation. If nonlinear parameter estimation is not possible, the best linearization technique in this situation is that of Hanes:

$$\frac{S_S}{\mu_H} = \frac{K_S}{\hat{\mu}_H} + \frac{S_S}{\hat{\mu}_H} \tag{8.20}$$

Thus, a plot should be made of S_S/μ_H versus S_S, with $\hat{\mu}_H$ being determined by the slope and K_S by the ordinate intercept. Because of the way the experiment is run, the substrate concentration is assumed to be free of error, i.e., the independent variable, and μ_H is assumed to contain error (because SOUR is the measured variable). Thus, linear least squares analysis can be applied for fitting purposes, even though it could not be earlier.

The other extant respirometric technique relies on a single substrate injection with determination of the kinetic parameters by fitting the theoretical oxygen consumption curve to the observed oxygen consumption curve, in a manner similar to that used for determination of the intrinsic parameter values.[10] However, as pointed

out in the preceding paragraph, the fitting routine is simpler because only the substrate mass balance equation must be solved. As in the intrinsic test, S_{SO} should also be equal to or greater than the expected K_S value. Details regarding both techniques are too involved to provide herein, but the reader is encouraged to consult the cited references.

8.5 COMPLEX SUBSTRATE MODEL WITH LYSIS: REGROWTH APPROACH TO DECAY AS PRESENTED IN CHAPTER 6 (INTERNATIONAL ASSOCIATION ON WATER QUALITY ACTIVATED SLUDGE MODEL NO. 1)

As we saw in Section 6.1, ASM No. 1 is very complex, with many parameters. Luckily, not all of them need to be evaluated for every wastewater. Rather, some parameter values appear to be about the same for all systems, and fixed values can be assumed.[19] These are listed in Table 8.1 and the values given in Table 6.3 are satisfactory for most purposes. In this section, we will briefly review the parameters that require evaluation to provide an overview of how to obtain them. We will also review procedures for characterizing the wastewater. Details can be found elsewhere,[19,29] as can a number of examples.[13,14,21,23,36]

Before proceeding, it is important to recognize that our knowledge of activated sludge modeling is still developing. As a result, one should not assume that it is necessary to assess all of the remaining parameters in all cases, or indeed that assessment of those parameters will ensure that extrapolations made with the model will be successful. Rather, one should recognize that many of the parameters are interrelated and that the value obtained for one will affect the value obtained for another. Furthermore, because of that it is likely that the minimum required parameter set may well turn out to be less than we now think. Consequently, the reader should consult the literature in this rapidly developing field.

8.5.1 Data to Be Collected

Because ASM No. 1 contains more elements than the simple soluble substrate model, evaluation of the parameters involved is more complicated. This requires more bioreactors to be operated.

Table 8.1 Parameters Which May Be Assumed

Symbol	Description
Y_A	Yield for autotrophic biomass
b_{LA}	Decay coefficient for autotrophic biomass
f_D'	Fraction of biomass leading to debris
$i_{N\,XB}$	Mass of nitrogen per mass of COD in biomass
$i_{N\,XD}$	Mass of nitrogen per mass of COD in biomass debris
$K_{O,H}$	Oxygen half-saturation coefficient for heterotrophic biomass
K_{NO}	Nitrate half-saturation coefficient for denitrifying heterotrophic biomass
$K_{O,A}$	Oxygen half-saturation coefficient for autotrophic biomass

Just as with the simple model, several CSTRs should be operated with steady-state feed over a range of SRTs. The SRTs employed should be in excess of 5 days, providing a mixed liquor that is low in undegraded particulate substrate and an effluent that is low in readily biodegradable substrate. These bioreactors will be important to the evaluation of the parameters related to biomass production and electron acceptor requirement. Consequently, the data collected from them should be similar to that listed in Section 8.2.1, with the exception of the active fraction, f_A, which is seldom measured. In addition, the total COD in the influent to the bioreactors should be measured.

A CSTR with a biomass separator should be operated at an SRT of 2 days while receiving a daily cyclic square wave feeding pattern with 12 hr of feed followed by 12 hr without feed. In addition to the typical data collected on the steady-state CSTRs, periodic measurements should be made of the oxygen utilization rate (OUR), as discussed later.

Finally, since information is needed about denitrification rates, a reactor system containing two CSTRs in series should be operated as an MLE system. The system SRT, the mixed liquor recirculation flow rate and the relative volumes of the aerobic and anoxic bioreactors should be selected to give stable nitrification and denitrification. Biomass from this system will be used to provide data on the nitrate utilization rate as outlined later.

8.5.2 Characterization of Wastewater and Estimation of Stoichiometric Coefficients

ASM No. 1 requires that the organic matter in a wastewater be partitioned into several components, as indicated in Tables 6.1 and 6.2: X_I, inert particulate organic matter; S_I, inert soluble organic matter; X_S, slowly biodegradable substrate; and S_S, readily biodegradable substrate. It is important to recognize that the distinction between the two types of biodegradable substrate is operationally defined and does not necessarily correspond to readily distinguishable physical characteristics, such as soluble and particulate, in spite of the symbols used in the model. Thus, characterization of the wastewater must be accomplished experimentally. One of the procedures for wastewater characterization, the determination of the concentration of inert soluble organic matter (SOM), is the same as the procedure presented in Section 8.2.4 and will not be repeated here. Rather, we will focus in this section on the quantification of the other three fractions. Before that can be done, however, Y_H must be known.

Determination of Y_H. The methods employed in Sections 8.2.2 and 8.3.3 for determining Y_H cannot be used here because of the presence of particulate organic matter in the MLSS. Thus, the approach that has been recommended is to observe Y_H directly as biomass is grown on only the soluble component of the watewater.[15] An aliquot of wastewater should be filtered to remove the particulate COD, placed into a batch bioreactor, and seeded with a small amount of biomass from one of the CSTRs. Care should be exercised to obtain a sample with as high a soluble COD concentration as possible and to make the concentration of seed biomass very small relative to the initial soluble COD (<1%). This is because Y_H is defined as the amount of biomass grown in the absence of decay and the only way this can be directly observed is to allow a small amount of biomass to grow rapidly on a large amount

of substrate. Samples should then be taken over time so that the total and soluble COD can be measured as biomass growth proceeds. The biomass COD can be calculated as the difference between the total COD and soluble COD, and the yield can be determined from its definition:

$$Y_H = \frac{\Delta \text{ Biomass COD}}{\Delta \text{ Soluble COD}} \tag{8.21}$$

Generally, the best way to determine Y_H is to plot the biomass COD as a function of the soluble COD removed and take the slope of the resulting line. Alternatively, the biomass can be measured directly by suspended solids measurements, rather than as COD, in which case the yield is expressed in terms of TSS or VSS, depending upon the measurement used. Every effort should be made to get an accurate estimate of Y_H because errors in its determination will influence the estimates of the various fractions of the wastewater organic matter.

Determination of Readily Biodegradable Chemical Oxygen Demand (S_{SO}). The procedure originally recommended for determination of the readily biodegradable COD concentration involves a bioassay.[9,10] Because of the difficulties associated with that assay, a number of alternatives have been proposed.[29,30] Space does not permit presentation of all of the procedures and experience is not yet extensive enough to make judgements among them. Consequently, only two procedures will be presented here, the original one and a rapid physical assay that is very simple.

In the original procedure, data are collected from the CSTR operated at an SRT of 2 days while receiving a daily cyclic square wave feeding pattern with 12 hr of feed followed by 12 hr without feed.[9,10] The short SRT is used to eliminate nitrification. After the bioreactor has been operated for a sufficient period to establish a stable response pattern, several measurements of the OUR should be made during a complete cycle. When the feed is stopped, the OUR will drop rapidly because the readily biodegradable substrate is quickly exhausted, as shown in Figure 8.7. The OUR will then decline slowly until the entrapped slowly biodegradable substrate is exhausted, at which point the OUR will reflect endogenous metabolism and decay. Use will be made of the latter characteristics later, but for determination of the readily biodegradable substrate concentration only the immediate drop in OUR (ΔOUR) is required:

$$S_{SO} = \frac{(\Delta OUR)V}{F(1 - Y_H)} \tag{8.22}$$

where F is the flow rate of the feed prior to its termination and V is the bioreactor volume. It is important to recognize that Y_H must be in COD units for use in Eq. 8.22. If it is expressed in TSS or VSS units, then it must be multiplied by $i_{O\,XBH}$ or $i_{O\,XBV}$, as discussed in Section 8.2.1.

The physical assay for S_{SO} developed by Mamais et al.[24] is much quicker than the original bioassay and gives results that correlate well with the original bioassay for domestic wastewater. In it, samples of raw wastewater are flocculated by adding $ZnSO_4$, mixing vigorously for 1 min, and adjusting the pH to 10.5 with NaOH. They are then allowed to settle quiescently before a sample of clear supernatant is withdrawn and filtered through a 0.45 μm membrane filter. The flocculation step removes colloidal organic matter that otherwise would pass through the filter and be measured

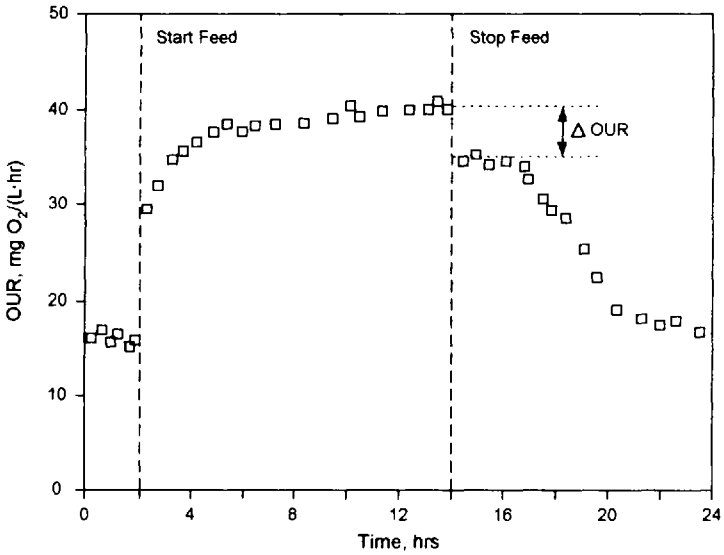


Figure 8.7 Response of a completely mixed activated sludge reactor to a 12 h square wave response as used to determine the concentration of readily biodegradable substrate. (From G. A. Ekama, et al., Procedures for determining influent COD fractions and the maximum specific growth rate of heterotrophs in activated sludge systems. *Water Science and Technology* **18**(6): 91–114, 1986. Copyright © Elsevier Science Ltd.; reprinted with permission.)

as "soluble" material. The COD of the filtrate is the total soluble COD of the wastewater. Subtraction of the inert soluble COD, S_I, provides the value of the readily biodegradable COD, S_{SO}. A major advantage of this technique is that more samples can be analyzed, giving a better measure of the long-term average concentration of the readily biodegradable COD in the wastewater than can be obtained with the bioassay.

Determination of Inert Particulate Chemical Oxygen Demand (X_{IO}). The total COD in a wastewater is given by:

$$\text{Total influent COD} = \text{COD}_{TO} = X_{SO} + X_{IO} + S_{SO} + S_{IO} \tag{8.23}$$

The total influent COD can be measured and S_{SO} and S_{IO} can be estimated by the procedures given previously. If either X_{SO} or X_{IO} is determined, the other can be calculated from Eq. 8.23. It is best to estimate X_{IO} as a parameter for fitting the model to data showing the effect of SRT on the concentration of mixed liquor suspended solids (MLSS) in the CSTRs. X_{SO} can then be calculated from the total influent COD using Eq. 8.23. Ignoring autotrophic biomass, which is usually negligible, the MLSS in the CSTRs comes from four major sources: (1) growth of heterotrophs, (2) production of microbial debris, (3) accumulation of inert suspended organic matter from the feed, and (4) accumulation of undegraded slowly biodegradable substrate. If the SRT is greater than 5 days, the values of X_S and S_S in a bioreactor will be negligibly small relative to X_{SO} and S_{SO}. This means that the MLSS concentration in a CSTR can be approximated by the simple soluble substrate model which includes inert particulate COD (X_{IO}), provided that the measurement of the

influent biodegradable COD includes the biodegradable particulate substrate. In other words, even though we are estimating parameters for ASM No. 1, Eq. 5.46 can be rewritten and used to estimate X_{IO}:

$$X_M = \left(\frac{\Theta_c}{\tau}\right)\left[X_{IO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H(COD_{TO} - S_{IO} - X_{IO})}{1 + b_H \cdot \Theta_c}\right] \tag{8.24}$$

Everything in Eq. 8.24 except X_{IO} is either known or can be assumed. The value of b_H can be evaluated independently by the technique in Section 8.3.2. It is important to recognize that the traditional decay coefficient, b_H, is used at this point. That is because Eq. 8.24 is from the simple model. The value of Y_H has already been evaluated. The value of f_D can be assumed to be 0.20, just as before with the simple model. Again, the value of f_D for the simple model is used in this computation because Eq. 8.24 is from that model. COD_{TO} and S_{IO} are measured values. The SRT (Θ_c) and the HRT (τ) are experimentally controlled variables. Generally, it is best to maintain a fixed HRT during the operation of the CSTRs with different SRTs, thereby making the MLSS concentration, X_M, a function of only the SRT. The value of X_{IO} can then be estimated by using a one-dimensional search routine which chooses X_{IO} to minimize the error sum of squares when the predicted MLSS concentration (X_M) is compared to the values measured in the CSTRs operated over a range of SRTs. If it is necessary to change the HRT as well as the SRT during the studies, then τ should be moved to the left side of Eq. 8.24 and predicted values of $X_M \cdot \tau$ should be compared to the measured values during the estimation of X_{IO}.

Characterization of Nitrogen-Containing Material. Activated sludge model No. 1 includes terms for nitrogen as well as COD. Thus, it is also necessary to characterize them during treatability studies. All soluble forms can be determined by appropriate chemical analyses on the wastewater. The concentration of inert soluble organic nitrogen in the wastewater can be determined by performing Kjeldahl nitrogen tests on aliquots of the samples used to determine the inert soluble COD (See Section 8.2.4). Subtraction of the inert soluble organic nitrogen from the soluble organic nitrogen in the influent gives the concentration of readily biodegradable organic nitrogen in the feed, S_{NSO}. Readily and slowly biodegradable organic nitrogen in the wastewater are assumed to be proportioned in the same way as readily and slowly biodegradable COD:

$$\frac{S_{NSO}}{X_{NSO} + S_{NSO}} = \frac{S_{SO}}{X_{SO} + S_{SO}} \tag{8.25}$$

Because everything in Eq. 8.25 except X_{NSO} is known, its value can be calculated. The sum of the biodegradable organic nitrogen species and the inert soluble organic nitrogen in the wastewater should be less than the total organic nitrogen, which can be measured. The difference between the two is the inert particulate organic nitrogen. Although the latter is not used in the model, its value should be calculated as a check on the procedures. A negative value is evidence of an error.

8.5.3 Estimation of Kinetic Parameters

Aerobic Growth of Heterotrophs. Because of all of the terms in ASM No. 1 and the generation of soluble substrate from slowly biodegradable substrate, it is not possible to use the approach described in Section 8.2 to obtain $\hat{\mu}_H$ and K_S. Conse-

quently, an alternative approach must be used. The batch techniques described in Section 8.4 may be applied using the soluble fraction of the wastewater, although as discussed in that section, it is currently unclear whether intrinsic or extant parameter estimates are more suitable when using the model. Intrinsic kinetic parameter estimates can be obtained by using coagulated and filtered wastewater in batch experiments like those described in Section 8.4.2. In this case, however, the substrate will be the readily biodegradable organic matter rather than a single organic compound. As a result, the K_S value obtained will be larger than that associated with single compounds. If the value of S_{SO} used in the batch test is larger than the resulting K_S value, then the parameters can be considered to be intrinsic. If this condition is not met, the resulting parameter values will be undefined and should not be used. Because the main function of $\hat{\mu}_H$ and K_S in ASM No. 1 is to allow the maximum oxygen uptake rate of an operating bioreactor to be calculated, it may be better to use one of the extant parameter techniques described in Section 8.4.3 to obtain $\hat{\mu}_H$ and K_S, provided that the biomass is taken from a bioreactor with a short SRT. The biomass concentration used in the estimation of the specific rates should correspond to that of the active heterotrophic biomass, $X_{B,H}$. Sufficient information in the form of stoichiometric and kinetic parameters is available at this point to allow its estimation for the bioreactor from which the biomass in the test is obtained, thereby allowing its estimation for the batch test reactor. Substrate injections should be made with coagulated and filtered wastewater, thereby providing only readily biodegradable substrate during the test.

Aerobic Growth of Autotrophs. As far as the design and control of nitrifying bioreactor systems are concerned, the maximum specific growth rate coefficient for autotrophic biomass, $\hat{\mu}_A$, is the most critical parameter value. There are two reasons for this. First, it determines the SRT at which nitrifying biomass will be lost from the system, thereby fixing the minimum acceptable SRT at which the system can be operated. Second, it can be affected strongly by chemicals in the wastewater; more strongly, in fact, than the half-saturation coefficient. Thus, it is important that an accurate assessment of $\hat{\mu}_A$ be obtained in an environment that represents the wastewater undergoing treatment. Several procedures have been proposed for measuring $\hat{\mu}_A$,[8,18,29] but the simplest involves a batch experiment started with a small amount of biomass. Effluent should be collected from one of the continuous CSTRs, preferably one with a short SRT so that little nitrification will have occurred, making the initial concentrations of nitrate-N and nitrite-N small. If necessary, ammonia-N should be added to the bioreactor to bring the concentration to approximately 40 mg/L. The bioreactor should then be seeded with biomass from a bioreactor with an active nitrifying population, but the initial nitrifying biomass concentration should be less than 1.0 mg/L. Because the production of nitrate-N and nitrite-N is proportional to the amount of nitrifying bacteria formed, the change in the concentration of oxidized nitrogen (NO_3-N + NO_2-N) can be used to estimate $\hat{\mu}_A$.[1,18] The concentration of oxidized nitrogen in the bioreactor should be measured over time as it increases through growth of the autotrophs and the natural log of (NO_3-N + NO_2-N) should be plotted versus time. The plot should give a straight line, with a slope of $\hat{\mu}_A - b_{t,A}$. The value of $b_{t,A}$ is an assumed value, as indicated in Table 8.1, and thus the value of $\hat{\mu}_A$ can be calculated. Examples of this procedure may be found in the literature.[1,14]

Although the half-saturation coefficient for nitrifying bacteria, K_{NH}, is not affected as strongly by organic contaminants as the maximum specific growth rate, it may be influenced and should be determined. This may be done with the extant kinetic parameter technique of Lamb et al.[22] and Chudoba and colleagues[4,5] discussed in Section 8.4.3. Biomass should be removed from a CSTR that is fully nitrifying and placed into a respirometer without dilution. Small quantities of ammonia-N should be injected and the net respiration rate measured in response to the injections. This provides data on OUR as a function of injected ammonia-N concentration which can be analyzed as discussed above for heterotrophic biomass. However, in this case the concentration of nitrifying biomass is unknown so that SOUR values cannot be calculated. As a result, the analysis will yield only a value of the maximum OUR in addition to the value of K_{NH}. However, the value of K_{NH} is valid for use with the value of $\hat{\mu}_A$ determined from the batch test described in the preceding paragraph.

Decay of Heterotrophs. The decay coefficient, $b_{L,H}$, is very important to predictions of biomass production and oxygen requirements, so it must be determined for the particular wastewater under study. Therefore, biomass should be removed from one of the CSTRs and used in the batch procedure of Section 8.3.2 to determine the traditional decay coefficient, b_H, which can be used to determine X_{IO} as discussed in Section 8.5.2. Biomass can be removed from any of the CSTRs, but correction for the effects of nitrification will be easier if the biomass is fully nitrifying as discussed in Section 8.3.1. Once the value of b_H has been obtained it can be used to calculate $b_{L,H}$ with Eq. 3.69:

$$b_{L,H} = \frac{b_H}{[1 - Y_H(1 - f'_D)]} \tag{3.69}$$

The value of f'_D should be assumed to be 0.08.

Correction Factors for Anoxic Conditions, η_g and η_h. Two important parameters in ASM No. 1 are η_g and η_h because they correct the rates of growth and hydrolysis reactions when they occur under anoxic conditions. As discussed in Section 6.1.2, this is required because only a portion of the biomass will be capable of functioning under anoxic conditions and the model needs some way to reflect that fact. Tests to measure η_g and η_h are performed at the same time by evaluating oxygen and nitrate consumption rates in two batch bioreactors which are equivalent in every respect except for the terminal electron acceptor.[19] The biomass for the tests should come from the MLE bioreactor run as part of the parameter estimation study because it will contain biomass capable of functioning under both aerobic and anoxic conditions. The rationale for the test is as follows. Immediately after biomass is brought into contact with wastewater in a batch bioreactor, the activity in the bioreactor will be dominated by growth of the heterotrophs on the readily biodegradable substrate. However, as soon as the readily biodegradable substrate is exhausted, the activity will be predominantly due to the use of substrate arising from hydrolysis of the slowly biodegradable substrate. Therefore, by comparing the activity of a biomass sample in both of these regions under both aerobic and anoxic conditions it is possible to estimate η_g and η_h.

Conceptually, the experiment is very simple. Biomass is removed from the MLE system and placed into two batch bioreactors, one of which is maintained under aerobic conditions with oxygen as the terminal electron acceptor and the other of

which is kept under anoxic conditions with nitrate as the terminal electron acceptor. The latter bioreactor should be constructed to minimize oxygen transfer to the liquid. Wastewater is then added to both bioreactors and the OUR and nitrate utilization rate (NUR) are measured in the appropriate bioreactors as the substrate is depleted. The OUR is normally measured by placing a dissolved oxygen (DO) probe in the aerobic bioreactor. By providing a mechanical mixer to keep the biomass in suspension, it is possible to turn off the air supply periodically and measure the OUR directly by the rate of decrease in the DO concentration over a short time period. The NUR is normally measured by manually removing samples from the bioreactor over time, stopping the reaction, removing the biomass, and measuring the nitrate-N concentration. The NUR in the two reaction regions is then determined from the slopes of the plot of nitrate-N over time as illustrated in Figure 8.8. Care should be exercised in the measurement of the NUR to ensure that nitrite is not accumulating in the bioreactor. If it is, then the results should be expressed in terms of the net amount of electron acceptor used expressed as the equivalent amount of oxygen used. Care should also be taken to ensure that an appropriate substrate to biomass (F/M) ratio is used in the tests. Figure 8.9 illustrates why this is important in terms of the OUR. If the F/M ratio is too low, the time required for readily biodegradable substrate removal will be too short to get a good measure of the rate. Conversely, if the ratio is too large, the difference in rate between the two zones will not be sufficiently large to clearly distinguish between them. The value of η_g can be calculated once data are available for OUR_g and NUR_g:

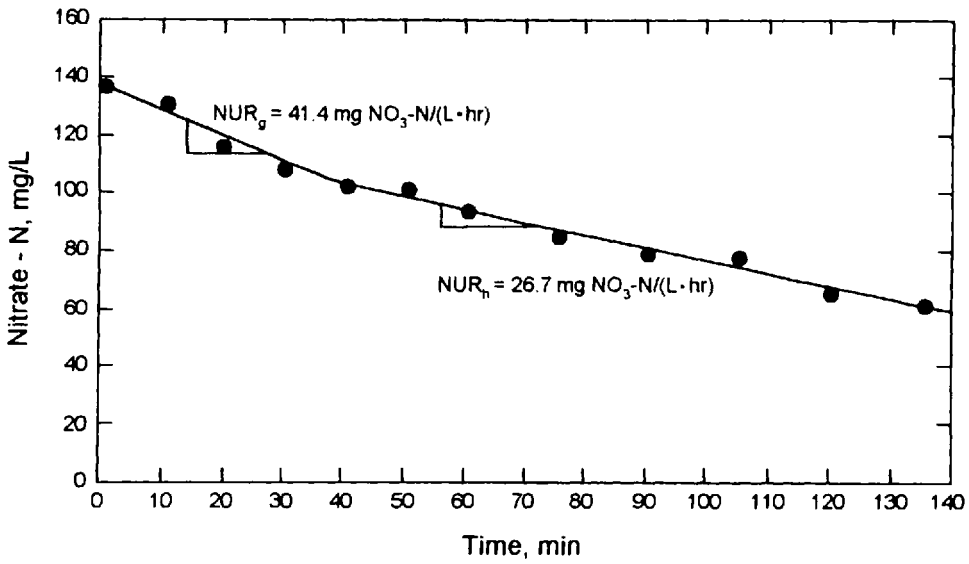


Figure 8.8 Nitrate utilization in a batch reactor. The faster rate corresponds to use of readily biodegradable substrate (NUR_g) whereas the slower rate corresponds to use of slowly biodegradable substrate (NUR_h). (From S. W. Givens, et al., Biological process design and pilot testing for a carbon oxidation, nitrification and denitrification system. *Environmental Progress* **10**:133–146, 1991. Copyright © American Institute of Chemical Engineers; reprinted with permission.)

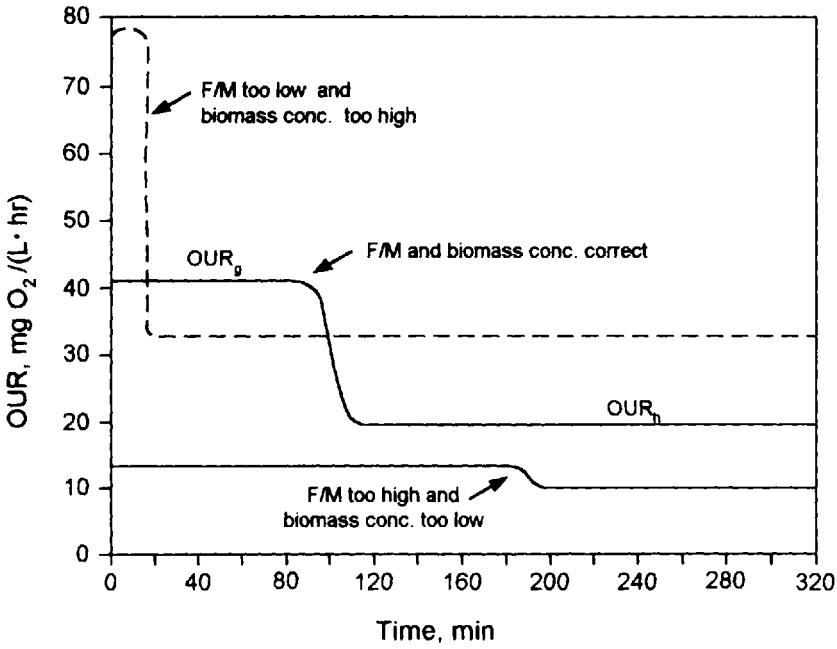


Figure 8.9 Effect of changing the substrate to biomass ratio (F/M) on the OUR in a batch reactor. The faster rate corresponds to use of readily biodegradable substrate (OUR_g) whereas the slower rate corresponds to use of slowly biodegradable substrate (OUR_h). (From G. A. Ekama, et al., Procedures for determining influent COD fractions and the maximum specific growth rate of heterotrophs in activated sludge systems. *Water Science and Technology* **18**(6): 91–114, 1986. Copyright © Elsevier Science Ltd.; reprinted with permission.)

$$\eta_g = \frac{2.86 \times NUR_g}{OUR_g} \tag{8.26}$$

Likewise, the value of η_h can be calculated from OUR_h and NUR_h:

$$\eta_h = \frac{2.86 \times NUR_h}{OUR_h} \tag{8.27}$$

Hydrolysis and Ammonification. Three parameters remain to be evaluated. Two are the parameters characterizing hydrolysis of slowly biodegradable substrate, k_h and K_X. The third is the parameter describing ammonification, k_a. These parameters can be evaluated with data collected with the CSTR receiving feed conforming to a daily cyclic square wave pattern.[19] This is done by nonlinear curve fitting of ASM No. 1 to the OUR data collected during one feed cycle. All other parameter values are known or assumed. The OUR pattern for such a bioreactor was shown in Figure 8.7. The plateau in the OUR after feed cessation is due to degradation of organic substrate released by hydrolysis of slowly biodegradable substrate. The existence of a sustained plateau is evidence that the biomass is saturated and that hydrolysis is occurring at the maximum rate, thereby allowing evaluation of k_h. Operation of the bioreactor at an SRT of two days ensures that this condition will exist. The pattern by which OUR declines with time is determined by K_X. The best

way to estimate k_h and K_X is by curve-fitting techniques to match the OUR response of the model to the OUR pattern in Figure 8.7. Because of the short SRT involved, nitrification will not be occurring. Furthermore, since the bioreactor is fully aerobic, denitrification need not be considered. Thus, a simplified form of the model may be used. Since all of the other parameters have been selected, the only unknowns for the curve-fit are k_h and K_X. In addition, because nitrification is excluded from the bioreactor, ammonia will build up as ammonification occurs. Consequently, estimation of k_a can be based on the release of ammonia during the nonfeed period.[19]

8.5.4 Order of Determination

Evaluation of the parameters and the wastewater characteristics must proceed in a particular order because the values of some are needed before others can be obtained. Table 8.2 summarizes the order of their determination.

Table 8.2 Parameters and Characteristics Which Must Be Evaluated and Information Needed

Symbol	Name	Prior information needed
S_{NOO}	Soluble nitrate-N concentration in wastewater	
S_{NHO}	Soluble ammonia-N concentration in wastewater	
S_{IO}	Soluble inert COD concentration in wastewater	
S_{NIO}	Soluble inert organic-N concentration in wastewater	
S_{NDO}	Soluble biodegradable organic-N concentration in wastewater	S_{NIO}
Y_H	Yield for heterotrophic biomass	
S_{SO}	Concentration of readily biodegradable COD in wastewater	Y_H
$\hat{\mu}_A$	Maximum specific growth rate for autotrophic biomass	$b_{I,A}$
K_{NH}	Ammonia-N half-saturation coefficient for autotrophic biomass	
$b_{I,H}$	Decay coefficient for heterotrophic biomass	Y_H, f'_D
X_{IO}	Inert suspended organic matter concentration in wastewater	f'_D, $b_{I,H}$, S_{SO}, S_{IO}
X_{SO}	Slowly biodegradable organic matter concentration in wastewater	
X_{NDO}	Slowly biodegradable organic-N concentration in wastewater	S_{SO}, X_{SO}, S_{NDO}
η_g	Correction factor for μ_H under anoxic conditions	
η_h	Correction factor for hydrolysis under anoxic conditions	
$\hat{\mu}_H$	Maximum specific growth rate for heterotrophic biomass	Y_H, X_{SO}, X_{IO}, S_{SO}, f'_D
K_S	Half-saturation coefficient for heterotrophic biomass	Y_H, X_{SO}, X_{IO}, S_{SO}, f'_D
k_h	Maximum specific hydrolysis rate	
K_X	Half-saturation coefficient for hydrolysis of slowly biodegradable substrate	
k_a	Ammonification rate	

8.6 USING TRADITIONAL MEASUREMENTS TO APPROXIMATE WASTEWATER CHARACTERISTICS FOR MODELING

As seen in the preceding section, characterization of a complex wastewater in a manner suitable for use with ASM No. 1 is quite involved and represents a significant investment of time and money. Consequently, such characterizations are not ordinarily done as part of the routine measurements made at wastewater treatment plants. Rather, in the United States, wastewaters are normally characterized in terms of the concentrations of TSS, VSS, five-day biochemical oxygen demand (BOD_5), ammonia-N, total Kjeldahl nitrogen (TKN), and alkalinity. Total COD is also commonly measured, but the frequency is usually less than that of the other characteristics, although it is increasing. Furthermore, distinction is seldom made between soluble and particulate phases during measurements of BOD_5, COD, and TKN. Because there are circumstances in which it would be advantageous to conduct preliminary modeling studies prior to conducting detailed treatability studies, it would be very useful to be able to translate the traditional data available in the records of wastewater treatment plants into a form that can be used with the models presented herein. Luckily, with a few simplifying assumptions, this can be done for domestic wastewaters. Such translations cannot be made for industrial wastewaters, however, because each is unique.

As indicated by Eq. 8.23, the total COD in a wastewater (COD_{TO}) is made up of four components: (1) particulate biodegradable COD (X_{SO}), (2) soluble biodegradable COD (S_{SO}), (3) particulate inert COD (X_{IO}), and (4) soluble inert COD (S_{IO}). (The subscript O in these symbols and the ones to follow designates that they are influent concentrations.) As a consequence, data on the total COD in the wastewater is essential to determination of the other constituents. If no COD data are available, the total COD of domestic wastewater can be approximated as:[26,37]

$$COD_{TO} \approx (2.1)(BOD_5) \tag{8.28}$$

The wastewater biodegradable COD (COD_{BO}) can be estimated from the ultimate BOD (BOD_u), which, in turn, can be estimated from the BOD_5:

$$BOD_u = \frac{BOD_5}{1 - 10^{-5k}} \tag{8.29}$$

where k is the BOD rate coefficient with units of day^{-1}. For domestic wastewater, the relationship between the ultimate BOD and the five-day BOD can be approximated as:[26]

$$BOD_u \approx (1.5)(BOD_5) \tag{8.30}$$

The biodegradable COD is greater than the ultimate BOD because the latter does not account for the electrons retained in the biomass debris formed during the BOD test. Consequently,

$$COD_{BO} = \frac{BOD_u}{(1 - f_D \cdot Y_H)} \tag{8.31}$$

where Y_H is in COD/COD units and f_D can be assumed to have a value of 0.20 mg

debris COD/mg biomass COD.[19] Substituting Eq. 8.29 for ultimate BOD into Eq. 8.31 allows the biodegradable COD to be estimated from the 5-day BOD:

$$COD_{BO} = \frac{BOD_5}{(1 - f_D \cdot Y_H)(1 - 10^{-5k})} \tag{8.32}$$

For wastewaters with a significant industrial component, both k and Y_H would have to be measured to allow the conversion, but for domestic wastewater, Y_H can be assumed to be 0.60,[19] and Eq. 8.30 can be used as the relationship between the two types of BOD, giving:

$$COD_{BO} \approx (1.14)(BOD_u) \approx (1.71)(BOD_5) \tag{8.33}$$

The division of biodegradable COD into slowly and readily biodegradable fractions requires specific knowledge of the wastewater in question. This is necessary because slowly biodegradable substrate is not necessarily the same as particulate biodegradable substrate, even though it is considered to be particulate in the models. Rather, some of the slowly biodegradable substrate may pass through the filters used to determine VSS concentrations. As discussed earlier, the physical assay of Mamais et al.[24] provides a simple means of determining the readily biodegradable COD in a wastewater. If no other basis is available for making the division, its use is encouraged, even during preliminary studies. If that can't be done, a "best guess" division must be made based on experience.

The nonbiodegradable, or inert, COD (COD_{IO}) is the difference between the total COD and the biodegradable COD:

$$COD_{IO} = COD_{TO} - COD_{BO} \tag{8.34}$$

It must be partitioned into soluble (S_{IO}) and particulate (X_{IO}) forms. Experience suggests that 35 to 40 percent of the particulate organic matter in domestic wastewater is nonbiodegradable.[19,20] Particulate organic matter is represented by the VSS. If one assumes that the elemental composition of the inert particulate organic matter is similar to that of protein, which has a COD equivalent of 1.5 g COD/g protein (Table 3.1), and that protein is totally volatile in a volatile suspended solids test, then:

$$X_{IO} \approx (0.375)(1.50)(VSS) = 0.56(VSS) \tag{8.35}$$

The soluble inert COD can be calculated by difference:

$$S_{IO} = COD_{IO} - X_{IO} \tag{8.36}$$

If one were going to use BOD as a measure of biodegradable organic matter, it would be better to measure the ultimate carbonaceous BOD (BOD_u) than to measure BOD_5, because the relationship expressed by Eq. 8.31 is subject to less error and variability than the relationship between biodegradable COD and BOD_5 expressed by Eq. 8.32, for which an assumed value of k is required. Although Eq. 8.33 can be used as a rough approximation for domestic wastewater, the relationship between biodegradable COD and BOD_5 depends on the rate of oxygen consumption in the BOD test, as indicated by Eq. 8.32. Since that rate will be influenced by the nature of the organic chemicals present in the sample being tested, the presence of industrial discharges to a municipal wastewater treatment plant may well change the relationship from that associated with a strictly domestic wastewater. Thus, there is a high degree of uncertainty associated with Eq. 8.33 under that circumstance and

it is better to use Eq. 8.32 with a measured value of the BOD rate coefficient, k. The best course of action, however, is to use COD as the measure of organic substrates, accounting for the nonbiodegradable material in the ways outlined earlier.

Some of the wastewater characteristics used in modeling are routinely measured directly: ammonia-N, nitrate-N, and alkalinity. The ammonia-N and nitrate-N concentrations can be used without conversion. Most domestic wastewaters contain no nitrate-N, although industrial wastewaters might. Alkalinity is typically measured as $CaCO_3$, but is expressed as mM/L in ASM No. 1 and No. 2. Since the molecular weight of $CaCO_3$ is 100, the conversion is simple.

Without treatability studies, the other nitrogen forms used in modeling must be deduced from the routine measurements made at wastewater treatment plants. The total organic nitrogen concentration (ON_{TO}) is the difference between the TKN and ammonia-N concentrations:

$$ON_{TO} = TKN - S_{NHO} \tag{8.37}$$

Furthermore, the wastewater total organic nitrogen can be divided into soluble, particulate, biodegradable, and inert fractions:

$$ON_{TO} = S_{NSO} + S_{NIO} + X_{NSO} + X_{NIO} \tag{8.38}$$

The concentration of soluble, inert organic nitrogen (S_{NIO}) in domestic wastewaters typically ranges from 1 to 2 mg/L as N,[30] suggesting that a value of 1.5 mg/L as N can be assumed for preliminary modeling without fear of gross error. The particulate inert organic nitrogen is associated with the particulate inert organic matter. The nitrogen content of this material can be assumed to be equal to $i_{N\,XD}$, the nitrogen content of biomass debris. Consequently, the concentration of particulate inert organic nitrogen (X_{NIO}) can be approximated as:

$$X_{NIO} \approx i_{N\,XD} \cdot X_{IO} \tag{8.39}$$

When possible, distribution of biodegradable organic nitrogen between the soluble and particulate phases should be based on data. In the absence of specific data, the biodegradable organic nitrogen is often distributed into particulate and soluble fractions in the same proportions as the slowly and readily biodegradable fractions of the COD, as indicated in Eq. 8.25.

Example 8.6.1

Conventional characterization of a domestic wastewater following primary clarification is given in the upper portion of Table E8.4. Translate that information into a form that can be used in ASM No. 1.

a. The first task is to estimate the concentration of biodegradable COD. This is done by using the BOD_5 and Eq. 8.33:

$$COD_{BO} = (1.71)(155) = 265 \text{ mg COD/L}$$

b. The inert COD concentration can then be calculated from the total COD using Eq. 8.34:

$$COD_{IO} = 325 - 265 = 60 \text{ mg COD/L}$$

Table E8.4 Translation of Traditional Wastewater[a] Characteristics into a Form Suitable for Modeling

Component	Concentration
Conventional wastewater characterization	
TSS	82 mg/L
VSS	61.5 mg/L
BOD_5	155 mg/L
Total COD	325 mg/L as COD
Ammonia-N	25 mg/L as N
Total Kjeldahl nitrogen (TKN)	43.5 mg/L as N
Nitrate-N	0.0 mg/L as N
Alkalinity	200 mg/L as $CaCO_3$
Characterization as required for use in ASM No. 1	
Particulate inert organic matter	35 mg/L as COD
Soluble inert organic matter	25 mg/L as COD
Slowly biodegradable substrate	150 mg/L as COD
Readily biodegradable substrate	115 mg/L as COD
Oxygen	0 mg/L as O_2
Soluble nitrate nitrogen	0 mg/L as N
Soluble ammonia nitrogen	25 mg/L as N
Soluble biodegradable organic nitrogen	6.5 mg/L as N
Particulate biodegradable organic nitrogen	8.5 mg/L as N
Alkalinity	2 mM/L

[a]The wastewater is considered to be typical of domestic wastewater that has undergone primary sedimentation.

c. The concentration of particulate inert COD can be estimated from the VSS concentration using Eq. 8.35:

$$X_{IO} = (0.56)(61.5) = 35 \text{ mg COD/L}$$

d. This, in turn allows the soluble inert COD concentration to be calculated from Eq. 8.36:

$$S_{IO} = 60 - 35 = 25 \text{ mg COD/L}$$

e. Partitioning of the biodegradable COD into slowly and readily biodegradable fractions requires some knowledge of the nature of the wastewater. Additional information suggests that 43% of the biodegradable COD is readily biodegradable. Consequently,

$$S_{SO} = (0.43)(265) = 115 \text{ mg COD/L}$$

and

$$X_{SO} = 265 - 115 = 150 \text{ mg COD/L}$$

f. The concentration of organic nitrogen in the wastewater can be obtained as the difference between the TKN and the ammonia-N concentrations as expressed in Eq. 8.37:

$$ON_{IO} = 43.5 - 25 = 18.5 \text{ mg N/L}$$

g. The biodegradable organic nitrogen concentration must be obtained by subtracting the concentrations of the soluble and particulate inert organic nitrogen. The concentration of soluble inert organic nitrogen can be assumed to be 1.5 mg N/L, as discussed previously. The concentration of particulate inert organic nitrogen can be estimated with Eq. 8.39 by assuming a value for $i_{N\,XD}$. A reasonable value 0.06 mg N/mg COD, as indicated in Table 6.3. Consequently:

$$X_{NID} = (0.06)(35) = 2 \text{ mg N/L}$$

Use of Eq. 8.38 gives the biodegradable organic nitrogen concentration:

$$S_{NSO} + X_{NSO} = 18.5 - 1.5 - 2.0 = 15 \text{ mg N/L}$$

h. Partitioning of the biodegradable organic nitrogen in accordance with Eq. 8.25 gives:

$$S_{NSO} = (15.0)\left(\frac{115}{150 + 115}\right) = 6.5 \text{ mg N/L}$$

Consequently,

$$X_{NSO} = 15.0 - 6.5 = 8.5 \text{ mg N/L}$$

The estimated characteristics of the wastewater are listed in the lower portion of Table E8.4. Comparison of them to the values in Table 6.6 shows that with the exception of alkalinity, they are the same. The alkalinity is a very site-specific characteristic, being dependent on the nature of the carriage water.

8.7 CONVERTING KINETIC AND STOICHIOMETRIC PARAMETER VALUES BETWEEN CHEMICAL OXYGEN DEMAND AND OTHER UNITS

So far in this book, the concentrations of all organic constituents, both substrate and biomass, and the values of all kinetic and stoichiometric parameters have been presented as COD. This has been done because it simplifies the equations and makes the computation of COD, i.e., electron, balances very straightforward. Furthermore, there are strong arguments for using biodegradable COD routinely as the measure of organic substrate concentrations, with the result that there is an increasing trend toward this practice throughout the world. Consequently, COD will be used as the measure of organic substrate concentration almost exclusively in this book, and the majority of the kinetic and stoichiometric coefficients will be presented in terms of it. We recognize that the BOD test is still widely used in the United States and that most historical records are expressed in terms of it. Thus, most historical data on kinetic and stoichiometric coefficients use BOD as the measure of organic substrate. Conversion of half-saturation coefficients to COD units can be done with the expressions presented in Section 8.6. In this section, we present the equations for converting yield values so that substrate is expressed as COD and biomass is expressed either as COD or as TSS.

Conversion between COD units and TSS or VSS units is straightforward for biomass and biomass debris because there is little variability associated with the COD equivalents of those constituents. Consequently, biomass concentrations may

be converted between the two units systems by using the coefficients $i_{O\,XB,T}$ and $i_{O\,XB,V}$ defined in Section 8.2.1 and quantified in Table 3.1. To assist in the interconversions for particulate constituents, the following conventions will be adopted throughout the remainder of this book. All symbols for particulate constituents defined in the simplified model of Chapter 5 and in ASM No. 1 will be assumed to be in COD units. When particulate constituents are expressed in TSS units, a subscripted T will be attached and when they are expressed in VSS units, a subscripted V will be used. For example, the concentration of active heterotrophic biomass will be denoted as $X_{B,H}$ when expressed in COD units and as $X_{B,H,T}$ when expressed in TSS units. Therefore,

$$X_{B,H} = i_{O\,XB,T} \cdot X_{B,H,T} \tag{8.40}$$

Likewise,

$$X_{B,H} = i_{O\,XB,V} \cdot X_{B,H,V} \tag{8.41}$$

Similar conversions apply for autotrophic biomass, biomass debris, etc.

Throughout the remainder of this book, biomass and MLSS concentrations will be expressed in either COD or TSS units, with the use of TSS units being indicated by a subscripted T, as indicated above. Although VSS units are frequently used by practitioners, we have adopted only those two to reduce the number of interconversions required and to make the examples easier to follow. If a reader has need to use VSS units, the interconversion is a straightforward one.

Yield values express the amount of biomass formed per unit of substrate used. Consequently, if there are three possible measures of biomass (COD, TSS, and VSS) and three possible measures of substrate (COD, BOD_u, and BOD_5), then there are nine possible yield values to describe the same situation. This is one reason we have limited biomass measurements to two techniques (COD and TSS) and substrate measurements to one (COD). In that way the reader will only have to routinely deal with two types of yield values, and the interconversion of one to the other is straightforward. If we employ a subscripted T to indicate that the biomass concentration is expressed in TSS units, then the relationship between a COD/COD yield and a TSS/COD yield is given by:

$$Y_H = i_{O\,XB,T} \cdot Y_{H,T} \tag{8.42}$$

Although they will not commonly be used herein, conversion factors are presented below for the other types of heterotrophic yield one is likely to encounter, thereby assisting the reader in converting historical yield values into the forms used herein. They were derived by applying the substrate and biomass relationships of Section 8.6. The additional subscripts UB and 5B will be used to represent ultimate and 5-day BOD, respectively, as the measure of substrate. For yields in which biomass is measured as VSS and substrate is measured as COD, the relationship is simple:

$$Y_H = i_{O\,XB,T} \cdot Y_{H,T} = i_{O\,XB,V} \cdot Y_{H,V} \tag{8.43}$$

Yields measured in TSS and ultimate BOD units can be converted to the yields used herein with:

$$Y_H = i_{O\,XB,T} \cdot Y_{H,T} = \frac{i_{O\,XB,T} \cdot Y_{H,T\,UB}}{1 + f_D \cdot i_{O\,XB,T} \cdot Y_{H,T\,UB}} \tag{8.44}$$

For yields measured in VSS and ultimate BOD units, the conversion can be done with:

$$Y_H = i_{O\,XB,T} \cdot Y_{H,T} = \frac{i_{O\,XB,V} \cdot Y_{H,V\,UB}}{1 + f_D \cdot i_{O\,XB,V} \cdot Y_{H,V\,UB}} \tag{8.45}$$

When the substrate is measured as BOD_5 and the biomass is expressed as TSS, the expression is:

$$Y_H = i_{O\,XB,T} \cdot Y_{H,T} = \frac{i_{O\,XB,T} \cdot Y_{H,T\,5B}(1 - 10^{-5k})}{1 + f_D \cdot i_{O\,XB,T} \cdot Y_{H,T\,5B}(1 - 10^{-5k})} \tag{8.46}$$

Finally, when biomass is expressed as VSS and substrate as BOD_5, conversion may be done with:

$$Y_H = i_{O\,XB,T} \cdot Y_{H,T} = \frac{i_{O\,XB,V} \cdot Y_{H,V\,5B}(1 - 10^{-5k})}{1 + f_D \cdot i_{O\,XB,V} \cdot Y_{H,V\,5B}(1 - 10^{-5k})} \tag{8.47}$$

For domestic wastewater, the term $(1 - 10^{-5k})$ often has a value of 0.67, as indicated by Eq. 8.30, allowing simplification of Eqs. 8.46 and 8.47.

Translation of autotrophic yields requires only conversion of the biomass concentration because the substrate is expressed as nitrogen in all situations. Thus,

$$Y_A = i_{O\,XB,T} \cdot Y_{A,T} = i_{O\,XB,V} \cdot Y_{A,V} \tag{8.48}$$

It should be noted that it is common for designers to use days as the unit of time in kinetic parameters, reflecting the fact that SRTs are normally longer than one day. Conversion between hours and days as the units of time is a trivial exercise, but it is important to recognize that it may be necessary.

Example 8.7.1

The parameter values in Table 6.3 represent typical ones for use in ASM No. 1 for domestic wastewater at neutral pH and a temperature of 20°C. You wish to use the simple model of Chapter 5 to make rough calculations about the size of a bioreactor system before conducting detailed simulations using ASM No. 1. Convert the parameters into a form that will allow this to be done. Express the parameters in both COD/COD and COD/TSS units, with days as the measure of time.

The results of the conversion are shown in Table E8.5. First compare the column in COD/COD units to Table 6.3. The differences between the two tables in $\hat{\mu}_H$, $\hat{\mu}_A$, and b_A (which is equal to $b_{L,A}$) are only in the change from hr^{-1} to day^{-1}. The conversion from $b_{L,H}$ to b_H requires application of Eq. 3.69 and the change of time units from hr^{-1} to day^{-1}, whereas the conversion of f'_D to f_D requires application of Eq. 3.68 as discussed in Section 3.3.2.

Conversion of the parameters of the simple model from COD/COD to COD/TSS units requires changes in Y_H, and Y_A. Conversion of Y_H to $Y_{H,T}$ requires use of Eq. 8.42:

$$Y_{H,T} = \frac{0.60}{(1.20)} = 0.50$$

Table E8.5 Stoichiometric and Kinetic Parameters from Table 6.3 After Conversion for Use in the Simple Model from Chapter 5

Parameters in COD/COD Units		Parameters in COD/TSS Units	
Parameters	Values	Parameters	Values
$\hat{\mu}_H$	6.0 day^{-1}	$\hat{\mu}_H$	6.0 day^{-1}
K_S	20 mg/L as COD	K_S	20 mg/L as COD
Y_H	0.60 mg biomass COD/mg COD removed	$Y_{H,T}$	0.50 mg biomass TSS/mg COD removed
b_H	0.18 day^{-1}	b_H	0.18 day^{-1}
f_D	0.20 mg debris COD/mg biomass COD	$f_{D,T}$	0.20 mg debris TSS/mg biomass TSS
		$i_{O\,XB,T}$	1.2 mg COD/mg TSS
$\hat{\mu}_A$	0.77 day^{-1}	$\hat{\mu}_A$	0.77 day^{-1}
K_{NH}	1.0 mg/L as N	K_{NH}	1.0 mg/L as N
$K_{O,A}$	0.75 mg/L as O_2	$K_{O,A}$	0.75 mg/L as O_2
Y_A	0.24 mg biomass COD/mg N oxidized	$Y_{A,T}$	0.20 mg biomass TSS/mg N oxidized
b_A	0.10 day^{-1}	b_A	0.10 day^{-1}

Conversion of Y_A to $Y_{A,T}$ requires use of Eq. 8.48:

$$Y_{A,T} = \frac{0.24}{1.20} = 0.20$$

Thus, it can be seen that conversion between the various models and unit systems is very straightforward as long as the basic definitions of terms are considered.

Example 8.7.2

Historical data from a domestic wastewater treatment plant suggest that the heterotrophic biomass yield is 0.70 mg VSS formed/mg BOD_5 removed. What would be the value of the yield if it were expressed in mg TSS formed/mg COD removed, i.e., as $Y_{H,T}$?

This conversion can be made by using Eq. 8.47. Since we are not interested in Y_H (which is in COD/COD units) it can be eliminated and $i_{O\,XB,T}$ can be moved to the right side. From Table 3.1, $i_{O\,XB,T}$ has a value of 1.20 g COD/g TSS and $i_{O\,XB,V}$ has a value of 1.42 g COD/g VSS. Furthermore, since the wastewater is domestic, the value of the term $(1 - 10^{-5k})$ can be assumed to be 0.67. Finally, f_D has a value of 0.20. Therefore:

$$Y_{H,T} = \frac{i_{O\,XB,V} \cdot Y_{H,V\,5B}(1 - 10^{-5k})}{i_{O\,XB,T}[1 + f_D \cdot i_{O\,XB,V} \cdot Y_{H,V\,5B}(1 - 10^{-5k})]}$$

$$= \frac{(1.42)(0.70)(0.67)}{(1.20)[1 + (0.20)(1.42)(0.70)(0.67)]}$$

$$= 0.49 \text{ mg TSS/mg COD}$$

The major difficulty in moving between COD/COD and TSS/COD units is the presence of fixed suspended solids (FSS) in wastewaters:

$$FSS = TSS - VSS \tag{8.49}$$

Fixed suspended solids are inorganic and thus have no COD. As a consequence, they are not considered when particulate concentrations are expressed on either a COD or a VSS basis. However, they must be considered when the concentrations of particulate materials, such as MLSS, are expressed in TSS units. Fixed suspended solids undergo no reactions in biochemical operations. Rather, they behave like inert organic solids, as discussed in Section 5.2.2. Consequently, when the MLSS concentration is being calculated in TSS units, the influent FSS concentration, given the symbol X_{FO}, should be used in addition to the influent inert organic solids, X_{IO}, in the appropriate equations. It should be handled in exactly the same manner as influent inert organic solids in all computations.

8.8 KEY POINTS

1. Care should be exercised in the design of a treatability study because the manner in which bioreactors are configured and operated has a strong effect on the microbial community that develops, thereby influencing the values of the kinetic and stoichiometric parameters obtained from the study.
2. During treatability studies to evaluate the parameters in the simple model of Chapter 5, data related to effluent quality and biomass concentrations should be collected from continuous stirred tank reactors (CSTRs) operated over a range of solids retention times (SRTs). Biomass concentrations may be expressed as chemical oxygen demand (COD), total suspended solids (TSS), or volatile suspended solids (VSS) as long as appropriate conversion factors are used to allow COD balances to be made.
3. Because of the interrelationships among the parameters in the simple model of Chapter 5, their values must be estimated in a certain order. Y_H and b_H are evaluated together, but b_H must be known before f_H and the Monod parameters ($\hat{\mu}_H$ and K_S) can be evaluated. The concentration of inert soluble COD, S_I, must also be known before the Monod parameters can be estimated.
4. Although yield and biomass concentrations are expressed in COD units in Chapters 5 through 7, they also may be measured and expressed in TSS or VSS units as long as the unit of expression is used consistently and an appropriate COD conversion factor ($i_{O\,XB,I}$ or $i_{O\,XB,V}$, respectively) is used in COD balances.
5. Nonlinear parameter estimation techniques are preferable, but several methods of linearizing the Monod equation are in common use for determining the values of the kinetic parameters in it. The efficacies of those linearizations depend on the nature of the error in the data set.

6. The active fraction of the biomass is difficult to measure and is not routinely measured during treatability studies. Luckily, f_D does not vary greatly, allowing a value of 0.20 to be assumed with little risk of error.
7. When data are not available concerning the active fraction of the biomass in the CSTRs used for treatability studies, a separate batch experiment must be performed to determine b_H. The experiment measures the change in oxygen uptake rate (OUR) of the biomass over time. Once the value of b_H is known, it may be used with an assumed f_D value to determine Y_H. The Monod kinetic parameters can then be evaluated in the same manner as previously described.
8. Consideration must be given to the physiological state of a culture and how it may change during batch experiments for determining biodegradation kinetics for single organic compounds. If the physiological state is not allowed to change, the resulting parameter values are called extant parameters because they reflect the conditions of the culture in the bioreactor from which they were removed prior to testing. If the test conditions allow the enzyme system of the culture to develop to the point that the microorganisms can grow at the fastest rate possible on the test substrate at the given temperature and pH, the kinetic parameters are called intrinsic.
9. Both intrinsic and extant biodegradation kinetic parameter values can be determined during batch experiments, although the experimental conditions required are quite different. To obtain intrinsic parameter estimates the initial substrate to relevant biomass ratio should be at least 20 when both are measured as COD. In contrast, to obtain extant parameter estimates the ratio should be less than 0.02. For both estimates, the initial substrate concentration should exceed the expected K_S value.
10. Because of the large number of parameters required by ASM No. 1, extensive treatability testing is required to fully evaluate them. Several CSTRs with biomass recycle should be run over a range of SRTs in excess of 5 days to provide data for evaluating biomass production and electron acceptor requirement. In addition, a similar bioreactor should be operated at an SRT of 2 days while receiving a daily cyclic square wave feeding pattern. Finally, two CSTRs in series should be operated as a modified Ludzack–Ettinger (MLE) system to provide biomass capable of denitrification.
11. Characterization of a wastewater and estimation of the stoichiometric coefficients requires data from CSTRs operated over a range of SRTs as well as a CSTR with a cyclic feeding pattern. The estimate of Y_H must be obtained from an experiment in which a small amount of biomass is used to seed a batch bioreactor containing only the soluble portion of the wastewater. The concentration of inert soluble COD is estimated in the same way as for the simple model. The concentration of readily biodegradable COD can either be obtained from OUR data from the CSTR with the cyclic feeding pattern or by an assay involving coagulation and filtration. The concentration of inert particulate organic matter is estimated by fitting a simplified model to data relating the mixed liquor suspended solids (MLSS) concentration in the CSTRs to their SRTs. The concentra-

tion of slowly biodegradable COD can be calculated from the total influent COD and the previously measured COD values.

12. The most important kinetic parameter in activated sludge model (ASM) No. 1 is $\hat{\mu}_A$, the maximum specific growth rate coefficient for nitrifying bacteria. Thus, it is important that it be determined in the wastewater matrix. This can be done by measuring the increase in nitrate-N and nitrite-N concentrations over time in a batch growth experiment and plotting $\ln(NO_3\text{-N} + NO_2\text{-N})$ versus time. The slope of the resulting plot is $\hat{\mu}_A - b_A$, allowing the value of $\hat{\mu}_A$ to be calculated.
13. Tests to measure η_g, the correction factor for heterotrophic growth under anoxic conditions, and η_h, the correction factor for hydrolysis of slowly biodegradable substrate under anoxic conditions, are performed at the same time by evaluating oxygen and nitrate consumption rates in two batch bioreactors which are equivalent in every respect except for the terminal electron acceptor.
14. The parameters characterizing hydrolysis of slowly biodegradable substrate, k_h and K_X, can be evaluated with OUR data collected during one feed cycle of the CSTR receiving the daily cyclic square wave feeding pattern. Because all other parameters are known, k_h and K_X can be estimated by nonlinear curve fitting of ASM No. 1 to the OUR data.
15. By using appropriate approximations, it is possible to use traditional wastewater characteristics (TSS, VSS, 5 day biochemical oxygen demand [BOD_5], ammonia-N, and total kjeldahl nitrogen [TKN]) to estimate the concentrations of the constituents required to use ASM No. 1. However, specific information on the particular wastewater in question is required to make the split between readily and slowly biodegradable substrate concentrations.
16. Through the use of simple conversion factors, it is possible to convert traditional parameters expressed in BOD, TSS and VSS units into the COD units used in the simple model of Chapter 5 and in ASM No. 1.

8.9 STUDY QUESTIONS

1. Explain why treatability studies are often run in stages, especially for design of more complex systems.
2. Describe how you would modify the procedures presented in Section 8.2 to provide the information required to design a system to achieve nitrification in addition to carbon oxidation. In your description, tell how you would estimate the kinetic parameters describing nitrification.
3. Discuss the efficacy of the various techniques for estimating the Monod parameters ($\hat{\mu}_H$ and K_S) from experimental data relating the effluent soluble biodegradable COD to the SRT of the bioreactors.
4. A treatability study was performed on a wastewater using a CSTR with cell recycle. The wastewater was totally soluble and had a COD of 474 mg/L. The studies were run in a lab-scale bioreactor which had a volume of 6.0 L. The flow rate was maintained at a constant value of 1.0 L/hr and the SRT was maintained at the desired values by wasting excess

Table SQ8.1 Data Collected from CSTRs with Cell Recycle During a Treatability Study of a Soluble Wastewater

SRT hrs	Soluble COD mg COD/L	Biomass mg COD/L	Active fraction
48	62.0	1,385	0.95
96	45.8	2,377	0.90
144	41.1	3,114	0.85
192	38.8	3,716	0.81
288	36.7	4,702	0.74
384	35.6	5,536	0.68

biomass directly from the bioreactors. A batch experiment with mixed liquor removed from the CSTR revealed that the concentration of inert soluble COD in the wastewater is 24 mg COD/L. Using the data provided in Table SQ8.1, determine the parameters $\hat{\mu}_H$, K_S, Y_H, b_H, and f_D. Use all three of the linearization techniques for estimating $\hat{\mu}_H$ and K_S, and compare their effectiveness. Could the first order approximation, i.e., k_c, be used to characterize this wastewater over the SRT range studied? Why?

5. Explain why nitrification does not interfere with the determination of b_H from OUR measurements in a batch bioreactor when nitrification is well established in the CSTR from which the biomass was obtained, but does interfere when only partial nitrification occurs in the CSTR.
6. A treatability study was performed on a wastewater using CSTRs with cell recycle. The wastewater was totally soluble and had a COD of 286 mg/L. The studies were run in lab-scale bioreactors with a volume of 8.0 L. The flow rate was maintained at a constant value of 2.0 L/hr and the SRT was maintained at the desired values by wasting excess biomass directly from the bioreactors. Data collected from the CSTRs are shown in Table SQ8.2. No data on the active fraction of the biomass were taken during the study. A batch experiment with mixed liquor removed from one CSTR revealed that the concentration of inert soluble COD in the

Table SQ8.2 Data Collected from CSTRs with Cell Recycle During a Treatability Study of a Soluble Wastewater

SRT hrs	Soluble COD mg COD/L	Biomass mg COD/L
48	41.7	1,530
96	39.5	2,580
144	38.8	3,390
192	38.5	4,070
288	38.2	5,240
384	38.0	6,280

wastewater is 36 mg COD/L. Another batch experiment with biomass taken from the CSTR with an SRT of 144 hr was performed for the determination of b_H. The results are shown in Table SQ8.3. Using the available data, determine the values of b_H and Y_H describing the biomass.

7. Explain what is meant by the term physiological state and why it will influence the values of the kinetic parameters describing biodegradation of an organic compound when those parameters are measured in batch experiments.
8. Explain why oxygen consumption measurements can be used as a surrogate for measurements of biomass growth or substrate utilization during batch tests for determining intrinsic or extant kinetic parameters.
9. Describe the types of bioreactors that should be operated and the type of data that should be collected during treatability studies to evaluate the parameters in ASM No. 1.
10. Describe how the various fractions of the wastewater COD (inert soluble, inert particulate, readily biodegradable, and slowly biodegradable) may be estimated during treatability studies for use of ASM No. 1.
11. Describe and contrast the procedures used to evaluate the kinetic parameters for aerobic growth of heterotrophs and aerobic growth of autotrophs for use in ASM No. 1.
12. Describe the procedure for estimating the correction factors for growth and hydrolysis under anoxic conditions, η_g and η_h, for use in ASM No. 1.
13. Describe the procedure for estimating the parameters describing hydrolysis of slowly biodegradable substrate under aerobic conditions.
14. A domestic wastewater has the characteristics listed in Table SQ8.4. Use those characteristics to estimate the concentrations of the various constituents required for using ASM No. 1. A listing of those constituents is

Table SQ8.3 Data Collected During a Batch Aeration Test for the Determination of b_H

Time hrs	OUR mg O_2/(L·hr)
0	37.0
12	33.4
24	30.2
36	27.2
48	24.6
72	20.1
96	16.4
120	13.3
144	10.9
192	7.2
240	4.8

Table SQ8.4 Characteristics of a Domestic Wastewater in Traditional Terms

Component	Concentration
TSS	125 mg/L
VSS	100 mg/L
BOD_5	225 mg/L
Total COD	475 mg/L as COD
Ammonia-N	35 mg/L as N
Total Kjeldahl nitrogen (TKN)	60 mg/L as N
Nitrate-N	0.0 mg/L as N
Alkalinity	200 mg/L as $CaCO_3$

provided in the lower portion of Table E8.4. Assume that the readily biodegradable substrate is 35% of the total biodegradable COD.

15. Historical data from a domestic wastewater treatment plant suggests that the heterotrophic biomass yield has a value of 0.85 mg TSS formed/mg BOD_5 removed. Express the yield in units of mg biomass COD formed/mg COD removed for use in ASM No. 1.

REFERENCES

1. Antoniou, P., J. Hamilton, B. Koopman, R. Jain, B. Holloway, G. Lyberatos, and S. A. Svoronos, Effect of temperature and pH on the effective maximum specific growth rate of nitrifying bacteria. *Water Research* **24**:97–101, 1990.
2. Berthouex, P. M. and D. R. Gan, Discussion of 'A comparison of estimates of kinetic constants for a suspended growth treatment system from various linear transformations.' *Research Journal, Water Pollution Control Federation* **63**:820–823, 1991.
3. Brown, S. C., C. P. L. Grady Jr., and H. H. Tabak, Biodegradation kinetics of substituted phenolics: demonstration of a protocol based on electrolytic respirometry. *Water Research* **24**:853–861, 1990.
4. Cech, J. S., J. Chudoba and P. Grau, Determination of kinetic constants of activated sludge microorganisms. *Water Science and Technology* **17**(2/3):259–272, 1985.
5. Chudoba, J., J. S. Cech, J. Farkac and P. Grau, Control of activated sludge filamentous bulking: experimental verification of a kinetic selection theory. *Water Research* **19**: 191–196, 1985.
6. Dang, J. S., D. M. Harvey, A. Jobbágy, and C. P. L. Grady Jr., Evaluation of biodegradation kinetics with respirometric data. *Research Journal, Water Pollution Control Federation* **61**:1711–1721, 1989.
7. Dowd, J. E. and D. S. Riggs, A comparison of estimates of Michaelis-Menten kinetic constants from various linear transformations. *Journal of Biological Chemistry* **240**: 863–869, 1965.
8. Drtil, M., P. Nemeth and I. Bodik, Kinetic constants of nitrification. *Water Research* **27**: 35–39, 1993.
9. Ekama, G. A., P. L. Dold and G. v. R. Marais, Procedures for determining influent COD fractions and the maximum specific growth rate of heterotrophs in activated sludge systems. *Water Science and Technology* **18**(6):91–114, 1986.

10. Ellis, T. G., D. S. Barbeau, B. F. Smets, and C. P. L. Grady Jr., Respirometric technique for determination of extant kinetic parameters describing biodegradation. *Water Environment Research* **68**:917–926, 1996.
11. Ellis, T. G., B. F. Smets, and C. P. L. Grady Jr., Influence of simultaneous multiple substrate biodegradation on the kinetic parameters for individual substrates. *Proceedings of the Water Environment Federation 68th Annual Conference and Exposition, Miami Beach, Florida, October 21–25, 1995, Vol. 1, Wastewater Treatment Research and Municipal Wastewater Treatment*, pp. 167–168, 1995.
12. Germirli, F., D. Orhon, and N. Artan, Assessment of initial inert soluble COD in industrial wastewaters. *Water Science and Technology* **23**(4/6):1077–1086 , 1991.
13. Germirli, F., D. Orhon, N. Artan, E. Ubay, and E. Görgün, Effect of two-stage treatment on the biological treatability of strong industrial wastes. *Water Science and Technology* **28**(2):145–154, 1993.
14. Givens, S. W., E. V. Brown, S. R. Gelman, C. P. L. Grady Jr., and D. A. Skedsvold, Biological process design and pilot testing for a carbon oxidation, nitrification and denitrification system. *Environmental Progress* **10**:133–146, 1991.
15. Grady, C. P. L. Jr., G. Aichinger, S. F. Cooper, and M. Naziruddin, Biodegradation kinetics for selected toxic/hazardous organic compounds. *Hazardous Waste Treatment: Biosystems for Pollution Control*, Proceedings of the 1989 A&WMA/EPA International Symposium, Air & Waste Management Association, Pittsburgh, PA, pp. 141–153, 1989.
16. Grady, C. P. L. Jr., B. S. Magbanua, R. L. Buddin, A. G. Rodieck, R. W. Sanders II, W. W. Sowers, and J. C. Stanfill, Relative efficacy of extant and intrinsic kinetic parameter estimates for predicting the removal of synthetic organic chemicals by activated sludge. *Proceedings of the Water Environment Federation 69th Annual Conference and Exposition, Dallas, Texas, October 5–9, 1996, Vol. 1, Wastewater Treatment Research and Municipal Wastewater Treatment*, pp. 103–114, 1996.
17. Grady, C. P. L. Jr., B. F. Smets, and D. S. Barbeau, Variability in kinetic parameter estimates: a review of possible causes and a proposed terminology. *Water Research* **30**: 742–748, 1996.
18. Hall, I. R., Some studies on nitrification in the activated sludge process. *Water Pollution Control* **73**:538–547, 1974.
19. Henze, M., C. P. L. Grady Jr., W. Gujer, G. v. R. Marais, and T. Matsuo, Activated sludge model No. 1. *IAWPRC Scientific and Technical Reports*, No. 1, 1987.
20. Henze, M., C. P. L. Grady, Jr., W. Gujer, G. v. R. Marais, and T. Matsuo, A general model for single-sludge wastewater treatment systems. *Water Research* **21**:505–515, 1987.
21. Kappeler, J. and W. Gujer, Estimation of kinetic parameters of heterotrophic biomass under aerobic conditions and characterization of wastewater for activated sludge modelling. *Water Science and Technology* **25**(6):125–139, 1992.
22. Lamb, J. C. III, W. C. Westgarth, J. L. Rogers, and A. P. Vernimmen, A technique for evaluating the biological treatability of industrial wastes. *Journal, Water Pollution Control Federation* **36**:1263–1284, 1964.
23. Lesouef, A., M. Payraudeau, F. Rogalla, and B. Kleiber, Optimizing nitrogen removal configurations by on-site calibration of the IAWPRC activated sludge model. *Water Science and Technology* **25**(6):105–123, 1992.
24. Mamais, D., D. Jenkins, and P. Pitt, A rapid physical-chemical method for determination of readily biodegradable COD in municipal wastewater. *Water Research* **27**:195–197, 1993.
25. Marais, G. v. R. and G. A. Ekama, The activated sludge process. Part I. Steady state behaviour. *Water SA* **2**:164–200, 1976.
26. Metcalf & Eddy, Inc., *Wastewater Engineering: Treatment, Disposal, Reuse*, 3rd Ed., McGraw-Hill, New York, 1991.

27. Moriarty, D. J. W., Measurement of bacterial growth rates in aquatic systems from rates of nucleic acid synthesis. *Advances in Microbial Ecology* **9**:245–292, 1986.
28. Ong, S. L., A comparison of estimates of kinetic constants for a suspended growth treatment system from various linear transformations. *Research Journal, Water Pollution Control Federation* **62**:894–900, 1990.
29. Orhon, D. and N. Artan, *Modeling of Activated Sludge Systems*, Technomic Publishing, Lancaster, PA, 1994.
30. Parkin, G. F. and P. L. McCarty, Sources of soluble organic nitrogen activated sludge effluents. *Journal, Water Pollution Control Federation* **53**:89–98, 1981.
31. Patterson, J. W., P. L. Brezonik, and H. D. Putnam, Sludge activity parameters and their application to toxicity measurements and activated sludge. *Proceedings of the 24th Industrial Waste Conference, 1969, Purdue University*, Engineering Extension Series No. 135, Purdue University, West Lafayette, Indiana, pp. 127–154, 1969.
32. Postgate, J. R., Viable counts and viability. *Methods in Microbiology* **1**:611–628, 1969.
33. Robinson, J. A. and W. G. Characklis, Simultaneous evaluation of V_{max}, K_m, and the rate of endogenous substrate production (R) from substrate depletion data. *Microbial Ecology* **10**:165–178, 1984.
34. SAS Institute, Inc., *SAS/STAT User's Guide*, Release 6.03 Edition, SAS Institute, Inc., Cary, North Carolina, 1988.
35. Servais, P., G. Billen, J. Martinez, and J. Vives-Rego, Estimating bacterial mortality by the disappearance of ^{3}H-labeled intracellular DNA. *FEMS Microbiology Ecology* **62**: 119–126, 1989.
36. Sollfrank, U. and W. Gujer, Characterization of domestic wastewater for mathematical modelling of the activated sludge process. *Water Science and Technology* **23**(4/6): 1057–1066, 1991.
37. Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Alexandria, Virginia, 1992.
38. Weddle, C. L. and D. Jenkins, The viability and activity of activated sludge. *Water Research* **5**:621–640, 1971.
39. Wentzel, M. C., A. Mbewe, and G. A. Ekama, Batch test for measurement of readily biodegradable COD and active organism concentrations in municipal waste waters. *Water SA* **21**:117–124, 1995.

This page intentionally left blank

Part III
Applications: Suspended Growth Reactors

Part I presents the fundamental principles upon which the design and evaluation of the biochemical operations used in wastewater treatment systems are based. These principles are then applied in Part II to the modeling of ideal suspended growth bioreactors. In Part III, we apply these principles to the practical design and operation of suspended growth biological wastewater treatment systems. Chapter 9 provides an overview of the design and evaluation of these systems, whereas the remaining chapters address specific suspended growth applications. Chapter 10 describes the design of activated sludge systems for the removal of biodegradable organic matter, the stabilization of particulate organic matter, and the oxidation of ammonia-N. The use of aerobic selectors to control the growth of certain types of filamentous bacteria is also considered. Chapter 11 addresses the design and operation of suspended growth biological nutrient removal systems. Single-sludge nitrogen removal, phosphorus removal, and combined nitrogen and phosphorus removal systems are considered, along with separate stage denitrification systems. The use of anoxic and anaerobic zones to control solids settleability is also addressed. The use of aerobic digestion to stabilize the waste solids (both primary and secondary) produced in the liquid process train of a wastewater treatment plant is the topic of Chapter 12. Conventional aerobic digestion systems are considered, along with anoxic/aerobic digestion systems and autothermal thermophilic aerobic digesters. Chapter 13 addresses the use of anaerobic processes for the treatment of high strength wastewaters and sludges. Both suspended growth and combined suspended and attached growth processes are considered. Finally, pond and lagoon systems are considered in Chapter 14. The environments in these systems are complex and deviate more than the environments in the other named biochemical operations from the ideal reactors considered in Part II. In spite of that, the fundamental principles developed in Parts I and II of this book can be applied to their design and operation.

This page intentionally left blank

9

Design and Evaluation of Suspended Growth Processes

Most of this book addresses the technical aspects of the design and evaluation of the biochemical operations used in wastewater treatment systems. This chapter, in contrast, addresses the process of designing and evaluating such operations, with particular emphasis on those that use suspended growth bioreactors. It has several purposes. First, it provides a transition between the fundamental principles presented in Parts I and II of this book and the detailed application of those principles to specific named biochemical operations in Part III. Second, it illustrates that design and evaluation of biochemical operations is iterative and provides a perspective on the typical steps involved. Third, it addresses the basic decisions that must be made to select among the various suspended growth biochemical operations, as well as those that are common to all of them. Finally, it contrasts the various levels of design and evaluation, from preliminary- to simulation-based. Even though this material is presented in the context of suspended growth bioreactors, much of it is also applicable to the design and evaluation of attached growth bioreactors. Consequently, the reader should also refer to this material while reading Part V.

The phrase "design and evaluation" is used here to reflect the range of tasks that biological process engineers must perform. In some instances, new facilities, or significant expansions of existing facilities, are needed to provide sufficient treatment capacity and/or capability. The term "design" refers to the process of determining the size and configuration of such new facilities. In other instances, a facility may already exist, but its treatment capacity and/or capability may not be known precisely. The term "evaluation" refers to the process of rationally determining that capacity and/or capability. The same process engineering principles are utilized in both situations.

9.1 GUIDING PRINCIPLES

Before investigating the design and evaluation of biochemical operations, it would be helpful to summarize the basic fundamental principles that arose in Parts I and II. These few essential principles, which provide the basis for all design and evaluation, are summarized in Table 9.1.

First, the biochemical environment imposed upon a bioreactor determines the nature of the microbial community that develops and the character of the biological reactions that they perform. If the engineer ensures that a high concentration of

Table 9.1 Guiding Principles for the Design and Evaluation of Suspended Growth Biochemical Operations

1. The biochemical environment determines the nature of the microbial community that develops in a bioreactor and the character of the reactions they perform.
2. The SRT is the most important design and control parameter available to the engineer.
3. A COD balance across a bioreactor provides valuable information about the amount of electron acceptor required and the amount of excess biomass produced.
4. The excess biomass production rate is essentially the same for all suspended growth systems with a given SRT and biochemical environment, regardless of the bioreactor configuration.
   a. The total mass of biomass in such systems will be the same, regardless of bioreactor configuration.
   b. The total mass of electron acceptor required for the removal of organic matter will be the same in such systems, regardless of bioreactor configuration, although the distribution of need will be different.
5. Only the mass of biomass in a bioreactor system is specified by the descriptive analytical expressions, not the concentration. The concentration is only specified after the bioreactor volume, or HRT, has been specified.

dissolved oxygen is maintained at all times, then organic substrates can be oxidized to carbon dioxide and water, providing energy for heterotrophic biomass growth and removing chemical oxygen demand (COD) from solution. In addition, ammonia-N can be oxidized to nitrate-N by autotrophic nitrifying bacteria, providing for their growth as well. The introduction of an anoxic zone in an otherwise aerobic bioreactor allows the nitrate-N formed by the autotrophic bacteria to be used as a terminal electron acceptor by facultative heterotrophic bacteria, converting it to nitrogen gas, thereby removing nitrogen from the wastewater. Furthermore, a properly positioned anaerobic zone in an otherwise aerobic bioreactor will allow phosphate accumulating organisms (PAOs) to effectively compete with ordinary heterotrophic bacteria for substrate, leading to a biomass enriched in phosphate. Wastage of that biomass removes phosphorus from the system. Finally, if the engineer provides a totally anaerobic environment in which neither oxygen nor nitrate-N is ever present, then an entirely different microbial community will develop in which methane is an important end product. The key point is that the engineer has control over the type of microbial community that may be present through the decisions that are made about the biochemical environment.

Second, the solids retention time (SRT) is the most important design and control parameter available to the engineer. This follows directly from its relationship to the specific growth rate of the biomass in the bioreactor, as reflected by Eq. 5.12 for a simple continuous stirred tank reactor (CSTR). Thus, while the biochemical environment provides the potential for growth of a given microbial population, the SRT, in concert with the bioreactor configuration, determines whether that potential will be realized. Furthermore, the SRT and the bioreactor configuration determine the extent of reaction in the system, thereby influencing the effluent substrate concentration, the excess biomass production rate, the rate at which electron acceptor must be provided, and the overall process performance.

Third, a COD balance across a bioreactor provides valuable information about the amount of electron acceptor required and the amount of excess biomass produced. As stated in Eq. 3.90, the COD removed in a bioreactor must equal the oxygen equivalents of terminal electron acceptor used plus the COD of biomass formed. Furthermore, at steady state, the amount of biomass formed is equal to the observed yield (Y_{obs}) times the amount of substrate used, as reflected by Eq. 5.31 for a simple CSTR. This means that the oxygen equivalent of terminal electron acceptor required is just equal to $1 - Y_{obs}$ times the amount of substrate used, as stated by Eq. 5.35. Thus, it can be seen that this simple balance is very useful for making initial estimates of excess biomass production and electron acceptor requirements.

Fourth, the excess biomass production rate is essentially the same for all suspended growth systems with a given SRT and biochemical environment, regardless of the bioreactor configuration. This was seen in the simulations of the various systems in Chapter 7 and has two important implications. First, the total mass of biomass in those systems will be the same, regardless of the system configuration. This means that expressions derived for a single CSTR, which can be solved analytically, can be used to estimate the total mass of biomass or mixed liquor suspended solids (MLSS) in a complex bioreactor system for which analytical solutions are not possible. Second, the total amount of electron acceptor required for removal of organic matter will be independent of the bioreactor configuration, although the distribution of the electron acceptor will not be. The same statement cannot be made about the amount of oxygen required for nitrification, however, because the extent of nitrification is much more dependent on the bioreactor configuration than is the extent of COD removal, as seen in Chapter 7. Nevertheless, these are very powerful tools, especially for preliminary designs.

Finally, Eq. 5.20 demonstrated that only the mass of biomass in a bioreactor system is specified by the descriptive analytical equations, not the concentration. Rather, the concentration is only specified after the bioreactor volume, or the hydraulic residence time (HRT), has been specified. The same is true for the MLSS as well. From a design and analysis perspective, this means that a designer has one free design variable that may be freely chosen, within reasonable limits. A large part of the design process is concerned with that choice because it strongly influences the potential interactions among the various unit operations that must be considered during system design.

9.2 ITERATIVE NATURE OF PROCESS DESIGN AND EVALUATION

The design of biological wastewater treatment systems is typically an iterative process. There are two reasons for this: (1) the definition of the problem to be solved evolves throughout the design, and (2) the database upon which the design is based improves as additional investigations are completed. Nevertheless, a "freeze" point must be reached during any project where the solution to the problem is fixed and then implemented. The steps which lead to this point are evolutionary in nature, with the problem statement continuously being redefined and potential solutions being evaluated and discarded until the best solution is selected. Most designs begin with an initial concept that is rather general in nature. For example, during early discus-

sions for a design, a decision may be made to treat a particular wastewater in an activated sludge system, even though the size of the system, its specific configuration, and the nature of any special features are unknown. Those questions, and many more, are addressed as the design proceeds, resulting in an ever more refined estimate of the required facilities. The increasing database that develops during the design process also allows refinement of the design. Typically the designer's understanding of the strength and nature of the wastewater, the characteristics of the treatment system, the effluent discharge standards, and the needs and desires of the treatment system owner and operator evolve as the project progresses.

Figure 9.1 illustrates this iterative nature of process design. The first step is to define the project objectives and requirements, allowing identification of the most reasonable potential solutions based on the current state of knowledge. The costs and scope of those potential solutions are then estimated using rough calculations of bioreactor size, oxygen requirements, solids wastage, etc. Next, the potential advantages and disadvantages of the alternative solutions are considered and a decision is made whether to more fully evaluate each one. When the potential advantages of a particular alternative solution are not sufficient to warrant further consideration, it is dropped and more study is devoted to those remaining. Each iteration around the loop results in more refined information, which allows better estimates to be made of the sizes and costs of the alternatives. Consequently, the use of more refined techniques is called for. This same logic can be applied to the refinement of a selected alternative. Additional studies to refine a given alternative are conducted only as long as the benefits derived from them outweigh their costs.

The iterative nature of process design and evaluation makes it clear that several levels of refinement are required. To begin the design process, a preliminary assessment must be made based on limited data. In spite of its preliminary nature, this assessment must be conceptually sound because important decisions will be based on it. In some instances, the preliminary assessment may be sufficiently precise to allow the project to proceed directly to implementation. This will usually occur for smaller projects where the cost of a conservative design is small compared to the cost of refining the estimates of facility requirements. It also occurs frequently for applications where significant experience already exists upon which to base the selection of the preliminary process design parameters. In other instances, little experience may exist with the subject wastewater or proposed treatment system, making the initial preliminary assessment quite uncertain. In that case, treatability studies as outlined in Chapter 8 must be performed, leading to parameters that may be used in models. In some cases, particularly for the activated sludge process, it may be possible to base design decisions on the simple stoichiometric model of Chapter 5 by incorporating a few broadening assumptions, such as grouping together slowly and readily biodegradable substrate. In other cases, such as large nutrient removal projects, even a small amount of uncertainty can result in significant over-expenditures. In those situations, additional testing to quantify the parameters in activated sludge model (ASM) No. 1 or No. 2 may be merited, allowing alternative designs to be considered by simulation, thereby reducing uncertainty. In the next section, we will consider the decisions that must be made in any design situation, regardless of its level. Then, in Section 9.4 we will consider the approaches used in the various levels of design.

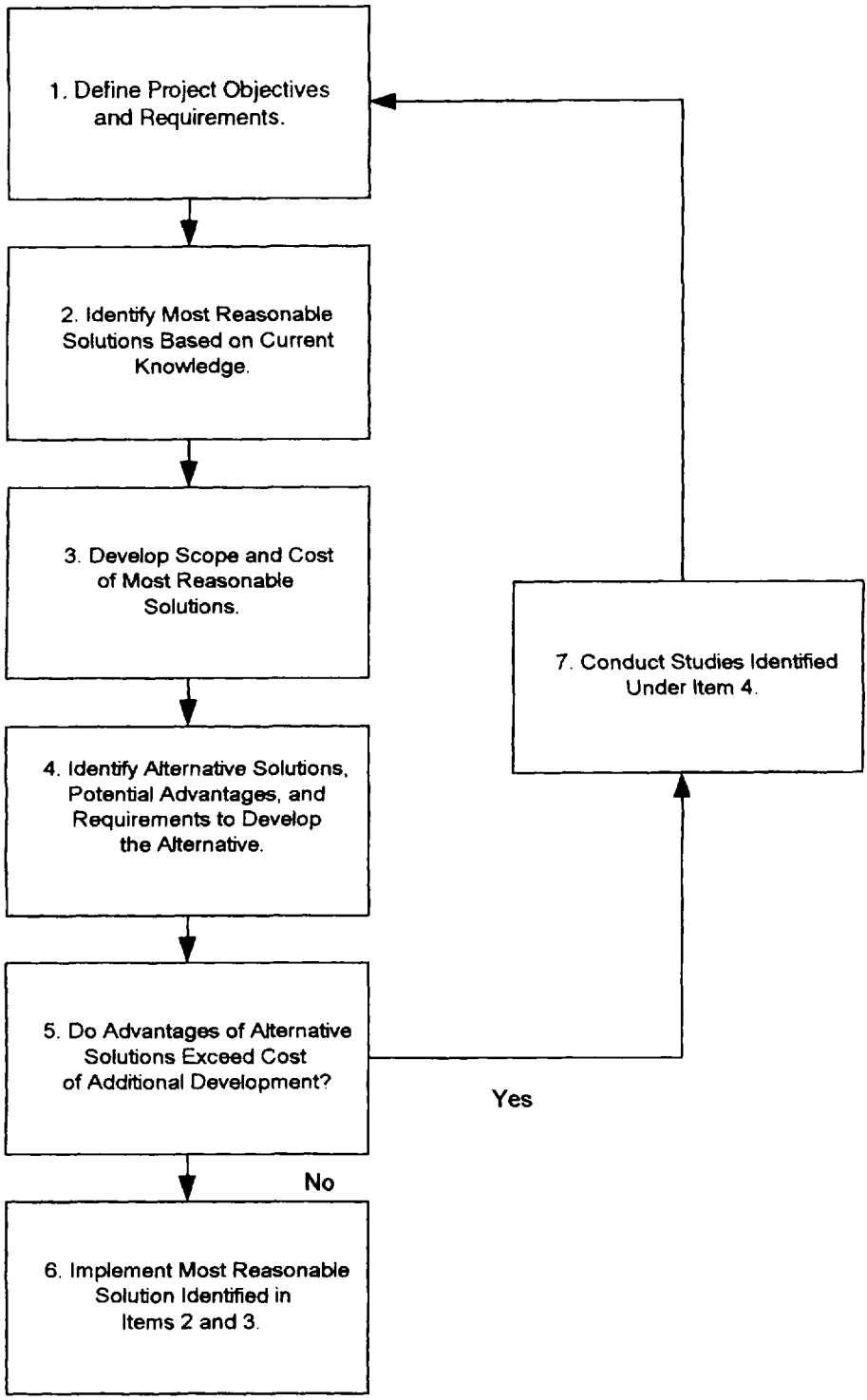


Figure 9.1 The iterative nature of process design and evaluation.

9.3 BASIC DECISIONS DURING DESIGN AND EVALUATION

Design of a biochemical operation requires that decisions be made that are consistent with the guiding principles summarized in Section 9.1. Some of those decisions establish the nature of the facility, whereas others determine its size. Since the biochemical environment has a profound effect, its choice is one of the earliest decisions that must be made. Then the SRT is chosen, determining the various factors that follow from it, such as the mass of biomass, electron acceptor requirement, etc. Finally, the interrelationships between the bioreactor and the other unit operations in the system must be considered.

9.3.1 Biochemical Environment

One of the fundamental decisions faced by a designer is whether to use an aerobic/anoxic or an anaerobic operation. As discussed in Section 2.3.1, in aerobic/anoxic operations, heterotrophic bacteria use oxygen or nitrate-N as their terminal electron acceptor while using biodegradable organic matter as an energy and carbon source for growth. Furthermore, the presence of dissolved oxygen in such systems allows for the growth of autotrophic nitrifiers, which use ammonia-N as an electron donor, producing nitrate-N. In contrast, as discussed in Section 2.3.2, when both dissolved oxygen and nitrate-N are absent, alternative electron acceptors must be used. In fermentative systems, the biodegradable organic matter itself serves as the terminal electron acceptor, yielding soluble fermentation products, whereas in methanogenic systems carbon dioxide is the major acceptor, yielding methane.

Table 9.2 compares the features of aerobic/anoxic and anaerobic wastewater treatment systems. Both systems are capable of achieving high organic removal efficiencies. However, the effluent quality from an aerobic/anoxic system will generally be excellent while that from an anaerobic system will be moderate to poor. Aerobic and anoxic conditions allow extensive removal of biodegradable organic matter, particularly soluble material. In addition, the biomass in aerobic/anoxic systems is generally well flocculated, resulting in low effluent suspended solids concentrations. In contrast, although a high percentage of the biodegradable organic matter is converted

Table 9.2 Comparison of Aerobic/Anoxic and Anaerobic Systems

	System	
Feature	Aerobic/anoxic	Anaerobic
Organic removal efficiency	High	High
Effluent quality	Excellent	Moderate to poor
Sludge production	High	Low
Nutrient requirements	High	Low
Energy requirements	High	Low to moderate
Temperature sensitivity	Low	High
Methane production	No	Yes
Nutrient removal	Possible	Negligible

to methane and carbon dioxide in anaerobic systems, the resulting concentrations of soluble biodegradable organic matter can still be relatively high and the produced solids may be poorly flocculated. As a result, the quality of the effluent from an anaerobic system does not generally equal that from an aerobic system.

Waste solids production is high in aerobic/anoxic systems due to the large amount of energy made available for the synthesis of new biomass, resulting in relatively high yield values. Consequently, nutrient requirements are also high. In contrast, the biomass production and associated nutrient requirements for anaerobic systems are low because the relatively small amount of available energy makes the yield low. Power requirements for aerobic systems are high because oxygen must be transferred to serve as the electron acceptor, although this need will be reduced when anoxic zones are present. In contrast, the power requirements for anaerobic systems are low to moderate and generally represent the energy required to heat and mix the bioreactor. Heating requirements can be significant, but energy for heating is typically provided by the methane produced. Temperature control is critical in anaerobic systems because the methanogens are quite sensitive to changes in temperature. The performance of aerobic systems, on the other hand, is much less sensitive to changes in temperature. Finally, removal of nitrogen and phosphorus is possible in aerobic/anoxic systems, whereas nutrient removal is negligible in anaerobic systems.

These features combine to provide advantages to aerobic systems for the treatment of low strength wastewaters and to anaerobic systems for the treatment of high strength wastewaters. Figure 9.2 presents the wastewater concentration ranges over which aerobic/anoxic and anaerobic bioreactors are typically applied and the ranges of HRT typically required. Both ranges are approximate and are provided only as general descriptors. The HRT range reflects both the range of SRTs required and the degree of separation between the SRT and the HRT achieved with each technology. Due to their ability to produce high quality effluents, aerobic/anoxic systems are typically used for wastewaters with biodegradable COD concentrations less than 1,000 mg/L. Although anaerobic systems can be applied to treat wastewaters in this concentration range, the effluent quality will generally not meet discharge standards, thereby requiring aerobic polishing. However, the combination of an anaerobic system followed by an aerobic system is usually not economical compared to a fully aerobic system for these wastewaters. In addition, low strength wastewaters typically result in insufficient methane production to heat the wastewater to the optimum temperature. Both aerobic/anoxic and anaerobic systems are used to treat wastewaters with biodegradable COD concentrations between 1,000 and 4,000 mg/L. Again, aerobic polishing of the anaerobic process effluent will be required if high quality is needed. Finally, in many instances the advantages of anaerobic systems outweigh the advantages of aerobic/anoxic systems for the treatment of wastewaters with biodegradable COD concentrations over 4,000 mg/L. The typical operating ranges for various anaerobic treatment systems are also presented in Figure 9.2. In general, low rate and high rate anaerobic systems employ biomass recycle to increase the SRT relative to the HRT, whereas anaerobic digestion systems do not, making the SRT and HRT identical.

Additional factors will also influence the relative economics of aerobic/anoxic versus anaerobic systems. Consequently, investigations must be conducted to distinguish the relative advantages and disadvantages of each biochemical environment for wastewaters with biodegradable COD concentrations near the overlap region in

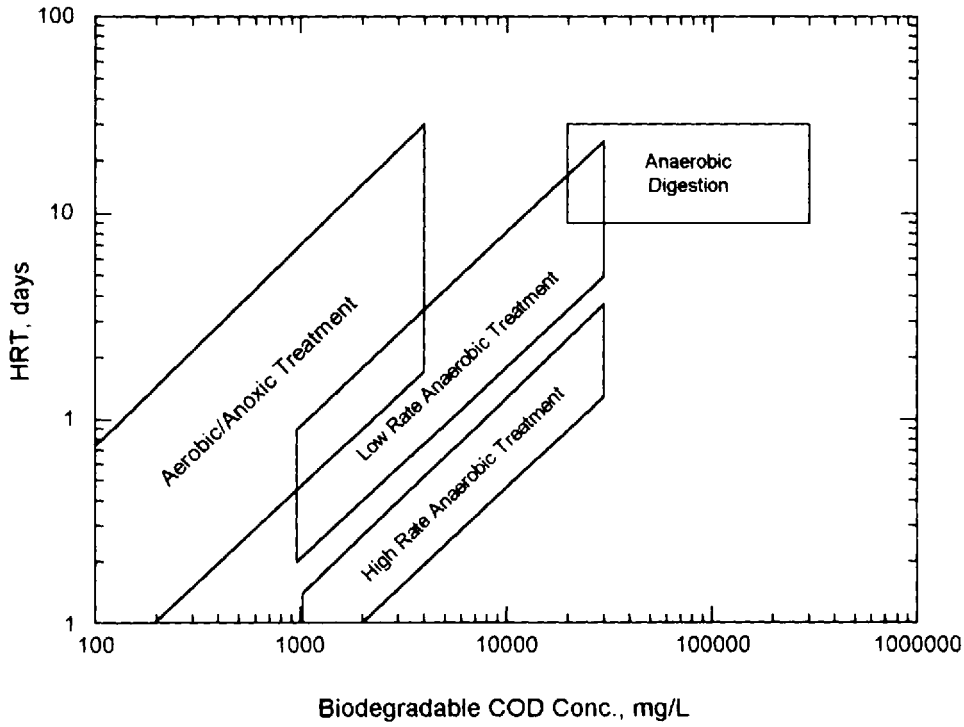


Figure 9.2 Typical operating ranges for aerobic/anoxic and anaerobic suspended growth biochemical operations. (Adapted from Hall[5]).

Figure 9.2. These include the wastewater temperature, flow rate, and composition, and they will be discussed in greater detail in Chapters 10, 11, and 13. Nevertheless, Figure 9.2 can be used for preliminary screening of biological wastewater treatment options.

9.3.2 Solids Retention Time

As illustrated in Part II of this book, the SRT exerts a dominant effect on the capabilities and performance of a biochemical operation. For example, it affects the types of microorganisms that can grow in a bioreactor, as well as their activity, thereby determining effluent quality. Because of the multiple effects associated with SRT, many factors must be considered during its selection. In fact, it is seldom possible to select the SRT based on a single criterion, such as effluent substrate concentration. The range of typical SRT values is already known for many applications, and often an appropriate SRT can be selected based on experience. In this section we will consider such situations.

Before addressing appropriate SRT values, it should be emphasized that the selected SRT must always exceed the minimum SRT associated with the microorganisms responsible for a particular required biochemical transformation. As discussed in Section 5.1.3, the minimum SRT is the value below which a particular group of microorganisms is unable to grow in a suspended growth bioreactor. As

expressed in Eq. 5.16, it is a function of the influent concentration of the limiting substrate for the microorganisms of interest and the kinetic parameters describing their growth on that substrate. The kinetic parameter that exerts the most pronounced effect is $\hat{\mu}$. Since the $\hat{\mu}$ value for heterotrophs growing on readily biodegradable substrate is high, the minimum SRT for them is very low. In contrast, because the $\hat{\mu}$ value for autotrophic nitrifying bacteria is very low, the minimum SRT associated with them may be quite high. The same may be true for heterotrophs growing on xenobiotic chemicals. If the SRT is maintained at a value less than the minimum SRT for the subject bacteria, they will be wasted from the bioreactor faster than they grow and a stable population will not develop. In other words, washout occurs, as discussed in Section 5.1.3. Conversely, if the operating SRT exceeds the minimum SRT, then the subject bacteria will be able to grow in the process and the reaction will occur. However, as seen in Chapters 5 and 7, the degree of conversion will depend on the operating SRT and the bioreactor configuration, and both must be chosen to meet effluent quality goals. This generally requires the operating SRT to be well above the minimum SRT. The ratio of the operating SRT to the minimum SRT is called the safety factor. To ensure that their washout does not occur, the safety factor for the most slowly growing microorganisms required in a bioreactor should always exceed 1.5, although larger values may be required in some circumstances. Furthermore, larger values may result when other factors control the choice of the SRT. Factors affecting the choice of the SRT for various named biochemical operations will be discussed in subsequent chapters. Only a brief overview is given here.

Aerobic/Anoxic Systems. Figure 9.3 illustrates the ranges of operating SRTs over which various events will occur in aerobic/anoxic systems. Because the ranges

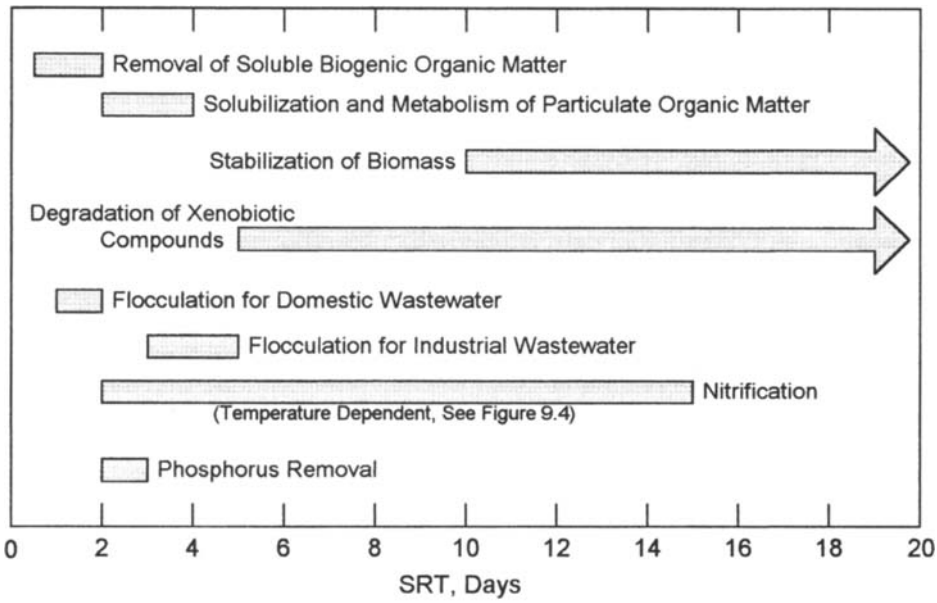


Figure 9.3 Typical SRT ranges for various biochemical conversions in aerobic/anoxic bioreactor systems at 20°C.

represent operating SRTs, the lower limits reflect the application of typical safety factors to the minimum SRTs associated with the microorganisms responsible for a particular event. The upper limits reflect SRT values above which little additional reaction occurs in a CSTR. Because heterotrophs can grow under both aerobic and anoxic conditions, the SRT values in Figure 9.3 can be thought of as any combination of aerobic and anoxic SRT values. On the other hand, both nitrifying bacteria and PAOs can grow only under aerobic conditions. Consequently, the SRT values in Figure 9.3 should be thought of as aerobic SRTs when considering nitrification and phosphorus removal.

The first thing to be noted in Figure 9.3 is that removal of biogenic soluble organic matter occurs at low SRTs, typically over a range of about 0.5 to 1.5 days. For SRT values in excess of this range, the degradation of soluble organic matter will be essentially complete. This follows from the fact that the $\hat{\mu}$ values for heterotrophic bacteria growing under aerobic/anoxic conditions on such substrates are relatively high, as discussed in Section 3.2.10. Furthermore, in municipal wastewater treatment systems, bacterial growth and substrate removal are assisted by the presence of microorganisms in the influent wastewater, which prevents washout, as discussed in Section 5.2.3. The solubilization and metabolism of particulate organic matter typically occurs over an SRT range of 2 to 4 days, with degradation being essentially complete at longer SRT values. Stabilization of biomass through decay and similar reactions will occur over a broad range of SRTs, but is generally not thought to be significant for SRT values less than about 10 days, as illustrated in Figure 5.10. Increasing stabilization is obtained as the SRT is increased beyond 10 days. Relatively long SRT values are often required to biodegrade xenobiotic compounds. As a general rule, the SRT should be at least 5 days to biodegrade some of these materials, but often an SRT in excess of 10 days is necessary for complete biodegradation.

The SRT must also be sufficiently long to allow flocculent growth of heterotrophic bacteria because biomass separation and recycle in suspended growth systems requires such a condition. A wide range of values has been reported in the literature, but practical experience indicates that when domestic wastewater is being treated, flocculation can be obtained at SRT values as short as 1.0 day. For industrial wastewaters, however, longer values may be required, typically ranging from 3 to 5 days. This difference may be due to differences in the nature of the substrates in the two types of wastewater, or to the higher concentration of bacteria typically present in domestic wastewater. A complicating factor is the growth of filamentous bacteria, which may be exacerbated by the use of low SRT values. Because of their importance, the topics of bioflocculation and filamentous organism growth will be discussed in detail in Section 10.2.1. For the present time, however, it is only necessary to recognize that a minimum SRT value exists below which bioflocculation will not occur and that the minimum value is a function of wastewater type and other factors.

As discussed in Section 3.2.10, the autotrophic nitrifying bacteria have lower $\hat{\mu}$ values than most heterotrophic bacteria, and they require a longer SRT to survive in an aerobic bioreactor, as reflected in Figure 9.3. In addition, their maximum specific growth rate coefficient is more sensitive to changes in temperature than that of heterotrophs. Consequently, a broad range of SRTs has been shown for nitrification. Denitrification is accomplished by heterotrophic bacteria, which have relatively high $\hat{\mu}$ values, allowing them to grow at relatively low SRTs if nitrate-N is present in the

influent wastewater. However, since most wastewaters contain ammonia-N rather than nitrate-N, a relatively long SRT is required for nitrification/denitrification systems to allow production of nitrate-N by the relatively slow growing nitrifiers.

The range of aerobic SRT values required for growth of the PAOs utilized in biological phosphorus removal and anaerobic selector systems is also presented in Figure 9.3. The $\hat{\mu}$ values for the PAOs are lower than those of ordinary heterotrophic bacteria, which are not capable of accumulating phosphorus. Consequently, the lower limit on the SRT for phosphorus removal is generally higher than that for soluble substrate removal.[13] On the other hand, the lower limit on the SRT for phosphorus removal is similar to the lower limit for nitrification, suggesting that it may be difficult to operate a bioreactor for phosphorus removal without experiencing the problems caused by the presence of nitrate, as discussed in Section 7.7.2.

Additional insight into the growth characteristics of nitrifying bacteria and PAOs can be gained by examining Figure 9.4, where the effects of temperature on the minimum SRT for each type of bacteria are shown. The values in the figure are computed values using typical kinetic parameter values and temperature correction coefficients.[9,12] It should be emphasized that the curves are for the minimum SRTs at which washout will occur, not operating SRTs. No safety factors are included. Examination of Figure 9.4 reveals that at temperatures below 20°C the minimum SRT for PAOs is sufficiently smaller than the minimum SRT for nitrifiers to allow operation of a phosphorus removal system without nitrification occurring, but that at temperatures above 24°C, this would be extremely difficult to do.

Two final points should be made about Figure 9.3. First, the design SRT must reflect the limiting event that is required in the system. For example, even though an SRT as low as 0.5 days would be sufficient to remove soluble organic matter, it

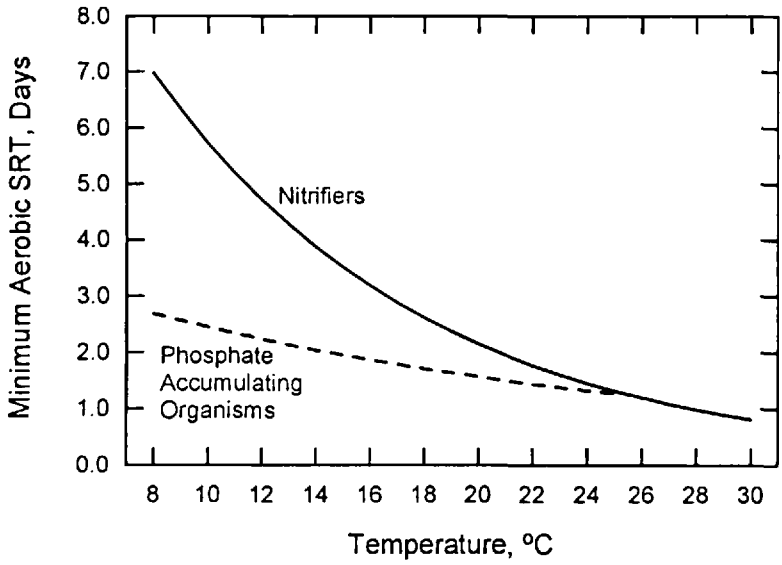


Figure 9.4 Effect of temperature on the minimum aerobic SRT required to grow nitrifiers and PAOs. The nitrifier curve was adapted from Sedlack[12] and the PAO curve was developed from data presented by Mamais and Jenkins.[9]

could not be used as the design SRT for a system using sedimentation to remove the biomass because it would not be sufficient to allow bioflocculation to occur. Rather, the SRT would have to be above one or two days to allow flocculation during treatment of domestic wastewater and above three to five days for an industrial wastewater. Second, any event that can occur at the chosen SRT will occur, provided the environmental conditions are adequate. For example, if an SRT of 15 days were used to achieve stabilization of biomass, provision should be made for nitrification since the SRT is long enough for it to occur. Figure 9.3 can be very helpful as a reminder of what is likely to occur at various SRTs, thereby helping a designer to consider all possible events.

Anaerobic Systems. A similar approach can be used for anaerobic systems. Figure 2.3 depicts the biochemical conversions that occur in such systems and Figure 9.5 indicates the SRT ranges over which they occur at 35°C. Longer SRT values will generally be required for lower temperatures.

Hydrolysis of particulate carbohydrates and proteins to produce simple sugars and amino acids is a relatively rapid reaction, which is essentially complete for SRT values in excess of about 3 days. In contrast, the hydrolysis of lipids to form long chain fatty acids and other soluble reaction products is a much slower reaction that does not generally occur for SRT values less than about 6 days. A significant difference also exists for the various acidogenic reactions that convert the hydrolysis products into acetic acid and hydrogen. Fermentation of amino acids and simple sugars occurs very rapidly and, generally, will not be rate limiting. In contrast, the anaerobic oxidation of fatty acids to acetic acid and hydrogen is much slower. The oxidation of propionic acid is particularly slow in comparison to the other anaerobic oxidations.

Various methanogenic reactions are also possible, depending on the SRT. Hydrogen oxidizing methanogens can grow quite rapidly; a complete population of such

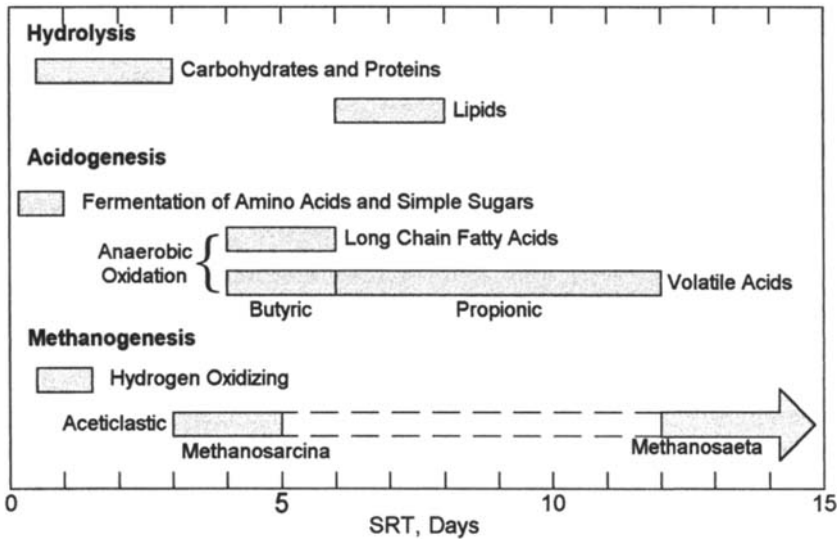


Figure 9.5 Typical SRT ranges for various biochemical conversions in anaerobic bioreactor systems at 35°C.

organisms will generally develop for SRT values in excess of about 1.5 days. In contrast, the aceticlastic methanogens grow much more slowly, and there are significant differences in the maximum specific growth rate coefficients of the two major types. *Methanosarcina* grows relatively rapidly, and a complete population will be available at SRTs in excess of about 5 days. In contrast, *Methanosaeta* grows relatively slowly and will not generally be present unless the SRT is in excess of about 12 days.

Analysis of the relative growth characteristics of the various anaerobic microorganisms results in some important observations. They are summarized in the three paragraphs that follow.

1. A relatively low SRT must be maintained if an anaerobic process is to achieve acidogenesis without significant methanogenesis. Hydrolysis of carbohydrates and proteins will generally be complete at SRTs of about 3 days, and the simple sugars and amino acids produced will be converted into acetic acid, other volatile acids, and hydrogen. Hydrogen oxidizing methanogens are capable of growing at these SRTs, so much of the hydrogen produced will be converted into methane. The quantity will be small, however, since only limited amounts of hydrogen are produced through fermentation reactions. The volatile acids will accumulate under these conditions because longer SRTs are required for growth of the acetogenic bacteria that convert them to acetic acid and hydrogen by anaerobic oxidation. An SRT of less than 3 days must be maintained to prevent the growth of *Methanosarcina*, which would convert acetic acid to methane and carbon dioxide. Such a short SRT will not be sufficient for hydrolysis of lipids, so they will remain unreacted.

2. Anaerobic treatment of a wastewater containing carbohydrate and protein with production of methane can be accomplished at SRT values of about 8 days. At this SRT, the carbohydrates and proteins will be hydrolyzed; the hydrolysis products will be converted by fermentation and anaerobic oxidation into acetic acid, carbon dioxide, and hydrogen; and the acetic acid and hydrogen will be utilized for methane production. In fact, significant methane formation will occur at SRT values as low as 5 to 6 days, but significant quantities of propionic acid will accumulate because this SRT is too short to allow the growth of bacteria which anaerobically oxidize propionic acid to acetic acid and hydrogen.

3. SRT values in excess of 8 days will be required to stabilize wastewaters containing significant quantities of lipids, such as primary sludges from domestic treatment systems. Generally, a minimum SRT of about 10 days is specified to ensure complete and reliable degradation of lipids in anaerobic bioreactors.

The performance of anaerobic bioreactors is affected by many factors in addition to the SRT, such as temperature, pH, and the presence of toxic materials. In addition, although anaerobic systems have demonstrated the ability to degrade xenobiotic materials, relatively long SRTs are often required. Nevertheless, the information presented in Figure 9.5 illustrates the relative effects of SRT on the growth of the various types of microorganisms found in anaerobic bioreactors and the resulting impact on the types of biochemical conversions that will occur.

9.3.3 Items from Process Stoichiometry

As seen from the guiding principles summarized in Section 9.1, once the biochemical environment and the SRT have been selected, a number of important items follow

directly from the stoichiometry of biomass growth and substrate utilization. First, for a given wastewater flow rate and concentration, the mass of biomass in the system is fixed. Consequently, the mass of MLSS is also fixed. Since only the mass of MLSS is specified, the designer may freely choose either the MLSS concentration or the bioreactor volume, thereby fixing the other. The values chosen will depend on the nature of the biochemical operation, the bioreactor configuration, and constraints that consider the interactions between the various unit operations in the system. Consideration of those factors is an important component of system design and will be discussed in detail in the chapters to follow.

Another item that is determined by the process stoichiometry is the mass rate at which solids must be wasted from the bioreactor system to maintain the desired SRT. It is very significant because it will be used to size the solids processing system for the facility.

The process stoichiometry also determines the quantity of electron acceptor required. This follows directly from the mass of biodegradable COD entering the bioreactor system and the mass of solids wasted from the system, as expressed by the COD balance that has been stressed throughout this book. If the biochemical environment is aerobic, the electron acceptor will be oxygen, and the calculated oxygen requirement will be central to the sizing of the oxygen transfer system. On the other hand, if the system is aerobic/anoxic, the computation will fix the sum of the oxygen and nitrate requirements, but additional information will be needed to determine the relative amounts of the two electron acceptors required. Finally, if the system is anaerobic, the electron acceptor requirement can be translated directly into the methane production rate, which can be used to estimate the quantity of energy available for use within the facility.

Finally, stoichiometry can be used to estimate the nutrient requirements for the bioreactor, as discussed in Sections 3.8.2 and 5.1.4. While it will seldom be necessary to add nutrients to bioreactors treating domestic wastewaters, many industrial wastewaters lack sufficient quantities of one or more macro- or micronutrients, and thus they must be added for successful operation of a bioreactor. Appropriate planning for such additions is an important component of design.

The quantitative information listed above can be developed with various degrees of precision, depending on the nature of the information available and the use to which it will be put. In Section 9.4 we will examine the various levels of design and evaluation, and their appropriate use.

9.3.4 Interactions Among Decisions

We have seen above that the designer can freely choose either the MLSS concentration or the bioreactor volume once the mass of MLSS in the system has been determined. While that choice is a "free" one in the sense that no equation specifies it, it cannot be made in isolation. Rather, it must be made with consideration of its impact on other unit operations in the system. Two such interactions are particularly important: those with the downstream biomass separation device, typically a gravity clarifier, and those with the mixing and/or aeration system.

For a given application, a wide range of MLSS concentrations can be used. Thus, it would appear that from an economic standpoint, the most cost-effective design would use the highest possible MLSS concentration, resulting in the smallest

possible bioreactor. However, this overlooks the fact that higher MLSS concentrations require the use of larger clarifiers to separate the applied solids and concentrate them for return to the upstream bioreactor. This occurs because the clarifier must be sized to avoid thickening failure, which is a function of the settleability of the solids, the applied MLSS concentration, and the recycle flow rate from the clarifier to the bioreactor. Thus, a trade-off exists between the use of a small bioreactor with a large secondary clarifier, and vice versa. If the settleability of the solids is known, then the trade-off between bioreactor and clarifier size can be analyzed by formulating a variety of feasible options and determining the most cost-effective one.[4] During preliminary design, however, the settling characteristics of the solids are seldom known. Fortunately, correlations have been developed between various solids settleability indices and the settling characteristics of activated sludge,[1,8] and these can be used to quantify expected settling characteristics. These relationships are the basis for secondary clarifier operating diagrams that can be used to select appropriate clarifier areas and recycle flow rates for preliminary design purposes.[1] A complete explanation of those diagrams and their use requires knowledge of solids flux theory,[8] which is beyond the scope of this chapter. Nevertheless, the reader should be aware of the existence of the diagrams and the fact that they can be used for both design and evaluation of final clarifiers associated with bioreactors.

If the proposed bioreactor is aerobic, provisions must be made to transfer the needed oxygen to the liquid phase for use by the biomass. In addition, regardless of the nature of the biochemical environment, provision must be made for keeping the MLSS in suspension without subjecting it to so much shear that it will not flocculate and settle properly. Generally, for economic reasons, in aerobic bioreactors the same equipment is used to transfer oxygen and to keep the MLSS in suspension. This results in certain constraints that must be considered during design and evaluation. Figure 9.6 illustrates those constraints.

The volumetric power input to a bioreactor is the power applied per unit volume for mixing and/or oxygen transfer, regardless of whether it comes from a mechanical mixer or from the movement of air discharged into the bioreactor. A minimum volumetric power input is required to provide the turbulence needed to keep solids in suspension. Its value depends on the type of biochemical operation being used, and appropriate values will be provided in subsequent chapters. Nevertheless, it is shown schematically in the figure as the lower horizontal line labeled, "limit for adequate mixing." Conversely, a maximum volumetric power input exists above which excessive floc shear will occur, making it very difficult to achieve adequate clarification of the treated wastewater. As with the mixing constraint, its value depends on the type of biochemical operation under consideration, but for illustrative purposes it is shown in the figure as the upper horizontal line labeled, "limit for floc shear."

The amount of oxygen required by biomass treating a given wastewater is fixed once the SRT of the system is fixed. For a given type of aeration device, transfer of that quantity of oxygen will require a total power input that is essentially independent of the volume of the bioreactor. Consequently, the larger the bioreactor, the smaller the volumetric power input for oxygen transfer, as illustrated by the curve in Figure 9.6. The intersections of that curve with the two horizontal limit lines define the upper and lower feasible bioreactor volumes, V_U and V_L, respectively. If the bioreactor volume were greater than V_U, the total power input to the bioreactor, as

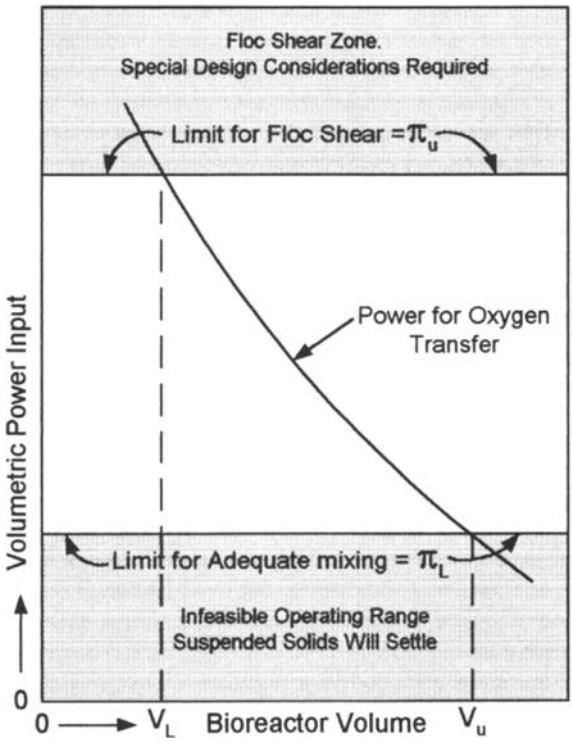


Figure 9.6 Effect of bioreactor volume on the volumetric power input required for oxygen transfer and its relationship to solids suspension and floc shear.

determined by the product of the volumetric power input limit for mixing times the bioreactor volume, would be greater than the power input needed for oxygen transfer. This is inefficient and wastes power. On the other hand, if the bioreactor volume were less than V_1, the biomass would be subjected to excessive shear forces, which will disrupt flocculation and clarification. Thus, when decisions are being made about the MLSS concentration and associated bioreactor volume, consideration should be given to the impact of the bioreactor volume on the volumetric power input. The quantitative aspects of this consideration will be presented in the chapters to follow.

9.4 LEVELS OF DESIGN AND EVALUATION

The iterative nature of process design and evaluation was discussed in Section 9.2. One consequence of it is that the level of detail required will change as the design progresses. We have described elsewhere how a series of process models can be applied at various stages in a project to provide a consistent and increasingly precise estimate of biological process requirements.[3] In this section we will present the essential elements required for each level of design.

9.4.1 Preliminary Design and Evaluation Based on Guiding Principles

In many instances, aerobic/anoxic and anaerobic systems can be sized on an approximate basis using the guiding principles set forth in Section 9.1. The resulting preliminary process design and evaluation provides an initial assessment of the capacity and capability of an existing biological treatment facility and of the changes required to expand it. It can also provide initial information about the nature of a proposed new facility. These preliminary estimates allow development of a preliminary scope and cost estimate for the project of concern which can subsequently be refined as further information is gathered.

After a decision has been made about the biochemical environment, information such as that provided in Figures 9.3 and 9.5 can be used to select the SRT. Both of these decisions require a degree of experience with the type of wastewater under consideration but the factors discussed in Sections 9.3.1 and 9.3.2 can provide guidance during the decision making process. In addition, the information in Figure 9.4 can be used in conjunction with Figure 9.3 to provide additional insight into the circumstances under which nitrification is likely to occur. Such an application is illustrated in the following example.

Example 9.4.1.1

Consider the design of an activated sludge system to treat a readily biodegradable industrial wastewater. The discharge standards require removal of biodegradable organic matter, but no limits have been set on effluent ammonia-N, total nitrogen, or phosphorus.

a. What SRT would be selected to provide reliable treatment and a high quality effluent while also minimizing the size of the bioreactor?
From Figure 9.3, it can be seen that an SRT is excess of 2 days is required to provide essentially complete removal of soluble, biodegradable organic matter. However, because the facility is to treat an industrial wastewater, an SRT of 3 to 5 days should be used to achieve good flocculation. Consequently, select an SRT of 5 days as a conservative measure for preliminary process sizing.

b. Will the system nitrify at the selected SRT?
Figure 9.3 suggests that nitrification will occur at an SRT of 5 days, provided that nothing in the wastewater inhibits the nitrifying bacteria. However, to gain more insight, the effects of temperature should be considered by using Figure 9.4. Nitrification will occur as long as the bioreactor SRT is above the minimum SRT for the nitrifying bacteria. Entering Figure 9.4 with the design SRT of 5 days reveals that nitrification will occur down to a temperature of about 12°C. At that temperature, however, the nitrifying bacteria will be on the verge of washing out. As a result, nitrification will not be stable. However, the warmer the temperature, the greater the design SRT will be relative the minimum SRT, and the more stable nitrification will be. This information can be combined with information about the annual variations in temperature to make a decision about when nitrification is likely to occur.

c. Could enhanced biological phosphorus removal capabilities be incorporated into the design of this facility without increasing the design SRT, if the need were to arise in the future?
Figure 9.3 indicates that the growth of PAOs requires an SRT of at least 2 to 3 days. Since the design SRT of 5 days exceeds this value, growth of phosphorus removing bacteria is possible. Furthermore, examination of Figure 9.4 reveals that it should be possible to sustain biological phosphorus removal down to quite low temperatures.

After selection of the SRT, estimates must be made of the various items related to process stoichiometry, as discussed in Section 9.3.3. This, too, can be done on the basis of the guiding principles. An important principle articulated in Section 9.1 is that for a given SRT, the mass of biomass in a bioreactor is fixed for a given treatment situation, i.e., flow and concentration of wastewater. This follows directly from combining Eqs. 5.20 and 5.28, and invoking the definition of the HRT:

$$X_{B,H} \cdot V = \Theta_c Y_{Hobs} F(S_{SO} - S_S) \tag{9.1}$$

The concept embodied in Eq. 9.1 can be carried another step, thereby giving a simple equation with general utility for preliminary process design. For most domestic wastewaters, both inert and biodegradable organic matter will be present in the influent, requiring us to be concerned about the MLSS concentration, rather than just the active biomass concentration. We saw in Section 5.2 that the simple model of Chapter 5 could be extended to include inert particulate organic matter and in Section 8.5.2 that it could even incorporate slowly biodegradable substrate, provided that the SRT was sufficiently long. Because of that, and recognizing that for the SRTs used in practice, $S_S \ll S_{SO}$, we can write an approximate equation that follows directly from Eq. 9.1 and the fundamental principle it embodies:

$$X_M \cdot V = \Theta_c Y_n F(S_{SO} + X_{SO}) \tag{9.2}$$

Examination of Eq. 9.2 reveals that it differs from Eq. 9.1 in three important ways. First, it expresses the MLSS concentration, X_M, and includes both readily and slowly biodegradable substrate, as discussed above. The most important difference is in the nature of the yield coefficient, however. Y_n is called the net process yield and it differs from the observed yield in that it incorporates both inert organic matter and undegraded slowly biodegradable organic matter into its value. In Eq. 5.28 we saw that Y_{Hobs} accounts for the effect of biomass decay on the net amount of heterotrophic biomass formed. As such it is a fundamental parameter that decreases in value as the SRT is increased. Y_n accounts for both that effect and the impact of inert solids and slowly biodegradable substrate on the MLSS in the bioreactor. As such, it is an empirical parameter that depends on the nature of the wastewater under treatment. However, if both X_{IO} and X_{SO} are zero, Y_n becomes equivalent to Y_{Hobs}.

In spite of its empirical nature, for many wastewaters, typical values of Y_n are known, or can be calculated using data from operating systems. An example of one relationship between Y_n and SRT is shown in Figure 9.7.[15] Several items should be noted. First, Y_n is affected by SRT in much the same way as Y_{Hobs}, as would be expected. (Compare Figure 9.7 to Figure 5.4.) Second, Y_n values are larger when the wastewater has not been subjected to primary treatment, i.e., settling, because the values of X_{IO} and X_{SO} are larger. Finally, Y_n has units of mg volatile suspended

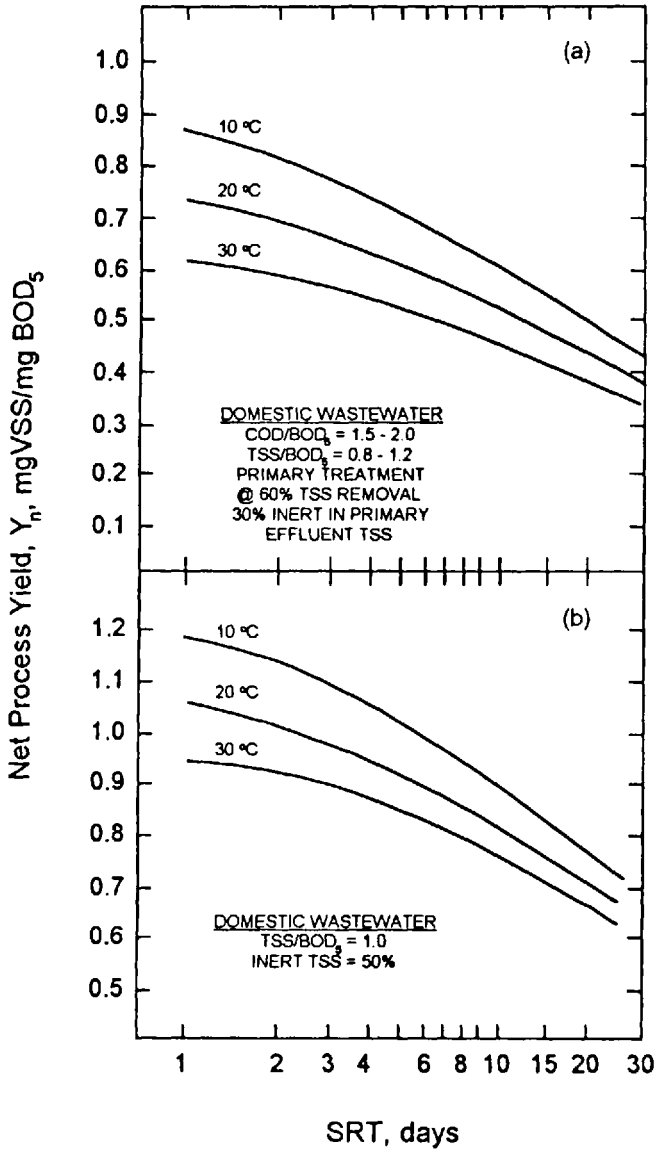


Figure 9.7 Net yield as a function of SRT and temperature for domestic wastewater (a) with primary treatment and (b) without primary treatment. (From *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, 1992. Copyright © Water Environment Federation; reprinted with permission.)

solid (VSS)/mg five-day biochemical oxygen demand (BOD_5). Actually, any unit system could be used to report Y_n, as long as it is consistent with the methods used to measure X_M and the substrate. The interconversion between various unit systems was discussed in Section 8.7. We have elected to use either COD or total suspended solid (TSS) units as the measure of MLSS in this book. Activated sludges typically

have a volatile solids content of approximately 75%, so the $X_M \cdot V$ values calculated with Eq. 9.2 can be converted to a TSS basis by dividing by 0.75.

The utility of Eq. 9.2 is that it can be used to estimate the mass of MLSS that will be present in the bioreactor, regardless of its configuration. The distribution of those solids will depend on the system configuration, as illustrated in Chapter 7 through simulation.

The net yield can also be used to calculate other information needed for a preliminary process design, such as the excess solids wastage rate, which is needed for design of the solids handling and disposal system. In Chapter 5, we derive Eq. 5.31 which expresses the excess biomass wastage rate in terms of $Y_{H,obs}$. By analogy to the development of Eq. 9.2, an expression can be developed from which the solids wastage rate can be calculated during preliminary design:

$$W_M = FY_n(S_{SO} + X_{SO}) \tag{9.3}$$

This equation has particular significance because it has another important use. If an existing treatment facility is to be upgraded, then information will be available in the plant records on the amount of solids wasted daily, as well as the quantity and strength of the wastewater. Such information allows Y_n to be calculated from the historical records with Eq. 9.3. It can then be used in Eq. 9.2 to estimate $X_M \cdot V$ for use in the design.

As discussed in Section 9.3.3, another factor that must be considered during design is the electron acceptor requirement. The basic COD balance states that the amount of electron acceptor required is equal to the biodegradable COD entering a bioreactor minus the COD of the solids wasted from the bioreactor. This is what led to Eq. 5.35, which gives the oxygen requirement in a simple CSTR receiving a soluble substrate. By analogy to the development of the previous preliminary design equations, an equation can be developed that relates the oxygen requirement to the flow and waste load in a manner similar to Eq. 9.3:

$$RO = FY_{O2}(S_{SO} + X_{SO}) \tag{9.4}$$

In this case, a new empirical coefficient, Y_{O2}, the process oxygen stoichiometric coefficient, has been defined, the units of which must be consistent with the units on S_{SO} and X_{SO}. Since BOD_5 is a commonly used measure of biodegradable substrate concentration in practice, Y_{O2} is commonly expressed as mg O_2/mg BOD_5. For domestic wastewater, the value of Y_{O2} is known as a function of the SRT, as illustrated in Figure 9.8.[14] Two things are of note about this figure. First, the values of Y_{O2} exceed 1.0. This follows directly from the fact that BOD_5 is not a measure of all of the electrons available in a substrate, as discussed in Section 8.6. Second, Y_{O2} increases as the SRT is increased in a manner consistent with Figure 5.6, illustrating that the empirical relationship depicted by Eq. 9.4 conforms to the fundamental principles developed earlier. For an existing facility Eq. 9.4 can be rearranged to calculate the process oxygen stoichiometric coefficient associated with a given waste load. Procedures for measuring the process oxygen requirement of such a facility are described elsewhere.[2,14]

The concept embodied in the COD mass balance and reflected in Eq. 5.35 can be generalized to any electron acceptor by expanding the substrate term to include slowly biodegradable substrate, as was done with Eq. 9.4, and incorporating an

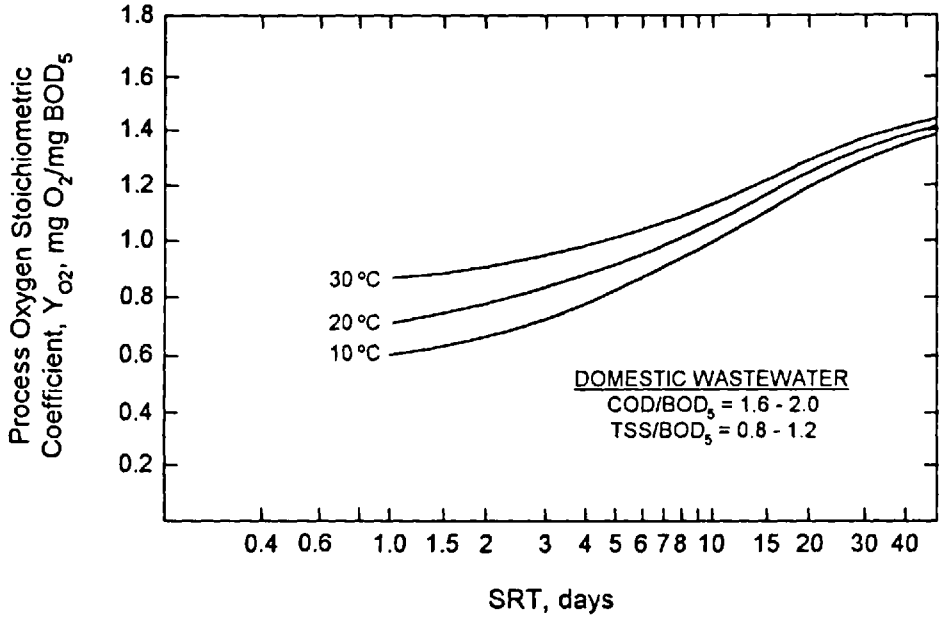


Figure 9.8 Process oxygen stoichiometric coefficient as a function of SRT and temperature for domestic wastewater. (From USEPA, *Design Manual—Fine Pore Aeration Systems*, EPA/625/1-89/023, Cincinnati, Ohio, 1989.)

appropriate conversion coefficient, $i_{O\,EA}$, reflecting the COD mass equivalent of the particular acceptor as listed in Table 3.1:

$$REA = \frac{F(S_{SO} + X_{SO})(1 - Y_{Hobs})}{-i_{O\,EA}} \tag{9.5}$$

where REA is the mass per time of electron acceptor required, and the substrate concentrations and observed yield are in COD units. If oxygen is the electron acceptor, $i_{O\,EA}$ has a value of -1.0 g COD/g O_2, whereas if nitrate-N is the acceptor, $i_{O\,EA}$ has a value of -2.86 g COD/g N. If carbon dioxide serves as the electron acceptor, leading to methane formation, the value of $i_{O\,EA}$ is -5.33 g COD/g C. In that case, however, it is more common to be interested in the amount of methane produced, rather than in the amount of carbon reduced. That quantity can be calculated by recognizing that all electrons removed end up in methane and that at standard temperature and pressure, the oxidation of 1.0 kg COD results in the formation of 0.35 m^3 of methane, as discussed in Section 2.3.2:

$$CH_4 \text{ production} = [F(S_{SO} + X_{SO})(1 - Y_{Hobs})] \left(0.35 \frac{m^3\ CH_4}{kg\ COD}\right) \tag{9.6}$$

As in Eq. 9.5, the substrate concentrations and observed yield are in COD units for Eq. 9.6.

Finally, for wastewaters that have a significant industrial component, the requirement for nutrients should be checked to ensure that adequate amounts are provided. A conservative estimate of the nutrient requirement can be obtained by as-

Table 9.3 Approximate Nutrient Requirements

Nutrient	Approximate requirement*	
	g/kg of VSS wasted	g/kg of TSS wasted
Nitrogen	125	104
Phosphorus	25	21
Potassium	14	12
Calcium	14	12
Magnesium	10	8
Sulfur	8.5	7
Sodium	4.3	3.6
Chloride	4.3	3.6
Iron	2.8	2.4
Zinc	0.3	0.2
Manganese	0.1	0.1

*Based on Table 3.3.

suming that all waste solids, as calculated from Eq. 9.3, are biomass and that the nutrients contained in those solids must be provided in the influent or by supplementation. Thus, the mass per day of nutrients required may be estimated by multiplying the solids wastage rate by the appropriate factor from Table 9.3, which was derived from Table 3.3. The supplementation requirement will be the difference between the amount required and the amount available in the influent.

All of the equations in this section are based on the guiding principles articulated in Section 9.1. Although they are approximate and incorporate several assumptions, they are sufficiently accurate for use during preliminary design when decisions are being made about the feasibility of various process alternatives. Thus, they have considerable utility in practice. Their use is illustrated in the following example.

Example 9.4.1.2

The preliminary design for a domestic wastewater treatment plant is being developed. The wastewater flow rate is 10,000 m^3/day, and the BOD, concentration is 200 mg/L (= 200 g/m^3). The plant discharge standards are 30 mg/L BOD, and 30 mg/L total suspended solids. Experience suggests that stable and reliable compliance with this discharge standard can be achieved using an SRT of 5 days. No primary clarifiers are to be provided. For the purpose of this assessment, the wastewater temperature can be assumed to be 20°C. Make a preliminary assessment of the size of bioreactor required if the MLSS concentration is maintained at 2500 mg/L as TSS. Assume that 75% of the MLSS is volatile. Also estimate the kg/day of solids that must be disposed of and the kg/day of oxygen that must be supplied. Because the wastewater is totally domestic, it can be assumed to contain adequate nutrients.

a. What would be an appropriate Y_n value for this application?
The value of Y_n can be estimated from Figure 9.7. Since primary treatment will not be used, part b of the figure is the appropriate part to use. For an SRT of 5 days, Y_n has a value of 0.92 mg VSS/mg BOD, at 20°C. Assuming

that the MLSS will be 75 percent volatile, the value of Y_n on a TSS basis is 1.2 mg TSS/mg BOD_u.

b. What is the $X_{M,t} \cdot V$ product for this application? From Eq. 9.2:

$$X_{M,t} \cdot V = (5)(1.2)(10{,}000(200) = 12{,}000{,}000 \text{ g TSS}$$
$$= 12{,}000 \text{ kg TSS}$$

c. If the allowable MLSS concentration is 2,500 mg/L (= 2,500 g/m^3), what size would the bioreactor be?
Since $X_{M,t} \cdot V$ = 12,000,000 g TSS,

$$V = \frac{12{,}000{,}000}{2{,}500} = 4{,}800 \text{ m}^3$$

d. What would the solids wastage rate be?
The solids wastage rate can be estimated with Eq. 9.3:

$$W_{M,t} = (10{,}000)(1.2)(200) = 2{,}400{,}000 \text{ g TSS/day}$$
$$= 2{,}400 \text{ kg TSS/day}$$

e. What would the process oxygen requirement be?
From Figure 9.8, at a 5-day SRT and a temperature of 20°C, Y_{O_2} has a value of 0.90 mg O_2/mg BOD_u. Using this, the oxygen requirement can be estimated from Eq. 9.4:

$$RO = (10{,}000)(0.9)(200) = 1{,}800{,}000 \text{ g } O_2\text{/day}$$
$$= 1{,}800 \text{ kg } O_2\text{/day}$$

The information provided in the example above would be adequate for a ball park estimate of the costs associated with a new treatment facility. However, they would not be adequate for more detailed estimates because such estimates require consideration of temperature variations, peak loadings, etc. While information like that in Figures 9.7 and 9.8 could be used when considering those effects, the analysis becomes much more detailed and is beyond the scope of this chapter. Such considerations will be covered in the chapters that follow, however.

In summary, even when little other information is available, it is often possible to develop a preliminary process design and evaluation for a wastewater treatment facility using the guiding principles developed in Parts I and II of this book. The general procedure is as follows:

1. Select the design SRT based on the treatment objectives and anticipated process operating conditions. An initially conservative approach is recommended to ensure that the bioreactor is adequately sized.
2. Estimate the net process yield (Y_n) and the process oxygen stoichiometric coefficient (Y_{O_2}) based on the best information available. The basis may be actual operating data from an existing facility, values from the literature, such as those in Figures 9.7 and 9.8, or an estimate based on knowledge of and experience with similar wastewaters.
3. Calculate the $X_M \cdot V$ product using Eq. 9.2.
4. Calculate the solids wastage rate using Eq. 9.3 and the process oxygen requirement using Eq. 9.4.

5. Use the value of the $X_M \cdot V$ product determined in item 3 to consider the interactions between the selected size of the bioreactor, the final clarifier, and the oxygen transfer/mixing system in order to arrive at a selection that meets all requirements.

For an existing facility, this will provide a preliminary estimate of existing capacity. For a new facility or a facility expansion, a preliminary estimate of capital and operating costs can be prepared for use in on-going project evaluation and planning. The selection of a conservative SRT value results in conservative estimates of each. It is important to note that no estimates of process effluent quality have been developed yet. Rather, the fact that effluent quality is largely determined by the process SRT has been exploited to allow selection of an SRT that will result in the desired degree of conversion. Design to achieve a specified effluent quality requires more refined techniques.

9.4.2 Stoichiometric-Based Design and Evaluation

As noted above, preliminary design and evaluation based on the guiding principles does not allow precise estimates of process effluent quality. That requires values of the kinetic parameters for biodegradation of the constituent of interest. For some situations, approximate or typical values of those parameters are known, allowing more precise estimates of process effluent quality to be made without the necessity for running treatability studies. In other cases, appropriate values of the kinetic parameters are unknown and they must be measured directly for the subject wastewater. Likewise, the mass of biomass required in the bioreactor, the solids wastage rate, and the process oxygen requirement are all functions of the characteristics of the wastewater to be treated and the microbial populations that develop in the treatment system. Consequently, more precise estimates of the bioreactor size, etc. require values for the kinetic and stoichiometric parameters that are specific to a particular wastewater. Again, for some situations, appropriate values of the wastewater constituent parameters can be selected based on experience, but in other instances the values of those parameters must be measured directly for the subject wastewater and the particular type of bioreactor under consideration.

Because the International Association on Water Quality (IAWQ) ASMs have been developed only recently, most of our experience in modeling bioreactors for design and evaluation has come from the simple model of Chapter 5. Although it was presented as a soluble substrate model, much experience exists with it when the organic matter is present in both a soluble and particulate state. In that case, influent concentrations are expressed as biochemical oxygen demand (BOD) or as biodegradable COD without regard to their physical state, whereas effluent concentrations are commonly expressed as soluble material alone. Such a practice assumes that bioflocculation will entrap the bulk of the undegraded particulate organic matter, making soluble material the main contributor of organic matter in the effluent. Treatability studies for quantification of the parameters in the simple model of Chapter 5 were outlined in Sections 8.2 and 8.3, while procedures for translating between the various measurement techniques and unit systems were discussed in Sections 8.6 and 8.7. By using such procedures it is possible to use the Chapter 5 model to design systems involving organic substrate removal and nitrification. Because most of the

computations with that model are based on the stoichiometry of microbial growth, it has been called a stoichiometric-based model.[3] It has severe restrictions for use with biological nutrient removal systems. Thus, one should strive to use the IAWQ models for them.

When a design or evaluation is sufficiently advanced to warrant the use of modeling, many factors are considered that were not considered during preliminary design or evaluation. Thus, the approach is more involved, although the same basic strategy is used. Consequently, a full discussion of this subject will be presented in the remaining chapters of Part III, with each chapter being devoted to a particular type of named biochemical operation. The material in this section will be limited to a few major points.

When modeling is being used, the choice of the SRT should consider the value required to meet a desired effluent quality. If the bioreactor is to be completely mixed, then a rearranged form of Eq. 5.13 may be used to select the minimum allowable design SRT because lower SRTs would not produce the required effluent quality:

$$\Theta_c = \frac{K_S + S_S}{S_S(\hat{\mu}_H - b_H) - K_S b_H} \tag{9.7}$$

This computation may be made on the basis of total soluble biodegradable organic matter, using either biodegradable COD or BOD as the measure of S_S, or on the basis of specific organic chemicals, in which case S_S would represent them. No matter what the basis for S_S, however, the same basis must be used for K_S. Alternatively, if the system must meet an ammonia-N criterion, Eq. 9.7 could be applied to the autotrophic bacteria by replacing the heterotrophic parameter values with autotrophic ones. In that case, however, appropriate safety factors may also be applied to the design. Once the minimum allowable design SRT has been computed with Eq. 9.7, it must be compared to the various constraints discussed in the context of Figure 9.3. In other words, even though modeling will be used in the design, the choice of SRT is still limited by the same constraints discussed previously, and they may control the choice. This subject will be discussed more in subsequent chapters.

Once the SRT has been selected, design or evaluation of a system based on the stoichiometric model of Chapter 5 proceeds in exactly the same way as the preliminary design presented in Section 9.4.1. The differences are in the equations used and the additional factors that are considered. If the waste organic constituents and MLSS concentrations are all expressed in COD units, the equations in Chapter 5 may be used directly, with one minor modification: the influent concentration is the sum of the easily and slowly biodegradable constituents entering the bioreactor. In other words, S_{SO} is replaced by the sum of S_{SO} and X_{SO}. Consequently, the mass of solids in the bioreactor can be calculated with a modified form of Eq. 5.46:

$$X_M \cdot V = \Theta_c F \left[X_{IO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c) Y_H (S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \Theta_c} \right] \tag{9.8}$$

The solids wastage rate can be calculated from a modified form of Eq. 5.49:

$$W_M = F \left[X_{IO} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c) Y_H (S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \Theta_c} \right] \tag{9.9}$$

The oxygen requirement can be calculated from a modified form of Eq. 5.33:

$$RO_H = F(S_{SO} + X_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_H}{1 + b_H \cdot \Theta_c}\right] \tag{9.10}$$

The nutrient requirement can still be calculated from the solids wastage rate as discussed above.

If the MLSS concentration is expressed in TSS units while the organic substrate is expressed in COD units, additional modifications of the equations are required. One is that the yield values must be expressed in TSS units. The relationships between yield values when the biomass concentration is expressed in COD and TSS units were presented in Section 8.7 and are straightforward. Another is that the build up of fixed solids from the influent must be accounted for. This was discussed in Section 5.2.2, but was not explicitly considered at that time. It needs to be accounted for now, however. Finally, the maintenance of a COD balance for computing oxygen requirements is greatly simplified if it can be assumed that the COD of biomass debris is the same as the COD of active biomass. Thus, that assumption will be made herein.

From a consideration of the factors in the preceding paragraph, the mass of MLSS in TSS units can be calculated with another modified form of Eq. 5.46 in which the concentration of inert particulate COD, X_{IO}, is replaced with the concentration of inert suspended solids, $X_{IO,T}$, and the yield factor is expressed in TSS units, as denoted by the subscript T:

$$X_{M,T} \cdot V = \Theta_c F\left[X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}(S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{9.11}$$

The concentration of inert suspended solids includes both nonbiodegradable VSS and the influent fixed suspended solids (FSS), which is just the difference between the influent TSS and VSS concentrations. The equation for the solids wastage rate must be changed similarly, giving:

$$W_{M,T} = F\left[X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}(S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{9.12}$$

It should be recognized that because of the presence of inert organic solids and fixed solids in the MLSS and waste solids, their COD equivalent, $i_{O\,XM,T}$ will be different from the COD equivalent of the biomass, $i_{O\,XB,T}$, and will depend on the magnitude of $X_{IO,T}$, as well as its COD equivalent. As a result, a measured value of $i_{O\,XM,T}$ will normally be required to calculate the oxygen requirement from a COD balance across the bioreactor system. Consequently, the simplest way of calculating the oxygen requirement is through a modified form of Eq. 5.33:

$$RO_H = F(S_{SO} + X_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T} i_{O\,XB,T}}{1 + b_H \cdot \Theta_c}\right] \tag{9.13}$$

The equations above can be used regardless of the bioreactor configuration, as

discussed in Section 9.1. However, by the time a design or evaluation has progressed to the point that a treatability study has been run to provide the parameter values for use in the model, it will be necessary to consider the distribution of MLSS and oxygen within the bioreactor system. This requires the use of heuristic rules based on experience. These will be discussed in subsequent chapters.

9.4.3 Simulation-Based Design and Evaluation

When a complex biochemical operation, such as a biological nutrient removal system, is being designed or evaluated, even a small degree of uncertainty can result in large consequences in terms of effluent quality or system cost. Thus, use of the simple stoichiometric-based model is usually not adequate, except as a starting point for more detailed studies. Because of the cost of treatability studies, particularly at the pilot-plant scale, the potential exists for significant cost savings through use of simulation. Computer codes implementing ASM No. 1 and ASM No. 2, listed in Table 6.4, provide excellent means for doing this.

Starting with the default parameters provided in the IAWQ reports,[9] one can use the simple model or the guiding principles of Section 9.1 to choose a starting bioreactor system. The performance of that system can then be evaluated by simulation and systematic changes in system configuration and operating conditions can be investigated, just as was done in Chapter 7. Through this approach, the most feasible bioreactor systems and the most sensitive parameter values can be identified for experimental study in the lab as discussed in Section 8.5. The results of the experimental study will provide new parameter estimates that can then be used in another round of simulations, leading to the one or two most feasible configurations, which can be investigated at pilot scale. In other words, the iterative process of design and evaluation illustrated to Figure 9.1 is applied to both the simulation and the testing phases of the project. The resulting system will have much less uncertainty associated with it.

9.4.4 Effluent Goals Versus Discharge Requirements

Collection of the kinetic and wastewater characterization data described above allows relatively precise estimates of process effluent quality to be developed. However, the inherent variability in the performance of such processes must be recognized. Precise data allow accurate prediction of the mean performance of a treatment process. In contrast, facility discharge requirements are typically expressed in "not to exceed" terms so that variability in process performance must be considered when establishing a process design. Numerous procedures exist for translating discharge requirements into process design requirements. One which is frequently used is to establish an effluent goal that is more stringent than the specified discharge requirement and incorporates an allowance for the expected variations in treatment performance.

A variety of procedures are available to select effluent goals to allow reliable compliance with specified discharge requirements. In some instances effluent goals are selected based on experience with a particular wastewater and application. Other procedures are based on the statistical characteristics of a process effluent, and their use requires prior knowledge of those characteristics.[10] Still other procedures are

general in nature and can be applied to a wide range of applications. The procedure of Roper, et al.[11] illustrates this latter approach and is presented here for illustrative purposes. Roper, et al.[11] collected effluent quality data from a wide variety of full-scale wastewater treatment facilities and plotted extreme values versus the mean process performance. They found that the two were well correlated and that the relationship was generally linear. Figure 9.9 presents several of the relationships they developed and illustrates the association between the extreme values typically considered in discharge requirements and the mean performance that can be predicted using process kinetics and mean process loadings. Relationships such as those presented in Figure 9.9 can be used to select treatment goals to allow reliable compliance with facility discharge standards.

9.4.5 Optimization

Further steps in the refinement of the design and evaluation of a biological process fall under the general heading of optimization. A wide variety of approaches can be considered here, including pilot-plant studies, further wastewater characterization and analysis, detailed analyses of the performance of existing facilities treating similar wastewaters or using similar processes, computer simulation of process performance to characterize the dynamic performance of a proposed process design, and detailed

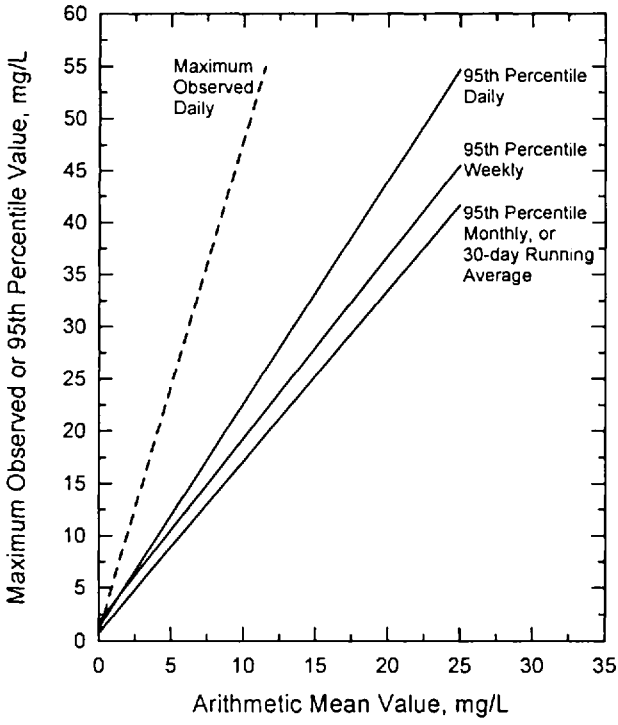


Figure 9.9 Relationship between arithmetic average and 95th percentile or maximum observed effluent BOD_5 and TSS concentrations for full-scale wastewater treatment plants. (Adapted from Roper, et al.[11])

engineering studies to evaluate cost trade-offs for the subject process. Detailed descriptions of these steps are beyond the scope of this book. However, the reader should recognize that many tools are available that can be applied to further refine the design and evaluation of a biological wastewater treatment process.

9.5 KEY POINTS

1. Five guiding principles provide the basis for the design and evaluation of all suspended growth bioreactors:
   - The biochemical environment imposed upon a bioreactor determines the nature of the microbial community that develops and the character of the biological reactions that they perform.
   - The solids retention time (SRT) is the most important design and control parameter available to the engineer.
   - A chemical oxygen demand (COD) balance across a bioreactor provides valuable information about the amount of electron acceptor required and the amount of excess biomass produced.
   - The excess biomass production rate is essentially the same for all suspended growth systems with a given SRT and biochemical environment, regardless of the bioreactor configuration. As a consequence, the total mass of biomass and the total amount of electron acceptor required for removal of organic matter in those systems will be the same, regardless of the system configuration.
   - Only the mass of biomass in a bioreactor system is specified by the descriptive analytical equations, not the concentration. The concentration is only specified after the bioreactor volume has been specified.
2. The design process is iterative in nature and contains feedback loops. This is due to two factors: (1) improved problem definition and (2) additional data, both of which become available as the design and evaluation proceeds. Consequently, the design is refined as more information becomes available. The iterative nature of design demands the use of a variety of process design procedures, which must be consistent and based on the guiding principles. The benefits derived from using a more sophisticated design procedure must justify its additional costs.
3. The operational and performance characteristics of aerobic/anoxic and anaerobic bioreactors differ significantly. Aerobic/anoxic systems are generally used to treat wastewaters with biodegradable COD concentrations less than 1,000 mg/L, while anaerobic systems are usually used to treat those with concentrations greater than 4,000 mg/L. Either can be used for wastewaters with concentrations between those limits, depending on the specific application. Effluents from an anaerobic bioreactor must generally be polished in an aerobic system if a high quality is needed.
4. The SRT selected for a bioreactor must always exceed the minimum SRT associated with the microorganisms responsible for a particular desired biochemical transformation.
5. Biochemical conversions that are controlled by the SRT in aerobic/anoxic bioreactor systems include metabolism of biodegradable soluble and par-

ticulate organic matter, stabilization of removed organic matter, biodegradation of xenobiotic compounds, bioflocculation, growth of nitrifying bacteria, and growth of phosphate accumulating organisms (PAOs).

6. Biochemical conversions controlled by the SRT in anaerobic bioreactor systems include hydrolysis of carbohydrates, proteins, and lipids; fermentation of simple sugars and amino acids; anaerobic oxidation of fatty acids; and methanogenesis.
7. Several important characteristics of bioreactors can be determined from process stoichiometry once the SRT has been fixed, regardless of the bioreactor configuration. These are the mass of biomass in the system, the mass rate of solids wastage, the quantity of electron acceptor that must be supplied, and the amount of nutrients needed.
8. Because only the mass of biomass in a bioreactor is fixed by process stoichiometry, the designer may freely choose either the mixed liquor suspended solids (MLSS) concentration or the bioreactor volume, thereby fixing the other. However, that decision must be made while considering its impact on the size of the final clarifier and the volumetric power input required for oxygen transfer and/or mixing.
9. During preliminary design and evaluation of a bioreactor system, Y_n, the net process yield may be used to calculate the mass of MLSS and the solids wastage rate for the system. A similar coefficient, Y_{O_2}, the process oxygen stoichiometric coefficient, can be used to estimate the oxygen requirement. They are based on experience at full-scale treatment facilities and incorporate the effects of SRT and wastewater characteristics.
10. More precise estimates of the bioreactor size, etc. require values for the kinetic and stoichiometric parameters that are specific to a particular wastewater. These may be obtained by treatability studies and used in stoichiometric models that are based on the simple model of Chapter 5. When a design or evaluation is sufficiently advanced to warrant the use of modeling, many factors may be considered that were not considered during preliminary design or evaluation. Thus, the approach is more involved, although the same basic strategy is used.
11. When a complex biochemical operation, such as a biological nutrient removal system, is being designed or evaluated, even a small degree of uncertainty can result in large consequences in terms of effluent quality or system cost. Thus, the potential exists for significant cost savings through use of simulation with active sludge model (ASM) No. 1 or No. 2.

9.6 STUDY QUESTIONS

1. One of the guiding principles summarized in Section 9.1 is that the excess biomass production rate is essentially the same for all suspended growth systems with a given SRT and biochemical environment, regardless of the bioreactor configuration. Discuss the implications and significance of this fact to the design and evaluation of bioreactor systems.
2. Explain why the design and analysis of bioreactor systems is an iterative process and why the approach used must change as the process proceeds.

3. Define the terms "minimum SRT" and "safety factor" as applied to bioreactor design and evaluation, discuss their significance, and explain their use.
4. An aerobic/anoxic bioreactor is to be designed, and the first step is the selection of the design SRT. Discuss the differences in the SRT required to achieve removal of biodegradable organic matter (both readily and slowly biodegradable), flocculation, biological phosphorus removal, nitrification, and stabilization of the produced microorganisms. What factors determine the SRT range for each treatment objective?
5. An anaerobic bioreactor is to be designed, and the first step is the selection of the design SRT. Discuss the differences in the SRT required to achieve fermentation of influent particulate matter into volatile fatty acids, stabilization of carbohydrates and proteins, and stabilization of lipids. What factors determine the SRT range for each treatment objective?
6. An industrial wastewater treatment plant treats a flow of 5,000 m^3/day with a COD concentration of 3,500 mg/L. The process operates at a 25 day SRT and produces an effluent with a soluble COD concentration of 125 mg/L. Solids are wasted from the system at a rate of 7,000 kg TSS/day. What is the net yield (Y_n) for this facility, expressed in units of mg TSS/mg COD?
7. A bioreactor system treats 60,000 m^3/day of wastewater with a BOD_5 concentration of 200 mg/L and a TSS concentration of 210 mg/L, which are typical of unclarified domestic wastewater. The average temperature is 20°C. The discharge standards are 30 mg/L BOD_5 and 30 mg/L TSS on a maximum monthly basis, i.e., typical secondary treatment. Calculate the mass of MLSS in the bioreactor (in TSS units), the process solids wastage rate (in TSS units), and the process oxygen requirements for this application. Neglect process oxygen requirements for nitrification. Justify your choice of SRT.
8. The addition of primary treatment to the domestic wastewater treatment plant described in Study Question 7 is being evaluated. Pilot studies demonstrate that primary clarification will remove 30 percent of the BOD_5 and 60 percent of the TSS contained in the influent wastewater. The discharge standards will remain the same at 30 mg/L BOD_5 and 30 mg/L TSS on a maximum monthly basis, and thus the SRT will remain the same. Calculate the mass of MLSS (in TSS units), the process solids wastage rate from the bioreactor (in TSS units), and the process oxygen requirements for this application. Neglect process oxygen requirements for nitrification.
9. Biological phosphorus removal is to be added to the domestic wastewater treatment plant described in Study Question 7. How does this affect the mass of biomass (in TSS units), the process solids wastage rate (in TSS units), and process oxygen requirements for this application? Neglect process oxygen requirements for nitrification.
10. The domestic wastewater treatment plant described in Study Question 7 is to be upgraded to provide nitrification. How does this affect the mass of biomass and the process solids wastage rate (both in TSS units)?

REFERENCES

1. Daigger, G. T., Development of refined clarifier operating diagrams using an updated settling characteristics database. *Water Environment Research* **67**:95–100, 1995.
2. Daigger, G. T. and J. A. Buttz, *Upgrading Wastewater Treatment Plants*, Technomic Publishing, Lancaster, Pennsylvania, 1992.
3. Daigger, G. T. and C. P. L. Grady, Jr., The use of models in biological process design. *Proceedings of the Water Environment Federation 68th Annual Conference & Exposition, Volume 1, Wastewater Treatment Research and Municipal Wastewater Treatment*, pp. 501–510, 1995.
4. Grady, C. P. L., Jr., Simplified optimization of activated sludge process. *Journal of the Environmental Engineering Division, ASCE* **103**:413–429, 1977.
5. Hall, E. R., Anaerobic treatment of wastewaters in suspended growth and fixed film processes. In: *Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes*, Malina, J. F. Jr., and F. G. Pohland eds. Technomic Publishing, Lancaster, Pennsylvania, pp. 41–118, 1992.
6. Henze, M., C. P. L. Grady Jr., W. Gujer, G. v. R. Marais, and T. Matsuo, Activated sludge model No. 1. *IAWPRC Scientific and Technical Reports*, No. 1, 1987.
7. Henze, M., W. Gujer, T Mino, T. Matsuo, M. C. Wentzel, and G. v. R. Marais, Activated sludge model No. 2. *IAWQ Scientific and Technical Reports*, No. 3, 1995.
8. Keinath, T. M., Diagram for designing and operating secondary clarifiers according to the thickening criterion. *Journal, Water Pollution Control Federation* **62**:254–258, 1990.
9. Mamais, D. and D. Jenkins, The effects of MCRT and temperature on enhanced biological phosphorus removal. *Water Science and Technology* **26**(5/6):955–965, 1992.
10. Niku, S., Schroeder, E. D., and F. J. Samaniego, Performance of activated sludge processes and reliability-based design. *Journal, Water Pollution Control Federation* **51**: 2841–2857, 1979.
11. Roper, R. E., Jr., R. O. Dickey, S. Marman, S. W. Kim, and R. W. Yandt, Design effluent quality. *Journal of the Environmental Engineering Division, ASCE* **105**:309–321, 1979.
12. Sedlak, R. I., *Phosphorus and Nitrogen Removal from Municipal Wastewater: Principles and Practice*, 2nd Ed. Lewis Publishers, Ann Arbor, Michigan, 1991.
13. Shao, Y. J., J. Crosse, E. Keller, and D. Jenkins, High rate air activated sludge operation at the City of Los Angeles Hyperion Wastewater Treatment Plant. *Water Science and Technology* **25**(4/5):75–87, 1992.
14. U. S. Environmental Protection Agency, *Design Manual–Fine Pore Aeration Systems*, EPA/625/1-89/023, U. S. Environmental Protection Agency, Cincinnati, Ohio, 1989.
15. Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992.

10
Activated Sludge

As indicated in Table 1.2, activated sludge is an aerobic suspended growth process in which microorganisms are grown in a variety of bioreactor configurations for the purpose of removing soluble organic matter. It is a flexible, reliable process capable of producing a high quality effluent. Soluble organic matter is reduced to low levels, and a clear effluent low in suspended solids is produced due to the flocculent nature of the biomass. High degrees of nitrification and of stabilization of insoluble organic matter can also be achieved by operation at a sufficiently long solids retention time (SRT). The process is highly controllable, and its operation can be adjusted in response to a wide range of conditions. It is also relatively resistant to short-term organic and hydraulic loading variations. Its main disadvantage is a result of its controllability; its operation is relatively complicated and requires the attention of qualified and experienced operators. Its dynamic response to transient loading variations can be sluggish, and the capability to respond to such variations must be built into the process. Capital and operating costs, although reasonable, are still significant. In short, successful implementation of an activated sludge system requires the commitment of sufficient capital, operating, and personnel resources to achieve success.

10.1 PROCESS DESCRIPTION

Since its inception by Arden and Lockett in 1914,[1] the activated sludge process has grown in popularity until today it is the most widely used biological wastewater treatment process. Much experimentation has occurred since its initial development, and many variations are known. Most of the variations can be grouped into the eight named operations listed under activated sludge in Table 1.2. Theoretical simulations of the performance of many of those variations are presented in Chapters 5, 6, and 7. The purpose here is to describe and compare them as they appear in practice.

10.1.1 General Description

Four factors are common to all activated sludge systems: (1) a flocculent slurry of microorganisms (mixed liquor suspended solids [MLSS]) is utilized to remove soluble and particulate organic matter from the influent waste stream; (2) quiescent settling is used to remove the MLSS from the process flow stream, producing an effluent that is low in suspended solids; (3) settled solids are recycled as a concentrated slurry from the clarifier back to the bioreactor; and (4) excess solids are wasted to control the SRT to a desired value. Figure 10.1 illustrates how those four functions

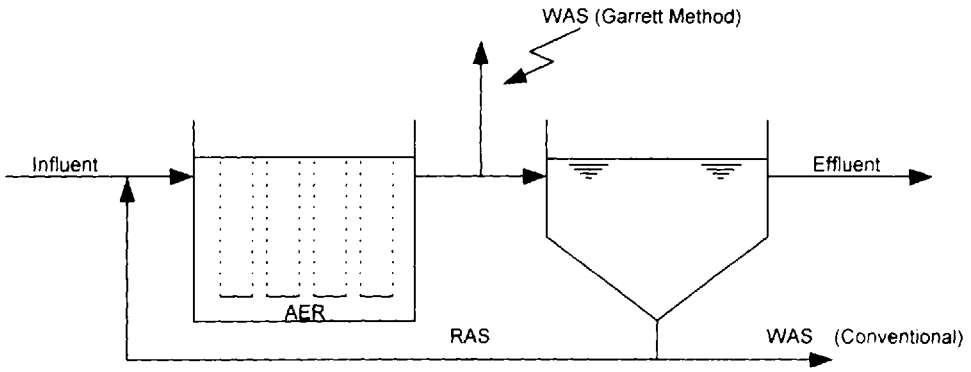


Figure 10.1 Typical activated sludge process.

are typically provided and is shown to allow the definition of some terms commonly used in practice. The basic similarity of Figure 10.1 to Figure 5.2 and the schematics of the systems used to simulate the various options in Chapter 7 should be noted. The bioreactor containing the MLSS is commonly called an aeration basin, and it is aerobic throughout, as indicated by AER in Figure 10.1. Sufficient mixing energy must be provided in the bioreactor to maintain the solids in suspension, as mentioned in Section 9.3.4. The MLSS concentration will depend on the characteristics of the influent wastewater, bioreactor hydraulic residence time (HRT), and process SRT, as discussed in Section 9.3.3. The stream of concentrated solids being recycled from the clarifier to the bioreactor is commonly called return activated sludge (RAS). The RAS suspended solids concentration will depend on the clarifier operating conditions, including the MLSS concentration and the influent and RAS flow rates. As discussed in Section 5.3, solids can be removed from the process at various points to maintain the desired SRT and the stream through which that is done is referred to as waste activated sludge (WAS). Figure 10.1 illustrates the two common solids removal points, from the secondary clarifier underflow (conventional method) and from the aeration basin (Garrett[20] method). The relative merits of the two alternatives are discussed in Section 5.3. As its name suggests, the conventional method is more common.

Aeration basins are typically open tanks containing equipment to transfer oxygen into solution and to provide mixing energy to keep the MLSS in suspension. The depth is determined largely by the characteristics of the oxygen transfer/mixing system and typically ranges from 3 to 7.5 m. In special cases, depths as low as 2 m or as great as 20 m may be used. The bioreactor may be constructed of concrete, steel, or as an earthen basin lined with clay or an impermeable membrane. Vertical sidewalls are typically used with concrete and steel structures, while sloping sidewalls are used with earthen structures. A wide variety of bioreactor configurations, i.e., length-to-width ratios, can be used. Bioreactor configuration, the characteristics of the oxygen transfer/mixing equipment, and the distribution of the RAS flow will affect the mixed liquor flow pattern within the bioreactor, which will affect process performance, as demonstrated in Chapter 7.

In many cases a single device is used both to transfer oxygen and to keep the MLSS in suspension. Typical devices include diffused air (both coarse and fine bub-

ble), floating or fixed mechanical surface aerators (both high speed and low speed), jet aerators, and submerged turbine aerators. Auxiliary mechanical mixers are used when the aeration device does not provide sufficient mixing energy to keep the MLSS in suspension.

The secondary clarifier provides two functions. One, called the clarification function, is removal of the MLSS to produce a clarified effluent that meets the effluent suspended solids goal. The other, called the thickening function, is the concentration of the settled solids for return to the bioreactor. A wide range of secondary clarifier configurations can be used, although circular and rectangular are the most common. Clarifier configuration has less impact on process performance than bioreactor configuration, as long as the clarifier is sized properly. Clarifiers contain an effluent collection device (typically an overflow weir and effluent collection launder) and a settled solids collection device. Most clarifiers are also provided with equipment to collect solids which float to the surface. The secondary clarifier can provide temporary storage of MLSS transferred from the bioreactor during periods of high flow. This occurs when suspended solids are transferred to the clarifier at a rate greater than they can be removed in the RAS and WAS streams. The solids storage capacity of a secondary clarifier is largely a function of its depth and the settling characteristics of the MLSS, but routine use of the secondary clarifier for solids storage is generally not recommended.

This book focuses on biological process design rather than on facility design. Readers interested in further information about the design of oxygen transfer devices, mixing systems, and final clarifiers are referred elsewhere.[40,71,79,82]

10.1.2 Process Options

Due to the long history and varied use of the activated sludge process, a great number of named process options exist. In this section eight that represent the range and capabilities of activated sludge are described. Six were introduced in Chapters 5 and 7 where their theoretical performance was investigated through simulation.

Conventional Activated Sludge. The first continuous flow activated sludge processes used long narrow aeration basins with influent and RAS being added at one end and flowing in a plug–flow[a] fashion to the opposite end. Diffused air with a uniform aeration pattern along the length of the basin was initially utilized. Later it was recognized that the oxygen requirement varies along the length of the bioreactor, as illustrated in Section 7.2.4. This led to the use of tapered aeration, which is the variation of oxygen input along the length of the basin in response to the oxygen demand.[41]

Conventional activated sludge (CAS) is the modern embodiment of those early processes. The bioreactor is generally rectangular in shape with influent and RAS

---

[a]It is apparent that the aeration and mixing in an activated sludge aeration basin prevent it from having the residence time distribution of a perfect plug–flow reactor (PFR) as defined in Chapter 4. Consequently, the use of the term "plug–flow" in the context of activated sludge and other biochemical operations means that the bioreactor is sufficiently long relative to its width to contain concentration gradients of soluble constituents as the fluid moves through the basin.

being added at one end and mixed liquor exiting at the opposite end. The flow pattern is quasi-plug–flow, with the residence time distribution depending on the length-to-width ratio of the basin, the mixing produced by the oxygen transfer equipment, and the details of the inlet and outlet.[48,71] As discussed in Section 4.4.1, flow patterns such as these can be modeled as a series of continuous stirred tank reactors (CSTRs). For example, the flow pattern within a bioreactor with a length-to-width ratio of 5:1 can be approximated as three tanks in series. Increasing the bioreactor length-to-width ratio will increase the degree of plug–flow as represented by an increase in the number of equivalent tanks in series. The HRT typically ranges from 4 to 8 hours, while the SRT ranges from 3 to 8 days, but seldom exceeds 15 days. As discussed in Section 7.2.4, the MLSS concentration and composition vary little through the bioreactor because the SRT is long relative to the HRT and the mixed liquor is recycled many times before it is wasted.

Step-Feed Activated Sludge. As covered in Section 7.3.1 and illustrated in Figure 7.10, step-feed activated sludge (SFAS) differs from CAS by having the influent wastewater distributed at several points. Figure 10.2 illustrates ways in which this is typically accomplished in practice. Figure 10.2a depicts a single narrow basin with influent added at various points along its length, while Figure 10.2b shows a series of such basins (each often referred to as a pass), with influent added to each. Many variations of the SFAS process exist,[31] with the fraction of the influent added at each feed point depending on the design objectives. Today SFAS is used primarily to redistribute the activated sludge inventory within the bioreactor, as discussed in Section 7.3.4. Compared to a CAS bioreactor of the same volume, this results in an increased activated sludge inventory, i.e, a longer SRT, and/or a lower concentration of MLSS entering the clarifier. SRTs are generally similar to those used with the CAS process.

Contact Stabilization Activated Sludge. The contact stabilization activated sludge (CSAS) process, discussed in Section 7.4 and illustrated in Figure 7.19, divides the bioreactor into two zones: the contact zone where removal of organic matter contained in the influent wastewater occurs, and the stabilization zone where RAS from the clarifier is aerated to allow stabilization of organic matter. Because of the relatively high MLSS concentration in the stabilization basin (equal to the RAS concentration), a smaller total bioreactor volume (contact plus stabilization) may be used relative to the CAS process, while maintaining the same SRT. As a consequence, CSAS can be used either to reduce the required bioreactor volume or to increase the capacity of an existing CAS facility. Contact zone HRTs are typically 0.5 to 2 hours, while stabilization zone HRTs are typically 4 to 6 hours based on the RAS flow. SRTs are similar to those used with the CAS process.

Completely Mixed Activated Sludge. Completely mixed activated sludge (CMAS) systems, described in Chapters 5 and 6, were developed in the late 1950s to treat high strength industrial wastewaters, particularly those containing inhibitory organic matter.[30,44] Treatment of such wastewaters in CAS systems is difficult due to the high concentrations of organic matter near the inlet, which can inhibit the biomass, causing poor performance. In contrast, in the CMAS process, influent wastewater is distributed uniformly throughout the bioreactor, maintaining a low concentration of biodegradable organic matter at all points. Thus, even if the organic matter is inhibitory at high concentration, inhibition is avoided, allowing biodegradation to

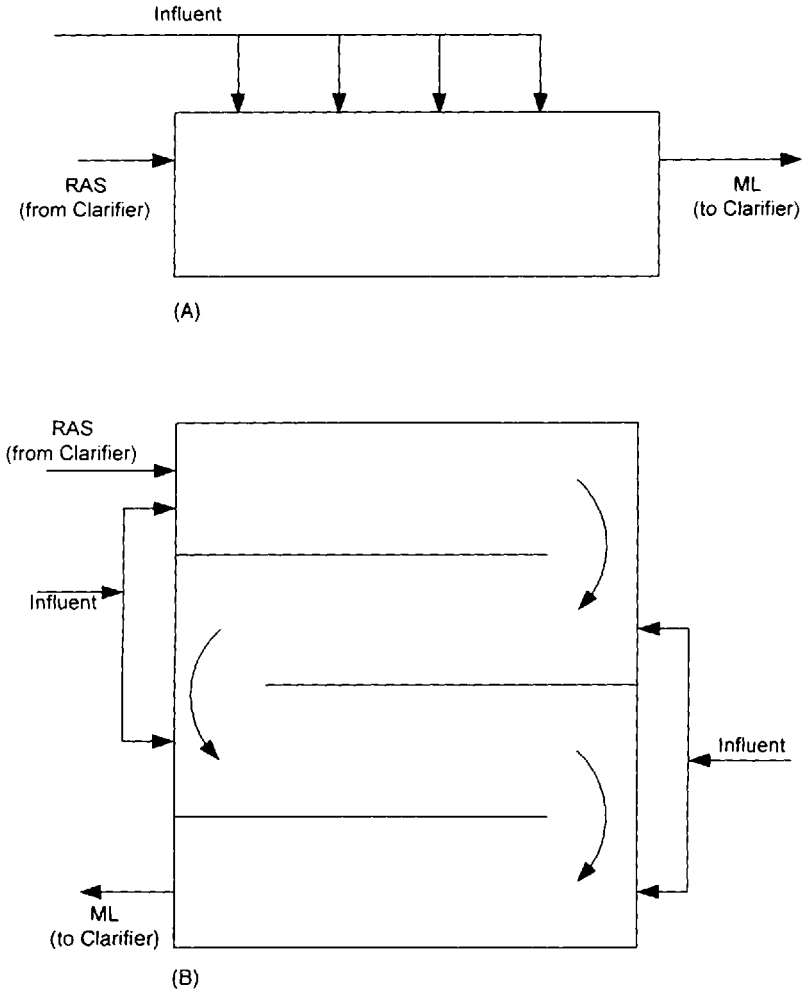


Figure 10.2 Step-feed activated sludge (SFAS) process.

proceed. It was also thought that process stability would be enhanced by maintaining the microorganisms under nearly constant conditions. The HRTs and SRTs for the CMAS process are similar to those used with CAS.

Figure 10.3 illustrates two bioreactor configurations commonly used to achieve completely mixed conditions. The first, in Figure 10.3a, has been used with diffused aeration systems; complete mixing is achieved by distributing the influent along one side of a long, narrow bioreactor, with effluent being taken from the opposite side. In other instances (Figure 10.3b) an essentially square shaped bioreactor is used with influent and effluent positioned to achieve completely mixed conditions. Mechanical surface aeration is typically used with the latter because it provides good overall circulation of basin contents.

Extended Aeration Activated Sludge. Extended aeration activated sludge (EAAS) processes utilize long SRTs to stabilize the biosolids resulting from the

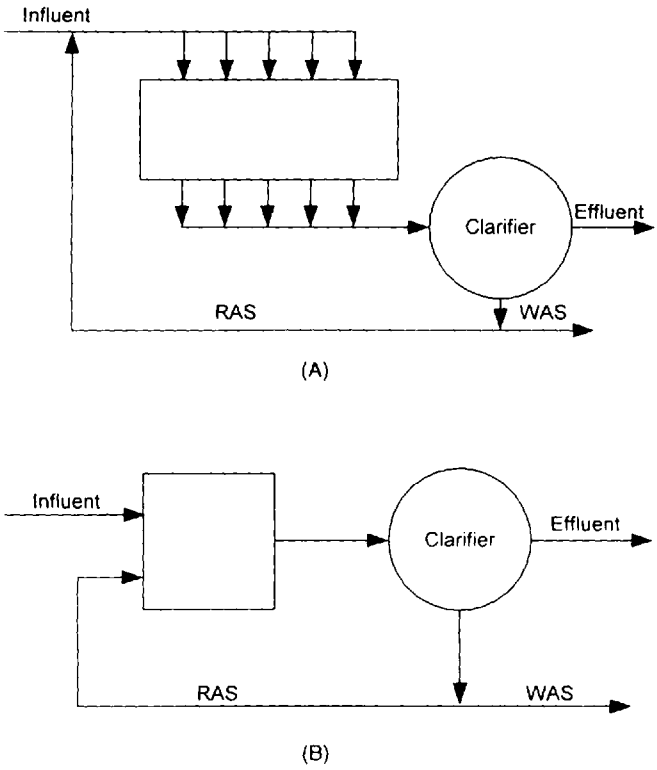


Figure 10.3 Completely mixed activated sludge (CMAS) process.

removal of biodegradable organic matter. SRTs of 20 to 30 days are typical, which means that HRTs around 24 hours are required to maintain reasonable MLSS concentrations. Long SRTs offer two benefits: reduced quantities of solids to be disposed of, and greater process stability. These benefits are obtained at the expense of the large bioreactors required to achieve the long SRTs, but for many small installations the benefits outweigh the drawbacks.

One of the primary drawbacks of large bioreactors is that they are typically mixing limited. In other words, as discussed in Section 9.3.4, the energy required to meet the oxygen requirement is less than the energy necessary to keep the MLSS in suspension. This limitation is minimized with the closed looped bioreactor, or oxidation ditch, illustrated in Figure 10.4. In this configuration an aeration device is used both to transfer oxygen and to provide motive velocity to the mixed liquor. Solids are maintained in suspension as the mixed liquor circulates around the ditch, typically at a velocity of 0.3 m/sec. Because the mixing energy is conserved in the form of fluid velocity, suspension of solids can be achieved at lower mixing energy inputs than typically required with other basin configurations. A variety of aeration devices can be used in oxidation ditches, including vertical and horizontal mechanical aerators, draft tube aerators, and jet aerators. EAAS designs using more conventional basin configurations are also possible. However, in those cases mixing generally controls the sizing of the oxygen transfer/mixing system, not oxygen requirements.

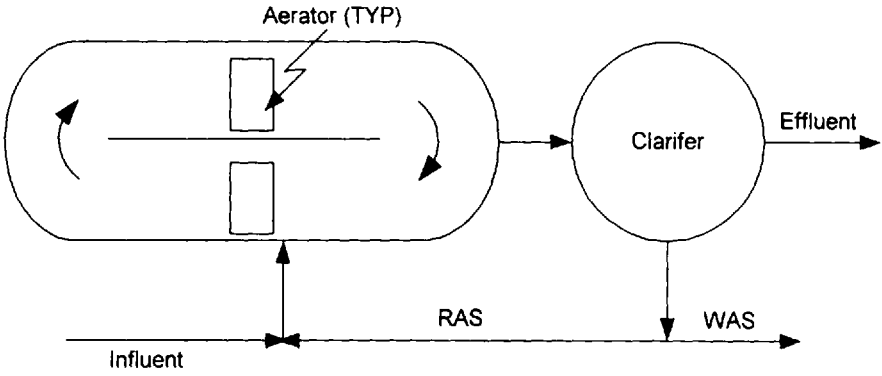


Figure 10.4 Oxidation ditch activated sludge system.

High-Purity Oxygen Activated Sludge. The high-purity oxygen activated sludge (HPOAS) process was developed in the late 1960s and was widely used during the 1970s and early 1980s.[45] As illustrated in Figure 10.5, the bioreactor is staged, enclosed, and provides for cocurrent flow of an oxygen enriched gas phase with the mixed liquor. Influent wastewater and RAS are added only to the first stage, along with oxygen (typically 98% pure). Each stage is a completely mixed cell and three or four are generally used, although as many as six have been utilized. Mixing for solids suspension and oxygen dissolution is provided in each stage. A variety of mechanical devices have been used for this purpose, including slow speed mechanical surface aerators and submerged turbines. The use of high-purity oxygen increases the oxygen partial pressure in the gas of each stage, thereby allowing higher volumetric oxygen transfer rates than are possible with systems using air. Theoretically, the volumetric oxygen transfer rate could be as much as five times higher than the rate with air at atmospheric pressure, but the practical increase is around two to three times. Higher volumetric oxygen transfer rates allow smaller bioreactor volumes to be used, so HRTs are generally in the 2 to 4 hour range. SRTs as low as 1 to 2 days are often used to treat municipal wastewaters, with somewhat longer values being used to treat industrial wastewaters.

Cocurrent flow of mixed liquor and gas results in a gradient in oxygen transfer rates which corresponds to the gradient in process oxygen requirements typical of staged bioreactors, as illustrated in Section 7.2.4. Consequently, oxygen requirements can be effectively met while maintaining high dissolved oxygen concentrations (6

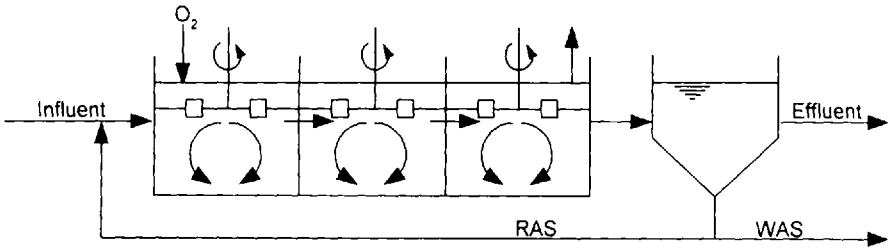


Figure 10.5 High-purity oxygen activated sludge (HPOAS) process.

mg/L or greater) throughout the bioreactor. Cocurrent gas flow also results in efficient use of the applied oxygen; typically over 90% can be transferred into solution. Offgas from the last bioreactor contains unused oxygen, impurities present in the influent oxygen, and carbon dioxide produced by the biological reactions. Retention of carbon dioxide within the system results in depression of the pH; values in the 6.0 to 6.5 range are not unusual unless pH control is practiced. Reduced pH values adversely impact the growth of nitrifying bacteria, as shown in Figure 3.4.

Selector Activated Sludge. The selector activated sludge (SAS) process is a recent development used to control excessive growths of filamentous bacteria,[16,39,78] which can be a nuisance, as discussed in Section 2.3.1. A selector is a portion of an activated sludge system that precedes the main bioreactor, receives the influent wastewater and RAS, and has a high process loading factor. It provides environmental conditions that favor growth of flocculent microorganisms at the expense of filamentous microorganisms, resulting in improved sludge settleability. Selectors use two mechanisms to accomplish microbial selection: kinetic and metabolic.[39,78] Kinetic selection is achieved by imposing a high process loading factor on the biomass, thereby providing a selective advantage for those microorganisms with the ability to take up readily biodegradable substrate at high rates. Metabolic selection is accomplished by controlling the terminal electron acceptor available within the selector. Aerated selectors, which are considered in this chapter, utilize kinetic selection. Anoxic and anaerobic selectors, which are considered in Chapter 11, use metabolic selection, and may or may not also use kinetic selection mechanisms. A selector is generally a small fraction of the entire bioreactor volume and is often staged, as illustrated in Figure 10.6. It may be constructed as a separate structure, or it may simply be a portion of the bioreactor baffled to provide the necessary process loading factor. The bioreactor downstream of the selector may be either completely mixed or plug-flow, although experience suggests better overall performance with plug-flow.

Sequencing Batch Reactor Activated Sludge. Sequencing batch reactor activated sludge (SBRAS) is another recent development.[37] The process, along with its typical operating cycle, is described in Section 7.8.1 and Figure 7.42. Each bioreactor in an SBRAS system is equipped with aeration and mixing equipment and an effluent decanting system.[6] Oxygen transfer may be provided by diffused air (both coarse and fine bubble), jet aerators, mechanical surface aerators (both high and low speed),

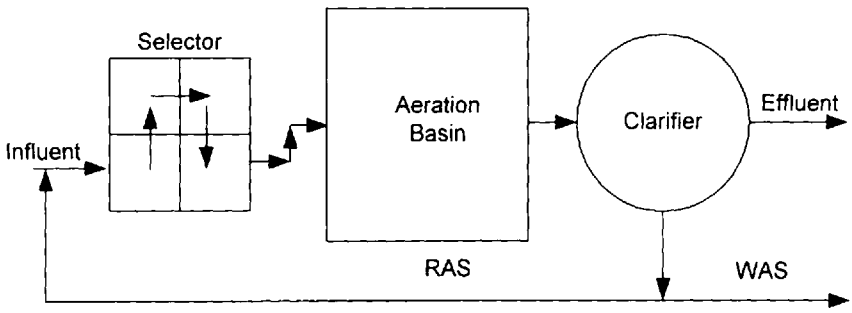


Figure 10.6 Selector activated sludge (SAS) process.

or combined diffused air and mechanical mixing. Decanters of various designs have been used.[6] Microprocessors control the influent flow, aeration, mixing, and effluent decanting functions. Solids are usually wasted during the idle time to take advantage of the increased solids concentration resulting from the settle period. Multiple bioreactors are generally provided to allow at least one to always be in the fill mode. This allows treatment to occur on a continuous basis despite the periodic nature of any individual bioreactor. Most systems also make provision for a continuous operational mode to improve flexibility. In this mode, the idle period is eliminated and influent is added to the reactor on a continuous basis throughout the entire cycle, including settle and decant. Although some deterioration in effluent quality will occur, a significant degree of wastewater treatment can still be achieved even though only one bioreactor may be operational.

10.1.3 Comparison of Process Options

Table 10.1 summarizes the benefits and drawbacks of the eight activated sludge process options described above. Conventional activated sludge and CMAS require similarly sized bioreactors and clarifiers, resulting in similar capital and operating costs. However, sludge settleability is generally better for CAS than for CMAS, due to greater growth of filamentous microorganisms in CMAS systems. Completely mixed activated sludge offers the benefit of greater resistance to inhibitory shock loads.

In comparison to CAS and CMAS, SFAS and CSAS offer the advantage of reduced bioreactor volumes and correspondingly lower capital cost. However, the reduced volumes and costs are accompanied by diminished treatment efficiency, particularly for nitrification. Although listed as a drawback, the ability of SFAS and CSAS systems to maintain stable but elevated effluent ammonia-N concentrations, i.e. to partially nitrify, can be a benefit when only partial ammonia-N oxidation is needed.[34]

Extended aeration activated sludge systems are simple to design and operate, produce a high quality effluent, and produce reduced quantities of more stabilized solids than other, comparable activated sludge systems. This is achieved at the expense of larger and more expensive bioreactors, increased oxygen requirements, and, in some instances, poor solids settling characteristics.

High-purity oxygen activated sludge offers the benefits of small bioreactor volumes, good resistance to excessive decreases in dissolved oxygen (DO) concentration caused by shock loads of biodegradable organic matter, and minimal air emissions. In contrast, they are mechanically complex and incompatible with situations requiring low process loading factors because then the bioreactor would be mixing, rather than oxygen transfer, limited. In such cases high-purity oxygen becomes unnecessary to meet oxygen requirements. Historically, the use of high-purity oxygen in activated sludge systems was claimed to result in an altered biological process with reduced sludge production rates, reduced oxygen requirements, and improved sludge settleability.[45] However, more thorough analysis demonstrated that the biological characteristics of HPOAS systems are the same as other activated sludge systems.[15,51]

Selector activated sludge systems were developed to improve activated sludge settleability, and they have proven capable of doing so.[22,39,78] Selector activated sludge is a relatively new activated sludge option and experience with and knowledge about

Table 10.1 Comparison of Activated Sludge Process Options

Process	Benefits	Drawbacks
Conventional activated sludge (CAS)	• Performance well characterized and predictable • Process and facility design well known • Operational parameters well characterized • Useful in a wide range of applications	• Moderate capital and operating costs • Moderate sludge settleability
Step-feed activated sludge (SFAS)	• Reduced bioreactor requirements compared to CAS • Reduced capital cost compared to CAS	• More complex operation • Reduced nitrification efficiency
Contact stabilization activated sludge (CSAS)	• Reduced bioreactor requirements compared to CAS • Reduced capital cost compared to CAS	• More complex operation • Reduced nitrification efficiency
Completely mixed activated sludge (CMAS)	• Simple design and operation • More resistant to toxic shock loads than others • Useful in a wide range of applications	• Moderate capital and operating costs • Susceptible to the growth of filamentous organisms
Extended aeration activated sludge (EAAS)	• Simple design and operation • Reduced sludge production • Organically stable waste sludge • High quality, well nitrified effluent	• Large bioreactor volumes required • Poor sludge settleability • Higher oxygen requirement
High-purity oxygen activated sludge (HPOAS)	• Small reactor volume • More resistant to organic shock loads than others • Minimal air emissions	• Mechanically complex • Incompatible with low process loading factors
Selector activated sludge (SAS)	• Excellent sludge settleability • Compatible with most activated sludge process options	• Relatively new option, limited experience • Increased mechanical complexity • Increased sensitivity to toxic organic compounds
Sequencing batch reactor activated sludge (SBRAS)	• Simple design and operation • High quality effluent • Process operational characteristics adjustable by adjusting operational cycles	• Discontinuous discharge • Relatively large reactor volumes

it are still being accumulated. It may be more susceptible to inhibitory shock loads because of the high initial process loading factor.

The SBRAS process offers the benefits of simple design and operation, the production of a high-quality effluent, and the capability to adjust process operational characteristics by adjusting operational cycles. However, effluent is removed on a discontinuous basis, which can negatively impact the operation of downstream treatment processes unless the effluent flow is equalized. Relatively large bioreactors are required to accommodate the addition of influent wastewater and the removal of treated effluent.

10.1.4 Typical Applications

Activated sludge is the most widely utilized biochemical operation, with many thousands of operating examples. It has been used to treat nearly every type of wastewater that contains biodegradable organic matter and has been applied to flows ranging up to 5,000,000 m^3/day. Extensive documentation of such applications exist in the literature and in standard references.[25,46,79,80]

Activated sludge is applicable to wastewaters with a wide range of concentrations. As illustrated in Figure 9.2, aerobic suspended growth bioreactors are typically used to treat wastewaters with biodegradable chemical oxygen demand (COD) concentrations up to about 4,000 mg/L. Above that, anaerobic bioreactors are often more cost-effective. Technically, activated sludge can be used to treat wastewaters with biodegradable COD concentrations ranging from about 50 to 10,000 mg/L. The technical constraints derive from sludge settleability and the ability to maintain a suitable SRT. For dilute wastewaters, the rate of generation of new biomass is so low that it is less than the rate at which biomass is lost into the effluent from the clarifier. As a consequence, it is impossible to accumulate biomass in the bioreactor and the necessary SRT cannot be achieved. Attached growth bioreactors, which employ alternate mechanisms for biomass retention, are typically used for these dilute wastewaters.

For high-concentration wastewaters, the constraint relates to the ability of the clarifier to consolidate the MLSS for recycle to the bioreactor. If a wastewater contains a high concentration of biodegradable COD, the concentration of MLSS produced could exceed the thickening capacity of the clarifier. For example, treatment of a wastewater with a biodegradable COD concentration of 20,000 mg/L could result in the production of about 8,000 mg/L of total suspended solids (TSS). This exceeds the typical MLSS concentrations achievable in activated sludge systems, as discussed in Section 9.3.4. Thus, it would not be possible to settle this mixed liquor in the clarifier and produce an even more concentrated RAS stream for recycle to the bioreactor, thereby making it impossible for the SRT to exceed the HRT. Consequently, either a lagoon, within which the HRT and the SRT are the same, or an anaerobic process, which has a lower yield, would have to be used for this application.

Plant size is one of the factors that affects the type of activated sludge process used, particularly for municipal applications. Extended aeration activated sludge and SBRAS are most often used in smaller municipal wastewater treatment plants, those with design flows up to about 20,000 to 60,000 m^3/day. Their simple operation and reliable performance more than offset the increased cost associated with the larger

bioreactors required. Primary clarifiers are also typically not provided to simplify the solids treatment and disposal system. Conventional activated sludge and CMAS are most often used in larger plants, typically with primary clarifiers. The cost savings associated with the smaller bioreactors needed for these processes are sufficient to justify the more intensive operation required. The use of primary clarifiers is an economic issue, not a process issue. Primary clarifiers can be used successfully in association with EAAS processes, whereas CAS or CMAS systems can be successfully designed and operated without primary clarifiers. The smaller bioreactors and reduced offgas volumes associated with HPOAS make this process useful for larger plants located in urban areas where compact designs are needed.

Extended aeration activated sludge is also often used to treat industrial wastewaters that contain slowly biodegradable, xenobiotic, and/or inhibitory materials. The long SRT used with this process is often necessary to treat these difficult to degrade materials. Furthermore, SBRAS is being used increasingly in industrial applications due to its simplicity and the periodic nature of some industrial discharges. The long SRT often used with the SBRAS process is another factor in its selection for these applications.

As discussed in Section 9.3.2, a sufficiently long SRT is needed when nitrification is to be maintained to control effluent ammonia-N concentrations. EAAS and SBRAS systems typically nitrify under all conditions due to the long SRTs associated with them. Conventional activated sludge and CMAS processes can also be designed with an SRT sufficiently long to achieve reliable nitrification. Step-feed activated sludge and CSAS, on the other hand, are not generally used for these applications because it may be difficult to achieve low effluent ammonia-N concentrations, as seen in Sections 7.3 and 7.4. They are used, however, for applications in which partial oxidation of ammonia-N is needed. High-purity oxygen activated sludge is not generally used for nitrification applications due to its low pH and short SRT.

In some instances, a wastewater treatment plant may be designed so that it can be operated in more than one activated sludge mode, depending on the nature of the influent wastewater. Many plants experience significant increases in influent flow during wet weather, either because of the infiltration of rain water into the sanitary sewer system or because the collection system is a combined sewer system. As a consequence, CAS may be used during dry weather to provide the maximum treatment efficiency and the greatest ease of operation, whereas either SFAS or CSAS may be used during wet weather to allow higher hydraulic loadings to be processed while maintaining adequate performance.[92] In fact, due to the reduced pollutant concentrations that often exist during high-flow periods, effluent quality may be no different than that achieved during dry weather using the CAS process.

10.2 FACTORS AFFECTING PERFORMANCE

Many factors affect the performance of activated sludge systems and the more important ones will be discussed in this section. They also affect design and operation, and their impacts will be discussed in those contexts in Sections 10.3 and 10.4.

10.2.1 Floc Formation and Filamentous Growth

Successful operation of activated sludge systems requires development of a flocculent biomass that settles rapidly and compacts properly in the clarifier; failure to do so can lead to process failure. Individual bacteria are colloidal solids, which will not settle in conventional clarifiers. Consequently, they must be aggregated, i.e., flocculated. An ideal activated sludge floc is strong and compact so that it settles rapidly, producing a dense sludge for recycle to the bioreactor and a clear, high-quality supernatant for discharge as treated effluent. Achieving this objective requires the proper proportion of floc-forming and filamentous bacteria.

Activated sludge settling problems can generally be related to inadequate bioflocculation and/or an improper balance between floc-forming and filamentous bacteria. Dispersed growth results from incomplete bioflocculation, leaving a high proportion of individual bacterial cells present in the culture. Because they will not settle, dispersed cells are carried into the effluent, resulting in poor effluent quality. Pin point floc results when bacteria are flocculated, but inadequate filamentous bacteria are present to strengthen the floc particles and make them resistant to mechanical abrasion. As a consequence, the floc particles are easily ruptured and a range of floc sizes exists. The larger flocs settle rapidly in the clarifier, forming a dense sludge for recycle to the bioreactor, but the smaller flocs settle slowly and are carried into the effluent. Filamentous bulking is characterized by excessive quantities of filamentous bacteria, forming floc with an open structure. Upon settling, these floc leave a clear supernatant, but their open structure causes them to settle slowly. Consequently, they compact poorly in the final clarifier, making it difficult to recycle the biomass without excessive RAS flow rates.

Sludge settleability and compaction are often quantified using the sludge volume index (SVI) measurement. It is performed by placing mixed liquor from a bioreactor into a one liter graduated cylinder and measuring the settled volume after 30 minutes of settling.[33] This volume is divided by the initial suspended solids concentration, and the result (with units of mL/g) represents the volume occupied by one gram of settled suspended solids. Table 10.2 summarizes the typical relationship between SVI and activated sludge settling characteristics. An SVI of 150 mL/g is often considered to be the dividing line between a bulking and a non-bulking sludge. In addition to the conventional SVI, several other sludge settleability tests are available, but somewhat different results are obtained with each. Consequently, caution must be exercised when comparing reported SVI values to ensure that comparable sludge settleability measurements were used. Table 10.3 compares the more common

Table 10.2 Relationship Between SVI and Activated Sludge Settling Characteristics

SVI range (mL/g)	Sludge settling and compaction characteristics
<80	Excellent
80–150	Moderate
>150	Poor

Table 10.3 Comparison of Sludge Settleability Indices

Index	Settling device	Low-speed stirring	MLSS concentration
Conventional SVI (SVI)	1 L graduated cylinder	No	Aeration basin MLSS concentration
Mallory SVI (SVI_m)	Mallory settleometer	No	Aeration basin MLSS concentration
Diluted SVI (DSVI)	1 L graduated cylinder	No	Mixed liquor diluted so that settled volume in graduated cylinder is less than or equal to 200 mL/L
Stirred SVI at 3.5 g SS/L ($SSVI_{3.5}$)	1 L graduated cylinder	Yes	Tests at several MLSS concentrations; value at 3.5 g/L obtained by interpolation

Adapted from Jenkins, et al.[19]

sludge settleability tests. The diluted SVI (DSVI) or the stirred SVI at 3.5 g/L ($SSVI_{3.5}$) produce the most reproducible results.[19,42] The SVI is correlated with activated sludge settling characteristics and the measured SVI can be used to determine allowable secondary clarifier suspended solids loading rates.[19]

Flocculation Theory. Flocculation is the first step in building the compact, readily settleable floc necessary to optimize activated sludge settling characteristics. Its absence results in dispersed growth, as discussed above. The mechanisms of bioflocculation are poorly understood, in spite of numerous studies, attesting to the complexity of the phenomenon. One point upon which there is agreement, however, is that exocellular polymers (ECP) are central to the aggregation of individual bacteria into floc particles.[17,63,73] Several types of ECPs are involved in bioflocculation. Polysaccharides have received the most study and are generally thought to be of major importance.[17,73] Nevertheless, it now appears that proteins also play an important role.[36] Possible sources of ECP are formation by microbial metabolism, release by cell lysis, and the wastewater itself.[73] Evidence for the role of the wastewater itself comes from the observation that flocculation in activated sludge systems treating industrial wastewaters, which contain a limited number of organic compounds, is often more difficult than in systems treating domestic wastewaters, which contain a rich variety of large molecular weight organic materials. Nevertheless, the most important sources of ECP are metabolism and cell lysis. Exocellular polymers are produced by both the protozoa[18] and the bacteria[10,53] found in activated sludge, although the relative importance of the two is unknown. Evidence for the role of protozoa comes from the observation that good bioflocculation and low effluent suspended solids concentrations are correlated with their presence. It is also known that floc-forming bacteria produce ECPs that are involved in aggregation,[17] although the factors that cause ECP formation are still a matter of speculation.[10]

The mere presence of ECPs is not sufficient to ensure bioflocculation. Rather, the milieu within which the activated sludge is growing also has an impact. For

example, both ionic strength[75,86] and divalent cations[36,50,73] play important roles. Bacteria are negatively charged. Consequently, the ionic strength must be sufficiently large to allow individual cells to approach closely enough together for bridging by the ECPs to occur, but not so large as to cause deflocculation. Furthermore, ionic strength will influence the conformation of the ECP. One study suggested that an ionic strength on the order of 0.005 to 0.050 resulted in optimum floc stability.[86] Because both the cell surface and ECPs are negatively charged, divalent cations are thought to act as bridges between the two, allowing aggregation to occur. Consequently, the proper level of divalent cations is essential. One study found that the minimum concentration of calcium and magnesium required to obtain an activated sludge with good settling properties was in the range of 0.7 to 2.0 meq/L of each (14 to 40 mg/L of calcium and 8 to 24 mg/L of magnesium).[36] However, the actual concentration required in a particular facility will depend on the ionic strength of the wastewater. Furthermore, the ratio of divalent to monovalent cations is also important because when that ratio is less than 0.5, deterioration of the settling characteristics results.[36] This is thought to be due to the competition between divalent and monovalent cations for binding sites on the cell surfaces and the ECP.

Empirical observations suggest that the SRT must exceed a minimum value to achieve bioflocculation. This observation is consistent with both the role of protozoa and ECP production by bacteria. Protozoa generally have a lower maximum specific growth rate than bacteria, and the SRT below which bioflocculation does not occur could correspond to the minimum SRT required for protozoan growth. Alternately, the observation that the amount of ECP per unit of biomass increases and approaches a maximum as the SRT is increased[14,63] suggests that the requirement for a minimum SRT could represent a balance between the rate of ECP production by the floc-forming bacteria and the rate of generation of new surface area by bacterial growth. Consequently, although ECP is produced on a continuous basis, at short SRTs the rate of generation of new bacteria exceeds the rate of ECP production and bioflocculation is incomplete. Only by reducing the rate of generation of bacterial surface area relative to the rate of ECP production by increasing the SRT can effective bioflocculation be achieved.

Figure 10.7 presents data illustrating the impact of SRT on bioflocculation in a pilot CMAS system receiving a synthetic wastewater consisting of glucose, yeast extract, and inorganic nutrients.[8] The proportion of activated sludge suspended solids that did not settle under quiescent conditions is plotted as a function of the SRT. A high proportion of the activated sludge solids (10% to 30%) would not settle when the process was operated at SRTs between 0.25 and 0.5 days, but that fraction was significantly reduced when the process was operated at an SRT of 1 day and it remained low as the SRT was increased up to 12 days. Settling velocity also increased with increasing SRT, as illustrated in Figure 10.8.[8] Microscopic analysis of the biomass produced at each operating SRT provided the results presented in Table 10.4.[8]

Based on data such as these, an SRT of at least 3 days is generally recommended for achieving good bioflocculation in activated sludge systems. Nevertheless, many municipal wastewater treatment plants have been successfully designed and operated at SRTs as low as one day.[25,62] Apparently, the presence of microorganisms in municipal wastewaters, which alters the relationship between SRT and specific growth rate as discussed in Section 5.2.3, and the presence of ECPs, as discussed

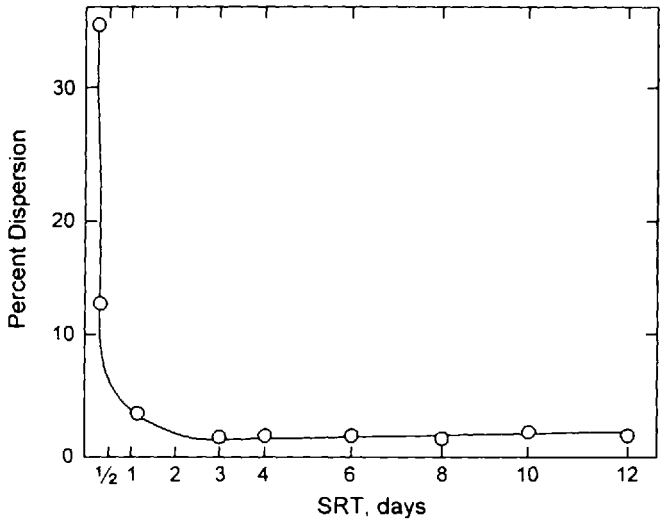


Figure 10.7 Effect of SRT on the amount of dispersed growth in activated sludge effluent. (From J. J. Bisogni and A. W. Lawrence, Relationship between biological solids retention time and settling characteristics of activated sludge. *Water Research* 5:753–763, 1971. Copyright © Elsevier Science Ltd. Reprinted with permission.)

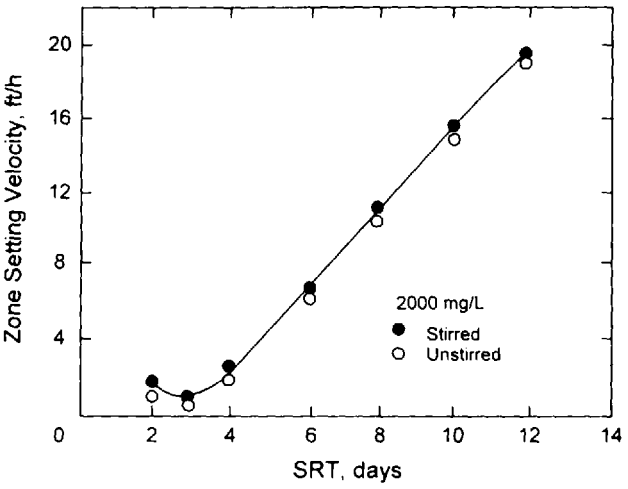


Figure 10.8 Effect of SRT on the settling velocity of activated sludge. (From J. J. Bisogni and A. W. Lawrence, Relationship between biological solids retention time and settling characteristics of activated sludge. *Water Research* 5:753–763, 1971. Copyright © Elsevier Science Ltd. Reprinted with permission.)

Table 10.4 Characteristics of the Biomass Produced in the CMAS Reactors of Bisogni and Lawrence[8]

SRT range (days)	Character of solids
0.25–2	Predominantly dispersed growth
2–9	Well formed average size floc of low to medium density
9–12	Pinpoint floc and irregularly shaped floc particles of low density that looked as though they had broken loose from larger floc particles (deflocculated)

Source: From Ref. 8.

above, result in a reduction in the SRT required to achieve good bioflocculation. This reduction is reflected in the typical SRT operating ranges presented in Figure 9.3.

Bioflocculation forms the microstructure of activated sludge floc.[39] The resulting flocs are relatively weak, however, and can readily be broken into smaller particles by turbulence. Consequently, if bioflocculation is the only mechanism of floc formation, a variety of particle sizes will be present, ranging from large flocs that settle rapidly to small particles that will not settle. A sludge of this type will compact well, but will leave a poor quality, turbid supernatant. A filament backbone is necessary to provide strength to the floc, resulting in a rapidly settling floc that is also strong enough to produce a clear supernatant.

Role of Filamentous Bacteria. The relative proportion of floc-forming and filamentous bacteria in floc determines its macrostructure,[61] as illustrated in Figure 10.9.[39] In an ideal activated sludge floc (Figure 10.9a), the filaments provide a strong backbone around which the well flocculated bacteria grow. This results in a large, dense, compact floc that settles rapidly and compacts well in the clarifier. A clear supernatant is also produced since few small, slowly settleable particles are present. An activated sludge composed of such flocs will have a low SVI, typically less than 100 mL/g. Figure 10.9b illustrates pin-point floc as described above, consisting primarily of individual floc particles with little or no filamentous bacteria present to provide floc strength. Figure 10.9c illustrates a filamentously bulking sludge. Excessive numbers of filamentous bacteria are present, causing them to extend beyond the activated sludge flocs. The resulting flocs are strong and produce a clear supernatant when they settle. However, because they are large and because the filamentous bacteria extend beyond them, the floc particles settle slowly and compact poorly. The slow settling rate negatively impacts the capacity of the clarifier, and the poor compaction results in a dilute settled sludge for recycle to the bioreactor.

Figure 10.10 illustrates that the extension of filaments beyond the floc particles results in poor sludge settling characteristics as measured by the SVI test.[50] It also illustrates why an SVI of 150 mL/g is used as an indicator of the onset of filamentous bulking. Beyond that point small increases in extended filament length lead to large increases in SVI, indicating significant deterioration of sludge settling and compaction characteristics. Although not shown by the figure, both impacts of filament

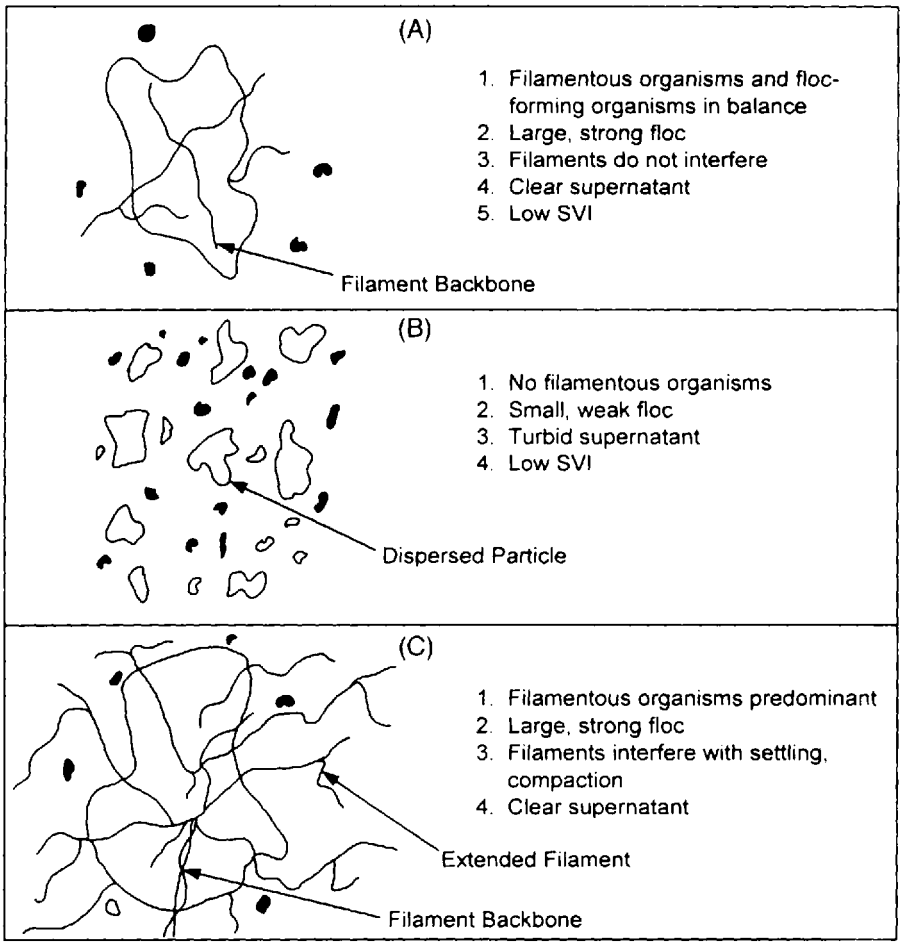


Figure 10.9 Effect of filamentous growth on activated sludge structure: (A) ideal, non-bulking activated sludge floc; (B) pinpoint floc; (C) filamentous bulking activated sludge. (From D. Jenkins, M. G. Richards, and G. T. Daigger, *The Causes and Cures of Activated Sludge Bulking and Foaming*, 2nd ed., Lewis Publishers, Ann Arbor, Michigan, 1993. Copyright © Lewis Publishers. Reprinted with permission.)

growth, expansion of the activated sludge floc and filament extension beyond the floc particle, negatively affect sludge settling and compaction.

The conceptual model presented in Figure 10.9 allows a more complete analysis of the results presented in Figures 10.7 and 10.8, and in Table 10.4. When the CMAS systems were operated at SRTs between 0.25 and 2 days, they produced a large proportion of dispersed growth, corresponding to failure to attain adequate bioflocculation. At SRTs between 2 and 9 days, well formed, average size flocs of low to medium density were produced, reflecting a balance between floc-forming and filamentous bacteria. Operation at SRTs between 9 and 12 days produced irregularly shaped pin point floc, which typically result from an inadequate proportion of filamentous bacteria. Thus, it seems likely that the continuous increase in settling ve-

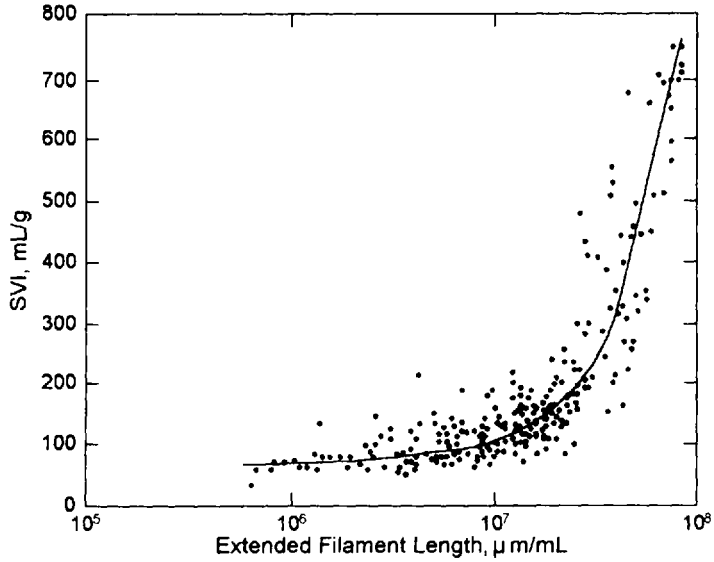


Figure 10.10 Effect of extended filament length on SVI. (From J. C. Palm, D. Jenkins, and D. S. Parker, Relationship between organic loading, dissolved oxygen concentration and sludge settleability in the completely-mixed activated sludge process. *Journal, Water Pollution Control Federation* **52**:2484–2506, 1980. Copyright © Water Environment Federation. Reprinted with permission.)

locity with SRT in Figure 10.8 was a result of a continuous decrease in the proportion of filamentous bacteria. Although observations such as these have led to the recommendation that SRTs be maintained in a moderate range, say from 3 to 15 days, operation outside of that range is entirely feasible, provided an appropriate balance is maintained between the floc-forming and filamentous bacteria.

Types of Filamentous Bacteria and Their Control. The preceding discussion emphasizes the need to control the relative populations of floc-forming and filamentous bacteria in activated sludge systems. As seen in Section 2.3.1, many types of filamentous microorganisms can exist in activated sludge systems, and the most common are listed in Table 2.1. Consequently, efforts to control their growth require knowledge of the particular type of filamentous bacteria that could potentially be present. Fortunately, the conditions that favor the growth of many of them are known, as shown in Table 2.2. This link between a specific environmental condition and a particular type of filamentous organism can be used to identify and correct activated sludge settling problems, as suggested by Table 10.5.

Individual types of filamentous bacteria have high affinities for different limiting nutrients, allowing them to out-compete floc-forming bacteria for them. Some filamentous bacteria have a high affinity for dissolved oxygen, some have a high affinity for readily biodegradable organic matter, and others have a high affinity for nitrogen and phosphorus. Furthermore, as indicated in Table 2.2, the filamentous bacteria *Thiothrix*, *Beggiatoa*, and 021N can also obtain energy from the oxidation of hydrogen sulfide, which provides a further advantage for them when it is present. Low pH will encourage the growth of filamentous fungi. Consequently, the key to

Table 10.5 Proposed Filamentous Organism Groups

Group I—Low DO Aerobic Zone Growers	
Features	• readily metabolizable substrates • low DO • wide SRT range
Organisms	*S. natans*, Type 1701, *H. hydrossis*
Control	• aerobic, anoxic, or anaerobic selectors • increase SRT • increase aeration basin DO concentration
Group II—Mixotrophic Aerobic Zone Growers	
Features	• readily metabolizable substrates, especially low molecular weight organic acids • moderate to high SRT • sulfide oxidized to stored sulfur granules • rapid nutrient uptake rates under nutrient deficiency
Organisms	Type 021N, *Thiothrix spp.*
Control	• aerobic, anoxic, or anaerobic selectors • nutrient addition • eliminate sulfide and/or high organic acid concentrations (eliminate septicity)
Group III—Other Aerobic Zone Growers	
Features	• readily metabolizable substrates • moderate to high SRT
Organisms	Type 1851, *N. limicola spp.*
Control	• aerobic, anoxic or anaerobic selectors • reduce SRT
Group IV—Aerobic, Anoxic, Anaerobic Zone Growers	
Features	• grow in aerobic, anoxic and anaerobic systems • high SRT • possible growth on hydrolysis products of particulates
Organisms	Type 0041, Type 0675, Type 0092, *M. parvicella*
Control	Largely unknown but: • maintain uniformly adequate DO in aerobic zone and stage the aerobic zone

Adapted from Jenkins, et al.[9]

controlling the growth of filamentous organisms is to control the concentration of the growth limiting nutrient. It is generally desirable to use permanent filament control methods such as those listed in Table 10.5. However, in certain instances it may be more economical to use nonspecific toxicants such as chlorine or hydrogen peroxide to control filament growth. The use of such techniques is discussed in Section 10.4.3.

The general objective of the activated sludge process is to remove biodegradable organic matter. This is achieved by creating conditions in which it is the limiting substance. Consequently, the presence of filamentous bacteria with a high affinity for nitrogen, phosphorus, or dissolved oxygen (Groups I and II) indicates that these nutrients may be limiting bacterial growth. The solution to problems caused by excessive growths of these filamentous bacteria is addition of the limiting nutrient. For nitrogen and phosphorus, residual concentrations of approximately 1 mg/L are desired. For DO, the required residual concentration is a function of the process loading factor, as illustrated in Figure 10.11.[50] This relationship exists because DO concentrations are measured in the bulk solution while bacterial growth occurs within the floc particle. As the process loading factor is increased, the biomass uses oxygen at a faster rate and a higher bulk DO concentration is required to ensure the penetration of DO throughout the floc particle.

Control of some filamentous bacteria results from an understanding of the relative growth kinetics of filamentous and floc-forming bacteria. Figure 10.12 illustrates the typical relationship. In general, for a particular substrate, floc-forming bacteria have higher $\hat{\mu}$ and K_s values than filamentous bacteria. In other words, the floc formers can grow faster when the substrate concentration is high, but the filamentous bacteria have a higher affinity for the substrate and can grow faster when its concentration is low. For example, if the substrate concentration is S_1 in Figure 10.12, the specific growth rate of the floc-forming bacteria is higher than that of the fila-

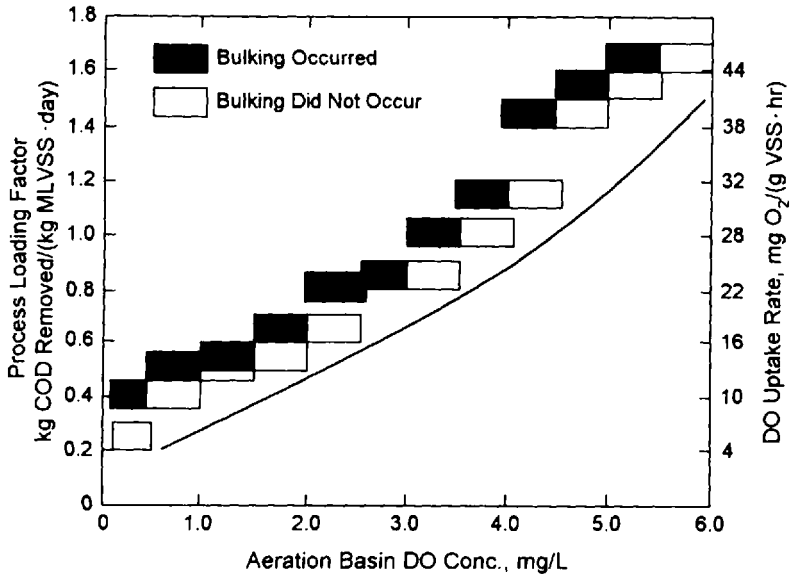


Figure 10.11 Combinations of process loading factors and aeration basin dissolved oxygen concentrations where bulking and nonbulking sludges occur in CMAS system. (From J. C. Palm, D. Jenkins, and D. S. Parker, Relationship between organic loading, dissolved oxygen concentration and sludge settleability in the completely-mixed activated sludge process. *Journal, Water Pollution Control Federation* **52**:2484–2506, 1980. Copyright © Water Environment Federation. Reprinted with permission.)

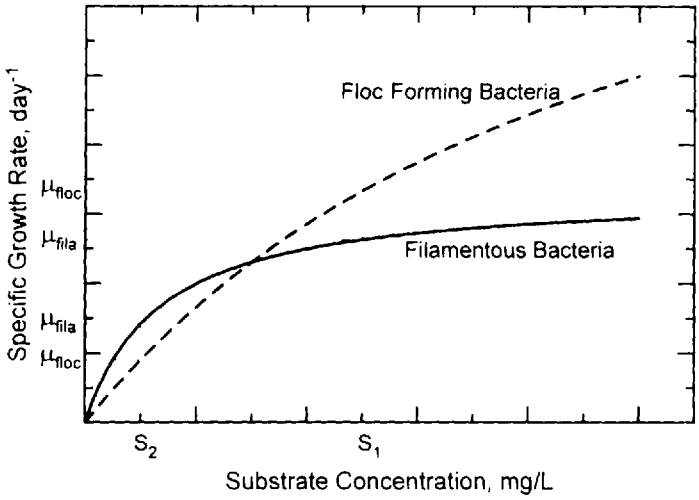


Figure 10.12 Comparison of the growth kinetics of floc forming and filamentous bacteria.

ments and the floc formers will out-compete the filaments. If, on the other hand, the substrate concentration is S_2, the specific growth rate of the filaments is higher than that of the floc formers and the filaments will out-compete the floc formers. This illustrates the characteristics of an environment that favors the growth of filamentous bacteria: the substrate must be supplied continuously in a manner that results in a low concentration. Continuous substrate supply is required so that biomass growth can occur. The residual concentration must be low to provide a competitive advantage for the filamentous organism. Said simply, filamentous organisms are good scavengers; they consume substrates more efficiently than floc-forming bacteria. Consequently, filaments proliferate under conditions that favor scavenging organisms. Such conditions typically occur in CMAS and sometimes occur in CAS and SFAS.

One approach for controlling the growth of filamentous bacteria with a high affinity for readily biodegradable organic matter is to reconfigure the bioreactor to create a substrate concentration gradient. This is the concept of kinetic selection, mentioned in Section 10.1.2. The goal is to produce a substrate concentration at the inlet to the bioreactor that favors the growth of floc-forming bacteria at the expense of the filamentous bacteria, as illustrated in Figure 10.12 by the concentration S_1. This can be accomplished by providing highly plug–flow conditions within the bioreactor. As illustrated in Figure 10.13,[10] the SVI of activated sludge is influenced by the residence time distribution in the bioreactor as characterized by the equivalent number of tanks in series. SVIs are generally low for bioreactors with a flow pattern characterized as five tanks in series or more. In a process with a low equivalent number of tanks in series, the residual readily biodegradable substrate concentration in the first tank will be relatively low, similar to S_2 in Figure 10.12. This low concentration favors the growth of filamentous bacteria. As the number of equivalent tanks in series is increased, the readily biodegradable substrate concentration in the first tank increases until it approaches S_1, which favors the growth of floc-forming bacteria.

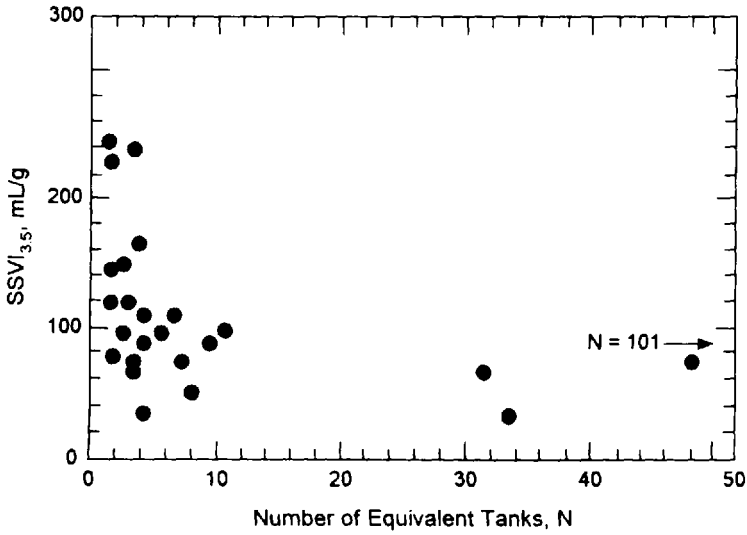


Figure 10.13 Relationship between activated sludge settling characteristics as expressed by the SVI and the bioreactor hydraulic characteristics as expressed by the equivalent number of tanks in series. (From E. J. Tomlinson and B. Chambers. The effect of longitudinal mixing on the settleability of activated sludge. Technical Report TR 122, Water Research Centre, Stevenage, England, 1978. Reprinted by permission of the Water Research Centre.)

The desired conditions in the initial equivalent bioreactor have been identified by calculating the process loading factor in that tank and correlating the activated sludge SVI with it. The process loading factor for the initial tank is calculated by using the mass flow rate of biodegradable organic matter in the process influent and the mass of biomass in the equivalent initial tank. As discussed in Section 5.1.5, the process loading factor is linearly related to specific growth rate. Thus, such correlations identify the specific growth rate in the initial equivalent tank required to produce a residual biodegradable substrate concentration comparable to S_1 in Figure 10.12. One such correlation is presented in Figure 10.14[60] for the full-scale wastewater treatment plants considered in Figure 10.13. The results indicate that, for these plants, an initial process loading factor of 2 kg five-day biochemical oxygen demand (BOD_5)/(kg volatile suspended solid [VSS] · day) produces a sufficiently high specific growth rate to encourage the growth of floc-forming bacteria over filamentous bacteria. The use of correlations such as these to design SAS systems will be discussed in Section 10.3.4.

Another approach for controlling the growth of filaments with a high affinity for readily biodegradable organic matter is by metabolic selection. Metabolic selection is accomplished by eliminating dissolved oxygen as a terminal electron acceptor in the selector and either providing nitrate-N to create an anoxic selector or excluding both dissolved oxygen and nitrate-N to create an anaerobic selector. Some strains of floc-forming bacteria are able to take up readily biodegradable organic matter under either anoxic or anaerobic conditions, whereas most filamentous bacteria cannot. Thus, control of the terminal electron acceptor provides a powerful selective pressure against filamentous bacteria. Metabolic selectors are discussed in Chapter 11.

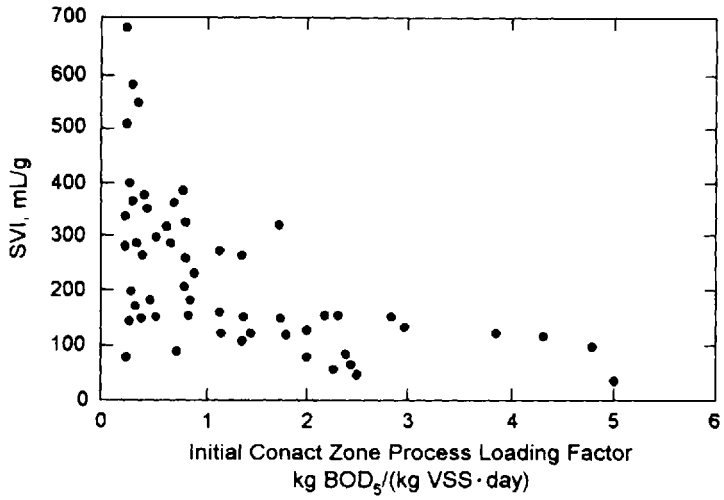


Figure 10.14 Effect of initial contact zone process loading factor on SVI. (From E. J. Tomlinson, Bulking—a survey of activated sludge plants. Technical Report TR35, Water Research Centre, Stevenage, England, 1976. Reprinted by permission of the Water Research Centre.)

Less understanding exists of the Group IV filamentous bacteria listed in Table 10.5. They are typically observed in processes with long SRTs, especially those with completely mixed bioreactors.[9,28] Some evidence suggests that the growth of at least some of these organisms is encouraged by cyclic low DO concentrations, which can occur with oxygen transfer systems such as mechanical surface aerators.[54] Their growth is controlled by maintaining plug-flow conditions and uniform DO concentrations throughout the bioreactor. The occurrence of these organisms in biological nutrient removal systems is discussed further in Chapter 11.

10.2.2 Solids Retention Time

It should be clear by now that SRT is a primary factor determining the performance of activated sludge systems. The theoretical impacts of the SRT on the concentration of soluble constituents for a variety of activated sludge processes were discussed in Chapters 5 through 7. The role of SRT in achieving bioflocculation was discussed in Section 10.2.1. Finally, the general factors that must be considered in the selection of the SRT for all biochemical operations were presented in Section 9.3.2. The purpose of this section is to emphasize certain common observations concerning the effect of SRT on activated sludge process performance.

Figure 9.3 illustrated that the design and operating SRT for activated sludge systems treating biogenic organic matter is generally controlled by bioflocculation, not the removal of soluble substrate. To demonstrate this point, Figure 10.15 shows the effect of SRT on the effluent soluble COD from CMAS bioreactors that received a feed with a soluble COD of 375 mg/L.[8] Soluble COD decreased rapidly as the SRT was increased from 0.25 to 1 day. At an SRT of 2 days, the COD reached a minimum value, which was maintained until the SRT exceeded 8 days. Beyond 8

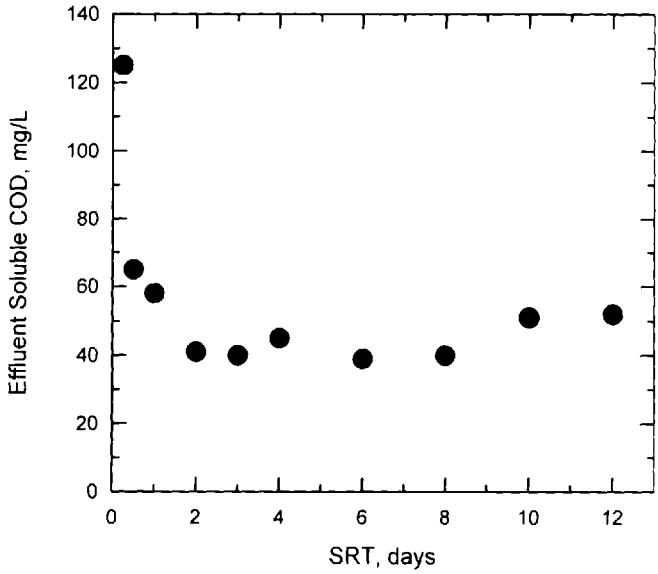


Figure 10.15 Effect of SRT on effluent soluble COD from a CMAS system. Data from Bisogni and Lawrence.[8]

days the COD increased again, probably as a result of the production of soluble microbial products.[21,55] Since those products are resistant to biodegradation, the biodegradable COD was essentially constant for SRTs in excess of 2 days. Figure 10.7 revealed, however, that the SRT for the same bioreactors had to exceed 2 days to obtain effective bioflocculation. Thus, once the SRT was long enough for effective bioflocculation to occur, further increases had only minor effects on soluble substrate removal. Consequently, for easily degradable substrates like those in domestic wastewaters, selection of the SRT is almost always controlled by factors other than soluble substrate removal.

The observation above helps to explain another common observation, that bioreactor configuration often has no observable impact on soluble effluent quality for many applications.[68] For example, although we saw in Section 7.2.2 that CAS systems theoretically have lower effluent soluble substrate concentrations than CMAS systems, little difference is observed in practice. Consequently, for treatment of easily biodegradable substrate, the choice of the activated sludge process variation is usually driven by factors other than soluble effluent quality.

Longer SRT values may be required for the treatment of industrial wastewaters containing more difficult to degrade materials, which may also be inhibitory to biological growth. This is illustrated in Figure 10.16, which presents data from a CMAS process treating a plastics manufacturing wastewater.[12] Note, however, that SRT had relatively little effect on effluent quality over the operating range for the facility. This is typical in that most activated sludge systems are designed and operated over a range of SRTs where little difference in the concentration of soluble organic matter in the effluent is observed. When the kinetic parameters for a particular application have been determined, Eq. 9.7 can be used to calculate the SRT required to achieve

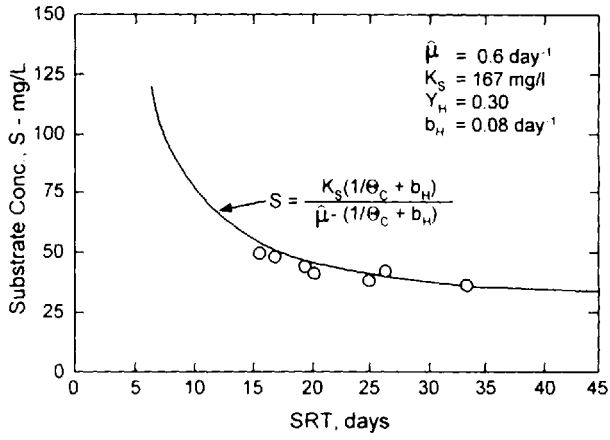


Figure 10.16 Effect of SRT on effluent COD from a full-scale CMAS reactor. (Adapted from Campbell and Rocheleau.[32])

a specified target effluent quality from a CMAS process. Because the substrate is slowly biodegradable, however, bioreactor configuration is likely to have more of an effect than it did with domestic wastewater. Consequently, consideration of other activated sludge process options will require pilot studies or simulations to arrive at a design SRT value, or the application of safety factors, much as is required for nitrification.

While operation at relatively short SRTs is possible for many wastewaters, it can result in high excess solids production rates because short SRTs do not permit hydrolysis of particulate organic matter,[47] as discussed in Section 6.2. This will result in a corresponding reduction in process oxygen requirements, consistent with the COD mass balance described by Eq. 3.18 and articulated in the third guiding principle of Section 9.1. The economic consequences of increased solids production and reduced oxygen requirement must be evaluated for each application.

Activated sludge systems are often designed to operate at long SRTs to achieve stabilization of entrapped particulate organic matter and heterotrophic biomass or to biodegrade xenobiotic and other slowly biodegradable organic compounds. This can lead to limited growth of filamentous bacteria resulting in pin-point floc, and measures should be taken to minimize its formation. One technique is to use completely mixed reactors, which will encourage the growth of filaments. Another is careful control of the DO concentration to encourage the growth of low DO filaments. Both techniques require care on the part of the designer and operator to prevent excessive growth of filaments, with their attendant problems. Nevertheless, it is important to recognize that pin-point floc is not an inevitable consequence of long SRT operation. On the other hand, inadequate bioflocculation occurs with some industrial wastewaters and is exacerbated by operation at long SRT. Thus, experience with particular wastewaters should be used to determine whether long SRT operation is possible and to establish the measures necessary for achieving adequate bioflocculation and growing sufficient filaments to build a stable floc.

Although nitrification may not be required at a particular facility, it will occur any time the SRT exceeds the minimum SRT for the nitrifying bacteria. As illustrated

in Figure 9.4, nitrification can occur at SRTs as low as 1 to 2 days when the mixed liquor temperature exceeds 20°C. To avoid the operational difficulties resulting from an inadequate oxygen supply, the oxygen transfer system should be designed for the oxidation of both carbon and nitrogen if the SRT could possibly be long enough to allow nitrification.

Both organic substrate removal and nitrification may be required in some situations, and an SRT must be selected that will allow both effluent quality goals to be met. As discussed in Section 9.3.2, this requires the design SRT to exceed the minimum SRT of the most slowly growing microorganisms by a sufficient degree to have stable performance. Nitrifying bacteria are usually the most slowly growing bacteria in activated sludge systems and thus they generally control the design in this situation. Selection of the design SRT for nitrification is considered in Section 10.3.1.

The nature and capacity of the excess solids processing system is influenced strongly by the SRT of an activated sludge process. Low SRTs, on the order of 1 to 3 days, require a solids processing system with the capability to continuously receive and process large and variable quantities of WAS. At such SRTs a substantial portion of the MLSS inventory is wasted each day. Consequently, little capacity exists within the activated sludge process to absorb variations in inventory caused by variations in process loadings or by interruptions in waste solids processing. Instead, the capacity to absorb such variations must be built into the solids processing system. In contrast, an activated sludge process operating at an SRT of 5 to 10 days generally has sufficiently large bioreactors to temporarily accumulate excess solids. Consequently, the solids processing system serving it need not have as much capacity to absorb variations as a system serving an activated sludge process with a low SRT. Finally, activated sludge systems with very long SRTs, such as EAAS, often practice periodic solids wasting, and the excess solids processing system should be sized accordingly. In order to foster process stability, any given incident of solids wasting should not result in a decrease in the bioreactor MLSS concentration of more than 10%. For a process operating at an SRT of 20 days, solids need to be wasted only every other day to meet this criterion.

10.2.3 Mixed Liquor Suspended Solids Concentration

Activated sludge processes can be successfully operated over a wide range of MLSS concentrations. In fact, the MLSS concentration itself does not affect the performance of the process; rather, performance is controlled by the mass of MLSS present. Furthermore, as expressed by the fifth guiding principle in Section 9.1, once the SRT for a biochemical operation has been selected, the mass of biomass in it becomes fixed. As discussed in Section 9.3.4, selection of the MLSS concentration (and consequently, the bioreactor volume) requires consideration of the interactions of the bioreactor with the final clarifier and the mixing/aeration system. Although a wide range of MLSS concentrations is possible, practical designs typically limit the MLSS concentration to between 2,000 and 5,000 mg/L.

A minimum MLSS concentration is necessary to allow the development of a flocculent biomass. If the process is operated at MLSS concentrations below this value, bioflocculation will be poor, entrapment of particulate organic matter will be inadequate, and a good settling activated sludge floc will not develop. The result

will be a turbid, poor quality effluent. Although actual experience must define the minimum MLSS concentration for a particular process, they typically fall between 500 and 1,000 mg/L. Thus, the maximum range in MLSS concentration typically used in activated sludge systems is between 500 and 5,000 mg/L.

10.2.4 Dissolved Oxygen

The primary effect of the DO concentration on activated sludge performance is on the growth of filamentous bacteria. Guidance on the selection of the appropriate DO concentration to control low DO filamentous bulking is presented in Figure 10.11. Many books recommend the maintenance of a minimum DO concentration of 2 mg/L in activated sludge processes. As indicated by Figure 10.11, however, 2 mg/L may not be sufficient in some cases and may be excessive in others. Rather, the required DO concentration depends on the process loading factor and specific oxygen uptake rate (SOUR) in a given bioreactor. Although it is considered prudent to design oxygen transfer systems to achieve a DO concentration of at least 2 mg/L, many activated sludge processes operate quite satisfactorily at lower DO concentrations, thereby achieving significant power cost savings.[82] It all depends on the process loading factor imposed.

10.2.5 Oxygen Transfer and Mixing

For economic reasons, the equipment used to transfer oxygen in activated sludge systems also provides the turbulence necessary to maintain solids in suspension. This results in constraints on process design and operation. One concern is the volumetric power input, Π, which is the power supplied per unit volume, either directly by mechanical aerators or indirectly by compression of the air for diffused aeration systems. As discussed in Section 9.3.4 and illustrated in Figure 9.6, the volumetric power input must be sufficiently large to keep solids in suspension, but not so large as to cause excessive floc shear. Another concern is the maximum volumetric oxygen transfer rate, i.e., the mass of oxygen transferred per unit time per unit volume, that can be attained economically with the equipment available.

As discussed in Sections 5.1.4 and 9.4, once the SRT has been selected for an activated sludge process, the oxygen requirement (RO) is fixed. This, in turn, fixes the amount of power that must be expended to supply the oxygen. Although this book does not address the design of oxygen transfer systems, it is necessary to approximate the power required to ensure that the volumetric power input is in the feasible region as indicated by Figure 9.6. For mechanical aeration systems, the required power input can be approximated from:

$$P = \frac{RO}{\eta_P} \tag{10.1}$$

where P is the power input in kW, RO is the oxygen requirement in kg/hr, and η_P is the in-process energy efficiency for the mechanical aeration system in kg O_2/(kW·hr). The value of η_P typically ranges from 0.7 to 1.2 kg O_2/(kW·hr).[82] For diffused air systems, the process air requirement can be calculated from the following dimensional expression:

$$Q = \frac{6.0\ RO}{\eta_O} \tag{10.2}$$

where Q is the air flow rate in m^3/min, RO is the oxygen requirement in kg/hr, and η_O is the field oxygen transfer efficiency expressed as the percent of the oxygen in the air actually transferred to the liquid. The value of η_O depends on the nature of the diffuser and the depth at which the air is released.[82] It typically lies in the range of 6 to 15%, but with values as low as 4% and as high as 80% observed under unusual circumstances. The volumetric power input required to meet the oxygen requirement is obtained by dividing P or Q by the bioreactor volume.

As illustrated in Figure 9.6, the lower limit on the volumetric power input, Π_L, is determined by the need to maintain solids in suspension. For mechanical aeration systems, this input, $\Pi_{L,P}$, is around 14 kW/1000 m^3. The manufacturer of a particular aeration device should be consulted for a more exact value. For spiral roll diffused aeration systems, a minimum air input rate, $\Pi_{L,Q}$, of 20 m^3/(min · 1000 m^3) is generally required. For full floor coverage aeration systems the corresponding value is 37 m^3/(min · 1000 m^2 of bioreactor floor area).[46,71,79,82] If these requirements exceed the volumetric power input required to meet the oxygen requirements, the power input required for solids suspension must be provided. This results in increased power usage and higher DO concentrations than required purely for process reasons. Consequently, the minimum volumetric power input can be used with the power required to meet the minimum oxygen requirement to establish the upper feasible bioreactor volume, V_U, in m^3:

$$V_U = \frac{1000\ P}{\Pi_{L,P}} \text{ or } \frac{1000\ Q}{\Pi_{L,Q}} \tag{10.3}$$

The volumetric power input must also be less than the value that causes excessive shear of the activated sludge floc, Π_U. Excessive shear will disperse a portion of the solids into a poorly settleable form that will not be removed in the clarifier and will, therefore, pass into the effluent. For mechanical aeration systems, the typical maximum volumetric power input, $\Pi_{U,P}$, is 60 kW/1000 m^3, while for diffused aeration systems $\Pi_{U,Q}$ is approximately 90 m^3/(min · 1,000 m^3).[46,71,79,82] To avoid excessive floc shear, the reactor volume should be no smaller than the lower feasible bioreactor volume associated with Π_U, which has been designated $V_{L,FS}$ to emphasize that it is determined by the floc shear criterion. Its value in m^3 is given by:

$$V_{L,FS} = \frac{1000\ P}{\Pi_{U,P}} \text{ or } \frac{1000\ Q}{\Pi_{U,Q}} \tag{10.4}$$

The necessity to indicate that the minimum feasible volume calculated with Eq. 10.4 comes from the floc shear criterion stems from the fact that there is a maximum volumetric rate at which oxygen can be transferred in activated sludge systems, and it also imposes a lower limit on the bioreactor volume. This maximum rate is device specific, and the manufacturer of the particular equipment of interest must be contacted to determine the appropriate maximum value for a given application. For example, for a floor coverage diffused air system, the limitation may be caused by the maximum number of diffusers that can be placed in the bioreactor per unit of floor area. Nevertheless, for the types of oxygen transfer systems typically

used today, the maximum volumetric oxygen transfer rate that can be achieved economically on a sustainable basis is around 100 g $O_2/(m^3 \cdot hr)$, which is equivalent to 0.10 kg $O_2/(m^3 \cdot hr)$. During short-term transients this rate can sometimes be pushed to 150 g $O_2/(m^3 \cdot hr)$, but typical mechanical aeration equipment should not be counted on to deliver oxygen at such a high rate on a sustainable basis because of excessive wear. Thus, if such high transfer rates are needed, specialized high-efficiency transfer systems must be used. As a result of this constraint, the lower limit on bioreactor volume based on oxygen transfer, V_{LOT}, should also be calculated:

$$V_{LOT} = \frac{RO}{0.10} \tag{10.5}$$

where RO is expressed as kg O_2/hr and the V_{LOT} is in m^3. The smallest allowable reactor volume is given by the larger of V_{LHS} and V_{LOT}.

Although floc shear has been correlated with the volumetric power input, a more fundamental parameter describing flocculation in general is the root mean square velocity gradient, G, and thus it is often used when examining flocculation in activated sludge systems. For diffused air systems, G (sec^{-1}) can be calculated as:

$$G = \left(\frac{Q \cdot \gamma \cdot h}{60 \cdot V \cdot \mu_w}\right)^{0.5} \tag{10.6}$$

where Q is the airflow rate in m^3/min, γ is the liquid specific weight in N/m^3, h is the liquid depth above the diffuser in m, V is the bioreactor volume in m^3, μ_w is the absolute viscosity in $N \cdot sec/m^2$, and 60 is the conversion from minutes to seconds. Note the direct relationship between the volumetric air flow rate and G. For mechanical aerators, G can be calculated as:

$$G = \left(\frac{1000\ P}{V \cdot \mu_w}\right)^{0.5} \tag{10.7}$$

where P is the aerator power input in kW ([W] = N·m/sec) and V and μ_w have the same units as above. Again, note the direct relationship between the volumetric power input and G.

For diffused air systems, effluent suspended solids concentrations have been correlated with G, with G values in excess of 125 sec^{-1} causing values to rise.[24] A G value of 125 sec^{-1} corresponds closely to a volumetric air input rate of around 20 $m^3/(min \cdot 1000\ m^3)$ and a volumetric power input of around 14 kW/1000 m^3, which are the minimum considered necessary to keep MLSS in suspension, i.e., II_1. At the other extreme, the G values associated with II_0 are on the order of 270 sec^{-1}, and above that value, excessive floc destruction occurs. Between those extremes, there is a continual rise in effluent suspended solids concentration with increasing G,[24] and a designer can use the calculated G value to get an idea about likely effluent suspended solids concentrations. A prudent designer will anticipate having effluent suspended solids concentrations above the minimum attainable unless provisions are made for reflocculation prior to clarification.

For mechanically aerated facilities, another factor that must be considered is the location of the aerator relative to the discharge to the final clarifier.[21] This is because mechanical aerators have very high localized velocity gradients. Consequently, in such systems the type and layout of the aerators has a stronger effect on effluent suspended solids concentrations than does the average G based on the overall volumetric power input.

Sheared floc can be reflocculated[76]. Thus, if the value of G exceeds 125 sec^{-1}, which will be true for most facilities, effluent quality can be improved by passing the activated sludge through a reflocculation zone prior to entering the final clarifier. A reflocculation time of 20 minutes at a G value of about 15 sec^{-1} may be appropriate.[76,83] The same thing can be accomplished in CAS systems by using low mixing energy in the latter stages where the oxygen requirement is low, but care must be exercised to keep all solids in suspension.

10.2.6 Nutrients

As discussed is Section 3.8.2, adequate nutrients are required to allow balanced growth of biomass in biochemical operations. Failure to provide them can have several consequences. For example, low nutrient concentrations can favor the growth of filamentous bacteria over floc-formers, as discussed in Section 10.2.1, resulting in a poor-settling activated sludge. More severe nutrient deficiencies can result in unbalanced growth of all bacteria, leading to the production of exocellular slime. In severe cases, the slime gives the activated sludge a jelly-like consistency, resulting in a sludge that settles slowly and compacts poorly.[39,78] Virtually no liquid–solids separation will occur in such cases.

Procedures to calculate nutrient requirements are described in Sections 5.1.4 and 9.4.1, and Tables 3.3 and 9.2 provide guidance as to the quantities needed. Experience suggests that such calculations should consider only the inorganic nitrogen and phosphorus available in the influent wastewater.[30] Organic nitrogen and phosphorus will be released into solution and become available to the biomass as organic matter is biodegraded. However, the rate of biodegradation of some of these materials can be relatively slow, making the associated nutrients unavailable to heterotrophic bacteria metabolizing readily biodegradable organic matter. Thus, limiting nutrient concentrations can occur within the process, even though the total mass of nutrients may be adequate. Consistent maintenance of residual inorganic nitrogen and phosphorus concentrations throughout the process of approximately 1 mg/l should be adequate.

10.2.7 Temperature

Temperature affects the performance of activated sludge systems as a result of its impact on the rates of biological reactions. Procedures for estimating the magnitudes of its effects are presented in Section 3.9. Two additional factors must be considered: the maximum acceptable operating temperature, and the factors that affect heat loss and gain by the process.

The maximum acceptable operating temperature for typical activated sludge systems is limited to about 35° to 40°C, which corresponds to the maximum temperature for the growth of mesophilic organisms. Even short-term temperature var-

iations above this range must be avoided since thermal inactivation of mesophilic bacteria occurs quickly. Successful operation can also be obtained if temperatures are reliably maintained above about 45° to 50°C, since a thermophilic population will develop, provided that thermophilic bacteria exist with the capability to degrade the wastewater constituents. Unacceptable performance will result for temperatures between about 40° and 45°C due to the limited number of microorganisms that can grow within this range. These considerations are particularly important for the treatment of high temperature industrial wastewaters.

One of the factors that affects heat gains in biological processes is the production of heat as a result of biological oxidation. As discussed in Section 2.4.1, the growth of bacteria requires that a portion of the electron donor be oxidized to provide the energy needed for biomass synthesis. Energy is also needed for cell maintenance. This oxidation and subsequent use of the energy results in the conversion of that energy into heat. Although this may seem surprising at first, it is directly analogous to the release of energy that occurs when material is burned; the only difference is the oxidation mechanism. The amount of heat released in the biooxidation of carbonaceous and nitrogenous material is directly related to the oxygen utilized by the process. For each gram of oxygen used, 3.5 kcal of energy are released.[40,43] Since 1 kcal is sufficient energy to raise the temperature of one liter of water 1°C, the impact of this heat release depends on the wastewater strength. For example, a typical domestic wastewater requires only one gram of oxygen for each 10 liters treated, therefore the temperature rise would be only 0.35°C, a negligible amount. On the other hand, it is not unusual for an industrial wastewater to require one gram of oxygen for each liter treated, in which case the temperature rise would be 3.5°C. This could be quite significant, particularly if the wastewater itself is warm.

Other heat gains and losses occur in biological systems. Heat inputs to the system include the heat of the influent wastewater, solar inputs, and mechanical inputs from the oxygen transfer and mixing equipment. Heat outputs include conduction and convection, evaporation, and atmospheric radiation. Models for accurately calculating heat balances across suspended growth bioreactors have been developed.[5,60,66] They are discussed in Section 14.2.5.

If experience or a heat balance suggest the likelihood of unsatisfactorily high or unstable temperatures, the bioreactor should be configured to maximize heat losses. Measures to accomplish this include the use of relatively large basins to increase the HRT, shallow sidewater depths to increase basin surface area, above ground construction to maximize conductive and convective heat loss, and the selection of an oxygen transfer device, such as mechanical surface aeration, which maximizes heat loss. Another solution is to provide mechanical cooling of the process influent or the bioreactor contents. Designs such as HPOAS or facilities using deep bioreactors with diffused aeration will have minimal heat loss and should be avoided in this situation. In fact, they may require mechanical cooling even when large heat inputs are not expected.

Although heat gain is not generally a concern during treatment of municipal wastewaters, heat loss can be, depending on the type of oxygen transfer systems used and the bioreactor HRT.[11] For example, submerged oxygen transfer systems, such as diffused aeration, have low heat losses, whereas mechanical surface aerators have high losses. This difference may affect the geographic region in which a par-

ticular type of oxygen transfer device can be used. When needed, heat loss can be minimized through proper facility design.

10.3 PROCESS DESIGN

10.3.1 Overview

The basic approach to the design of suspended growth biochemical operations is presented in Chapter 9. In this chapter, we focus that approach on activated sludge systems and examine the types of decisions that are required in their design. As discussed in Chapter 9, design is an iterative procedure and can take place at several levels of sophistication, depending on the information available to the designer. At the simplest level, in which little information is available, a preliminary design can be accomplished by applying the guiding principles articulated in Table 9.1. This approach is illustrated in Section 9.4.1, and the steps involved are summarized at the end of that section.

The next level, stoichiometric-based design, uses the simple model of Chapter 5, as extended in Section 9.4.2, and requires that specific information be available about the nature of the wastewater and the parameter values describing its biodegradation. That information sometimes can be obtained from historical records at a facility that is to be expanded, or from facilities treating similar wastewaters when a new system is being designed. In this case, it will usually be necessary to convert the information from traditional measurements, such as BOD_5, into the more descriptive measurements, such as biodegradable COD, in use today. The procedures for doing this are presented in Sections 8.6 and 8.7. In other cases, treatability studies will be required to provide the necessary information. The procedures for conducting them are presented in Sections 8.2 and 8.3. Stoichiometric-based design provides quantitative information about the mass of MLSS to be contained in the activated sludge process, the steady-state oxygen requirement, and the mass of excess solids to be disposed of daily. The equations are for a single completely-mixed bioreactor, such as in CMAS, but as guiding principle No. 4 in Table 9.1 states, the calculated values are applicable to any of the activated sludge variants. Thus, they can be used as the basis for decisions about bioreactor configuration and the distribution of MLSS and oxygen supply within the bioreactors. These decisions require heuristic approaches, which are presented in the material that follows. It should be recognized, however, that all of the calculated values are based on the daily average flow and substrate concentration entering the facility, even though wastewater treatment facilities are subject to diurnally variable inputs, as illustrated in Figure 6.2. Thus, unless the activated sludge process is to be preceded by equalization, the impact of those variations on the design must be considered. This also requires the application of heuristically derived approaches.

As discussed in Section 9.4.3, the third level of design is simulation-based design. It is the most precise way to consider the impact of dynamic loadings on activated sludge systems and requires the use of a suitable dynamic model, such as activated sludge model (ASM) No. 1, and a computer code that implements it. Several such codes are listed in Table 6.4; all are simulation programs, not design programs. This means that the designer must choose a particular activated sludge process and provide the sizes of the component bioreactors as input to the programs. Sim-

ulations are then run to examine the performance of the process and the output is used to assess its acceptability. Should the design be unacceptable, the process must be modified and another simulation run. This procedure must be repeated in a logical manner until an acceptable design is arrived at, with the output providing needed information about the distribution of oxygen, MLSS, etc. Thus, the designer must have already accomplished a basic process design before beginning the simulations. This can be done with either of the first two approaches. However, characterization of the wastewater constituents and the parameters describing treatment is much more complex than that for the other design levels. The techniques for performing that characterization are described in Section 8.5.

The primary focus of this section is on stoichiometrically-based design. There are several reasons for doing this. First, preliminary design based on the guiding principles is discussed in sufficient detail in Section 9.4.1 to allow its application. It need not be expanded upon here. Second, as discussed above, simulation-based design requires the designer to provide a basic process flow diagram as input to the simulation program. The most effective way of doing this is by stoichiometrically-based design. Third, execution of a stoichiometrically-based design requires the designer to understand the most important aspects of activated sludge design. Thus, it provides an excellent framework within which to present them.

As discussed in Sections 8.6 and 8.7, many types of measurements have been used to express the concentrations of wastewater and activated sludge constituents. However, to allow us to focus on the decisions to be made during design and not distract the reader with multiple unit conversions, we use only biodegradable COD as the measure of organic substrate and TSS as the measure of MLSS. We have chosen the former because of its fundamental importance as a measure of available electrons, and the latter because of its widespread use in the profession. Sections 8.6 and 8.7 provide the information needed to convert between unit systems. In addition, to provide continuity in the examples, we use a standard wastewater throughout that is typical of domestic wastewater after primary treatment. It is the one in Table E8.4, and the translation between the conventional characterization in the top of the table and the more complete characterization in the bottom is explained in Example 8.6.1. Finally, we must emphasize that the calculations presented in this book are meant only to illustrate the procedures and decisions the process designer must make. The numerical results should not be considered to be typical of the application of biochemical operations to any particular real wastewater.

10.3.2 Factors to Be Considered During Design

Selection of the Appropriate Process Option. The selection of a particular activated sludge process is based on many considerations, including the wastewater characteristics, effluent quality goals, facility capital and operating costs, facility operational objectives, other processes at the facility, and the desires of the owner. Consequently, a full discussion of the selection of the process option is beyond the scope of this book. Nevertheless, a few generalizations are possible. CAS is popular for treatment of domestic wastewater because of its proven reliability and ability to achieve high effluent quality, including complete nitrification. However, for treatment of industrial wastewaters containing inhibitory organic compounds, CMAS has advantages, although special consideration must be given to the settling properties of

the resulting biomass. If a wastewater contains a high percentage of readily biodegradable organic matter and no inhibitory materials, SAS may be required to control filamentous sludge bulking. On the other hand, if the wastewater contains a high percentage of colloidal organic matter that can be removed by entrapment in the biofloc, and the readily biodegradable substrate can be removed at an SRT shorter than that associated with good bioflocculation, then CSAS and SFAS have distinct advantages relative to system volume, although full nitrification may be difficult to achieve. For small communities wishing to minimize the number of operational personnel and types of unit operations on site, EAAS is popular. Finally, when little space is available and the emission of volatile organic compounds must be minimized, situations commonly associated with industrial facilities, HPOAS is often used. For additional information on the selection of the process option the reader should consult design manuals, such as MOP No. 8.[79]

Selection of the Solids Retention Time. The effects of SRT on activated sludge performance are discussed in Section 10.2.2, while the factors that must be considered during its selection are covered in Section 9.3.2. Consideration of the information in those sections makes it clear that selection of the SRT is a multifaceted decision requiring input from a number of sources. An important consideration of course, is the SRT needed to meet the required effluent quality. If a single completely mixed bioreactor is to be used, such as in CMAS or EAAS, the SRT required to produce a particular effluent COD is given by Eq. 9.7:

$$\Theta_c = \frac{K_S + S_S}{S_S(\hat{\mu}_H - b_H) - K_S b_H} \tag{9.7}$$

As discussed in Section 9.4.2, the parameters associated with autotrophs can be substituted for the heterotrophic parameters and S_{NH} can be substituted for S_S to determine the SRT required to achieve a required ammonia-N concentration through nitrification. In either case, the SRT calculated from Eq. 9.7 would not necessarily produce the required effluent quality. There are several reasons for this. One is uncertainty in the kinetic parameters, influent characteristics, natural variability in the microbial community, and other factors. Such factors cause statistical variability in the effluent quality as illustrated in Figure 9.9.[58] Thus, the calculated SRT must be multiplied by an appropriate safety factor, ζ_{Θ_c}, to account for that uncertainty, or the value of S_S (or S_{NH}) must be chosen with the uncertainty in mind.

Because Eq. 9.7 represents only steady-state performance, another factor that must be considered is the impact of loading variations on process performance. Loading variations take two forms, day to day variations caused by seasonal and other events, and diurnal loading variations within a day, such as those illustrated in Figure 6.2. Discussion of the factors that go into decisions about design loadings and the use of equalization to dampen them is beyond the scope of this book. However, it is important to recognize that seasonal loading variations must be considered by the designer in selecting the SRT. With regard to typical diurnal loading variations, examination of the curves labeled continuous stirred tank reactor (CSTR) in Figure 7.4 reveals that COD removal is much less subject to their effects than is nitrification. Consequently, the safety factor for uncertainty is often sufficient to guard against unsatisfactory effluent organic substrate concentrations as a result of diurnal loading variations, but not against high effluent ammonia-N concentrations. Rather, an ad-

ditional safety factor is required to account for the effects of ammonia-N loading variations on nitrification. It is the peak load safety factor, ς_{PL}:

$$\varsigma_{PL} = \frac{(F \cdot S_{NHO})_{Peak}}{(F \cdot S_{NHO})_{Avg}} \tag{10.8}$$

The choice of the period over which $(F \cdot S_{NHO})_{Peak}$ is defined, i.e., peak diurnal within the average day, maximum month, maximum week, etc., depends on the nature of the discharge standard that must be met, the plant process flow diagram, etc., and as such, is also beyond the scope of this book. Consequently, the reader should consult other sources, such as the U. S. EPA Nitrogen Control Manual[72] for more information.

Finally, nitrification is very sensitive to DO concentration, as discussed in Section 6.3.1 and illustrated in Figure 6.7. That sensitivity can be expressed by a double Monod expression, such as Eq. 3.46, which is used in ASM No. 1 (Table 6.1). Since DO was not included in Eq. 9.7, it can be included by defining another safety factor, ς_{DO}, that is the reciprocal of the DO term in Eq. 3.46:

$$\varsigma_{DO} = \frac{K_{O,A} + S_O}{S_O} \tag{10.9}$$

A similar safety factor does not need to be applied to organic substrate removal because the half-saturation coefficient for DO for heterotrophic bacteria is sufficiently small to make Eq. 10.9 approach a value of 1.0 at typical bioreactor DO levels.

Considering all of these safety factors, the required value of the SRT, $\Theta_{c,r}$, can be determined from the computed SRT, Θ_c, as:

$$\Theta_{c,r} = \Theta_c \cdot \varsigma_T \cdot \varsigma_{PL} \cdot \varsigma_{DO} \tag{10.10}$$

For organic substrate removal, ς_{PL} and ς_{DO} are normally set at 1.0, as discussed above, simplifying the expression. However, the full expression is typically used for nitrification. Application of Eq. 10.10 results in a very conservative design for nitrification because the safety factors are multiplicative. Consequently, application of simulation-based design techniques, which can explicitly account for dynamic conditions, often allows reductions in the SRT required to achieve adequate nitrification performance from a CMAS system.

Application of Eq. 10.10 to nitrification in a CAS system, which behaves like a plug–flow or tanks-in-series system, would result in selection of a longer SRT than is actually necessary. This can be seen in Figure 10.17 where the effluent ammonia-N concentration is plotted as a function of SRT for activated sludge systems with various numbers of CSTRs in series. The curves in the figure were developed by simulation with ASM No. 1, using the parameter values in Table 6.3. The DO concentration was set at 4.0 mg/L so that it would not be a factor. To illustrate the impact of bioreactor configuration, consider a situation in which the desired ammonia-N concentration is 1.0 mg/L as N. Examination of the figure reveals that the calculated SRT for a CMAS system (N = 1) would be 4.4 days, whereas the SRT required for a CAS system (N = 9) would be only 2.36 days. Thus, if the SRT calculated with Eq. 9.7 (which is for a CMAS system) were substituted into Eq. 10.10 to determine the required SRT for a CAS system, it is clear that the CAS system would be badly overdesigned. Furthermore, the overdesign becomes worse

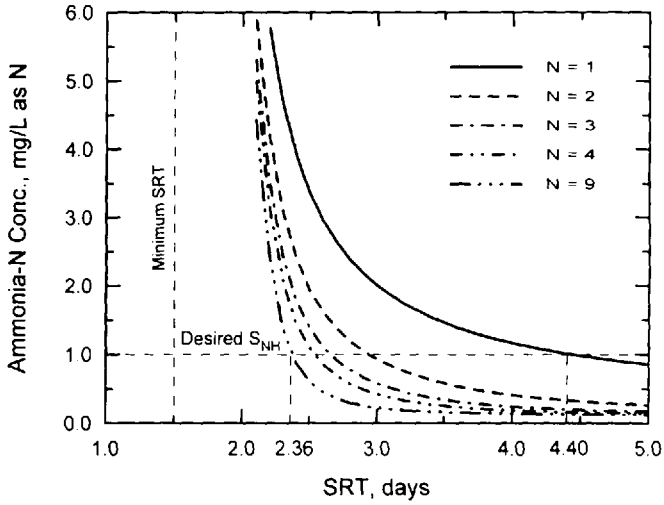


Figure 10.17 Effect of SRT and the hydraulic characteristics of an activated sludge bioreactor, as expressed by the equivalent number of tanks in series, N, on the effluent ammonia-N concentration as simulated with ASM No. 1. The parameter values used are listed in Table 6.3.

the lower the desired ammonia-N concentration. Consequently, a different approach is used for CAS and other systems that approach plug–flow.

The approach to selection of the required SRT for nitrification in CAS systems is based on another observation from Figure 10.17; when a system behaves in a plug–flow manner, the SRT needed to achieve a low effluent ammonia-N concentration is only slightly larger than the minimum SRT. This suggests a convenient way of determining the required SRT for a CAS system. Since the SRT associated with a given effluent ammonia-N concentration from a CAS system cannot be calculated directly for substitution into Eq. 10.10, designers use the minimum SRT instead, recognizing that application of the multiple safety factors will make the required SRT sufficiently long to achieve the desired effluent quality in a reliable manner. Consequently, for nitrification in CAS, and similar, systems:

$$\Theta_{c,r} = \Theta_{c,min} \cdot S_T \cdot S_{PF} \cdot S_{DO} \tag{10.11}$$

The minimum SRT for nitrification can be estimated with Figure 9.4, or it can be calculated with Eq. 5.16 if the appropriate kinetic parameters are known.

$$\Theta_{c,min} = \frac{K_{NH} + S_{NH0}}{S_{NH0}(\hat{\mu}_A - b_A) - K_{NH} \cdot b_A} \tag{5.16a}$$

The equation has been modified to include the autotrophic parameters and thus is renumbered as Eq. 5.16a. The influent ammonia-N concentration is the appropriate nitrogen concentration to use in Eq. 5.16a because organic nitrogen is unlikely to have been metabolized by the heterotrophic bacteria and made available at short SRTs. Regardless of the technique used to determine the minimum SRT, the value should be based on the coldest wastewater temperature at which nitrification is re-

quired. The application of Eq. 10.11 provides a much more realistic required SRT than Eq. 10.10 for CAS and other plug–flow systems.

Example 10.3.2.1

A CAS system is to be designed to fully nitrify, i.e., produce an effluent ammonia-N concentration less than 1.0 mg/L, at a temperature of 15°C when the DO concentration in the bioreactor is 2.0 mg/L. The wastewater has been sufficiently characterized to allow the safety factor for uncertainty to be set to 1.0. The characterization also revealed that the peak to average ammonia-N loading for the wastewater is 1.75 and that the half-saturation coefficient for DO for nitrifiers is 0.75 mg/L. Determine the required SRT for the design.

Since a CAS system is to be used, Eq. 10.11 is the appropriate expression with which to calculate the required SRT. The value of the minimum SRT can be obtained from Figure 9.4, which gives a value of 3.5 days. The peak load safety factor, ζ_{PL}, has a value of 1.75. The safety factor for uncertainty, ζ_U, has a value of 1.0. The DO safety factor can be calculated with Eq. 10.9:

$$\zeta_{DO} = \frac{0.75 + 2.0}{2.0} = 1.375$$

Thus,

$$\Theta_{x,r} = (3.5)(1.0)(1.75)(1.375) = 8.4 \text{ days}$$

Application of Eqs. 9.7 and 10.8 to selection of the SRT for a CMAS system producing an effluent ammonia-N concentration of 1.0 mg/L as N results in a value of 17.3 days. Thus, it can be seen that use of CAS results in a smaller SRT, and thus a smaller system size.

Consideration of the Effects of Temperature. Because temperature affects the kinetic parameters describing biological reactions, as described in Section 3.9, it has important impacts on process design. Three of these are illustrated in Figure 5.11. First, effluent quality is worst at the coldest temperature. Thus, the parameter values used in the selection of the SRT should always reflect the lowest sustainable temperature anticipated in the bioreactor. This is particularly important when nitrification is required because of the extreme sensitivity to temperature exhibited by the maximum specific growth rate coefficient for nitrifying bacteria. Second, at SRTs encountered in practice, excess biomass production will be greatest at the lowest temperature. This has two important impacts; it implies that excess solids handling and processing systems should be designed for winter operation, and it suggests that the mass of MLSS in the system will be greatest then as well. Consequently, the value of $X_{M,T} \cdot V$ for use in bioreactor sizing should be calculated for the lowest sustainable temperature anticipated in the bioreactor. Finally, Figure 5.11 shows that the oxygen requirement will be greatest at the highest operating temperature. Consequently, the oxygen transfer system must be designed for summer operation. With regard to this last point, it is important to recognize that the occurrence of nitrification should be checked for summer temperatures, even if it will not occur in winter. The minimum SRT for nitrification should be calculated with Eq. 5.16 using temperature corrected parameter values, or it should be estimated with Figure 9.4. If the design SRT exceeds that value, provision must be made for supplying the additional needed oxygen

since it may equal that required for organic substrate removal, as seen in Section 6.3.2. On the other hand, for domestic wastewater, the nitrifiers generally will not make a significant contribution to the mass of MLSS in the system, and need not be considered. However, such a generalization cannot be made about industrial wastewaters; each of them is unique.

The most commonly used temperature adjustment technique for the kinetic and stoichiometric parameters characterizing activated sludge is Eq. 3.95:

$$k_1 = k_2 \cdot \theta^{(T_1 - T_2)} \tag{3.95}$$

where k represents any parameter. Generally the reference temperature, T_2, is 20°C, and that is the case with the parameters used herein. Typical values of the temperature coefficient θ are discussed in Section 3.9.2. Values selected for use in the examples of this chapter are given in Table E10.1.

Consideration of the Effects of Transient Loadings. Figure 6.3 illustrates two important points about the effect of typical diurnal loadings on an activated sludge process; the oxygen requirement is influenced quite strongly, whereas there is little impact on the mass of biomass present. This means that bioreactor sizing, which is based on the value of $X_{M,T} \cdot V$, can be done on the basis of average loads. Design of the oxygen transfer system, on the other hand, and the impact of that system on the mixing energy input, must be based on the expected peak loading.

Additional insight into the sizing of the oxygen transfer system comes from further examination of Figure 6.3. The figure shows the oxygen consumption associated only with carbon oxidation, since the maximum specific growth rate coefficient for autotrophs was set to zero during the simulation. As discussed in Section 6.2.2, the use of soluble substrate causes a more severe transient response than the use of particulate substrate because the latter must be hydrolyzed, which is a slow reaction. Nevertheless, the hydrolysis reactions are rapid enough to cause some of the particulate substrate applied during peak loading periods to be used. Thus, the transients in both readily and slowly biodegradable substrate must be considered when estimating peak oxygen requirements. Because of the complexities of the reactions involved, dynamic simulation is the only truly accurate way to assess the transient oxygen requirement in an activated sludge process. Nevertheless, it would be ad-

Table E10.1 Temperature Correction Factors

Parameter	θ
$\hat{\mu}_H$	1.08
b_H and $b_{1,H}$	1.04
K_S and Y_H	1.00
$\hat{\mu}_A$	1.11
b_A and $b_{1,A}$	1.04
K_{NH}	1.14
Y_A	1.00
k_h	1.08
k_a	1.08

vantageous to have a way to approximate the peak oxygen requirement for a single tank system like CMAS.

During a short-term transient loading, biodegradable organic matter will be oxidized to synthesize new cell mass, but little additional decay will occur because the decay rate is proportional to the active biomass concentration, which changes little during the transient. Consequently, the additional oxygen requirement associated with the short-term increase in loading will be proportional to $(1 - Y_{H,T} i_{O\,XB,T})$, which is equivalent to $(1 - Y_H)$. Thus, the fractional transient increase in oxygen requirement will be less than the fractional increase in the biodegradable organic matter loading.

Simulations conducted using wastewater characteristics and parameters similar to those in Tables 6.6 and 6.3, respectively, demonstrated that transient peak oxygen requirements correspond to oxidation of all of the additional readily biodegradable organic matter applied, but only a portion of the additional slowly biodegradable organic matter applied.[2] This was true for a broad range of SRTs and load peaking factors. The load peaking factor is the peak mass loading divided by the average mass loading. Furthermore, the fraction of the additional slowly biodegradable organic matter oxidized, $f_{XS,H}$, decreased as the load peaking factor increased. Based on these considerations, the transient state oxygen requirement for the growth of heterotrophic bacteria ($RO_{H,TS}$) may be estimated as follows:

$$RO_{H,TS} = (1 - Y_{H,T} i_{O\,XB,T})[\Delta(F \cdot S_{SO}) + f_{XS,H}\Delta(F \cdot X_{SO})] \qquad (10.12)$$

where $\Delta(F \cdot S_{SO})$ is the transient increase in the loading of readily biodegradable organic matter above the average loading, and $\Delta(F \cdot X_{SO})$ is the transient increase in the loading of slowly biodegradable organic matter above the average. The value of $f_{XS,H}$ will generally range from 0.5 to 1.0, with smaller values being associated with larger transient loading increases.[2]

The peak oxygen requirement due to heterotrophic activity is the sum of the steady-state oxygen requirement, as given by Eq. 9.13, and the transient-state oxygen requirement as given by Eq. 10.12. The oxygen transfer rate to the system must be capable of meeting both requirements, in addition to any oxygen utilization by the autotrophic bacteria.

The transient increase in the oxygen requirement due to nitrification is more complicated for a number of reasons. The first is that SRT has a much stronger effect than it does on the heterotrophic oxygen requirement.[2] At SRTs that are above the minimum SRT for the autotrophic bacteria, but below that required for full nitrification at steady-state, transient loadings will have no effect on the rate of nitrification, and hence on the oxygen consumption associated with it, because nitrification will already be occurring close to its maximum rate. At long SRTs where full nitrification can occur even during the transient, the increase in oxygen consumption rate will be proportional to the increase in loading, just as it is for heterotrophic bacteria, although the proportion oxidized will be different. At SRTs that are just sufficient to give full nitrification at steady-state, the ammonia nitrogen concentration may rise sufficiently during the transient to allow the rate of nitrification to reach its maximum value, thus causing the oxygen consumption rate to rise, but by a smaller amount than the loading increase. A second complicating factor is that not all nitrogen in the influent is in a form that is available to the autotrophic bacteria. Some will be in the form of biodegradable organic nitrogen. This nitrogen will become available only as the

organic matter containing it undergoes decomposition. Based on the arguments in the preceding paragraph, we would expect all of the nitrogen associated with the readily biodegradable substrate to be made available as ammonia-N, but only the fraction $f_{XS,H}$ of that associated with the slowly biodegradable substrate. Finally, some of the ammonia-N entering during the transient will be incorporated into the extra biomass formed during the transient as the additional organic matter is removed. This, too, must be accounted for.

For the situation in which the SRT is sufficiently long to allow full nitrification during the peak loading period, the transient state oxygen requirement associated with the autotrophic bacteria, $RO_{A,TS}$ can be calculated with an equation analogous to Eq. 10.12:

$$RO_{A,TS} = (4.57 - Y_{A,T}i_{O\,XB,T})[\Delta(F \cdot S_N)_{a,TS}] \tag{10.13}$$

in which $\Delta(F \cdot S_N)_{a,TS}$ is the transient increase in ammonia-N available to the autotrophic bacteria. It is given by:

$$\Delta(F \cdot S_N)_{a,TS} = \Delta(F \cdot S_{NHO}) + \Delta(F \cdot S_{NSO}) + f_{XS,H}\Delta(F \cdot X_{NSO}) - 0.087Y_{H,T}i_{O\,XB,T}[\Delta(F \cdot S_{SO}) + f_{XS,H}\Delta(F \cdot X_{SO})] \tag{10.14}$$

where $\Delta(F \cdot S_{NHO})$, $\Delta(F \cdot S_{NSO})$, and $\Delta(F \cdot X_{NSO})$ are the transient increases in the loadings of ammonia-N, soluble biodegradable organic nitrogen, and particulate biodegradable organic nitrogen above the average. The negative term in Eq. 10.14 accounts for the additional use of nitrogen associated with synthesis of the heterotrophic bacteria during the transient organic loading. The result from Eq. 10.14 must be added to the steady-state autotrophic oxygen requirement to determine the peak autotrophic requirement for this situation.

For the situation in which nitrification is not complete during the transient, causing the ammonia-N concentration to rise high enough to allow the autotrophic bacteria to grow at their maximal rate, the maximum autotrophic oxygen utilization rate, $RO_{A,max}$, can be calculated from a modified form of Eq. 5.32:

$$RO_{A,max} = \left[\left(\frac{4.57 - Y_{A,T}i_{O\,XB,T}}{Y_{A,T}i_{O\,XB,T}}\right)\hat{\mu}_A + (1 - f_D)b_A\right](X_{B,A,T} \cdot V)i_{O\,XB,T}\left(\frac{S_O}{K_{A,O} + S_O}\right) \tag{10.15}$$

The mass of autotrophic bacteria in the system, $X_{B,A,T} \cdot V$, should be that associated with the average loading on the system. The last term is included because of the sensitivity of the autotrophic nitrifying bacteria to the dissolved oxygen concentration and the likelihood of that concentration falling during the transient. Consequently, the DO concentration used should be the concentration expected during the transient. In situations where this condition occurs, the peak autotrophic oxygen requirement will just be $RO_{A,max}$, because that value cannot be exceeded. Consequently, the determination of which situation controls is made by comparing the two potential peak requirements. The smaller of the two controls.

There will be circumstances, particularly during preliminary design, where insufficient information is available to allow the procedures above to be used. In that case it may be satisfactory to multiply the heterotrophic and autotrophic steady-state oxygen requirements by an oxygen peaking factor to arrive at the transient state

oxygen requirements. Figure 10.18 provides oxygen peaking factors as a function of the load peaking factor. It was generated from simulations conducted using wastewater characteristics and parameters similar to those used in Tables 6.6 and 6.3, respectively.[2] Because the oxygen peaking factor for the removal of organic matter was not influenced strongly by SRT, the curve for carbon oxidation should be safe for a broad range of SRTs. The curve for nitrification, on the other hand, is only valid for SRTs above 10 days.

A final point about transient loadings concerns their impact on mixing energy input and floc shear. It will be recalled from Section 10.2.5 that there is both a lower and an upper limit on the volumetric power input to an activated sludge bioreactor, with the lower limit being the minimum energy required to keep the MLSS in suspension, and the upper limit being set to prevent floc shear. The ratio of the upper to the lower limit is around 4.5. It is not unusual, however, for the ratio of the maximum loading to the minimum loading within a day at a wastewater treatment plant to be greater than 4.5, particularly for small plants.[2] This suggests that the ratio of the maximum to minimum oxygen requirements associated with diurnal loadings can be greater than 4.5. In that situation, since the volumetric power input required for oxygen transfer is directly proportional to the oxygen requirement, it would be impossible to meet both the upper and lower limits on power input. Consequently, most designers use the upper limit and the peak oxygen requirement during sizing of the bioreactor and then limit the turn-down on the aeration system to meet the lower limit during low loading, recognizing that the DO concentration in the bioreactor will be higher than needed then. The other alternative is to include flow equalization in the process flow diagram.

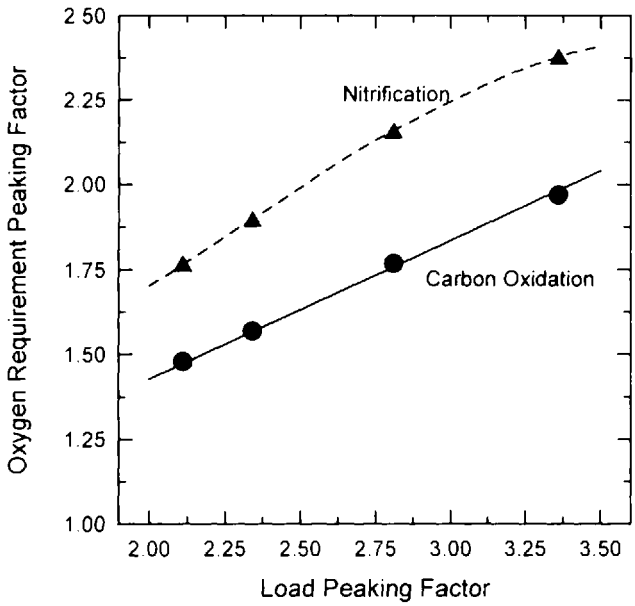


Figure 10.18 Effect of the load peaking factor on the oxygen peaking factor for a CMAS system receiving a diurnally varying input. Data from Amalan.[2]

Distribution of Volume, Mixed Liquor Suspended Solids, and Oxygen in Non-uniform Systems. As indicated by guiding principle 4 in Table 9.1, the total mass of biomass and the total oxygen requirement in the various alternative activated sludge systems will all be essentially the same, provided they all have the same SRT. Thus, they can be calculated by the simple model of Chapter 5 as modified in Section 9.4.2. However, when the design involves the distribution of flows or volumes into reactors in series, the biomass and oxygen requirement must also be distributed appropriately. This can be done for the steady-state case by the application of mass balances and appropriate heuristics. Distribution of the transient-state oxygen requirement is more difficult. Because the procedures involved are unique to each activated sludge variation, they will be considered individually in the sections that follow.

10.3.3 Design of a Completely Mixed Activated Sludge System—The General Case

Because it is the simplest, the basic design process will be outlined for a CMAS system. All examples will be developed for the wastewater characteristics given in Table E8.4 and the kinetic and stoichiometric parameters given in Table E8.5. Those values are for a temperature of 20°C. As indicated earlier, all organic substrate concentrations will be expressed as biodegradable COD and all MLSS concentrations will be expressed as TSS. Thus, the parameters on the right side of Table E8.5 apply. Two situations will be considered. First, to illustrate basic principles, the case of full equalization, i.e., the steady-state case, will be considered. Then we will consider the impacts of diurnal variations in loading, i.e., the case without equalization.

Basic Process Design for the Steady-State Case. The first task in a process design is to establish the maximum and minimum sustained temperatures likely to be encountered in the activated sludge system. The stoichiometric and kinetic parameters are then adjusted to those temperatures using Eq. 3.95, as discussed in Section 10.3.2. The temperature adjusted parameters are used in selection of the design SRT. Because we have already discussed the selection of the SRT, it will not be considered further here. Rather, we will assume that the decision has already been made.

Example 10.3.3.1

A CMAS system is to be designed to remove organic matter from a wastewater with the characteristics given in Table E8.4. Removal of ammonia-N is not required, so the system does not have to nitrify. Consequently, an SRT of three days has been chosen for the design. The average design wastewater flow rate is 40,000 m^3/day and full equalization will maintain the loading at the average value throughout the day. The oxygen transfer system will be sized to maintain the DO concentration above 1.5 mg/L under all conditions. The parameter values characterizing the wastewater at 20°C are given in Table E8.5. However, the lowest sustained temperature anticipated is 15°C and the highest is 25°C. Prepare a table of temperature adjusted parameter values by using the temperature coefficients in Table E10.1.

All temperature adjustments are made with Eq. 3.95, in which k_2 is the value of the parameter at reference temperature T_2. Using b_H as an example,

$b_{H,15} = (0.18 \text{ day}^{-1})\ 1.04^{(15-20)} = 0.15 \text{ day}^{-1}$

$b_{H,25} = (0.18 \text{ day}^{-1})\ 1.04^{(25-20)} = 0.22 \text{ day}^{-1}$

The values of the other parameters are given in Table E10.2.

The next step in the process design is to calculate the oxygen requirement for the system. As mentioned in Section 10.3.2, that should be done for the highest expected sustained temperature because that is when the highest oxygen requirement will occur. The information has two uses. First, it provides the base requirement for design of the oxygen transfer system. That aspect of design will not be covered in this book, so the reader is referred to other sources for it.[46,71,79,82] Second, the maximum oxygen requirement will be used with the upper limit on the volumetric power input, Π_U, to select the lower feasible reactor volume based on floc shear, $V_{L,FS}$, as given by Eq. 10.4. The maximum oxygen requirement will also be used in Eq. 10.5 to calculate the lower limit on bioreactor volume based on oxygen transfer, $V_{L,OT}$. The minimum oxygen requirement, which will occur at the lowest, sustained operating temperature, must also be calculated. It will be used with the lower limit on the volumetric power input, Π_L, to select the upper feasible bioreactor volume, V_U, as given by Eq. 10.3. Those volume limits will then be used to make the final selection of the bioreactor volume and the associated MLSS concentration.

The oxygen requirement for removal of organic matter can be calculated with Eq. 9.13:

$$RO_H = F(S_{SO} + X_{SO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T} i_{O\,XB,T}}{1 + b_H \cdot \Theta_c}\right] \tag{9.13}$$

If nitrification will occur, then the oxygen requirement associated with it can be calculated with a slightly modified version of Eq. 6.2, shown below as Eq. 10.16:

$$RO_A = F(S_{N,a} - S_{NH})\left[4.57 - \frac{(1 + f_D \cdot b_A \cdot \Theta_c)Y_{A,T} i_{O\,XB,T}}{1 + b_A \cdot \Theta_c}\right] \tag{10.16}$$

In this expression the influent nitrogen concentration is designated as $S_{N,a}$, repre-

Table E10.2 Stoichiometric and Kinetic Parameter Values in COD/TSS Units from Table 8.5 After Adjustment for Temperature

Parameter	Units	Value at 15°C	Value at 25°C
$\hat{\mu}_H$	day^{-1}	4.1	8.8
K_S	mg/L as COD	20	20
$Y_{H,T}$	mg TSS/mg COD	0.50	0.50
b_H	day^{-1}	0.15	0.22
f_D	mg TSS/mg TSS	0.20	0.20
$i_{O\,XB,T}$	mg COD/mg TSS	1.2	1.2
$\hat{\mu}_A$	day^{-1}	0.46	1.3
K_{NH}	mg/L as N	0.52	1.9
$K_{O,A}$	mg/L as O_2	0.75	0.75
$Y_{A,T}$	mg TSS/mg N	0.20	0.20
b_A	day^{-1}	0.08	0.12

senting the nitrogen available to the nitrifiers, rather than the influent ammonia-N concentration as given in Eq. 6.2. There are two reasons for this. First, at SRTs long enough to allow nitrification, essentially all biodegradable COD will be used, releasing all organic nitrogen as ammonia-N. Thus, $S_{N,a}$ must include all biodegradable organic nitrogen in addition to the ammonia-N. Second, the heterotrophic bacteria will use some of the nitrogen in the synthesis of their biomass, making it unavailable to the nitrifiers. Thus, the available nitrogen concentration is given by:

$$S_{N,a} = S_{NHO} + S_{NSO} + X_{NSO} - NR(S_{SO} + X_{SO} - S_S) \tag{10.17}$$

where NR is the heterotrophic nitrogen requirement given by Eq. 5.36:

$$NR = 0.087\left[\frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T} i_{O\,XB,T}}{1 + b_H \cdot \Theta_c}\right] \tag{5.36a}$$

Equation 5.36 has been modified to reflect the use of total suspended solid (TSS) as the measure of MLSS in this chapter, and thus is renumbered as Eq. 5.36a to indicate this. The effluent concentration of ammonia-N can be calculated with Eq. 5.13 by substituting the kinetic parameters for the autotrophic bacteria in place of the heterotrophic parameters. Because nitrification is so sensitive to the DO concentration, however, it would be wise to consider its effect on the effluent ammonia-N concentration. This can be done by using the double Monod equation, Eq. 3.46, in place of the Monod equation when deriving Eq. 5.13. The resulting equation can be simplified, however, by recognizing that the term $S_O/(K_{O,A} + S_O)$ is just the reciprocal of the DO safety factor, as given by Eq. 10.9. Substituting for it gives:

$$S_{NH} = \frac{K_{NH}(1/\Theta_c + b_A)}{(\hat{\mu}_A/\varsigma_{DO}) - (1/\Theta_c + b_A)} \tag{10.18}$$

If nitrification is not a design objective during selection of the SRT, then the minimum SRT for nitrification should be checked to determine whether it is likely to occur, because if it does, it will have a large impact on the oxygen requirement. This can be done either with Figure 9.4 or with Eq. 5.16a. The effect of DO concentration on the SRT required for nitrification can be considered by multiplying the minimum SRT by the DO safety factor, ς_{DO}, as was done in Eq. 10.11. However, neither of the other two safety factors should be used in this application.

The following example illustrates the technique for determining the oxygen requirement and the upper and lower limits on the bioreactor volume.

Example 10.3.3.2

Continue with the CMAS design started in Example 10.3.3.1 and determine the maximum and minimum steady-state oxygen requirements, and the lower and upper limits on the feasible bioreactor volumes. The design SRT is three days. Assume that bubble aeration will be used, with an oxygen transfer efficiency, η_O, of 10%.

a. What is the maximum steady-state heterotrophic oxygen requirement?
This can be calculated with Eq. 9.13 using the parameter values in Table E10.2 for 25°C, since the oxygen requirement is maximum at the warmest temperature. We saw earlier that almost all readily biodegradable COD will

be removed at SRTs in excess of one day, so the value of S_S can safely be assumed to be negligible. As given in Example 10.3.3.1, the flow rate is 40,000 m^3/day. Thus, for consistency in units, concentrations should be expressed with m^3 as the measure of volume. From Table E8.4, S_{SO} = 115 mg/L = 115 g/m^3, and X_{SO} = 150 mg/L = 150 g/m^3. Furthermore, all time dependent coefficients in Table E10.2 are expressed with days as the unit of time, so the SRT should be expressed in days for consistency.

$$RO_H = (40{,}000)(115 + 150)\left[1 - \frac{[1 + (0.2)(0.22)(3.0)](0.50)(1.20)}{1 + (0.22)(3.0)}\right]$$

$$= 6{,}260{,}000 \text{ g } O_2/\text{day} = 6{,}260 \text{ kg } O_2/\text{day}$$

b. Will the system nitrify in the summer when the temperature is 25°C?
This can be determined either with Figure 9.4 or with Eq. 5.16a. Because the needed parameter values are available, we will use Eq. 5.16a with a value of S_{NHO} of 25 mg/L from Table E8.4. Again, the values of the parameters are for 25°C from Table E10.2.

$$\Theta_{x,min} = \frac{(1.9 + 25)}{(25)(1.30 - 0.12) - (1.9)(0.12)}$$

$$= 0.92 \text{ day.}$$

This value is for high DO concentrations. Because the lowest expected DO concentration is 1.5 mg/L, the DO safety factor, as given in Eq. 10.9 should be applied, using the value of $K_{O,A}$ from Table E10.2:

$$\zeta_{DO} = \frac{(0.75 + 1.5)}{1.5} = 1.50$$

Therefore, the minimum SRT is:

$$\Theta_{x,min} = (0.92)(1.5) = 1.38 \text{ days.}$$

Since the design SRT is 3 days, nitrification will occur in the summer.

c. What is the concentration of nitrogen available to the autotrophic bacteria for nitrification?
This can be calculated with Eq. 10.17. From Table E8.4, S_{NHO} = 25 mg/L as N = 25 g/m^3, S_{NSO} = 6.5 mg/L as N = 6.5 g/m^3, X_{NSO} = 8.5 mg/L as N = 8.5 g/m^3, S_{SO} = 115 mg/L as COD = 115 g/m^3, and X_{SO} = 150 mg/L as COD = 150 g/m^3. As in Part a above, S_S may be neglected. All that is needed is the value of NR, the nitrogen requirement of the heterotrophs. This can be calculated with Eq. 5.36a using parameter values from Table E10.2 for 25°C:

$$NR = 0.087\left[\frac{[1 + (0.2)(0.22)(3)](0.5)(1.20)}{1 + (0.22)(3)}\right]$$

$$= 0.036 \text{ mg N used/mg COD removed}$$

Substituting this into Eq. 10.17 gives:

$$S_{NH} = 25 + 6.5 + 8.5 - 0.036(115 + 150)$$

$$= 30.5 = \text{g/m}^3 \text{ as N}$$

d. What is the steady-state autotrophic oxygen requirement?
This can be calculated with Eq. 10.16, which requires knowledge of the

effluent ammonia-N concentration. Since nitrification is not likely to be complete at an SRT of 3 days, the effluent ammonia-N concentration must be calculated with Eq. 10.18. Using the parameter values from Table E10.2 for 25°C and the value of the DO safety factor calculated in Part b above,

$$S_{NH} = \frac{1.9(1/3.0 + 0.12)}{(1.30/1.5) - (1/3.0 + 0.12)}$$

$$= 2.1 \text{ mg/L as N} = 2.1 \text{ g/m}^3 \text{ as N}$$

Substitution of this value into Eq. 10.16 gives the autotrophic oxygen requirement:

$$RO_A = (40{,}000)(30.5 - 2.1)\left[4.57 - \frac{[1 + (0.2)(0.12)(3.0)](0.20)(1.20)}{1 + (0.12)(3.0)}\right]$$

$$= 4.980{,}000 \text{ g } O_2\text{/day} = 4{,}980 \text{ kg } O_2\text{/day}$$

e. What is the maximum steady-state oxygen requirement?
The maximum steady-state oxygen requirement is the sum of the heterotrophic and autotrophic oxygen requirements:

$$RO = 6{,}260 + 4{,}980 = 11{,}240 \text{ kg } O_2\text{/day} = 468 \text{ kg/hr}$$

The oxygen transfer system must be designed to transfer this amount, plus an appropriate factor of safety. Note that the autotrophic oxygen requirement is almost as much as the heterotrophic requirement. This points out why it is so important to determine whether it is likely that nitrification will occur, even when the system is not being designed with nitrification as an objective.

f. What is the lower limit on the bioreactor volume based on the mixing energy constraint to avoid floc shear?
This lower limit on the bioreactor volume can be calculated with Eq. 10.4, after estimation of the required airflow rate, Q, with Eq. 10.2. The oxygen requirement in Eq. 10.2 has units of kg/hr, giving Q in m^3/min. The same units are used for Q in Eq. 10.4.

$$Q = \frac{(6.0)(468)}{10} = 281 \text{ m}^3\text{/min}$$

The lower limit on the CMAS bioreactor volume based on floc shear comes from application of Eq. 10.4 using 90 m^3/(min · 1000 m^3) as an appropriate value for $\Pi_{t,Q}$:

$$V_{LFS} = \frac{(1000)(281)}{90} = 3{,}120 \text{ m}^3$$

g. What is the lower limit on the bioreactor volume based on the maximum sustainable volumetric oxygen transfer rate?
This lower limit on the bioreactor volume can be calculated with Eq. 10.5 using the total oxygen requirement of 468 kg/hr:

$$V_{LOT} = \frac{468}{0.10} = 4{,}680 \text{ m}^3$$

This value is larger than the volume associated with the constraint on floc shear, and thus it controls.

h. What is the minimum steady-state heterotrophic oxygen requirement?
This must be calculated for winter conditions when the temperature is 15°C. The procedure is exactly the same as in Part a above, except that the parameter values for 15°C from Table E10.2 are used. The result of that computation is:

$$RO_H = 5{,}820{,}000 \text{ g/day} = 5{,}820 \text{ kg/day}$$

i. Will the system nitrify in the winter when the temperature is 15°C and the DO concentration is 1.5 mg/L?
Using the same procedure as in Part b above, but with the parameter values for 15°C from Table E10.2, the minimum SRT for nitrification is found to be 4.0 days. Thus, nitrification will not occur in the winter.

j. What is the minimum steady-state oxygen requirement?
Since nitrification will not occur, the minimum steady-state oxygen requirement is just the minimum heterotrophic oxygen requirement:

$$RO = 5{,}820 \text{ kg/day} = 242.5 \text{ kg/hr}$$

This is only slightly more than half of the summer requirement. Since this is all that will be required during winter operation, the oxygen transfer system must be designed with sufficient turn-down capacity to allow this amount to be delivered in an economic manner.

k. What is the upper limit on the bioreactor volume based on the mixing energy constraint to keep all biomass in suspension?
The upper limit on the bioreactor volume can be calculated with Eq. 10.3, after estimation of the required airflow rate, Q, with Eq. 10.2. The oxygen requirement in Eq. 10.2 has units of kg/hr, giving Q in m^3/min. The same units are used for Q in Eq. 10.3.

$$Q = \frac{(6.0)(242.5)}{10} = 146 \text{ m}^3\text{/min}$$

The upper limit on the CMAS bioreactor volume comes from application of Eq. 10.4 using 20 $m^3/(min \cdot 1000\ m^3)$ as an appropriate value for $\Pi_{t,O}$:

$$V_t = \frac{(1000)(146)}{20} = 7{,}300 \text{ m}^3$$

l. Any bioreactor volume between 4,680 and 7,300 m^3 can be used as long as it results in a MLSS concentration that is acceptable.

After the range of feasible bioreactor volumes has been calculated, the next step in the design of a CMAS system is to determine the MLSS concentration associated with each extreme reactor volume. This will provide a range of feasible MLSS concentrations from which a design value can be chosen after consideration of the size of the final settler, as discussed in Section 9.3.4. Since we are using COD and TSS units in this chapter, Eq. 9.11 is the appropriate equation with which to calculate the mass of MLSS, $X_{M,T} \cdot V$:

$$X_{M,T} \cdot V = \Theta_c F \left[X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c) Y_{H,T} (S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \Theta_c} \right] \tag{9.11}$$

This computation is made for the coldest anticipated sustained wastewater temperature because the decay coefficient will be smallest then, resulting in the highest quantity of biomass, as discussed in Section 10.3.2. The value of the effluent soluble biodegradable COD, S_S, is generally small enough to be neglected in the computation, even for cold conditions. The equation does not include the contribution of autotrophic biomass to the mass of MLSS in the bioreactor. This is because for domestic wastewater, their contribution will be negligible, as discussed in Section 6.3.2. This may not be true for wastewaters containing a high nitrogen content, and in those circumstances another term should be added to account for the contribution of the nitrifying bacteria. It is similar to the right term within the brackets, except that the parameters and the influent concentrations would represent nitrification:

$$X_{B,A,T} \cdot V = \Theta_c F \left[\frac{(1 + f_D \cdot b_A \cdot \Theta_c) Y_{A,T} (S_{N,o} - S_{NH})}{1 + b_A \cdot \Theta_c} \right] \tag{10.19}$$

Example 10.3.3.3

Continuing with the CMAS design begun in Example 10.3.3.1, determine the range of feasible MLSS concentrations.

a. What value of $X_{IO,I}$ should be used in the calculation?
$X_{IO,I}$ includes both the fixed suspended solids (FSS) and the nonbiodegradable VSS. In Section 8.6 it was stated that 35 to 40% of the particulate organic matter in domestic wastewater is nonbiodegradable. As a result, a nonbiodegradable fraction of 0.375 was used in Example 8.6.1, where Table E8.4 was developed. From that table, the VSS concentration was 61.5 mg/L. This suggests that the nonbiodegradable VSS concentration is 0.375 × 61.5, or 23 mg/L. The FSS concentration is the difference between the TSS and VSS concentration, or 82 − 61.5 = 20.5 mg/L. Therefore, the value of $X_{IO,I}$ is:

$$X_{IO,I} = 23 + 20.5 = 43.5 \text{ mg/L} = 43.5 \text{ g/m}^3$$

b. What is the mass of MLSS present in the CMAS system at 15°C?
This can be calculated with Eq. 9.11. As given in Example 10.3.3.1, the flow rate is 40,000 m^3/day. Thus, for consistency in units, concentrations should be expressed with m^3 as the measure of volume. From Table E8.4, $S_{SO} = 115$ mg/L = 115 g/m^3, and $X_{SO} = 150$ mg/L = 150 g/m^3. Furthermore, all time dependent coefficients in Table E10.2 are expressed with days as the unit of time, so the SRT should be expressed in days for consistency. The mass of MLSS is calculated for winter conditions using the parameters from Table E10.2.

$$X_{M,T} \cdot V = (3.0)(40{,}000) \cdot \left[43.5 + \frac{[1 + (0.20)(0.15)(3.0)](0.5)(115 + 150)}{1 + (0.15)(3.0)} \right]$$
$$= 17{,}170{,}000 \text{ g MLSS}$$

c. What are the upper and lower limits on the MLSS concentration?
The highest feasible MLSS concentration, $X_{M,U,C}$, is associated with the smallest feasible bioreactor volume, 4,680 m^3:

$$X_{M,LC} = \frac{17{,}170{,}000}{4{,}680} = 3{,}670 \text{ g/m}^3 = 3{,}670 \text{ mg/L}$$

The smallest feasible MLSS concentration, $X_{M,LL}$, is associated with the largest feasible bioreactor volume, 7,300 m^3:

$$X_{M,LL} = \frac{17{,}170{,}000}{7{,}300} = 2{,}350 \text{ g/m}^3 = 2{,}350 \text{ mg/L}$$

d. The choice of MLSS concentration between those limits must be made by considering the cost trade-offs between the bioreactor and the final settler, as discussed in Section 9.3.4. It is important to note that the constraints on mixing energy input and oxygen transfer limit the range of values that need to be considered.

The final step in the design of the activated sludge process is the calculation of the solids wastage rate, $W_{M,T}$. It is very straightforward and is simply an extension of the procedure used to calculate $X_{M,T} \cdot V$. For the stoichiometric approach with the simplified model of Chapter 5, the appropriate equation is Eq. 9.12:

$$W_{M,T} = F\left[X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \Theta_c)Y_{H,T}(S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \Theta_c}\right] \tag{9.12}$$

This computation is generally made for winter conditions since that is when the most solids must be disposed of. As with the computation of $X_{M,T} \cdot V$, no term is included for the contribution of the autotrophic bacteria. If their contribution to the waste solids is likely to be significant, then an appropriate term should be added to the equation. The actual flow rate of waste solids will depend on whether solids are wasted from the settler underflow or directly from the bioreactor. Both the flow rate and the daily mass of waste solids are used to size the solids handling system.

Example 10.3.3.4

Continue the design begun in Example 10.3.3.1 by calculating the solids wastage rate from the process. This is done for the low temperature condition using Eq. 9.12:

$$W_{M,T} = (40{,}000)\left[43.5 + \frac{[1 + (0.20)(0.15)(3.0)](0.5)(115 + 150)}{1 + (0.15)(3.0)}\right]$$

$$= 5{,}720{,}000 \text{ g TSS/day} = 5{,}720 \text{ kg TSS/day}$$

Consideration of the Effects of Transient Loadings. The basic design given above considered a system that used equalization to dampen transient loadings. Quite frequently, however, facilities are designed without equalization, and consideration must be given to the impacts of typical diurnal loadings or other transients on the system. As discussed in Section 10.3.2, diurnal loadings have little impact on the mass of MLSS in the system because of the dampening effect of the SRT relative to the HRT. Consequently, the main consideration during design is on the oxygen requirement and the feasible bioreactor volumes while meeting the mixing energy and oxygen transfer constraints. Generally, the peak loading is most important because of its impact on the size of the oxygen transfer system and the potential for

floc shear. Little consideration is given to the minimum daily loading because of its short duration. If aeration rates are turned down to the lower limit on mixing energy during that period, and that provides more oxygen than is needed, the penalty in power costs will be small.

The basic procedure for calculating the additional oxygen requirement as a result of a transient load was discussed in Section 10.3.2, where Eqs. 10.12–10.15 were presented. The peak oxygen requirement for the process, RO_P, is then the sum of the average and transient-state oxygen requirements for both the heterotrophic and the autotrophic biomass. However, as discussed in Section 10.3.2, consideration must be given to whether the autotrophic bacteria have reached their maximum possible growth and oxygen consumption rates. Thus, the peak oxygen requirement is given by the smaller of the two expressions:

$$RO_P = RO_H + RO_{H,TS} + RO_A + RO_{A,TS} \tag{10.20}$$

or:

$$RO_P = RO_H + RO_{H,TS} + RO_{A,max} \tag{10.21}$$

In the example that follows, we will examine the impact of transient loads on the design of the CMAS system considered in Examples 10.3.3.1–10.3.3.4.

Example 10.3.3.5

Continue with the design of the CMAS system begun in Example 10.3.3.1, which has an average daily flow rate of 40,000 m^3/day. In this case, however, no equalization will be employed so that the system routinely experiences a peak loading 2.5 times the average daily loading. What will the peak oxygen requirement be? What oxygen peaking factor does the peak oxygen requirement represent?

a. What is the peak transient oxygen requirement for the heterotrophic bacteria? This is calculated with Eq. 10.12 in which $\Delta(F \cdot S_{SO})$ and $\Delta(F \cdot X_{SO})$ represent the transient increases in the readily and slowly biodegradable organic matter respectively. Since the peaking factor, which is the ratio of the peak to the average loading, is 2.5, the transient increase of the biodegradable materials above the average loading is 1.5 times that average. Thus,

$$\Delta(F \cdot S_{SO}) = 1.5(40{,}000)(115) = 6{,}900{,}000 \text{ g COD/day}$$
$$= 6{,}900 \text{ kg COD/day}$$
$$\Delta(F \cdot X_{SO}) = 1.5(40{,}000)(150) = 9{,}000{,}000 \text{ g COD/day}$$
$$= 9{,}000 \text{ kg COD/day}$$

A value must be assumed for $f_{XS,H}$, the fraction of additional slowly biodegradable substrate oxidized during the transient. We stated earlier that values typically lie between 0.5 and 1.0, with smaller values being associated with larger transient increases. Since the magnitude of the transient load is neither extremely high nor low, we will assume that $f_{XS,H}$ has a value of 0.75. The peak oxygen requirement will occur at the warmest temperature. Consequently, substituting the parameter values for 25°C from Table E10.2 into Eq. 10.12, along with the transient increases calculated above, gives:

$$RO_{H,TS} = [1 - (1.20)(0.5)][6{,}900 + (0.75)(9{,}000)] = 5{,}460 \text{ kg } O_2\text{/day}$$

b. What is the peak transient oxygen requirement for the autotrophic bacteria assuming that they are able to oxidize the additional ammonia load?
We use the same procedure as above to calculate the transient nitrogen loadings:

$$\Delta(F \cdot S_{NH0}) = 1.5(40{,}000)(25) = 1{,}500{,}000 \text{ g N/day} = 1{,}500 \text{ kg N/day}$$

$$\Delta(F \cdot S_{NSO}) = 1.5(40{,}000)(6.5) = 390{,}000 \text{ g N/day} = 390 \text{ kg N/day}$$

$$\Delta(F \cdot X_{NSO}) = 1.5(40{,}000)(8.5) = 510{,}000 \text{ g N/day} = 510 \text{ kg N/day}$$

Use Eq. 10.14 and $f_{XN,H} = 0.75$ to calculate the transient increase in the ammonia-N concentration available to the autotrophic bacteria:

$$\Delta(F \cdot S_N)_{A,TS} = 1{,}500 + 390 + (0.75)(510) - (0.087)(1.20)(0.50) \cdot [6{,}900 + (0.75)(9{,}000)] = 1{,}560 \text{ kg N/day}$$

Using Eq. 10.13 and the appropriate parameters for 25°C from Table E10.2,

$$RO_{A,TS} = [4.57 - (1.2)(0.20)](1560) = 6{,}750 \text{ kg } O_2\text{/day}$$

c. What is the maximum potential autotrophic oxygen requirement assuming that the autotrophic bacteria are not able to oxidize all of the transient nitrogen input, causing the ammonia-N concentration to rise sufficiently for them to reach their maximal growth rate? As in the other examples, assume that a DO concentration of at least 1.5 mg/L is maintained.
This is calculated using Eq. 10.15 with parameter values for the maximum temperature of 25°C. Use of this equation requires knowledge of the mass of autotrophic bacteria in the system based on average loading conditions, $X_{B,A,1} \cdot V$. This can be calculated for average conditions with Eq. 10.19 using the autotrophic parameter values for 25°C from Table E10.2 and the values of $S_{N,a}$ and S_{NH} from Example 10.3.3.2:

$$X_{B,A,1} \cdot V = (3.0)(40{,}000)\left[\frac{[1 + (0.20)(0.12)(3.0)](0.20)(30.5 - 2.1)}{1 + (0.12)(3.0)}\right]$$

$$= 537{,}000 \text{ g autotropic MLSS}$$

Substitution of this mass into Eq. 10.15 gives:

$$RO_{A,max} = \left[\left(\frac{4.57 - (1.2)(0.20)}{(1.2)(0.20)}\right)(1.3) + (1 - 0.20)(0.12\right] \cdot (1.2)(537{,}000)\left(\frac{1.5}{0.75 + 1.5}\right) = 10{,}100{,}000 \text{ g } O_2\text{/day}$$

$$= 10{,}100 \text{ kg } O_2\text{/day}$$

d. What is the peak autotrophic oxygen requirement?
The peak autotrophic oxygen requirement is the smaller of $RO_A + RO_{A,TS}$ and $RO_{A,max}$. The value of RO_A determined in Example 10.3.3.2, is 4,980 kg O_2/day. Thus, $RO_A + RO_{A,TS} = 4{,}980 + 6{,}750 = 11{,}730$ kg O_2/day. However, as seen above, $RO_{A,max} = 10{,}100$ kg O_2/day. Thus, ammonia-N breakthrough occurs and the peak autotrophic oxygen requirement is 10,100 kg O_2/day.

e. What is the peak oxygen requirement, RO_P?
Since $RO_{A,max} < (RO_A + RO_{A,TS})$, Eq. 10.21 gives the peak oxygen requirement. Recalling the value of RO_H from Example 10.3.3.2:

$$RO_p = 6{,}260 + 5{,}460 + 10{,}100 = 21{,}820 \text{ kg } O_2/\text{day} = 909 \text{ kg } O_2/\text{hr}$$

f. What is the oxygen peaking factor?
The peaking factor for the oxygen requirement is the peak requirement divided by the average from Example 10.3.3.2, which was 11,240 kg O_2/day:

$$\text{Oxygen peaking factor} = \frac{21{,}820}{11{,}240} = 1.94$$

This occurred for a transient peak in the organic loading on the process of 2.5. This value is slightly higher than the value obtained from Figure 10.18. The difference is due entirely to the difference in the peaking factor for the heterotrophic activity and is caused primarily by the approximations associated with Eq. 10.12.

After the peak oxygen requirement has been estimated, it may be used to refine the range of feasible bioreactor volumes and MLSS concentrations. For the steady-state case, the lower limit on the bioreactor volume, V_1, was calculated for the average summer oxygen requirement. When transient loadings occur, that volume must be calculated on the basis of the peak summer oxygen requirement, thereby raising the value of V_1. The impact of the peak oxygen requirement will be examined in the example below for the design we have been considering.

Example 10.3.3.6

What is the range of allowable bioreactor volumes and MLSS concentrations for the CMAS system considered in Examples 10.3.3.2, 10.3.3.3, and 10.3.3.5?

a. What is the lower limit on the bioreactor volume based on the mixing energy constraint to avoid floc shear?
This lower limit on the bioreactor volume can be calculated with Eq. 10.4, after estimation of the required airflow rate, Q, with Eq. 10.2. The oxygen requirement in Eq. 10.2 should be the peak value with units of kg/hr, giving Q in m^3/min. The same units are used for Q in Eq. 10.4. Using a peak oxygen requirement of 909 kg/hr from Example 10.3.3.5, Q is:

$$Q = \frac{(6.0)(909)}{10} = 545 \text{ m}^3/\text{min}$$

The lower limit on the CMAS bioreactor volume based on floc shear comes from application of Eq. 10.4 using 90 $m^3/(\text{min} \cdot 1000 \text{ m}^3)$ as an appropriate value for Π_{LO}:

$$V_{LIS} = \frac{(1000)(545)}{90} = 6{,}060 \text{ m}^3$$

This value is almost twice the steady-state value because the peak oxygen requirement was almost twice the steady-state.

b. What is the lower limit on the bioreactor volume based on the maximum oxygen transfer rate?
The lower limit on the CMAS bioreactor volume based on the maximum oxygen transfer rate comes from application of Eq. 10.5:

$$V_{LOL} = \frac{909}{0.10} = 9{,}090 \text{ m}^3$$

This value is also much larger than the steady-state value. Because it is the larger of the two lower limits, it will control.

c. What is the upper limit on bioreactor volume based on the mixing energy constraint to keep all biomass in suspension?
The upper limit does not change for the reasons given earlier. It remains 7,300 m^3 as calculated in Example 10.3.3.2.

d. What range of bioreactor volumes can be used?
Examination of the upper and lower limits on the bioreactor volume reveals that they both cannot be met since the lower limit is greater than the upper limit. Thus, the designer must consider the consequences of violating one or both of them. The lower limit of 9,090 m^3 was chosen to keep the volumetric oxygen transfer rate no greater than 100 g $O_2/(m^3 \cdot hr)$ at all times, including the diurnal peak, which does not last long. Thus, this volume could be considered to be conservative. Another option would be to use the lower limit associated with floc shear, which is 6,060 m^3. This requires a peak oxygen transfer rate of 150 g $O_2/(m^3 \cdot hr)$, which is achievable, but would require special attention to the design of the oxygen transfer system. Alternatively, the bioreactor volume could be set equal to the upper limit based on the mixing energy required to keep biomass in suspension during the period when the average oxygen requirement is minimum, which is 7,300 m^3. If that volume is chosen, the peak oxygen transfer rate will be 125 g $O_2/(m^3 \cdot hr)$, which will be easier to attain than 150. Thus, while any volume between 6,060 and 7,300 m^3 could be chosen, it would be more prudent to choose a value near the upper end of the range.

e. What is the new range of MLSS concentrations?
The imposition of the transient load on the system doesn't change the mass of MLSS in the system, $X_{MLSS} \cdot V$. It is the same as calculated in Example 10.3.3.3, 17,170,000 g MLSS. Thus, the lower limit remains 2,350 mg/L. The new upper limit is:

$$X_{MLSS} = \frac{17{,}170{,}000}{6060} = 2{,}800 \text{ g/m}^3 = 2{,}800 \text{ mg/L}$$

f. The transfer of oxygen and the avoidance of floc shear during peak loading conditions greatly limits the designer's options while balancing the costs of the bioreactor and the final settler. This is necessary, however, because of the nature of activated sludge oxygen transfer devices and because the effluent from a CMAS bioreactor passes directly to the settler. More latitude with respect to floc shear can be gained in the design by adding a flocculation chamber prior to the settler, as discussed in Section 10.2.5, or by adding equalization prior to the CMAS system, as discussed above.

This section has focussed on CMAS design as a means for presenting the basic factors that must be considered in the design of any activated sludge system. Like CMAS, EAAS usually has both the biomass and the oxygen requirement distributed uniformly throughout the bioreactor. Consequently, the design procedure for an

EAAS system is essentially the same. The only difference is that due to the long SRTs typically used with the EAAS process, floc shear is not usually a factor limiting the bioreactor size. Rather, the major issue is to maintain MLSS in suspension in an economical manner.

10.3.4 Conventional Activated Sludge, High-Purity Oxygen Activated Sludge, and Selector Activated Sludge—Systems with Uniform Mixed Liquor Suspended Solids Concentrations, but Variations in Oxygen Requirements

The basic approach to the design of all other activated sludge systems is the same as that presented in Section 10.3.3, although CMAS is the only variation for which that approach can be used directly. However, because of guiding principle No. 4 in Table 9.1, much of the information can be used for the other variations, with appropriate modifications. The activated sludge variations for which the least modification is required are those which have uniform MLSS concentrations throughout. These are conventional activated sludge (CAS), high-purity oxygen activated sludge (HPOAS), and selector activated sludge (SAS). As discussed in Sections 7.2 and 10.1.2, CAS and HPOAS can both be considered to behave as a number of completely mixed tanks in series. In HPOAS the separate tanks are real, since the bioreactor is staged by partitioning it, whereas in CAS the tanks are imaginary, representing the residence time distribution of the bioreactor. For purposes of design and analysis, however, both can be considered to be made of equal size tanks in series. Examination of the simulations presented in Figure 7.6 shows that the variation in the MLSS concentration from tank to tank is insignificant, justifying the assumption of uniform concentration throughout the system. An SAS system generally is designed with a series of small completely mixed tanks preceding the main bioreactor, as shown in Figure 10.6. As indicated in Section 10.1.2, the main bioreactor may be one large completely mixed basin, or it may be like CAS, in which case it can be considered to behave as a number of tanks in series. In either case, however, the SAS system can be modeled for design and analysis as several unequal tanks in series, but with uniform MLSS concentration throughout.

For all of these systems, decisions about the design SRT are made in the same way as previously discussed. Moreover, the mass of MLSS in the system, $(X_{M,I} \cdot V)_{System}$, can be calculated directly with Eq. 9.11 and the total steady-state oxygen requirement can be estimated with Eqs. 9.13 and 10.16. The total transient-state oxygen requirement can be estimated with the techniques discussed in Section 10.3.2 and illustrated in Example 10.3.3.5 for a CMAS system. The added level of complexity arises from the need to distribute that oxygen requirement appropriately throughout the reactor system, and to size the oxygen transfer system in a corresponding manner.

The need to spatially distribute the oxygen requirement in CAS and HPOAS systems can be seen by examining Figure 7.5, which shows both the steady-state and range of diurnal oxygen requirements in such systems. There it can be seen that the variation from the first to the last tank is large, particularly when diurnal loading variations are imposed on the system. Similar variations will occur in SAS systems,

with peak requirements occurring in the selector. Failure to properly account for such spatial variations will lead to poor performance and/or an uneconomic system. Simulation is the most accurate way to predict the variation in oxygen requirement, and its use is encouraged, but sufficient information for simulation is often unavailable, requiring approximations. Thus, before we consider the design approach for these systems, we need to consider how to approximate the required spatial oxygen distribution.

Approximate Technique for Spatially Distributing Oxygen Requirements. The spatial distribution of the oxygen requirement requires consideration of the different events contributing to oxygen utilization and the rates at which they occur. First consider a system operating at steady-state with a uniform loading. Utilization of readily biodegradable substrate is very rapid and will usually be complete even in systems with very short SRTs. It is the major contributor to the high utilization rate in tank one of Figure 7.5. Utilization of slowly biodegradable substrate, on the other hand, is slower because it is limited by the rate of hydrolysis reactions. However, we saw in Figure 9.3 that it is often complete in systems with SRTs as short as two days. This suggests that most slowly biodegradable substrate will have been used in the first third to one-half of a tanks-in-series system, depending on the system SRT. Its use contributed to a substantial portion of the oxygen consumption in the first two tanks in Figure 7.5. Biomass decay, on the other hand, is a very slow reaction that occurs at a constant rate throughout the entire activated sludge process because its rate is driven solely by the biomass concentration, which can be considered to be uniform, as discussed earlier. It contributed the base rate seen in tanks four and five of Figure 7.5, and the same contribution also occurred in all of the preceding tanks. Finally, nitrification is also a slow process, but faster than decay. Furthermore, its rate is driven by the ammonia-N concentration, which is soluble, and attains a maximum at moderate ammonia-N concentrations. Consequently, nitrification will occur at its maximal rate in the first few tanks in the system, but often will be completed before the last tank is reached, depending on the system SRT. In Figure 7.5, nitrification contributed to oxygen consumption primarily in the first three tanks.

Use can be made of the generalizations in the preceding paragraph for partitioning the total oxygen requirement into its component parts. Consider first the heterotrophic oxygen requirement. Rearrangement of Eq. 9.13, after neglecting S_S, allows the oxygen requirement to be divided into two component parts, that associated with biomass synthesis and that caused by decay:

$$RO_H = F(S_{SO} + X_{SO})(1 - Y_{H,T}i_{O\,XB,T}) + F(S_{SO} + X_{SO}) \cdot (Y_{H,T}i_{O\,XB,T}) \left[\frac{(1 - f_D)(b_H \cdot \Theta_c)}{1 + b_H \cdot \Theta_c}\right] \tag{10.22}$$

The first term on the right side of the equation is the oxygen used for synthesis of new biomass, whereas the second term is the oxygen utilization for biomass decay. The synthesis term can be further subdivided into oxygen utilization for biomass synthesis from readily biodegradable substrate,

$$RO_{H,SS} = F \cdot S_{SO}(1 - Y_{H,T}i_{O\,XB,T}) \tag{10.23}$$

and oxygen utilization for biomass synthesis from slowly biodegradable substrate:

$$RO_{H,XS} = F \cdot X_{SO}(1 - Y_{H,T}i_{O\,XB,T}) \tag{10.24}$$

This subdivision is desirable because of the differences in the rates of utilization of the two substrate types, as discussed above. The oxygen utilization due to decay of biomass may be given by one term because the type of substrate from which biomass was grown has no effect on its decay rate:

$$RO_{H,D} = F(S_{SO} + X_{SO})(Y_{H,T}i_{O\,XB,T})\left[\frac{(1 - f_D)(b_H \cdot \Theta_c)}{1 + b_H \cdot \Theta_c}\right] \tag{10.25}$$

These component oxygen utilization terms can be used to determine the profile of heterotrophic oxygen utilization through a CAS, HPOAS, or SAS system. The distribution of oxygen utilization for decay is the easiest. Because biomass is distributed evenly throughout these systems and because the rate of decay is proportional to the biomass concentration, the rate of oxygen utilization due to decay is the same throughout the system. As a result, the mass of oxygen required for decay in any tank, i, of a multitank system is just:

$$RO_{H,Di} = RO_{H,D}\left(\frac{V_i}{V_T}\right) \tag{10.26}$$

where V_i is the volume of tank i and V_T is the total system volume.

The distribution of the oxygen requirement for the utilization of slowly biodegradable substrate is less exact and is dependent on an approximation. Let $\Theta_{c\,XS}$ be the SRT at which slowly biodegradable substrate utilization would be essentially complete in a CMAS system. If we then recognize that biomass is uniformly distributed in a tanks-in-series system, and neglect any kinetic benefits to hydrolysis of having a tanks-in-series configuration, then we can approximate the fraction of the system volume within which biomass synthesis occurs on slowly biodegradable substrate, $f_{V,XS}$, as:

$$f_{V,XS} = \frac{\Theta_{c\,XS}}{\Theta_c} \tag{10.27}$$

We saw earlier that slowly biodegradable substrate can be fully utilized at SRTs as short as two days. If we considered that figure to be applicable in a system with an SRT of four days, we might expect oxygen utilization for synthesis of biomass from slowly biodegradable substrate to occur in the first half of the system. Furthermore, if the system could be characterized as being equivalent to five tanks in series, we would expect 40% of that oxygen utilization to occur in each of the first two tanks and 20% in the third. None would occur in the last two tanks. Alternatively, if the system behaved like three tanks in series, we might expect 67% of the utilization to occur in the first tank, 33% in the second, and none in the third. While this technique is crude, it at least provides a means to approximate where oxygen utilization is likely to occur.

Distribution of the oxygen requirement for biomass synthesis from readily biodegradable substrate requires computation of the volume of a fictitious completely mixed bioreactor, V_F, that receives the influent stream and the RAS flow, and reduces the substrate concentration to a desired level. The oxygen requirement would then be apportioned to the initial fraction of the activated sludge bioreactor that contained an equivalent volume. If we let S_{SF} be the desired readily biodegradable substrate concentration in the fictitious completely mixed bioreactor, then it follows from

Monod kinetics, Eq. 3.36, that the specific growth rate of the biomass in that reactor must be:

$$\mu_{H,F} = \hat{\mu}_H \frac{S_{SF}}{K_S + S_{SF}} \tag{10.28}$$

A mass balance on readily biodegradable substrate in that reactor, neglecting the contribution of hydrolysis of slowly biodegradable substrate, gives:

$$\mu_{H,F} = \frac{Y_{H,F}[F(S_{SO} - S_{SF}) - F_I(S_{SF} - S_S)]}{f_A \cdot V_F \cdot X_{M,F,F}} \tag{10.29}$$

where S_S is the readily biodegradable substrate concentration in the effluent from the entire activated sludge process and $X_{M,F,F}$ is the MLSS concentration in the fictitious bioreactor. Because the composition of the MLSS in any activated sludge system is the same throughout, the active fraction in the MLSS of the fictitious bioreactor will be the same as the active fraction in the activated sludge system under consideration, which is governed by the system SRT. Making use of this fact, and assuming that $F_I(S_{SF} - S_S) << F(S_{SO} - S_{SF})$, that $S_{SF} << S_{SO}$, and that $S_S << (S_{SO} + X_{SO})$, it can be shown that:

$$f_{XM,F} = \frac{V_F \cdot X_{M,F,F}}{(V \cdot X_{M,F})_{System}} = \left(\frac{1/\Theta_x + b_H}{\mu_{H,F}}\right)\left(\frac{S_{SO}}{S_{SO} + X_{SO}}\right) \tag{10.30}$$

In other words, the fraction of the system MLSS in the fictitious bioreactor, $f_{XM,F}$, is proportional to the ratio of the average net specific growth rate in the process ($1/\Theta_x + b_H$) relative to the required specific growth rate in the fictitious bioreactor, $\mu_{H,F}$, as given by Eq. 10.28. For the case under consideration here, the MLSS concentration in the fictitious bioreactor will be the same as the concentration throughout the process. Therefore, the fraction of the total system volume in which the readily biodegradable substrate is removed, $f_{V,SS}$, is given by:

$$f_{V,SS} = \frac{V_F}{V_T} = \left(\frac{1/\Theta_x + b_H}{\mu_{H,F}}\right)\left(\frac{S_{SO}}{S_{SO} + X_{SO}}\right) \tag{10.31}$$

where V_T is the total system volume.

The fraction given by Eq. 10.31 represents the smallest possible fraction of the system volume within which the readily biodegradable substrate concentration could be reduced to S_{SF}. The possibility exists that the fraction will be larger because hydrolysis of slowly biodegradable substrate will be occurring, a contribution that was not considered in the derivation of Eq. 10.29. The exact contribution of hydrolysis is difficult to estimate without using a model like ASM No. 1, which would defeat the purpose for which these equations are given. However, the largest possible fraction of the system within which the readily biodegradable substrate might be removed can be calculated by assuming that all of the slowly biodegradable is hydrolyzed and contributes to substrate removal in the fictitious bioreactor. When that assumption is made, the right parenthetical term, $S_{SO}/(S_{SO} + X_{SO})$, goes to one. Thus, the largest possible fraction can be calculated with Eq. 10.31 with that term set equal to 1.0. The two fractions give the designer bounds within which to apportion the oxygen consumption associated with biomass synthesis from readily biodegradable substrate.

In using Eq. 10.28 to calculate $\mu_{H,T}$ for substitution into Eq. 10.31, the choice of $S_{S,T}$ is very important. It must be large enough to represent the rapid rate of removal of readily biodegradable substrate, but low enough for the assumptions to be valid. Generally, a value of 10% of $S_{S,0}$ should be adequate.

The total oxygen requirement for nitrification can be calculated with Eq. 10.16. It can be partitioned into two components, synthesis and decay, just as the heterotrophic oxygen requirement was partitioned. However, as shown in Figure 10.17, the effluent ammonia-N concentration from a tanks-in-series system will be less than the concentration from a CMAS system. This will make the oxygen requirement for nitrification slightly larger. Because the effluent ammonia-N concentration cannot be easily predicted without simulation, the effluent ammonia-N concentration should be assumed to be zero. This will provide a slightly conservative estimate. The oxygen requirement associated with synthesis of the autotrophic biomass can be calculated from:

$$RO_{A,SN} = F \cdot S_{N,a}(4.57 - Y_{A,T} i_{O,XB,T}) \tag{10.32}$$

The oxygen requirement associated with decay of the autotrophic biomass can be calculated with an equation like Eq. 10.25:

$$RO_{A,D} = F \cdot S_{N,a}(Y_{A,T} i_{O,XB,T}) \left[\frac{(1 - f_D)(b_A \cdot \Theta_c)}{1 + b_A \cdot \Theta_c} \right] \tag{10.33}$$

If the wastewater undergoing treatment is a domestic wastewater, the contribution of autotrophic decay to the total oxygen requirement will typically be negligible. However, this may not be the case for industrial wastewaters, and thus both autotrophic oxygen requirements should be calculated explicitly.

The distribution of the oxygen requirement associated with synthesis of the nitrifiers within a tanks-in-series system is very straightforward since nitrification can be assumed to behave as a zero-order reaction in the first part of such a system. The maximum mass nitrification rate can be estimated from a modified form of Eq. 5.7 that is analogous to Eq. 10.15 for the maximum autotrophic oxygen requirement:

$$RN_{A,max} = \left(\frac{\hat{\mu}_A}{Y_{A,T}} \right) (X_{B,A,T} \cdot V) \left(\frac{S_O}{K_{A,O} + S_O} \right) \tag{10.34}$$

in which $X_{B,A,T} \cdot V$ is calculated with Eq. 10.19. Because nitrification behaves as a zero-order reaction over most of the system, the fraction of the system volume over which autotrophic nitrification will occur, $f_{V,A}$, is just given by:

$$f_{V,A} = \frac{F \cdot S_{N,a}}{RN_{A,max}} \tag{10.35}$$

If the value of $f_{V,A}$ is greater than 1.0, then nitrification will not be complete, which for the tanks-in-series configuration suggests that the SRT is too short and washout will occur (recall Figure 10.17). Actually, this never should occur since the value of $X_{B,A,T} \cdot V$ is required to get $RN_{A,max}$. If the SRT were too low, that value could not be calculated. Thus, a value of $f_{V,A}$ greater than 1.0 suggests that an error has been made somewhere in the computations. Once the value of $f_{V,A}$ is known, the oxygen requirement for nitrification, RO_A, is apportioned proportionally over the fraction of the system within which nitrification occurs, just as the oxygen requirement for biomass synthesis from slowly biodegradable substrate was apportioned.

As was done for the oxygen requirement associated with decay of heterotrophs, that associated with decay of the autotrophs should be distributed evenly throughout the entire system.

Example 10.3.4.1

Consider the activated sludge system that was the subject of all of the examples in Section 10.3.3. Instead of a CMAS system, however, the system is to be configured as a CAS system with hydraulic characteristics equivalent to four tanks-in-series. Distribute the steady-state oxygen requirement associated with the summer equalized loading.

a. How should the oxygen requirement associated with biomass synthesis from readily biodegradable substrate be distributed?
Calculate the oxygen requirement with Eq. 10.23. The readily biodegradable substrate concentration is 115 mg/L (= 115 g/m^3) and the flow rate is 40,000 m^3/day.

$$RO_{H,SS} = (40{,}000)(115)[1 - (0.50)(1.20)] = 1{,}840{,}000 \text{ g } O_2/\text{day}$$
$$= 1{,}840 \text{ kg } O_2/\text{day}$$

Calculate the specific growth rate in a fictitious bioreactor capable of removing 90% of the readily biodegradable substrate. In that case, S_{S1} = 11.5 mg/L. Utilization of Eq. 10.28 gives:

$$\mu_{H,1} = (8.8)\left(\frac{11.5}{20 + 11.5}\right) = 3.21 \text{ days}^{-1}$$

Use it in Eq. 10.31 to calculate the smallest possible fraction of the total system volume within which the readily biodegradable substrate is removed. The SRT of the CAS system is 3 days.

$$f_{V,SS} = \left(\frac{1/3.0 + 0.22}{3.21}\right)\left(\frac{115}{115 + 150}\right) = 0.075$$

The largest possible fraction within which the readily biodegradable substrate is removed can be calculated by setting the right parenthetical term equal to 1.0, giving a value of 0.17. Thus, the readily biodegradable substrate will be removed in 7.5 to 17 percent of the system volume. Since the CAS system behaves as four equal tanks-in-series, 25% of the system volume is in each tank. Since the readily biodegradable substrate will be removed in less than 25% of the volume, all of the oxygen requirement associated with it should be apportioned to tank No. 1.

b. How should the oxygen requirement associated with biomass synthesis from slowly biodegradable substrate be distributed?
Calculate the oxygen requirement with Eq. 10.24. The slowly biodegradable substrate concentration is 150 mg/L (= 150 g/m^3) and the flow rate is 40,000 m^3/day.

$$RO_{H,SS} = (40{,}000)(150)[1 - (0.50)(1.20)] = 2{,}400{,}000 \text{ g } O_2/\text{day}$$
$$= 2{,}400 \text{ kg } O_2/\text{day}$$

Use Eq. 10.27 to estimate the fraction of the system volume within which

biomass growth on slowly biodegradable substrate occurs. Since the total system SRT is 3 days, all of the substrate must be fairly easy to degrade. Therefore, it seems reasonable to assume that $\Theta_{c,XS}$ has a value of 2 days.

$$f_{v,XS} = \frac{2.0}{3.0} = 0.67$$

Each tank in the system contains 25% of the system volume. Since 67% of the system volume has 100% of the oxygen requirement, 25% of the volume will have 37% of the oxygen requirement, i.e., 0.25/0.67 = 0.37. Therefore, tanks No. 1 and No. 2 will both receive 37% of the oxygen requirement and tank No. 3 will receive the remainder, or 26%. Therefore,

$$RO_{H,XS,1} = (0.37)(2{,}400) = 888 \text{ kg } O_2/\text{day}$$
$$RO_{H,XS,2} = (0.37)(2{,}400) = 888 \text{ kg } O_2/\text{day}$$
$$RO_{H,XS,3} = (0.26)(2{,}400) = 624 \text{ kg } O_2/\text{day}$$

Tank No. 4 will have no oxygen requirement associated with use of slowly biodegradable substrate.

c. How should the oxygen requirement associated with heterotrophic biomass decay be distributed?
The total oxygen requirement for heterotrophic decay can be calculated with Eq. 10.25:

$$RO_{H,D} = (40{,}000)(115 + 150)(0.50)(1.20)\left[\frac{(1 - 0.20)(0.22)(3.0)}{1 + (0.22)(3.0)}\right]$$
$$= 2{,}020{,}000 \text{ g } O_2/\text{day} = 2{,}020 \text{ kg } O_2/\text{day}$$

The oxygen requirement associated with decay should be distributed equally to all reactors. Since each reactor contains 25% of the system volume, each will have 25% of the oxygen requirement. Thus, each tank will require 505 kg O_2/day for biomass decay.

d. How should the oxygen requirement associated with synthesis of autotrophic biomass be distributed?
Calculate the oxygen requirement for nitrification with Eq. 10.32. The value of $S_{N,a}$ is obtained from Eq. 10.17, and can be assumed to be the same as for the CMAS system, or 30.5 mg/L as N:

$$RO_{A,NN} = (40{,}000)(30.5)[4.57 - (0.2)(1.20)] = 5{,}280{,}000 \text{ g } O_2/\text{day}$$
$$= 5{,}280 \text{ kg } O_2/\text{day}$$

The fraction of the system volume over which nitrification will occur can be calculated with Eq. 10.35, which requires knowledge of the maximum mass nitrification rate from Eq. 10.34. That equation, however, requires the mass of autotrophic biomass in the system, which can be calculated with Eq. 10.19. Just as the oxygen requirement was slightly higher in the CAS system because of the lower effluent ammonia-N concentration, so will the mass of autotrophic biomass be higher. In this case it will be:

$$X_{B,A,T} \cdot V = (3.0)(40{,}000)\left[\frac{[1 + (0.20)(0.12)(3.0)](0.20)(30.5)}{1 + (0.12)(3.0)}\right]$$
$$= 577{,}000 \text{ g autotrophic MLSS}$$

It can now be substituted into Eq. 10.34. Since the minimum DO concentration in the system is to be 1.5 mg/L, this gives:

$$RN_{N,max} = \left(\frac{1.3}{0.20}\right)(577{,}000)\left(\frac{1.5}{0.75 + 1.5}\right) = 2{,}500{,}000 \text{ g N/day}$$

$$= 2{,}500 \text{ kg N/day}$$

This can now be used with Eq. 10.35 to find the fraction of the system volume over which nitrification occurs:

$$f_{V,N} = \frac{(40{,}000)(30.5)}{2{,}500{,}000} = 0.49$$

Each tank in the system contains 25% of the system volume. Since 49% of the system volume receives 100% of the oxygen requirement, 25% of the volume will receive 51% of the oxygen requirement, i.e., 0.25/0.49 = 0.51. Therefore, tank No. 1 will receive 51% of the oxygen requirement and tank No. 2 will receive the remainder, or 49%. Therefore,

$$RO_{N,SS,1} = (0.51)(5{,}280) = 2{,}690 \text{ kg } O_2\text{/day}$$

$$RO_{N,SS,2} = (0.49)(5{,}280) = 2{,}590 \text{ kg } O_2\text{/day}$$

e. How should the oxygen requirement associated with autotrophic biomass decay be distributed?
The oxygen requirement associated with decay of the autotrophic biomass can be calculated with Eq. 10.33:

$$RO_{N,D} = (40{,}000)(30.5)(0.20)(1.20)\left[\frac{(1 - 0.20)(0.12)(3.0)}{1 + (0.12)(3.0)}\right]$$

$$= 62{,}000 \text{ g } O_2\text{/day} = 62 \text{ kg } O_2\text{/day}$$

This will be distributed equally to all bioreactors, so each has an oxygen requirement of 15.5 kg O_2/day, which is negligible compared to the other oxygen requirements in the system.

f. The total oxygen requirement in each equivalent tank of the CAS system is given in Table E10.3, where the component oxygen requirements are summarized.
Examination of Table E10.3 reveals several things. First, the total heterotro-

Table E10.3 Distribution of Steady-State Oxygen Requirement in CAS System of Example 10.3.4.1

	Oxygen requirement, kg O_2/day			
Component	Tank 1	Tank 2	Tank 3	Tank 4
Readily biodegradable organic matter	1,840	0	0	0
Slowly biodegradable organic matter	888	888	624	0
Heterotrophic decay	505	505	505	505
Nitrification	2,690	2,590	0	0
Autotrophic decay	16	16	16	16
Total	5,939	3,999	1,145	521

phic oxygen requirement is 6,260 kg O_2/day, which is the same as in the CMAS system, as calculated in Example 10.3.3.2. This is to be expected since the equations for distributing the oxygen requirement were derived by partitioning the equations for a CMAS system. Nevertheless, the agreement between the values serves as a convenient check on the calculations. Second, the total autotrophic oxygen requirement is 5,340 kg O_2/day, which is slightly higher than the 4,980 kg O_2/day calculated for the CMAS system in Example 10.3.3.2. The higher value is due to the greater extent of nitrification that will occur in a CAS-type system. Third, the majority of the oxygen requirement occurs in the first two equivalent tanks of the CAS-type system, which is in agreement with the simulations in Section 7.2.2 and experience in the field. This means that the volumetric oxygen transfer rate in the first equivalent tanks of a CAS-type system will be greater than the rate in a CMAS system if the total bioreactor volumes are the same. Fourth, the oxygen requirements in the last equivalent tanks of a CAS-type system are small, suggesting that the required volumetric oxygen transfer rate is smaller than that in an equivalent CMAS system.

g. How accurate is the approximate technique for distributing the oxygen requirement?
Simulations conducted with ASM No. 1 gave smaller oxygen requirements in the first two equivalent tanks and larger requirements in the last two, than the values indicated in Table E10.3. The values from the simulation are more accurate because they come from a true dynamic model that has been shown to mimic well real-world activated sludge bioreactors.[7] The reasons for this difference are the assumptions made about biodegradation of slowly biodegradable substrate and nitrification in the approximate technique. Thus, when the approximate technique of this section is used to distribute the steady-state oxygen requirement in a CAS, HPOAS, or SAS system, its tendency to overpredict the oxygen requirement at the start of the system and to underpredict the requirement at the end of the system should be kept in mind.

The distribution of the transient-state oxygen requirement is even more difficult to approximate than the distribution of the steady-state requirement. Consequently, simulation with a dynamic model like ASM No. 1 or No. 2 is the only way to accurately determine how the oxygen supply should be distributed to a CAS, HPOAS, or SAS system receiving diurnal loads. The reason for this is the uncertainty associated with the biodegradation of the slowly biodegradable organic matter and with nitrification. If it is necessary to make such a distribution without simulation, then it should be done by applying the same principles used to distribute the steady-state requirement. Since that distribution is based on dividing the oxygen requirement into its component parts, it is directly applicable to the transient-state case, making it unnecessary to use the approach used for a CMAS system. In applying the principles presented above, one need only consider the impact of the transient loading on the synthesis oxygen requirement. The oxygen requirements associated with decay will not change because the mass of MLSS in the system changes little in response to the transient loading, as discussed earlier.

Design of Conventional Activated Sludge Systems. The design of a CAS system is an iterative process because of the need to distribute the oxygen requirement

throughout the bioreactor. This requires knowledge of its hydraulic characteristics, expressed as the number of equivalent tanks-in-series, which are determined by its size and configuration. As a consequence, the designer must select a bioreactor volume and shape, determine the number of equivalent tanks-in-series, and distribute the oxygen requirement among those equivalent tanks. Only then is it possible to check the mixing intensities in the equivalent tanks. If they are not adequate, then another bioreactor volume and shape must be selected and the process repeated until an acceptable combination is obtained. Luckily, the selection process is simplified somewhat by a relaxation of the mixing constraints as a result of the tanks-in-series behavior of a CAS bioreactor. The mixing intensity in the first equivalent tank may exceed the floc shear limitation, Π_U, as long as the mixing intensity in subsequent equivalent tanks is sufficiently small to allow for reflocculation of the biomass.[76] However, consideration must still be given to the maximum practical oxygen transfer rate. At the other end, where the oxygen requirement is small, the mixing intensity in the last equivalent tank can be set equal to the limit for adequate mixing, Π_L, without significant sacrifice in power costs because only a small percentage of the total system volume will be involved.

To start the process, Eq. 9.11 may be used to calculate the mass of MLSS in the system, $(X_{M,T} \cdot V)_{System}$. Using the procedures in Section 10.3.3, the total steady-state oxygen requirement can then be used to select the upper and lower limits on bioreactor volume that would exist if the system were configured as a CMAS system. The design volume for the CAS system should be intermediate between these values because the oxygen requirement in the first equivalent tank will be greater than the rate in the CMAS system and the rate in the last equivalent tank will be smaller, as illustrated in Example 10.3.4.1. Generally, however, the selected volume should be closer to the upper limit for the CMAS system because of limitations on the maximum rate of oxygen transfer in the first equivalent tank. Since the choice of this volume fixes the target MLSS concentration (through the calculated $(X_{M,T} \cdot V)_{System}$ value), it should be based on consideration of the anticipated solids settling characteristics and the proposed final settler design, including the biomass recycle ratio, α. The bioreactor configuration, i.e., its length, L, width, W, and depth, H, is then selected by considering the site characteristics. The next task is to estimate the number of equivalent tanks-in-series, N, for the chosen dimensions. This may be done with an empirical relationship developed at the Water Research Centre in England:[77]

$$N = 7.4 \left[\frac{F(1 + \alpha)L}{W \cdot H} \right] \tag{10.36}$$

in which the units of the flow rate, F, are m^3/sec, and L, W, and H are all expressed in meters. N must be an integer value and should be rounded appropriately. Once N is known, then the volume of each equivalent tank, V_i, may be obtained by dividing the chosen system volume by N. The volumetric oxygen transfer rate in any tank, $r_{SO,i}$, can then be obtained from:

$$r_{SO,i} = \frac{RO_i}{V_i} \tag{10.37}$$

where:

$$RO_i = RO_{H,SS,i} + RO_{H,XS,i} + RO_{H,D,i} + RO_{A,SN,i} + RO_{A,D,i} \quad (10.38)$$

For average load conditions, the highest practical sustained volumetric oxygen transfer rate is approximately 100 g $O_2/(m^3 \cdot hr)$ for conventional oxygen transfer systems, as discussed previously. Values as high as 150 g $O_2/(m^3 \cdot hr)$ can be obtained under peak load conditions, but values greater than 100 should not be used for prolonged operation of mechanical aeration systems because of rapid wear of the components. Thus, when a conventional oxygen transfer system is to be used, if the calculated volumetric oxygen transfer rate exceeds 100 g $O_2/(m^3 \cdot hr)$ a larger total tank volume must be chosen or special attention should be given to the design of the oxygen transfer system. Use of a larger total tank volume will require alteration of the tank dimensions, which will require recalculation of the number of equivalent tanks-in-series, N. Because the use of Eq. 10.36 requires rounding to arrive at an integer value for N, judgement must be exercised in deciding whether a change in N is justified. If N is changed, the oxygen requirement must be redistributed and the procedure repeated. After a system size has been determined based on oxygen transfer in the first equivalent tank, the mixing energy input into the last few equivalent tanks must be calculated using RO_i for them and the techniques presented for a CMAS system. If the mixing energy input is less than Π_L, additional mixing must be provided to maintain the MLSS in suspension. The objective is to arrive at a system size that will allow the high oxygen requirements in the front of the system to be met while minimizing the amount of mixing energy in excess of Π_L required in the back of the system. Consideration must be given to both the maximum oxygen requirement for summer operation and the minimum for winter operation. Consequently, this exercise will require judgement and compromise.

Example 10.3.4.2

Continue with the problem of Example 10.3.4.1. A tentative decision has been made to use a reactor volume of 7,500 m^3 for the CAS system, configured in such a way that the hydraulic characteristics will be equivalent to four tanks-in-series. Is this an acceptable choice?

a. What is the required volumetric oxygen transfer rate in the first equivalent tank?

From Table E10.3 of Example 10.3.4.1 the required oxygen transfer rate to the first tank is 5,939 kg O_2/day, which is 247 kg O_2/hr. The volume of the first equivalent tank is one fourth of the total system volume, or 1,875 m^3. The volumetric oxygen transfer rate can be calculated with Eq. 10.37, giving a value of:

$$r_{SO,1} = \frac{247}{1,875} = 0.132 \text{ kg } O_2/(m^3 \cdot hr) = 132 \text{ g } O_2/(m^3 \cdot hr)$$

This value is higher than can be achieved on a sustained basis with conventional mechanical oxygen transfer equipment, but can be attained with diffused aeration by giving careful consideration to design of the oxygen transfer system. Thus, the selected volume is acceptable from the standpoint of the maximum oxygen transfer rate.

b. What air flow rate will be required in the last equivalent tank if the oxygen transfer efficiency is ten percent?
From Table E10.3, the required oxygen transfer rate in the last equivalent tank is 521 kg O_2/day (22 kg O_2/hr). The air flow rate required to achieve this can be calculated with Eq. 10.2 in which RO must be expressed as kg O_2/hr:

$$Q = \frac{(6.0)(22)}{10} = 13.2 \text{ m}^3\text{/min}$$

The air flow rate required to keep the MLSS in suspension can be calculated from Eq. 10.3 using the equivalent tank volume of 1,875 m^3:

$$Q = \frac{(20)(1{,}875)}{1000} = 37.5 \text{ m}^3\text{/min}$$

The air flow rate required to keep the biomass in suspension is larger than that required to deliver the needed oxygen, and therefore it controls. The tank volume cannot be decreased to reduce this amount because then it would be impossible to deliver the needed oxygen to the first equivalent tank. This commonly occurs in CAS systems. The design can be considered to be acceptable.

Design of High-Purity Oxygen Activated Sludge Systems. Design of a HPOAS system is essentially the same as design of a CAS system, with two significant differences. First, the system is actually staged, so that the number of tanks-in-series is chosen by the designer. Thus, there is no need to iterate when distributing the oxygen requirement from tank to tank. Second, because high-purity oxygen is introduced into the first stage and the gas moves sequentially from stage to stage from the inlet to the outlet, the partial pressure of oxygen in the gas phase decreases as it moves through the system. As a result, the gas phase is most heavily enriched in oxygen in that part of the system where the required oxygen transfer rate is highest, allowing high oxygen transfer rates to be achieved with lower power inputs than would be required if the gas phase were air. This makes the power inputs more uniform throughout the system and alleviates some of the problems associated with balancing the power inputs into CAS systems. Of course, mixing must still be checked to ensure that sufficient energy is expended to maintain the MLSS in suspension.

Design of Selector Activated Sludge Systems. SAS systems are used when either the wastewater or the bioreactor configuration of the chosen activated sludge variation have a tendency to favor the growth of filamentous bacteria. A selector counteracts that tendency. As described in Section 10.1.2, a selector is a highly loaded section at the inlet end of an activated sludge bioreactor. The high loading condition creates an environment that favors the growth of floc-forming rather than filamentous bacteria, as discussed in Section 10.2.1. This results in improved solids settling characteristics, allowing the use of higher MLSS concentrations and/or higher secondary clarifier solids loading rates.

The selector must be properly sized if kinetic selection of floc-forming bacteria is to occur. First, it must be large enough to remove the majority of the readily biodegradable organic matter applied. If it is not, some of that organic matter will pass through the selector into the remaining portion of the bioreactor where envi-

ronmental conditions favor the growth of filamentous bacteria. Second, it must be small enough to maintain the readily biodegradable substrate concentration on the right side of the cross-over point on a plot of specific growth rate versus substrate concentration for the two bacterial types, as shown in Figure 10.12. If it isn't, environmental conditions favoring the growth of floc-forming bacteria will not be established in the selector, resulting in the preferential growth of filamentous bacteria in it.

Several factors must be considered when choosing the size and configuration of a selector. Because it is difficult to generalize about the kinetic parameters describing growth of floc-forming and filamentous bacteria, it is impossible to identify the readily biodegradable substrate concentration associated with the crossover point on the μ:S_s curve. However, empirical evidence has suggested the minimum initial process loading factor that will lead to good settling biomass, as shown in Figure 10.14 and discussed in Section 10.2.1. Since the process loading factor is proportional to the specific growth rate, specification of the process loading factor for the selector is analogous to specifying the crossover point on the μ:S_s curve. Consequently, selector design is typically based on specification of a process loading factor for it. However, because influent loadings vary, it is difficult to specify a single selector volume that is large enough to remove all of the readily biodegradable organic matter while also providing a sufficiently high process loading factor under all loading conditions. This problem can be solved by using a staged selector, as illustrated in Figure 10.6. During periods of low loading, the process loading factor will be sufficiently high in the first stage to favor the floc-forming bacteria, yet the bulk of the readily biodegradable substrate will be removed. During periods of high loading, on the other hand, the additional selector stages ensure that the readily biodegradable organic matter is removed prior to the main bioreactor.

Our understanding of the organism selection and organic substrate removal mechanisms occurring within SAS systems is evolving and, consequently, so are the approaches used to design selector systems. Based on current knowledge, however, the following general approach has been successful:[39]

1. A minimum of three stages should be used. The first two stages should each contain 25% of the total selector volume, while the third stage should contain the remaining 50%.
2. The total selector volume is chosen to give an overall process loading factor of 3 kg total biodegradable COD/(kg MLSS · day), which is analogous to about 1.75 kg BOD_5/(kg MLSS · day). Note that even though the purpose of the selector is to remove readily biodegradable organic matter, the process loading factor is based on the total biodegradable COD. This is because the empirical evidence upon which it is based, i.e., Figure 10.14, did not distinguish between the readily and slowly biodegradable organic matter in the wastewater.

This relatively low overall process loading factor results in good removal of readily biodegradable organic matter, even during periods when the organic loading is temporarily higher. Furthermore, use of the selected overall process loading factor will result in a process loading factor in the initial selector stage of 12 kg COD/(kg MLSS · day). Reference to Figure 10.14 indicates that this high initial stage process

loading factor provides a factor of safety to ensure that sufficiently high process loading factors are achieved regardless of the influent load.

Based on the above considerations, the selector volume is chosen from rearrangement of the definition of the process loading factor as given by Eq. 5.37:

$$V_s = \frac{F(S_{SO} + X_{SO})}{U_S \cdot X_{M,T}} \tag{10.39}$$

where the influent COD includes both readily and slowly biodegradable substrate, and V_s and U_s are the volume and process loading factor for the selector, respectively. The MLSS concentration is the same as in the rest of the activated sludge system, and is determined from $(X_{M,T} \cdot V)_{System}$ as calculated with Eq. 9.11, where V is the selected total system volume. The volume of the rest of the system is just $V - V_s$.

Determination of the required oxygen transfer rate within a selector is not straightforward. Experience with some selector installations indicates that respiration rates in the first stage may be as high as 40 to 60 g O_2/(kg MLSS · hr).[23,34] These values are high, but are less than would be expected if the heterotrophic bacteria were growing at their maximum specific growth rate. It appears that, in some instances, the specific growth rates in the selector are sufficiently high to trigger a substrate storage response called the selector effect.[30] In those cases, selector oxygen requirements corresponding to the oxidation of only about 20% of the COD removed have been observed.[30] It has been hypothesized that the remainder of the removed substrate is stored as intracellular carbon storage polymers such as glycogen and/or PHB. Because COD is conserved when storage polymer formation occurs, it is not necessary to supply sufficient oxygen in the selector to oxidize all of the removed substrate. Rather, the stored substrate is oxidized in the main bioreactor, and the oxygen requirement associated with it must be met there. Further research is needed to refine our understanding of the conditions under which this phenomenon occurs. In the meantime, it is prudent to design aerobic selector systems with significant flexibility and capacity in terms of the oxygen that can be transferred.

The determination of the oxygen requirement in the selector can be made by the same techniques used to spatially distribute the oxygen requirement in multitank systems, as discussed earlier in this section. From a conservative perspective, the oxygen requirement in the selector can be calculated by assuming that all of the readily biodegradable organic matter is removed and used for biomass synthesis in the selector. In other words, the selector effect is assumed to not occur. The oxygen requirement associated with the removal of slowly biodegradable organic matter can be distributed in proportion to the selector volume as a fraction of the system volume in which biomass synthesis occurs on slowly biodegradable substrate as determined with Eq. 10.27. The oxygen requirement for decay is given by Eq. 10.26 in which V_1 is the selector volume, V_s. Finally, nitrification will be occurring at its maximal rate in the selector, as given by Eq. 10.15. Because the selector is highly loaded, the volumetric oxygen transfer rate is likely to be quite high, particularly during periods of peak loading. As a consequence, special care should be given to the design of the oxygen transfer system in it to ensure that the needed transfer rate can be achieved.

The oxygen transfer system for the main bioreactor should be designed as if the selector were not present. There are two reasons for doing this. First, it will be necessary to bypass the selector periodically for maintenance and other purposes, placing the entire oxygen requirement into the main bioreactor. Second, the extent

to which the selector effect will occur is usually unknown. However, the larger it is, the more the oxygen requirement is shifted to the main reactor. Thus, designing the main bioreactor to handle all of the oxygen requirement ensures that any situation can be handled.

Example 10.3.4.3

Consider the CMAS system that was the subject of the examples in Section 10.3.3. In Example 10.3.3.6 the range of acceptable bioreactor volumes and associated MLSS concentrations was determined for the unequalized case, and was found to be small. It was also found that it would be preferable to design the system for the maximum feasible volume. Thus, assume that a volume of 7,300 m^3 was chosen, giving a MLSS concentration of 2,350 mg/L. Size a three-compartment selector for the system and determine the maximum potential oxygen requirement under average loading conditions.

a. What is the size of the selector?
   As seen above, aerobic selectors are typically sized with an overall process loading factor of 3.0 kg total biodegradable COD/(kg MLSS·day). Adopting that value, and recognizing that the MLSS concentration in the selector will be the same as the MLSS concentration in the main bioreactor, i.e., 2,350 mg/L = 2.35 kg/m^3, the selector volume can be calculated with Eq. 10.39. Recalling that the average flow rate is 40,000 m^3/day, and that the average readily and slowly biodegradable substrate concentrations are 115 mg COD/L (0.115 kg/m^3) and 150 mg COD/L (0.150 kg/m^3), respectively, gives:

$$V_S = \frac{(40{,}000)(0.115 + 0.150)}{(3.0)(2.35)} = 1{,}500 \text{ m}^3$$

b. How is the selector configured?
   It should be configured as three tanks-in-series, with the first two each being 25% of the total volume, and the third 50%. Thus, the first two selectors each have a volume of 375 m^3 and the third a volume of 750 m^3.

c. What is the volume of the main CMAS bioreactor?
   The total system volume is not changed by the addition of the selector, so the main reactor volume is reduced by the volume of the selector. Therefore:

$$V_{CMAS} = 7{,}300 - 1{,}500 = 5{,}800 \text{ m}^3$$

d. What is the oxygen requirement in the selector under average load conditions in the summer?
   The oxygen requirement associated with biomass synthesis from readily biodegradable substrate was determined in Example 10.3.4.1 to be 1,840 kg O_2/day by using Eq. 10.23. All of it will occur in the selector.

   The oxygen requirement for biomass synthesis from slowly biodegradable substrate was determined in Example 10.3.4.1 to be 2,400 kg O_2/day by using Eq. 10.24. Because the rest of the system is not compartmentalized, the utilization of slowly biodegradable substrate will occur uniformly throughout it. Therefore, the utilization of slowly biodegradable substrate in the selector can just be assumed to be in proportion to its volume as a fraction of the total system volume. (If the remainder of the system was like a CAS system, rather than being a CMAS system, the oxygen requirement would

have to be distributed in the same way as in Example 10.3.4.1.) Therefore,

$$RO_{HSS} = \left(\frac{1{,}500}{7{,}300}\right) 2{,}400 = 490 \text{ kg } O_2/\text{day}$$

The oxygen requirement associated with decay of heterotrophic biomass was determined in Example 10.3.4.1 to be 2,020 kg O_2/day by using Eq. 10.25. The amount of oxygen required in the selector for heterotrophic decay will be in proportion to its volume:

$$RO_{HDS} = \left(\frac{1{,}500}{7{,}300}\right) 2{,}020 = 415 \text{ kg } O_2/\text{day}$$

The oxygen requirement associated with autotrophic biomass synthesis was determined in Example 10.3.4.1 to be 5,280 kg O_2/day by using Eq. 10.32. If the DO concentration in the selector is maintained at 1.5 mg/L, then nitrification will occur at the maximum rate, as calculated in that example. Using the logic used in Example 10.3.4.1 to determine the fraction of the nitrification oxygen requirement occurring in each tank of a CAS system, the fraction of the nitrification oxygen requirement occurring in the selector is (1,500/7,300) ÷ 0.49 = 0.42. Therefore,

$$RO_{ASS} = (0.42)(5{,}280) = 2{,}210 \text{ kg } O_2/\text{day}$$

Finally, the oxygen requirement associated with decay of autotrophic biomass was determined in Example 10.3.4.1 to be 62 kg O_2/day by using Eq. 10.34. The amount of oxygen required in the selector for autotrophic decay will be in proportion to its volume:

$$RO_{ADS} = \left(\frac{1{,}500}{7{,}300}\right) 62 = 13 \text{ kg } O_2/\text{day}$$

The total oxygen requirement in the selector is the sum of all the above components:

$$RO_s = 1{,}840 + 490 + 415 + 2{,}210 + 13 = 4{,}968 \text{ kg } O_2/\text{day}$$
$$= 207 \text{ kg } O_2/\text{hr}$$

e. What is the required volumetric oxygen transfer rate?
The required volumetric oxygen transfer rate is determined by dividing RO_s by V_s, giving a value of 138 g $O_2/(m^3 \cdot hr)$. This can be achieved by giving special attention to the design of the oxygen transfer system.

Techniques for designing aerobic selectors will continue to evolve as our understanding of microbial competition and metabolic selection increase. Consequently, the approach presented here can be expected to be modified or replaced in the future.

10.3.5 Step-Feed Activated Sludge and Contact Stabilization Activated Sludge—Systems with Nonuniform MLSS Concentrations

Examination of Figures 7.15 and 7.26 reveals that SFAS and CSAS have two characteristics in common: (1) the MLSS concentration is not uniform throughout the system, and (2) the concentration in the tank that discharges to the final settler is the

lowest of all of the tanks in the system. These are a direct result of the influent and recycle flow distributions as discussed in Sections 7.3.4 and 7.4.4. Furthermore, they provide the justification for the choice of these activated sludge variations for a particular installation. Guiding principle No. 4a in Table 9.1 states that all activated sludge variants with the same SRT contain about the same mass of biomass. As a consequence, when a SFAS or CSAS process is designed to have the same MLSS concentration entering the final settler as one of the activated sludge processes with uniform biomass concentration, the SFAS or CSAS process will always have a smaller total volume. Furthermore, the savings in system volume will be greater the longer the system SRT and the easier the organic substrate is to remove. Thus, SFAS and CSAS are often chosen for situations where space is limited. Another reason for designing a system so that it can be operated as SFAS was discussed in Section 7.3.4, and the same reason also applies to CSAS. If a CAS or CMAS system receives extremely high flows periodically, the high flow rate may cause the solids loading on the final settler to exceed allowable values, leading to loss of MLSS, poor effluent quality, and process failure. However, if the operating configuration can be switched to SFAS or CSAS, the MLSS concentration entering the final settler will be reduced, thereby keeping the solids loading on the clarifier within an acceptable range.[6] Thus, another reason wastewater treatment plants are designed with the ability to operate in the SFAS or CSAS mode is for operational flexibility.

One consequence of the hydraulic characteristics of a CSAS system is that its performance is much more dependent on the recycle ratio (and on the RAS flow rate) than other activated sludge variations, as illustrated in Figures 7.24 and 7.25. This is because of the effect that the recycle ratio has on the MLSS concentration gradient through the system, with higher recycle ratios diminishing the gradient. For the same reason, the performance of a SFAS system is also somewhat sensitive to the recycle ratio, but much less so, as illustrated in Figures 7.17 and 7.18. Nevertheless, for both systems, more consideration must be given to the impacts of the RAS flow rate during system design, particularly when its impact on the final settler size is taken into consideration. As far as effluent quality is concerned, it is often acceptable to consider only the minimum anticipated recycle ratio because it is the critical one, causing the maximum MLSS gradient and minimizing the amount of biomass in the last bioreactor. If the system produces an acceptable effluent soluble substrate concentration at that recycle ratio, it will at all higher ones, as shown in Figures 7.17 and 7.24. On the other hand, the oxygen requirement in the contact tank of a CSAS system can increase significantly as the recycle ratio is increased, particularly when nitrification is occurring, as shown in Figure 7.25. Therefore, the distribution of oxygen requirements should be examined at both the upper and lower anticipated recycle ratios.

For an existing system with uniform MLSS concentration, the impact of switching to a SFAS or CSAS operational configuration can be calculated by using mass balance techniques based on a selected RAS flow rate, the anticipated distribution of influent and RAS flows to the various tanks, and the mass of MLSS in the system as calculated with Eq. 9.11. For the design of a new system in either SFAS or CSAS mode, the situation is more complicated because of the large number of choices involved, particularly for CSAS as illustrated in Section 7.4.4. Nevertheless, in all situations the overriding criterion is that the specific growth rate of the biomass in the tank discharging to the final settler must be low enough to allow effluent quality

objectives to be met. Furthermore, all of the criteria discussed earlier about oxygen transfer and mixing energy input must also be met. This requires distribution of the oxygen requirement among the various tanks. Because it is the simpler of the two, we will first investigate SFAS. Then we will consider CSAS.

Design of Step-Feed Activated Sludge Systems. As shown in Figure 10.2 and illustrated schematically in Figure 7.10, SFAS systems are usually configured so that they behave as equally sized tanks-in-series with flow distributed equally to each tank. Other configurations can, and often are, used. For example, the equivalent tanks may not all be of equal size or influent may not be distributed to all, particularly the last. Furthermore, it some cases it may be advantageous to distribute the influent in unequal portions to the various tanks. Because it would be impossible to quantitatively describe all possible configurations, this presentation is limited to the case of equally sized tanks with equal distribution of flows to all of them. The concepts presented can easily be extended to other configurations by the reader should the need arise. The design of a SFAS system follows the same general approach as the design of the other alternatives presented earlier, with some additional steps. The emphasis here will be on the additional steps.

The first task in the design is to choose the bioreactor configuration, fixing the number of equivalent tanks in series, N, and the SRT, choices which are made by considering the factors presented earlier. Once those choices have been made, the mass of biomass in the system, $(X_{M,T} \cdot V)_{System}$, can be calculated by using Eq. 9.11 for a CMAS system.

The next task is to ensure that the selected configuration and SRT will result in the desired effluent quality. For that to occur, the specific growth rate in the last tank, μ_N, must be equal to or smaller than the specific growth rate as calculated by the Monod equation, Eq. 3.36. The approach used to calculate that specific growth rate is similar to the approach for determining the specific growth rate in the fictitious bioreactor used in the distribution of the oxygen requirement in CAS, HPOAS, and SAS systems, as discussed in Section 10.3.4. Thus, Eq. 10.28 may be used in which S_{SF} is set equal to the desired effluent substrate concentration. The chosen effluent concentration can be either the readily biodegradable substrate concentration or the ammonia-N concentration, depending on the type of standard that must be met. A lower specific growth rate will be required to meet an ammonia-N standard than a soluble biodegradable COD standard. Nevertheless, for the development that follows, organic substrate removal will be assumed to be the objective, making the heterotrophic specific growth rate controlling. The logic is similar to that used to derive Eqs. 10.29 and 10.30, except that in this case the production of soluble substrate by hydrolysis of slowly biodegradable substrate will make a significant contribution to biomass growth in the last tank. If we assume that utilization of both readily and slowly biodegradable substrate is important, it can be shown that the specific growth rate in the last tank is given by:

$$\mu_{H,N} = \frac{(X_{M,T} \cdot V)_{System}(1/\Theta_c + b_H)}{N(X_{M,T,N} \cdot V_N)} \tag{10.40}$$

in which V_N is the volume of the last tank (which is the same as V_T/N since all tanks have the same volume) and $X_{M,T,N}$ is the MLSS concentration in the last tank. The derivation of this equation is dependent on the assumption that the mass flow

rate of substrate into the last tank from the preceding tank is much less than the mass flow rate from the influent, which will generally be true. Since the value of $\mu_{H,N}$ as calculated with Eq. 10.40 must be less than or equal to the value of $\mu_{H,N}$ associated with the desired effluent substrate concentration as calculated with Eq. 10.28, then:

$$\frac{X_{M,T,N} \cdot V_N}{(X_{M,T} \cdot V)_{System}} = f_{XM,N} \geq \frac{1/\Theta_c + b_H}{N \cdot \mu_{H,N}} \tag{10.41}$$

where $\mu_{H,N}$ has been calculated with Eq. 10.28. In other words, as long as the fraction of MLSS in the Nth tank, $f_{XM,N}$, is greater than or equal to the right side of Eq. 10.41, the effluent quality will be acceptable.

The fraction of MLSS in any given tank of a SFAS system is determined totally by the hydraulics of the system. If both growth and wastage are neglected, then the fraction of MLSS in any tank i of an SFAS system in which the influent flow is split equally among N equal size tanks is given by:

$$\frac{X_{M,T,i} \cdot V_i}{(X_{M,T} \cdot V)_{System}} = f_{XM,i} = \frac{\dfrac{1.0 + \alpha}{(i/N) + \alpha}}{\displaystyle\sum_{i=1}^{N} \frac{1.0 + \alpha}{(i/N) + \alpha}} \tag{10.42}$$

where α is the biomass recycle ratio, which is the RAS flow rate, F_r, divided by the influent flow rate to the system, F. Furthermore, since the last tank is tank N, the fraction of MLSS in the last tank is:

$$\frac{X_{M,T,N} \cdot V_N}{(X_{M,T} \cdot V)_{System}} = f_{XM,N} = \frac{1.0}{\displaystyle\sum_{i=1}^{N} \frac{1.0 + \alpha}{(i/N) + \alpha}} \tag{10.43}$$

Figure 10.19 is a plot of Eq. 10.43 over the range of recycle ratios and number of equivalent tanks likely to be encountered in practice. To determine whether a proposed SFAS system will work, the fraction of MLSS in the last tank should be determined with Eq. 10.42 or Figure 10.19 for the smallest anticipated recycle ratio and that value should be used to determine if Eq. 10.41 is satisfied. As long as that fraction is greater than the right side of Eq. 10.41, the desired effluent quality will be met and a SFAS system can be used.

Example 10.3.5.1

Consider the wastewater that was the subject of the examples in Section 10.3.3 and 10.3.4. Consideration is being given to using a SFAS system with an SRT of 3 days that is equivalent to four tanks in series with equal distribution of the influent to all tanks. The effluent quality objective is 10 mg/L as COD of readily biodegradable organic matter. Can that objective be met if the recycle ratio is 0.5?

a. What specific growth rate is required in the last tank of the SFAS system? Calculate the specific growth rate in the last tank, $\mu_{H,N}$, using Eq. 10.28 and the kinetic parameters for winter conditions, since they will control. The

desired substrate concentration is 10 mg/L. Using the kinetic parameters from Table E10.2 gives:

$$\mu_{H,S} = (4.1)\left(\frac{10}{20 + 10}\right) = 1.37 \text{ days}^{-1}$$

b. What is the smallest fraction of the MLSS that can be in the last tank?
Use of $\mu_{H,S}$ in Eq. 10.41 gives the smallest fraction of the MLSS that can be in the last tank.

$$f_{XM,S} \geq \frac{1/3.0 + 0.15}{(4)(1.37)} = 0.088$$

Thus, as long as more than 8.8% of the MLSS is in the last tank, the effluent quality goal can be met.

c. Will the proposed system be capable of meeting the effluent quality goal?
The actual fraction of MLSS in the last tank can be calculated with Eq. 10.43, or read from Figure 10.19. For a system equivalent to four tanks in series with a recycle ratio of 0.5, the fraction of MLSS in the last tank can be seen from the figure to be 0.175. Thus, the SFAS system is capable of meeting the effluent quality goal. In fact, examination of the figure reveals that the goal can be met no matter what the recycle ratio is.

The next task in the design is to distribute the steady-state oxygen requirement to each of the equivalent tanks. This can be done using the techniques described for spatially distributing the oxygen requirement in CAS systems, but in a simpler manner. The oxygen requirements for heterotrophic biomass synthesis from slowly and readily biodegradable substrate can be calculated with Eqs. 10.23 and 10.24, respectively. Likewise, the requirement for synthesis of autotrophic biomass can be cal-

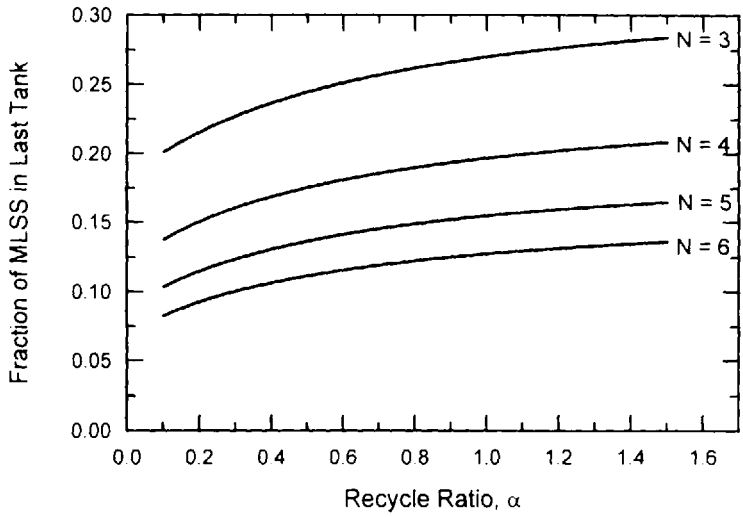


Figure 10.19 Effect of the biomass recycle ratio on the fraction of MLSS in the last tank of a SFAS system, as calculated with Eq. 10.43.

culated with Eq. 10.32. As seen in Figure 7.11, the extent of nitrification in SFAS is essentially the same as in a CMAS system. However, the ammonia-N concentration in the early tanks will be lower, as shown in Figure 7.15. Thus, to be conservative, Eq. 10.32, which assumed that the residual ammonia-N concentration was negligible, should be used without modification. Because the influent to type of SFAS system considered here is distributed evenly among the tanks, all of the synthesis oxygen requirements should be apportioned evenly as well. If some other influent flow distribution were used, the synthesis oxygen requirements should be apportioned in the same manner as the flow. The oxygen requirement due to heterotrophic decay can be calculated with Eq. 10.25, while the requirement for autotrophic decay can be calculated with Eq. 10.33. Both should be apportioned to each tank in accordance with the fraction of the MLSS in each as given by Eq. 10.42. Although the recycle ratio will influence that distribution, the effect will be relatively small so the distribution can be made on the basis of the most commonly used ratio. Because the MLSS concentration decreases from tank to tank down the chain, so will this oxygen requirement. The transient-state oxygen requirement can be distributed in a similar manner by using the approaches developed for CMAS systems and considering the distribution of influent to all tanks.

For equal influent flow distribution to all tanks, the first tank will have the highest oxygen requirement, and thus it is used to determine the minimum acceptable bioreactor volume based on oxygen transfer. This can be calculated with Eq. 10.5. Even though the last tank will have a somewhat lower oxygen requirement than the first, it should be used to calculate the lower limit based on floc shear because its effluent goes to the final settler. That limit can be calculated with Eq. 10.4. Both calculations should be made on the basis of the summer time oxygen requirement, with either the steady- or transient-state requirement being used, depending upon the nature of the influent flow. The last tank will have the lowest oxygen requirement, and it should be used to determine the maximum acceptable bioreactor volume. The upper limit on the size of last tank can be calculated with Eq. 10.3 by using the minimum steady-state or transient-state oxygen requirement (as needed) as calculated for winter operating conditions.

Once the limits on tank volume are known, they can be used with the mass of MLSS in the last tank to establish an acceptable range of MLSS concentrations for that tank. The mass of MLSS in the last tank is just the fraction of biomass in that tank multiplied by $(X_{M,T} \cdot V)_{System}$ as calculated with Eq. 9.11 for winter conditions. Since the MLSS concentration in the last tank is the same as the concentration entering the final settler, various concentrations within the allowable range can be investigated for their effects on final settler size and operation. Once one is selected, it, in turn, establishes the total system volume, just as in the other designs.

Sometimes it may be desirable to alter the configuration of an existing activated sludge system to that of SFAS. In that situation, the planned change must be evaluated to be sure that it can be made while still accomplishing the treatment objectives. The steps for doing that are similar to those for designing a SFAS system, except that the system volume and number of equivalent tanks-in-series are known, thereby fixing the volume of each tank. Thus, the major questions are whether the desired effluent quality can be met with the existing system SRT and whether the needed amount of oxygen can be transferred to each tank while meeting the con-

straints on mixing energy input and oxygen transfer. The steps outlined above can be followed to answer those questions.

Design of Contact Stabilization Activated Sludge Systems. A schematic diagram of a CSAS system is shown in Figure 7.19. There it can be seen that the influent enters the contact tank, from which it flows directly to the final settler. The biomass recycle from the settler flows to the stabilization basin where additional reactions can occur before the biomass is returned to the contact tank. As far as organic substrates are concerned, removal from the wastewater undergoing treatment must occur in the contact tank, where the majority of the soluble substrate is metabolized. Particulate organic matter is entrapped in the mixed liquor, with its degradation occurring in both the contact and stabilization tanks. As discussed in Section 7.4, one characteristic of CSAS is that partial nitrification can occur, with the extent depending on both the system SRT and the fraction of the biomass in the contact tank. CSAS systems are seldom designed specifically to achieve partial nitrification, and thus the design approach presented here is based on biodegradable COD removal. However, because some nitrification will occur if the proper environmental conditions are achieved, its occurrence must be considered or the estimation of the amount of oxygen required in the system will be incorrect. While it is possible to derive solvable analytical equations for organic substrate removal with a few simplifying assumptions that are unlikely to be violated, the same is not true for nitrification. Only approximations can be achieved. This means that the only truly accurate way to estimate the degree of nitrification and the distribution of the associated oxygen requirement is through simulation with a model like ASM No. 1 or No. 2. However, because hand calculations are very useful during preliminary design we will present an approach for using them to evaluate the degree of nitrification likely to occur. Their approximate nature should be recognized, however, and appropriate caution should be exercised in their use.

The first task in the design of a CSAS process is the selection of the system SRT, which requires the consideration of many factors as discussed previously. After that, the relative amount of biomass in the two tanks must be selected to ensure that the desired effluent quality is attained. This, in turn, requires selection of the biomass recycle ratio and the fraction of the system volume allocated to each tank. Because system performance depends on both of those factors, as illustrated in Figures 7.24–7.27, CSAS systems should not be built with separate vessels for the contact and stabilization tanks because such a design fixes their relative volumes. Rather, both should be in the same vessel, with a non–load-bearing, curtain wall between them. This will allow their relative sizes to be changed as circumstances require.

Selection of the fraction of biomass in the contact tank requires the following. The heterotrophic specific growth rate in the contact tank, $\mu_{H,C}$, must be consistent with the desired effluent readily biodegradable substrate concentration. Thus, as in SFAS design, Eq. 10.28 may be used to calculate $\mu_{H,C}$ by setting S_{SI} equal to the desired effluent substrate concentration. The kinetic parameters should be those for the coldest expected operating condition since it will control. Following the same logic used to derive Eqs. 10.29 and 10.30, it can be shown that the specific growth rate in the contact tank is given by:

$$\mu_{H,C} = \left[\frac{(X_{M,T} \cdot V)_{System}(1/\Theta_c + b_H)}{(X_{M,T,C} \cdot V_C)}\right]\left(\frac{S_{SO}}{S_{SO} + X_{SO}}\right) \tag{10.44}$$

when it is assumed that no utilization of slowly biodegradable substrate occurs. In Eq. 10.44, V_C is the volume of the contact tank and $X_{M,T,C}$ is the MLSS concentration in it. Since the value of $\mu_{H,C}$ as calculated with Eq. 10.44 must be less than or equal to the value of $\mu_{H,C}$ associated with the desired effluent substrate concentration as calculated with Eq. 10.28, then:

$$\frac{X_{M,T,C} \cdot V_C}{(X_{M,T} \cdot V)_{System}} = f_{XM,C} \geq \left(\frac{1/\Theta_c + b_H}{\mu_{H,C}}\right)\left(\frac{S_{SO}}{S_{SO} + X_{SO}}\right) \tag{10.45}$$

where $\mu_{H,C}$ has been calculated with Eq. 10.28. In other words, as long as the fraction of MLSS in the contact tank, $f_{XM,C}$, is greater than or equal to the right side of Eq. 10.45, the effluent quality will be acceptable. Since the system contains only two tanks, the remainder of the biomass is in the stabilization basin. The derivation of Eq. 10.44 is based on the assumption that no slowly biodegradable substrate is used in the contact tank. Some hydrolysis will occur, however, allowing a portion of the slowly biodegradable substrate to be used. The effect of slowly biodegradable substrate utilization in the contact tank is to increase the required fraction of biomass in that tank, with the right parenthetical term in Eq. 10.45 approaching one as all of the slowly biodegradable substrate is used there, an event that is unlikely to occur. Because of the uncertainty associated with the fraction of the slowly biodegradable substrate that will be used in the contact tank, a designer should calculate the required minimum value of $f_{XM,C}$ with Eq. 10.45 twice, once as written and once with the right parenthetical term set to one. A design value for $f_{XM,C}$ between those two extremes should then be chosen.

The fraction of the biomass in the contact tank is determined solely by the system hydraulics. If both growth and wastage are neglected, as they were for the SFAS system, then:

$$f_{XM,C} = \frac{\alpha \cdot \upsilon}{1 + \alpha - \upsilon} \tag{10.46}$$

where α is the biomass recycle ratio and υ is the fraction of the system volume in the contact tank:

$$\upsilon = \frac{V_C}{V_C + V_S} = \frac{V_C}{V_T} \tag{10.47}$$

where V_S is the volume of the stabilization basin and V_T is the total system volume. Many combinations of α and υ can result in the same value of $f_{XM,C}$, as shown in Figure 10.20, where Eq. 10.46 is plotted in two ways. Figure 10.20a shows that for any value of υ the fraction of MLSS in the contact tank will increase as the recycle ratio is increased. Thus, the selection of υ should be made for the smallest anticipated recycle ratio. Because of the criterion given in Eq. 10.45, the needed effluent quality will be obtained for any larger value of the recycle ratio.

Example 10.3.5.2

Consider the wastewater that has been the subject of all of the examples in this chapter. Consideration is being given to using a CSAS process with an SRT of 3 days. The effluent quality objective is 10 mg/L as COD of readily biodegradable organic matter. What fraction of the biomass and what fraction of the system

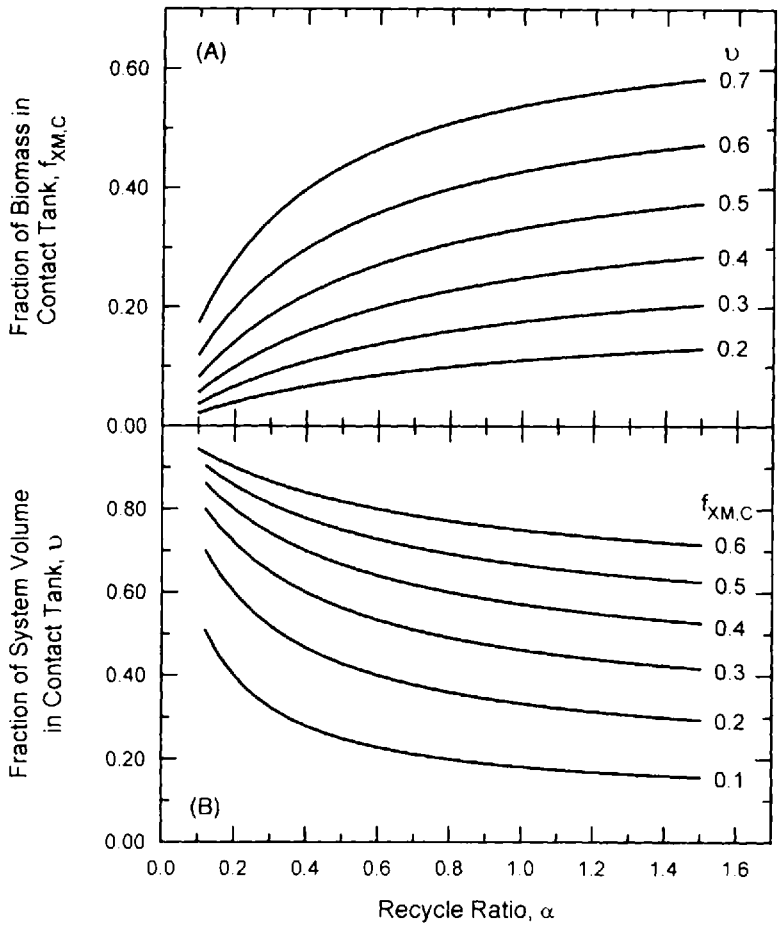


Figure 10.20 Effect of the biomass recycle ratio on: (a) the fraction of biomass in the contact tank, and (b) the fraction of the system volume in the contact tank, for a CSAS system, as calculated with Eq. 10.46.

volume should be in the contact tank if the smallest recycle ratio that will be used is 0.30?

a. What specific growth rate will be required in the contact tank?
Calculate the specific growth rate in the contact tank, $\mu_{H,C}$, using Eq. 10.28 and the kinetic parameters for winter conditions, since they will control. The desired substrate concentration is 10 mg/L. Using the kinetic parameters from Table E10.2 gives:

$$\mu_{H,C} = (4.1)\left(\frac{10}{20 + 10}\right) = 1.37 \text{ days}^{-1}$$

b. What is the smallest fraction of the MLSS that can be in the contact tank?
Use of $\mu_{H,C} = 1.37$ in Eq. 10.45 gives the smallest fraction of the MLSS that can be in the contact tank.

$$f_{XM,C} \geq \left(\frac{1/3.0 + 0.15}{1.37}\right)\left(\frac{115}{115 + 150}\right) = 0.15$$

Thus, assuming that no utilization of slowly biodegradable substrate occurs in the contact tank, as long as 15% or more of the MLSS is in that tank, the effluent quality goal can be met. Recalculation of the fraction with the right parenthetical term set to one shows that at least 35% of the MLSS would have to be in the contact tank if all of the slowly biodegradable substrate was used there. This is unlikely to occur, but to provide a factor of safety, choose a fraction of 25% for design.

c. What fraction of the system tank volume should be in the contact tank with a recycle ratio of 30%?
The required fraction of the system volume in the contact tank, υ can be calculated with a rearranged form of Eq. 10.46:

$$\upsilon = \frac{1 + 0.30}{1 + 0.30/0.25} = 0.59$$

This value could also have been read from Figure 10.20b. Thus, 59% of the system volume must be in the contact tank to ensure that 25% of the system biomass is there. This large fraction is due to the short SRT of the system. Generally, the larger the SRT, the smaller the contact tank as a fraction of the total system.

After the fraction of biomass and the fraction of total system volume in the contact tank have been selected, the oxygen requirement must be distributed between the two reactors. This must be done for summer and winter conditions, and for the maximum and minimum anticipated recycle ratios. The minimum oxygen requirement will occur in the winter, and it will be used to determine the upper limit on the total system volume, just as with the other systems. Likewise, the summer conditions will give the maximum oxygen requirement, which determines the lower limit on the system volume, as discussed previously. Increasing the recycle ratio shifts some of the oxygen requirement from the stabilization basin to the contact tank, and also increases the oxygen requirement due to nitrification, as shown in Figures 7.24 and 7.25. Consequently, the extreme conditions must be examined so that the controlling situation can be identified.

The heterotrophic oxygen requirement can be distributed by a technique similar to that used to distribute the oxygen requirement in a CAS system, with special consideration of the unique characteristics of the CSAS system. The oxygen requirement for synthesis of biomass from readily biodegradable substrate can be calculated with Eq. 10.23. All of it will occur in the contact tank. The oxygen requirement for synthesis of biomass from slowly biodegradable substrate can be calculated with Eq. 10.24. Because slowly biodegradable substrate is entrapped in the MLSS where it undergoes hydrolysis, which is a slow reaction, it is likely to occur in both tanks. Although the relative amount of slowly biodegradable substrate utilization occurring in each tank can only be determined by simulation with a model like ASM No. 1 or 2, in the absence of such simulations it seems reasonable to distribute the oxygen requirement to the two tanks in proportion to the mass of MLSS in them. The oxygen requirement due to heterotrophic biomass decay can be calculated with Eq. 10.25. It, too, should be distributed in proportion to the mass of MLSS in each tank. Because

of the effects of the recycle ratio on the distribution of MLSS in the system, its extreme values should be considered. The maximum value will maximize the heterotrophic oxygen requirement in the contact tank and minimize it in the stabilization basin, whereas the minimum recycle ratio will have just the opposite effect.

Distribution of the oxygen requirement associated with nitrification is more difficult and cannot be done precisely with analytical equations. However, it is possible to approximate it for those situations in which substantial nitrification will occur, i.e., when the system SRT is well in excess of the minimum SRT for nitrification. The rationale used to derive the equations is similar to that used to derive Eqs. 10.44 and 10.45, as well as 10.29 and 10.30, with additional criteria. First, nitrification will occur in the stabilization basin, making the concentration of ammonia-N entering the contact tank in the RAS flow approach zero. Second, when stable partial nitrification is occurring in the contact tank, it cannot be assumed that the mass of nitrogen oxidized in the stabilization tank is much less than the mass oxidized in the contact tank. In fact, it may be more. Furthermore, neither can it be assumed that the change in ammonia-N concentration across the contact tank is approximately equal to change in concentration across for an equivalent CMAS system, since more nitrification will occur in the latter. Rederiving Eq. 10.30 with these changes, and with the assumption that all forms of nitrogen are equally available in both tanks, gives the following approximate expression for the fraction of autotrophs in the contact tank, $f_{XA,C}$:

$$f_{XA,C} \approx \frac{1/\Theta_c + b_A}{\mu_{A,C}} \left[1 - \alpha \left(\frac{S_{NHC}}{S_{N,a} - S_{NHC}} \right) \right] \tag{10.48}$$

where $\mu_{A,C}$ is the specific growth rate of the autotrophs in the contact tank as calculated with Eq. 10.28 using S_{NHC}, which is the ammonia-N concentration in that tank. The fraction of autotrophs in the contact tank is the same as the fraction of the MLSS in the contact tank, $f_{XM,C}$, because the mixed liquor is homogeneous throughout the system. Substitution of Eq. 10.28 into Eq. 10.48 and rearrangement gives a quadratic equation allowing the effluent ammonia-N concentration, S_{NHC}, to be calculated for any given biomass recycle ratio, α, with its associated value of $f_{XM,C}$, as determined by Eq. 10.46:

$$\left(1 + \alpha - \frac{f_{XM,C} \cdot \hat{\mu}_A}{1/\Theta_c + b_A} \right) S_{NHC}^2 + \left[K_{NH}(1 + \alpha) - S_{N,a} \left(1 - \frac{f_{XM,C} \cdot \hat{\mu}_A}{1/\Theta_c + b_A} \right) \right] S_{NHC} - K_{NH} \cdot S_{N,a} = 0 \tag{10.49}$$

where $S_{N,a}$ is the available nitrogen concentration, as given by Eq. 10.17. The value obtained from Eq. 10.49 is only approximate because of the approximate nature of Eq. 10.48. Use of the highest anticipated value of α in the solution of Eq. 10.49 will minimize the ammonia-N concentration in the contact tank, which will maximize the oxygen requirement associated with nitrification in that tank, while minimizing the oxygen requirement in the stabilization basin. Use of the lowest anticipated recycle ratio will have just the opposite effect.

Once the ammonia-N concentration in the contact tank has been estimated, the oxygen requirement due to nitrification can be apportioned to the two vessels. Because particulate organic nitrogen will have a different fate than soluble nitrogen, a

distinction must be made between the two, even though that was not done in the derivation of Eq. 10.48. Let $S'_{N,a}$ be the available soluble nitrogen:

$$S'_{N,a} = S_{N,a} - X_{SNO} \tag{10.50}$$

Then the oxygen requirement due to synthesis of autotrophic biomass from soluble nitrogen in the contact tank can be calculated from a modified form of Eq. 10.32:

$$RO_{A,SN,C} = F(S'_{N,a} - S_{NH,C})(4.57 - Y_{A,T}i_{O\,XB,T}) \tag{10.51}$$

The oxygen requirement in the stabilization tank associated with synthesis of autotrophic bacteria from soluble nitrogen is then:

$$RO_{A,SN,S} = \alpha \cdot F \cdot S_{NH,C}(4.57 - Y_{A,T}i_{O\,XB,T}) \tag{10.52}$$

The value of α used in Eq. 10.52 should be consistent with the value used to solve Eq. 10.49. The total oxygen requirement associated with synthesis of autotrophic biomass from particulate organic nitrogen can be estimated as:

$$RO_{A,XNS} = F \cdot X_{NSO}(4.57 - Y_{A,T}i_{O\,XB,T}) \tag{10.53}$$

It is apportioned to each of the vessels in proportion to the fraction of biomass in each. That apportionment will depend on the recycle ratio, as discussed previously. The oxygen requirement due to decay of autotrophs can be ignored because its contribution is less than the uncertainty associated with the other determinations.

Transient-state oxygen requirements can be estimated by combining the techniques presented in Section 10.3.2 with those presented above. As with the other multitank activated sludge systems, there will be considerable uncertainty associated with the estimates because of the need to distribute the transient oxygen requirement spatially. Consequently, simulation with a model like ASM No. 1 is a much more certain way of estimating such requirements.

Example 10.3.5.3

Continue considering the CSAS system whose analysis was begun in Example 10.3.5.2. What would the summer time steady-state oxygen requirement be if 59% of the system volume was in the contact tank and the system was being operated with a recycle ratio of 0.30? As determined in Example 10.3.5.2, the contact tank of such a system would contain 25% of the MLSS.

a. What is the heterotrophic oxygen requirement in each tank?
The oxygen requirement for heterotrophic synthesis from readily biodegradable substrate can be determined with Eq. 10.23. Since the SRT is the same as in all previous examples, this computation is the same as the one made in Part a of Example 10.3.4.1. Thus, $RO_{H,SS}$ = 1,840 kg O_2/day. All of this will occur in the contact tank.

The oxygen requirement for heterotrophic synthesis from slowly biodegradable substrate can be determined with Eq. 10.24. This computation is the same as the one made in Part b of Example 10.3.4.1. Thus, $RO_{H,XS}$ = 2,400 kg O_2/day. Since the contact tank contains 25% of the biomass, 25% of this oxygen requirement will occur in it and 75% in the stabilization tank. Thus,

$$RO_{H,XS,C} = (0.25)(2{,}400) = 600 \text{ kg } O_2\text{/day}$$

$$RO_{H,NS,S} = (0.75)(2{,}400) = 1{,}800 \text{ kg } O_2/\text{day}$$

The oxygen requirement associated with heterotrophic decay can be estimated with Eq. 10.25. This was done in Part c of Example 10.3.4.1. Thus, $RO_{H,D} = 2{,}020$ kg O_2/day. It should also be distributed to the two vessels in proportion to the fraction of the MLSS that each contains. Thus,

$$RO_{H,D,C} = (0.25)(2{,}020) = 505 \text{ kg } O_2/\text{day}$$

$$RO_{H,D,S} = (0.75)(2{,}020) = 1{,}515 \text{ kg } O_2/\text{day}$$

Summing these values gives the heterotrophic oxygen requirement in each tank:

$$RO_{H,C} = 1{,}840 + 600 + 505 = 2{,}945 \text{ kg } O_2/\text{day}$$

$$RO_{H,S} = 1{,}800 + 1{,}515 = 3{,}315 \text{ kg } O_2/\text{day}.$$

b. What is the autotrophic oxygen requirement in each tank?
The first task is to estimate the concentration of ammonia-N leaving the contact tank using Eq. 10.49. In that equation, $\alpha = 0.30$, $f_{XM,C} = 0.25$, $\hat{\mu}_A = 1.3$ day^{-1}, $\Theta_c = 3$ days, $b_A = 0.12$ day^{-1}, $K_{NH} = 1.9$ mg N/L, and $S_{NH} = 30.5$ mg N/L. Setting up the quadratic equation gives:

$$0.583\, S_{NH,C}^2 - 6.162\, S_{NH,C} - 57.950 = 0$$

Solution of this equation reveals that $S_{NH,C}$, the ammonia-N concentration leaving the contact tank, is 16.6 mg /L as N. This compares to a concentration of 2.1 mg/L from the CMAS system (see Example 10.3.3.2). This difference is consistent with the differences between the two systems shown in Figure 7.20.

The oxygen requirement from oxidation of soluble nitrogen in the contact tank can be calculated with Eq. 10.51 after $S'_{N,e}$ has been estimated with Eq. 10.50 using the value of $S_{N,e}$ of 30.5 calculated in Example 10.3.3.2:

$$S'_{N,e} = 30.5 - 8.5 = 22.0 \text{ mg/L as N}$$

$$RO_{A,NS,C} = (40{,}000)(22.0 - 16.6)[4.57 - (0.2)(1.20)]$$

$$= 935{,}000 \text{ g } O_2/\text{day} = 935 \text{ kg } O_2/\text{day}$$

The oxygen requirement associated with oxidation of soluble nitrogen in the stabilization basin can be calculated with Eq. 10.52 using a value of 16.6 mg/L as the ammonia-N concentration entering the basin through the RAS flow, which is 30% of the system influent flow:

$$RO_{A,NS,S} = (0.30)(40{,}000)(16.6)[4.57 - (0.2)(1.20)]$$

$$= 862{,}000 \text{ g } O_2/\text{day} = 862 \text{ kg } O_2/\text{day}$$

The oxygen requirement from oxidation of particulate organic nitrogen can be calculated with Eq. 10.53:

$$RO_{A,NS} = (40{,}000)(8.5)[4.57 - (0.2)(1.20)]$$

$$= 1{,}470{,}000 \text{ g } O_2/\text{day} = 1{,}470 \text{ kg } O_2/\text{day}$$

Since 25% of the biomass is in the contact tank, 25% of this oxygen requirement will be exerted there, and 75% in the stabilization basin. Consequently, the oxygen requirement from oxidation of particulate organic nitrogen in the two vessels is:

$RO_{A,NSS,C}$ = (0.25)(1,470) = 368 kg O_2/day

$RO_{A,NSS,S}$ = (0.75)(1,470) = 1,102 kg O_2/day

Summing these values gives the autotrophic oxygen requirement in each tank:

$RO_{A,C}$ = 935 + 368 = 1,303 kg O_2/day

$RO_{A,S}$ = 862 + 1,102 = 1,964 kg O_2/day.

c. What is the total oxygen requirement in each tank?
The total oxygen requirement in each tank is obtained by summing the heterotrophic and autotrophic requirements:

RO_C = 2,945 + 1,303 = 4,248 kg O_2/day

RO_S = 3,315 + 1,964 = 5,279 kg O_2/day

The total oxygen requirement in the whole system is 9,530 kg O_2/day, which is less than that required by the CMAS system as calculated in Example 10.3.3.2. The difference is due to the lower degree of nitrification in the CSAS system.

The contact tank comprises 59% of the system volume and experiences 45% of the oxygen requirement, whereas the stabilization basin contains 41% of the system volume and experiences 55% of the oxygen requirement. These percentages are specific to the operating conditions imposed and depend on the SRT, the recycle ratio, and the fraction of system volume in each tank, as illustrated in Figures 7.25 and 7.27.

After the oxygen requirements have been estimated for winter and summer and for both the maximum and minimum planned recycle ratios, the next task in the design of a CSAS system is to determine the total bioreactor volume, thereby fixing the volume of each tank, the MLSS concentration in each, and the final settler size. This is done in the same way as for the other activated sludge designs. An extra level of complexity is involved, however, because it is not apparent before the exercise whether the contact tank or the stabilization basin will control the design. For instance, in the example above, the stabilization basin had the greater requirement per unit volume, with 55% of the oxygen requirement in 41% of the volume. Thus, for that operational situation (summer with minimum recycle flow), the stabilization basin will control the minimum system volume. The engineer must consider all of the possible scenarios to determine which actually control the upper and lower limits on the system volume. Once those volumes have been determined, the range of permissible MLSS concentrations is fixed by the mass of MLSS in the system as calculated with Eq. 9.11. Finally, that information is used with the anticipated recycle ratios and projections of solids settling properties to arrive at an economic combination of bioreactor and settler sizes.

10.3.6 Batch Reactors—Sequencing Batch Reactor Activated Sludge

As discussed in Section 7.8, batch processes offer opportunities for flexibility that can be advantageous in some circumstances. For example, when flows are highly variable or when the character of the contaminants changes on a regular and periodic

basis, batch processes allow those special needs to be met. Furthermore, by changing the length of the fill period within a cycle, batch processes can be made to behave like continuous flow processes with hydraulic characteristics anywhere between a perfect CSTR and a perfect plug–flow reactor (PFR). Consequently, sequencing batch reactor activated sludge (SBRAS) systems have found increasing popularity in recent years.

Although SBRAS systems are often treated as if special design techniques are required, they actually operate according to the same principles as other activated sludge processes, as discussed in Section 7.8.2, and can be designed according to them. Consequently, the mass of MLSS in the system is defined by Eq. 9.11, the overall daily solids wastage rate is given by Eq. 9.12, the total daily heterotrophic oxygen requirement is given by Eq. 9.13, and the total daily autotrophic oxygen requirement by Eq. 10.16, provided that the effective SRT is used in their computation. The three primary differences between SBRAS and other activated sludge systems are that the oxygen requirement must be distributed to each of the discrete operating cycles and then apportioned within each cycle in a manner consistent with the length of the fill period, the fraction of each operating cycle that is not devoted to biological reaction must be accounted for in the design, and the interaction between the bioreactor and the secondary clarifier must be analyzed differently because biological reaction and sedimentation occur in the same vessel (although at different times in the cycle). The first of these can be accomplished by the same techniques used to apportion steady- and transient-state oxygen requirements in the various continuous-flow activated sludge systems. The second can be accounted for through use of the effective HRT, τ_c, and the effective SRT, Θ_{cc}, when using the CMAS equations listed above. The third can be considered through simple procedures to be presented below.

The effective HRT and effective SRT account for the portion of the SBRAS cycle that is not utilized for biological reaction, i.e., that portion devoted to settling and decanting (draw), as illustrated in Figure 7.42. In Section 7.8.2, ζ is defined as the fraction of the total cycle devoted to fill plus react. Using those definitions, the effective HRT is defined as:

$$\tau_c = \frac{\zeta \cdot V}{F} \tag{7.8}$$

and the effective SRT is defined as:

$$\Theta_{cc} = \frac{\zeta \cdot V \cdot X_M}{F_w \cdot X_{Mw}} \tag{7.9}$$

Selection of the effective SRT is governed by the same considerations as the selection of the SRT for any other activated sludge system. Relatively low values can be used if the removal of biodegradable organic matter is the primary objective. However, SBRAS systems are often used in small wastewater treatment plants where sludge stabilization is also an important consideration. Thus, effective SRTs of 10 days or more are often used.

Selection of the system volume, and hence the effective HRT, must consider not only mixing and oxygen transfer, but also the anticipated settling properties of the MLSS. That is necessary because the bioreactor volume must be sufficient to

contain not only the flow added per cycle, F_c, but the volume retained after effluent is decanted to contain the recycled biomass, V_{br}, where:

$$F_c = \frac{F}{N_c} \tag{7.11}$$

and

$$V_{br} = \alpha \cdot F_c \tag{7.12}$$

As discussed in Section 7.8.2, N_c is the number of cycles per day and α is analogous to the recycle ratio in a continuous flow system. Consequently, when an SBRAS system is being designed as a simple activated sludge system with no nutrient removal capabilities:

$$V = F_c(1 + \alpha) \tag{10.54}$$

or:

$$V = F_c + V_{br} \tag{10.55}$$

The primary factors considered in the selection of both N_c and ζ are adding influent wastewater and removing treated effluent, allowing a sufficient time for settling and decanting in the operating cycle, and allowing for peak hydraulic flows through the bioreactor. These factors, in turn, are affected by the total number of bioreactors selected and the desired operational cycle. Values of N_c often range from 4 to 6 cycles per day. Since the time required for settling and decanting is relatively constant, whereas the length of a cycle gets shorter as the number of cycles per day is increased, the value of ζ decreases as N_c increases. Nevertheless, values of ζ often range from 0.5 to 0.7. Considerable flexibility exists in the selection of both parameters.

The minimum bioreactor volume associated with a given number of cycles per day will result when V_{br} is just large enough to contain the MLSS retained for use in the next cycle after the effluent has been decanted. The mass of MLSS in the bioreactor after solids settling and effluent decanting is the same as the mass of MLSS at the end of the react period, which is given by Eq. 9.11. Therefore:

$$X_{M,Tr} \cdot V_{br} = (X_{M,T} \cdot V)_{System} \tag{10.56}$$

where $X_{M,Tr}$ is the settled MLSS concentration. Thus, it can be seen that when the settled MLSS concentration is as large as possible, the retained volume, V_{br}, will be as small as possible. Although experience with treatment of a particular wastewater is the best way to select the maximum achievable value of $X_{M,Tr}$, it can also be estimated from the definition of the sludge volume index (SVI). As discussed in Section 10.2.1, the SVI is defined as the volume in mL occupied by a gram of solids after 30 minutes of quiescent settling. If we take the SVI as being indicative of the highest concentration to which the MLSS can be settled, then the maximum attainable settled solids concentration, $X_{M,Tr,max}$, (in mg/L) will be given by:

$$X_{M,Tr,max} \approx \frac{10^6}{SVI} \tag{10.57}$$

It follows from Eq. 10.56 that the smallest retained volume, $V_{br,min}$, is given by:

$$V_{bf,min} = \frac{(X_{M,T} \cdot V)_{System}}{X_{M,T,max}} \tag{10.58}$$

Consequently, the lower limit on the sequencing batch reactor (SBR) volume is obtained by substituting Eq. 10.58 into Eq. 10.55:

$$V_T = F_c + \frac{(X_{M,T} \cdot V)_{System}}{X_{M,T,max}} \tag{10.59}$$

Although care should be exercised in selection of the $X_{M,T,max}$, SBRAS systems offer some flexibility in the control of the MLSS settling properties. Recall that SBRs can achieve any hydraulic characteristic between a CSTR and a PFR simply by changing the length of the fill period. This means that the instantaneous process loading factor can be changed simply by changing the length of the fill period. Since the process loading factor influences the competition between filamentous and floc-forming bacteria, as discussed in Section 10.2.1, changing the length of the fill period allows control of the settling properties, provided that the oxygen transfer system is capable of meeting the imposed oxygen requirement. Thus, steps can be taken during operation to ensure that the selected SVI is achieved, as long as a realistic value was selected during design.

With the above information in mind, we can now set forth the steps in the design of an SBR system. After selection of the effective SRT, the mass of MLSS in the system is calculated with Eq. 9.11. The number of cycles per day is then selected, taking into account the length of time in each cycle to be devoted to settling and decanting. Since the flow to be treated is known, this fixes both F_c and ζ. Estimation of an attainable SVI value allows computation of $X_{M,T,max}$ with Eq. 10.57, which allows calculation of the minimum possible bioreactor volume with Eq. 10.59. Selection of the design bioreactor volume requires consideration of the oxygen requirement and the constraints on the rate of oxygen transfer and floc shear, just as with all of the other activated sludge systems. Equations 9.13 and 10.16 can be used to calculate the total daily oxygen requirement. While the mass of oxygen required per cycle will just be the total daily requirement divided by the number of cycles per day, the oxygen transfer rate will depend on the length of the fill period during that cycle. If the fill period lasts throughout the entire react cycle, then the system will behave like a CSTR and the oxygen transfer rate will be equivalent to that in a CMAS system, allowing the techniques for that system to be used. On the other hand, the shorter the fill period relative to the react period, the more the system will behave like a PFR, requiring the techniques used during CAS design to be used. No matter which technique is employed, however, the design reactor volume must be selected to ensure proper mixing and oxygen transfer. Floc shear is seldom a problem with an SBRAS process because flocculation will generally occur later in the react period when aeration rates are lower. Suspension of solids is not a problem in facilities which provide separate mixing equipment. Nevertheless, the designer should verify, either by calculation or logical analysis, that these constraints will be satisfied. After the volume is known, the anticipated MLSS concentration can be calculated by dividing $(X_{M,T} \cdot V)_{System}$, as calculated with Eq. 9.11, by the selected bioreactor volume. The effective HRT can then be determined with Eq. 7.8 and the flow rate of waste solids can be calculated with Eq. 7.9 after it has been determined when in the cycle solids will be wasted, thereby determining $X_{M,Tw}$. The required solids mass

wastage rate is also given by Eq. 9.12, so a system check can be made. The following example illustrates the unique aspects of SBRAS design.

Example 10.3.6.1

An SBRAS facility is to be constructed to treat the wastewater considered throughout Section 10.3. A partial degree of sludge stabilization is important, so an effective SRT of 10 days is to be maintained. Four cycles per day are to be utilized at average flow to allow more cycles to be used to treat peak flows. A value of ζ of 0.5 will be used for design purposes, again to allow operational flexibility during peak flows. Experience indicates that the SVI in the process will generally not exceed 120 mL/g. What is the minimum possible bioreactor volume that could be used and what MLSS concentration would be associated with its use? What effective HRT would the SBRAS have at average flow?

a. What value of $(X_{M,I} \cdot V)_{System}$ should be used for design purposes?
$(X_M \cdot V)_{System}$ is calculated using Eq. 9.11 at the winter operating temperature of 15°C. Values of all of the parameters in Eq. 9.11 are given in Table E10.2. Using these values and Θ_c = 10 days:

$$(X_{M,I} \cdot V)_{System} = (10.0)(40{,}000) \cdot \left[43.5 + \frac{[1 + (0.20)(0.15)(10.0)](0.5)(115 + 150)}{1 + (0.15)(10.0)} \right]$$
$$= 45{,}000{,}000 \text{ g MLSS}$$

b. What value of $X_{M,I,max}$ is appropriate for this design?
As noted above, the SVI for this application is expected to be less than 120 mL/g. Using Eq. 10.57,

$$X_{M,I,max} = \frac{10^6}{SVI} = \frac{10^6}{120} = 8{,}300 \text{ mg/L} = 8{,}300 \text{ g/m}^3$$

c. What is the minimum possible bioreactor volume?
Since there will be four cycles per day, and the average daily flow rate is 40,000 m^3/day, the flow per cycle, F_c, is 10,000 m^3. Substitution of this value into Eq. 10.59 gives:

$$V_I = 10{,}000 + \frac{45{,}000{,}000}{8{,}300} = 15{,}400 \text{ m}^3$$

d. What is the bioreactor MLSS concentration?
This is calculated by dividing the mass of MLSS in the system, as calculated in Part a, by the bioreactor volume.

$$X_{M,I} = \frac{45{,}000{,}000}{15{,}400} = 2{,}900 \text{ g/m}^3 = 2{,}900 \text{ mg/L}$$

e. What is the effective HRT of the bioreactor.
From Eq. 7.8,

$$\tau_c = \frac{\zeta V}{F} = \frac{(0.5)(15{,}400)}{40{,}000} = 0.19 \text{ day}$$

10.3.7 Process Optimization Using Dynamic Models

As discussed in Section 9.4.3, dynamic simulation is an important tool that allows an engineer to refine the design of an activated sludge system or to evaluate alternative operating strategies for an existing facility. The necessity for using this tool increases as the system to be designed is further removed from the assumptions inherent in the simple stoichiometric models that are the basis for the analytical expressions in this chapter. For example, if a CMAS system were to be used to treat a wastewater flowing at a constant rate and containing a constant concentration of soluble pollutants, the approach presented in this chapter would be very accurate because the assumptions in the model are totally consistent with the nature of the problem. In fact, the equations would even do a very reasonable job under dynamic loading conditions. If part of the pollutants were particulate, on the other hand, the equations would still be accurate for a CMAS system operated under constant loading conditions, but they would be more approximate as dynamic loads were applied because they contain no rate expressions for hydrolysis of particulate substrates. However, by using approximations based on experience, it is still possible for the approach to give adequate information about transient-state oxygen requirements in a single tank system. The approaches presented in this chapter are weakest for design of multitank systems treating wastewaters containing both soluble and particulate pollutants, regardless of the nature of the influent flow. This is because the apportionment of the oxygen requirement among the various tanks requires assessment of the rates of degradation of both soluble and particulate constituents, information that is not incorporated into the simple model upon which the approaches are based. However, experience can help the engineer make the decisions required, although the information will always be approximate. Consequently, as the system configuration becomes more complex and as the influent conditions become more dynamic, the engineer must exercise more caution in the application of the approaches presented and should rely more and more on experience to make decisions.

Given the situation described in the preceding paragraph, what should be done when the experience base is small, either for an individual engineer or for the profession? The answer to that question is to rely more on dynamic simulation. As discussed in Part II of this text, IAWQ ASM Nos. 1 and 2 adequately represent a number of suspended growth biological treatment systems, and are particularly effective for activated sludge systems of the type presented here. Furthermore, software packages implementing them are readily available, as indicated in Table 6.4. Using the techniques described in Chapter 8, the parameters in ASM No. 1 can be assessed with sufficient accuracy to allow it to adequately mimic the performance of real systems.[7] Therefore, an engineer can use the approaches presented in this chapter to decide on tentative sizes for the various bioreactors in an activated sludge system, and then use simulation to investigate the oxygen requirement in each vessel in the system under a variety of anticipated dynamic loading scenarios. That output can then be used to evaluate the ability to transfer the needed oxygen while meeting the constraints on floc shear and mixing, allowing modification of the design as needed. Additional rounds of simulations can then be used to further refine the design. Because the model is known to reflect reality,[7] the engineer can have more confidence in the proposed design than he/she could have in a design based on the approximate approaches presented here. This allows smaller factors of safety to be used, resulting in more economic designs, etc.

In addition to the benefits presented above, simulation allows a neophyte engineer to build an experience base in a relatively short time. By investigating a variety of activated sludge variations by simulation, and comparing the output to the approximations obtained through the analytical equations, an engineer can quickly learn how much uncertainty is associated with the approximate analytical approach for a given configuration. That information then becomes part of the engineer's experience base, which allows better use to be made of the approximate techniques when they are suitable. Returning to the concepts presented in Chapter 9, a good engineer will use the tool that is appropriate for the job at hand. Simulation is one of those tools, but it should not be viewed as something that is only justified for large and complex systems. Rather, the ease with which it can be done suggests strongly that it should be used to learn more about systems, just as it was used in Chapters 6 and 7, thereby increasing the engineer's experience base. Thus, it should be thought of as an important extension of this book.

10.4 PROCESS OPERATION

Control of activated sludge systems is accomplished by the application of the principles described throughout this book. While common sense and a good understanding of the process are keys to successful activated sludge operation, some specialized techniques have been developed. These specialized techniques are discussed in this section.

10.4.1 Solids Retention Time Control

Throughout this book, we have emphasized the importance of SRT in determining the performance and operating characteristics of suspended growth bioreactors. Consequently, control of the SRT is the key to achieving reliable, consistent performance from activated sludge systems.[38] It is important to recognize, however, that control of the SRT is a long-term control strategy, allowing the system to respond to long-term changes in process loadings and to seasonal differences in required operating conditions. This is because a time equivalent to 2 to 3 SRTs will be required before a performance change is observed as a result of a change in the SRT.

Determination of Solids Wastage Rate. The equation defining SRT (Eq. 5.1) was based on the assumption that the biomass separator, i.e., the final clarifier, was perfect, so that no biomass was lost in the final effluent. However, as discussed in Section 8.1, all pilot- and full-scale clarifiers lose some suspended solids to the effluent, and this must be accounted for in the computation of the SRT. For that situation, Eq. 8.1 must be used to calculate the SRT:

$$\Theta_c = \frac{V \cdot X_M}{F_w \cdot X_{Mw} + (F - F_w)X_{Me}} \tag{8.1}$$

Examination of Eq. 8.1 reveals that suspended solids are lost from the process in two ways: (1) in the WAS ($F_w \cdot X_{Mw}$), and (2) in the process effluent ($(F - F_w)X_{Me}$). Loss of solids in the WAS is often referred to as intentional wastage, while loss in the process effluent is often called unintentional wastage. Under certain circumstances, unintentional wastage can represent a sizable proportion of the total solids

wastage from the system, and exclusion of it from the SRT calculation can result in a significant error.[56] For the more general situation in which the MLSS concentration is not uniform throughout the activated sludge system, the following expression applies when wastage is from a single point:

$$\Theta_c = \frac{\sum_{i=1}^{n} V_i \cdot X_{Mi}}{F_w \cdot X_{Mw} + (F - F_w)X_{Me}} \tag{10.60}$$

In this case, the volume V_i with its associated MLSS concentration, X_{Mi}, may represent a distinct bioreactor, such as in CSAS, or it may represent a zone within a bioreactor, such as in SFAS. Rearranging Eq. 10.60 provides a general expression from which required sludge wastage rate can be calculated:

$$F_w = \frac{\dfrac{\left(\sum_{i=1}^{n} V_i \cdot X_{Mi}\right)}{\Theta_c} - (F - F_w)X_{Me}}{X_{Mw}} \tag{10.61}$$

Because the SRT is expressed in days, F_w, is the volume of sludge that must be wasted every day, expressed as an average daily flow rate. Solids may be wasted continuously at rate F_w, or they may be wasted discontinuously so that the total volume wasted in a day equals F_w. Continuous wastage maintains a more constant solids inventory.

Two different values of the SRT can be calculated, depending on whether the activated sludge in the clarifier is included in the calculation. Unless denitrification occurs, resulting in sludge flotation and the uncontrolled loss of suspended solids in the clarifier effluent, the solids in the clarifier do not influence the performance of an activated sludge system.[55] That is why they were excluded from the definition of SRT used in Part II of this book. However, in practice, inclusion of the solids within the clarifier is necessary to achieve proper control over the solids inventory within the activated sludge system. For those systems in which the mass of solids in the clarifier is typically a small portion of the total system inventory, exclusion of the clarifier from the SRT calculation is acceptable. On the other hand, for those systems in which the solids in the clarifier are a significant and/or variable portion of the total system inventory, the clarifier must be included in the SRT calculation. In this case, the SRT based on the solids within the bioreactor alone must also be calculated and monitored to ensure that sufficient biomass is in contact with the incoming wastewater to achieve the necessary treatment.

Control of the SRT provides the required mass of biomass in an activated sludge system. For systems in which the MLSS concentration is not uniform throughout, such as CSAS and SFAS, it is important that the solids be distributed appropriately to achieve the desired performance. This requires adjustment of the solids recycle ratio, or the relative volumes allocated to the various zones, to achieve the desired specific growth rate at the point of effluent discharge to the clarifier, as discussed in Chapter 7 and Section 10.3.

Because maintenance of the SRT is a long-term control strategy, it is typically done by using running averages, in which the desired SRT itself is the time period

over which the averages are calculated.[74] Two techniques are typically used. In one, values of the SRT are calculated on a daily basis and used to calculate the running average. In the other, running averages are calculated for the various solids concentrations and flow rates, which are measured on a daily basis, and those running averages are used to calculate the SRT. Examples of both approaches can be found in activated sludge operations manuals.[80,84] The fact that the SRT provides long-term control of the activated sludge process is reflected in the typical rule-of-thumb that no operating parameter should be changed by more than 20% on any given day. This suggestion is particularly appropriate for the solids wastage rate.

Although continuous wastage of activated sludge maintains the most constant solids inventory, it is not practiced at all plants. Rather, wastage is done discontinuously, for a variety of reasons. In that case, the solids wasting frequency should be selected relative to the SRT so that it does not result in more than a 5 to 10% variation in the MLSS concentration. Thus, if a plant is being operated at a short SRT, solids wastage should occur several times a day, such as once a shift. On the other hand, if the SRT is long, it may be possible to waste solids only once a day while still keeping the variation in MLSS concentration less than 10%.

When solids are not wasted continuously, care must be exercised to ensure that the measured waste solids concentration is truly representative. Collection of a composite sample of the waste activated sludge is critical because the solids concentration during the period when wastage is occurring may be quite different from the daily average solids concentration at the wastage location. This can be the case when solids are wasted from the RAS stream and RAS flow rates are not proportional to the influent flow rate. It is especially true if wasting is done from a solids hopper within the clarifier. Due to the critical nature of the waste solids concentration in determining the SRT, waste activated sludge sampling and analysis procedures should be verified for each system.

Solids Retention Time Control Based on Direct Analysis of Mixed Liquor Suspended Solids Concentration. Using this approach, the mass of suspended solids in the activated sludge system and the mass wasted intentionally and unintentionally are measured directly. The SRT is then calculated from the measured values. Flow rates are measured on a continuous basis, and daily totalized values are determined. Those values give the average daily flow rate for each relevant stream. Composite samples are collected from each stream, and the suspended solids concentration is measured for each. MLSS concentrations are typically expressed on a TSS basis, but they can also be measured and expressed on a COD or VSS basis. The basis of measurement has no impact on the calculation of the SRT as long as only one basis is used. This can be seen by examining Eqs. 10.60 and 10.61.

A major problem associated with SRT control based on MLSS analysis is that information about the waste solids concentration is not available until after the solids have been wasted, making it necessary to base the wastage calculation on an estimate of that concentration. There are two reasons for this. One is the need to use a composite sample of the waste solids stream, as discussed above. The other is the lag time required for a suspended solids analysis. Time is required to take a sample, filter it, dry it, and weigh it. As a consequence, estimates of the suspended solids concentrations required in Eqs 8.1, 10.60 and 10.61 are historical, not current, which is one reason running averages are usually used in the computation. This is another

reason for never changing the solids wastage rate by more than 20% in any given day.

Solids Retention Time Control Based on Centrifuge Analysis of Mixed Liquor Suspended Solids Concentration. Centrifuge analysis is often used to reduce the lag time associated with MLSS analysis.[80,81] A sample is placed in a graduated centrifuge tube and centrifuged for a standard time at a standard speed, after which the volume of the resulting solids pellet is determined. Suspended solids concentrations are measured on parallel samples, and a relationship between the solids pellet volume in the centrifuge tube and the MLSS concentration of the sample is developed. This relationship can then be used as a calibration curve for estimation of the MLSS concentration when the centrifuge analysis is applied to other samples. The benefit of this approach is that it can be accomplished very quickly, thereby reducing the lag time between sample collection and analysis. It also reduces the number of time consuming MLSS analyses that must be conducted, although periodic analyses are required to verify the calibration.

Hydraulic Control of Solids Retention Time. Hydraulic control of the SRT utilizes relationships between the suspended solids concentrations in the bioreactor and the wastage to eliminate suspended solids concentrations from the SRT calculation.[59,77] As a consequence, only the bioreactor volume and process flow rates are required to calculate the SRT for activated sludge process variations with uniform MLSS concentrations. If the mass flow rate of suspended solids in the effluent is negligible compared to the mass flow rate of suspended solids in the wastage, then, Eq. 8.1 may be simplified to:

$$\Theta_c \approx \frac{V \cdot X_M}{F_w \cdot X_{Mw}} \tag{10.62}$$

This expression is equivalent to Eq. 5.1, except that an approximately equal to sign has been used to emphasize that the expression is approximate and depends on the validity of the assumption concerning the effluent suspended solids. For the Garrett wastage approach, in which solids are wasted directly from the bioreactor, X_{Mw} is equal to X_M. Substituting this into Eq. 10.62 gives an expression equivalent to Eq. 5.2:

$$\Theta_c \approx \frac{V}{F_w} \tag{10.63}$$

For the conventional wastage approach, in which solids are wasted from the clarifier underflow, X_{Mw} is related to X_M through the RAS flow rate, F_r. Substituting the appropriate relationship into Eq. 10.62 gives an expression equivalent to Eq. 5.70:

$$\Theta_c \approx \frac{V}{F_w}\left(\frac{F_w + F_r}{F + F_r}\right) \tag{10.64}$$

Since only bioreactor volume and process flow rates are included in Eqs. 10.63 and 10.64, suspended solids concentrations need not be measured to allow the SRT to be calculated on an approximate basis. The term hydraulic control of the SRT is based on the fact that only flow rates must be set to control the SRT. Thus, for the Garrett wastage approach:

$$F_w \approx \frac{V}{\Theta_c} \tag{10.65}$$

and for the conventional wastage approach:

$$F_w \approx \frac{V \cdot F_r}{(F + F_r)\Theta_c - V} \tag{10.66}$$

which is equivalent to Eq. 5.71, except for the approximately equal to sign.

This approach offers significant potential for simplifying activated sludge operation and the concepts have been applied to systems with uniform MLSS concentrations.[57] Use is made of a calibration chart to account for suspended solids in the process effluent and consideration is given to the constraints imposed by the thickening limitations of the clarifiers. The concepts can also be used with activated sludge systems with nonuniform MLSS concentrations.

10.4.2 Qualitative Observations

Qualitative observations provide valuable information on the actual operating conditions in activated sludge systems, thereby helping operators to make decisions about alterations in process control parameters.[80,84,85] They are a necessary adjunct to long-term SRT control and application of the process fundamentals described above.

Bioreactor. Important information can be gained from observing the color and appearance of the biomass in an activated sludge bioreactor. The mixed liquor in a well operating bioreactor will be brown in color, and a small amount of fresh, crisp, white foam will be present on the liquid surface. An earthy, musty aroma will be present which represents normal biological degradation products. A black color, or the presence of the "rotten egg" odor of hydrogen sulfide, indicates that inadequate aeration is being provided. The presence of a voluminous, billowing white foam indicates that inadequate treatment is occurring. The foam is a result of incomplete degradation of surfactants contained in the influent wastewater and/or the production of surface active agents by rapidly growing bacteria. Such a condition indicates either that the SRT is too short or that a toxicant has entered the plant and inactivated a portion of the biomass. This condition can be remedied by increasing the SRT and/or by locating and eliminating the source of the toxicant.

Under some conditions, a thick, viscous brown foam may accumulate on the surface of the bioreactor. This is typically caused by the growth of actinomycetes such as *Nocardia*, although other filamentous bacteria such as *Microthrix parvicella* can also cause it.[39] In extreme cases, the foam can accumulate sufficiently to overflow the bioreactor, resulting in unsightly and unsafe conditions. Accumulations within the clarifier can result in carryover of solids to the effluent, degrading it. The factors that cause such foams are complex and are not yet fully understood. The causative organisms are relatively slow growing and can, in some instances, be controlled by lowering the SRT. Their growth and accumulation in activated sludge systems are facilitated by physical designs that trap accumulated foam in the bioreactor. Consequently, good design practice requires foam to flow easily out of the bioreactor to the clarifier, where it is collected and removed from the system.[39] Once collected, the foam should be directed to the solids handling system in a way that minimizes recycle; direction to the plant headworks or to a gravity thickener from which the

overflow is recycled should be avoided. Some experience suggests that selectors can provide control over the growth of the causative bacteria. Another approach is a classifying selector, where aeration is used to produce foam, which will be selectively enriched with the causative bacteria.[13,49] The foam is then removed, resulting in their selective removal. If all else fails, then a high concentration chlorine spray can be applied directly to the foam on the bioreactor.[1] A fine mist is used so that the chlorine will contact the foam without entering the mixed liquor itself. The chlorine oxidizes the foam and the microorganisms in it, thereby destroying them. Since the foam-causing bacteria will be enriched in the foam, this practice selectively removes them from the system.

Clarifier. Visual observations of the clarifier indicate its operational status and also provide clues about the settling characteristics of the activated sludge. Denitrification in the clarifier solids blanket causes large clumps of solids to rise to the liquid surface, break, and spread over it. This phenomena, known as clumping, is caused when nitrate-N is present, DO is depleted in the solids blanket, and the residence time of the solids in the blanket is long enough to allow the generation of sufficient nitrogen gas bubbles to cause the solids to float. It can be controlled by altering the SRT of the activated sludge and/or by increasing the RAS flow rate to reduce the size and residence time of the solids blanket. The SRT may be decreased to reduce the degree of nitrification, or it may be increased to reduce the respiration rate of the activated sludge, thereby reducing the production rate of nitrogen gas. Alternatively, an anoxic zone may be incorporated into the bioreactor to remove nitrate-N, thereby reducing its concentration in the clarifier. The design and operation of activated sludge systems with anoxic zones are discussed in Chapter 11.

Two problems are often associated with long SRTs and/or excessive DO concentrations. They are ashing and pin floc. Both result from an inadequate filament backbone in the activated sludge floc. Growth at long SRTs and high DO concentrations prevents the growth of filamentous microorganisms, which are necessary to produce a strong floc that is resistant to mechanical turbulence. Ashing is the term used when small dark brown to gray particles rise in the clarifier. Pin floc refers to dense, granular solids that flocculate poorly and result in particles that settle rapidly but leave behind a turbid supernatant. Such a situation can be alleviated by reducing the SRT and/or the DO concentration to encourage the growth of a moderate amount of filamentous bacteria.

Straggler floc and dispersed growth are often observed at short SRTs. Straggler floc are relatively large floc particles (0.25 to 0.5 mm) that appear to be light, fluffy, and almost buoyant within the clarifier. They are caused by excessive quantities of filamentous microorganisms, which expand the activated sludge floc, resulting in slowly settling floc particles that can be carried out of the clarifier by hydraulic currents. As discussed in Section 10.2.1, dispersed growth is caused by inadequate flocculation and can often be corrected by increasing the SRT.

Observation of the clarifier solids blanket thickness, i.e., the vertical distance between the bottom of the clarifier and the top of the blanket, can provide an early indication of settleability problems. However, care must be exercised to properly interpret such observations because a blanket that is increasing in thickness may simply indicate improper system operating conditions. For example, an inadequate RAS flow rate will cause the blanket to increase in thickness as more solids are applied to it than can be removed by the RAS. An unintended increase in the acti-

vated sludge SRT can also cause the blanket to rise by increasing the MLSS concentration beyond that which can be handled by the clarifier. On the other hand, if an increase in the solids blanket thickness is coupled with an increase in the SVI, then a deterioration in solids settling and compaction characteristics is probably the cause. The term "bulking" should only be applied to that situation and not to the accumulation of solids within the clarifier. Bulking is generally caused by the growth of excessive quantities of filamentous bacteria, and its causes and cures are discussed in Section 10.2.1. If the corrective measures discussed there cannot be successfully applied or if an immediate remedy is required to prevent permit violations, then either oxidants such as chlorine or hydrogen peroxide may be applied to the activated sludge to reduce the population of filamentous bacteria, or polymers or other coagulants may be added to improve sludge settleability. The addition of oxidants to control the relative populations of filamentous and floc-forming bacteria is discussed in Section 10.4.3.

During Sludge Volume Index Measurement. Changes in activated sludge settling characteristics, referred to as "sludge quality" in operations manuals,[80,84] can often be observed in the vessel used to measure the SVI (settleometer) before they have an impact on the clarifier solids blanket thickness. Phenomena such as ashing, pin floc, straggler floc, and dispersed floc can be observed more precisely in the settleometer than in the clarifier. Denitrification will also often be observed first in the settleometer because its small hydraulic head allows the development of sufficient gas to float the solids before it occurs in the full-scale clarifier.[35] An increase in the SVI value indicates an increase in the population of filamentous bacteria.

Some operators measure the settling velocity of the activated sludge in the settleometer during the SVI test and use it as a control parameter.[80,84] They then adjust the DO concentration and the SRT in the activated sludge system (sometimes called "oxidation pressure" by operators) until the relative populations of low DO filaments and floc-forming bacteria are balanced to attain the desired sludge settling velocity. The settling characteristics of the sludge can be adjusted relatively quickly using this technique. Longer times are required, however, to recover acceptable settleability for sludge with a high SVI because of the need to waste a large inventory of filamentous bacteria from the system. This emphasizes the need to carefully monitor solids settling characteristics and maintain them within acceptable ranges.

Microscopic Examination. Regular microscopic examination of an activated sludge can provide insight into the factors affecting system performance and greatly assist with process optimization.[81] One type of microscopic examination uses a relatively inexpensive microscope to determine the types of Eucarya present and to characterize the overall structure of the activated sludge floc. Such an examination can supplement observations of the bioreactor, the clarifier, and the characteristics of the sludge in the settleometer. The types of Eucarya present tend to correlate with the operating SRT and can provide rapid, visual confirmation that the proper value is being achieved. A rapid change in the types of Eucarya present can be an indication of an inadequate DO concentration or the presence of a toxicant. A second type of microscopic examination is used to characterize the types of filamentous bacteria present using the technique of Eikelboom[26,27] as modified by Jenkins et al.[30] This examination requires a research grade microscope and a high degree of specialized training. The information obtained can be used to identify corrective actions required

to alleviate filamentous bulking problems, as discussed in Section 10.2.1 and indicated in Table 10.5. Facilities that do not generally experience filamentous sludge bulking problems need not conduct this type of examination on a regular basis.

10.4.3 Activated Sludge Oxidation to Control Settleability

The application of oxidizing agents to activated sludge can be used to control the growth of filamentous bacteria. Because such chemicals oxidize filamentous bacteria faster than floc-forming ones, they reduce the relative population size of the filaments in the activated sludge, thereby influencing its settling characteristics.[39] Due to chlorine's low cost and ready availability, it is the oxidant used most often for this purpose, although others, such as hydrogen peroxide, can be used with equal effect. Three factors are important in the use of chemical oxidation to control activated sludge settling characteristics: (1) proper control of the oxidant dose, (2) selection of an appropriate dose point, and (3) mixing at the dose point.

Because the purpose of oxidant addition is to destroy part of the activated sludge, and because the oxidant should be added continuously, the dosing rate should be expressed as the mass of oxidant added per day per mass of activated sludge in the system. Typical units are g of oxidant/(kg MLVSS·day), with MLVSS being used to represent the activated sludge because only the organic fraction of the sludge will react with the oxidant. The dose required will depend on the severity of the filamentous bulking and the speed with which it is desirable to reduce the SVI, with dosing rates typically ranging from a low of about 2 g Cl_2/(kg MLVSS·day) to a high of about 10. Prior to the initiation of chlorination, a target SVI is selected that will give the desired overall system performance, and a target SVI range is also chosen, often ± 20 mL/g. The dosing rate is then selected, chlorination is initiated, and the response of the system is monitored in terms of changes in the SVI and the abundance of filaments, as revealed by periodic microscopic examination. The dose is adjusted in accordance with the response until the SVI is within the target range. The dose is reduced if the SVI is within the target range but decreasing, while it is increased if the SVI is increasing. Chlorination is terminated when the SVI falls below the target range. If time is available to reduce the SVI slowly, then chlorination can begin at a relatively low dose of about 2 g Cl_2/(kg MLVSS·day) and be slowly increased until a downward trend in the SVI is established. If the SVI must be reduced rapidly, then a more aggressive dose, on the order of 6 g Cl_2/(kg MLVSS·day), should be selected, with the dose being reduced when a downward SVI trend is established.

The dose point should be selected to avoid contact with the influent wastewater while achieving a desired dosing frequency. Convenience and adequate mixing should also be considered. Direct contact with the influent wastewater must be avoided because the oxidant will react with any organic matter present at the dose point, thereby reducing its effectiveness against the activated sludge. Dosing frequency is important because biomass circulates through activated sludge systems and the entire inventory must be exposed to the oxidant about three times per day. If the dosing frequency is less than this, significant filament growth can occur in the fraction of the biomass that is not being dosed. For some systems, such as CAS, CMAS, CSAS, and SFAS, the RAS stream is an excellent location to add the oxidant. Little

influent organic matter is present in the RAS stream and the bioreactor HRT is short enough that biomass will be circulated through the clarifier several times per day, thereby ensuring the needed dosing frequency. For other processes, such as EAAS, the bioreactor HRT is much longer and biomass circulates to the clarifier less than once per day. In such cases, the dosing frequency will be insufficient and the oxidant may not be fully effective in controlling filament growth. In this case, the oxidant should be added directly to the bioreactor to obtain the needed dosing frequency.

Mixing at the dose point is important because the reaction between oxidant and biomass is very fast. Thus, rapid mixing must be provided to avoid over-oxidation of a portion of the biomass and under-oxidation of the remainder. A diffuser is generally required to ensure good contact between the added oxidant and the process flow. In addition, mechanical or other mixing may also be needed to ensure adequate dispersion of the oxidant. When the oxidant is added directly to the bioreactor, adequate mixing is often provided by the aeration equipment itself. The reader is referred to the book by Jenkins et al.[39] for further discussion of the use of oxidants to control activated sludge settling characteristics.

10.4.4 Dynamic Process Control

Regulation of the SRT provides long-term control of an activated sludge system. Furthermore, visual observations provide feedback on the success of the SRT control strategy and allow fine-tuning of certain operating parameters, such as the DO concentration. While this approach will permit operational adjustments in response to seasonal variations in process operating conditions and long-term changes in process loadings, most activated sludge systems are also subject to short-term loading variations. Loading variations occur on a diurnal and a day-to-day basis, and process operation must be adjusted accordingly. Facilities may also be subject to shock organic loadings due to industrial discharges and shock hydraulic loadings due to the inflow and infiltration of precipitation into the wastewater collection system. Process operation must be adjusted in response to these variations as well. Activated sludge systems possess some ability to respond to short-term loading variations, particularly if they occur regularly and over a period of several hours. However, this capacity is limited by the capability of bacteria to rapidly adjust their enzyme levels and by the time required to grow significant quantities of additional biomass. Thus, some deterioration in performance should be expected in response to dynamic loadings. Dynamic simulation, as illustrated in Chapters 6 and 7, is one technique that can be used to define the deterioration in effluent quality that will result from variations in process loadings. It can be used to determine the treatment limits for a particular process configuration and the benefits to be gained from the use of equalization to dampen load variations. Dynamic simulation can also be used to identify alternative process design and operating conditions that will improve the dynamic response of the system.

Several factors must be considered to ensure that an activated sludge system has adequate dynamic response capability. One is selection of an appropriate SRT.[32,64] The procedure used to select the design SRT for a nitrifying activated sludge system illustrates the principle; a safety factor for the design is selected with consideration of process loading variations. Equalization can be applied to smooth the organic and hydraulic loading variations. It will generally be a necessity for industrial waste-

waters in which significant short-term organic loading variations occur as a result of the operation of the production facility that generates the wastewater. On the other hand, experience indicates that the activated sludge process is generally quite resilient and can accept the dry weather variations in organic and hydraulic loadings typical of domestic wastewaters, while still providing acceptable performance. Even in that situation, however, flow equalization can improve the performance of the entire treatment train. Improved response to organic and hydraulic variations can also be achieved by providing alternative activated sludge process operating modes. For example, step feeding capability can be provided to allow peak hydraulic loadings to be processed without causing clarifier thickening failure.

While not yet in routine use, much research is being directed at the development and application of automatic control technology.[3,52] Process instrumentation is available to monitor suspended solids concentrations, DO concentrations, oxygen uptake rates, and residual nutrient concentrations (ammonia, nitrate, and phosphate). The data collected by such sensors, supplemented with laboratory and flow rate data, can be interfaced with dynamic process models to allow investigation of alternative operating strategies. Such approaches have been used off-line to analyze the performance of existing activated sludge systems,[20] and may become available for on-line monitoring and process optimization. Expert systems are also being developed to assist with activated sludge process operation. All in all, these developments offer significant potential to optimize activated sludge operation, thereby improving performance and reliability, and reducing costs.

10.5 Key Points

1. Four factors are common to all activated sludge systems: (1) biomass grows as a flocculent slurry by oxidizing organic matter under aerobic conditions; (2) quiescent sedimentation removes the biomass, producing an effluent low in suspended solids; (3) settled biomass is recycled as a concentrated slurry back to the bioreactor; and (4) excess biomass is wasted to control the solids retention time (SRT).
2. Activated sludge bioreactors are typically open basins containing mechanical equipment to transfer oxygen and maintain the flocculent biomass in suspension. Several devices are used to do this, including coarse and fine bubble diffused air, mechanical surface aerators, jet aerators, and submerged turbines.
3. The clarifiers used in activated sludge systems provide two functions: (1) separation of the flocculent biomass to produce a clarified effluent (clarification), and (2) concentration of the biomass for recycle to the upstream bioreactor (thickening).
4. Eight major activated sludge process options exist: (1) conventional (CAS), (2) step feed (SFAS), (3) contact stabilization (CSAS), (4) completely mixed (CMAS), (5) extended aeration (EAAS), (6) high purity oxygen (HPOAS), (7) selector (SAS), and (8) sequencing batch reactor (SBRAS).
5. CAS and CMAS contain uniform mixed liquor suspended solids (MLSS) concentrations and typically have hydraulic retention times (HRTs) rang-

ing from 4 to 8 hours and SRTs ranging from 3 to 15 days. CAS uses a plug–flow bioreactor, while CMAS uses a single, completely mixed bioreactor. Sludge settleability is usually better for CAS facilities due to better control of the growth of filamentous bacteria. CMAS offers greater resistance to upset by inhibitory organic chemicals.

6. The return activated sludge (RAS) is sequentially diluted as it flows through SFAS and CSAS systems, resulting in an outlet MLSS concentration that is lower than the average concentration in the bioreactor. As a consequence, SFAS and CSAS bioreactors can be smaller than CAS or CMAS bioreactors with the same SRT. Effluent quality may be somewhat poorer from SFAS and CSAS systems, however.
7. EAAS systems use long SRTs (20 to 30 days) to partially stabilize the biomass produced. This requires large bioreactors (HRTs of 24 hours or more), but results in excellent process stability and the production of a high-quality effluent.
8. In HPOAS systems, biomass and oxygen enriched gas flow cocurrently through a staged, covered bioreactor. This allows use of high volumetric organic loading rates and short HRTs (generally 2 to 4 hours). Short SRTs are also generally used (1 to 5 days).
9. The SAS system uses a highly loaded section at the inlet end of the bioreactor (the selector) to create conditions favorable to the growth of floc-forming bacteria relative to filamentous bacteria. Selectors can be incorporated into the other activated sludge options.
10. SBRAS incorporates biological reaction and sedimentation into a single vessel. Microprocessors are used to automatically control the influent flow, aeration, mixing, and effluent decanting functions. They are used most often in smaller wastewater treatment plants.
11. Sludge settleability is typically quantified by the sludge volume index (SVI), which is the volume (in mL) occupied by one gram of settled solids. A value greater than 150 mL/g represents a poor settling, bulking sludge. Observation of settling rate and the clarity of the supernatant produced in the SVI test can provide valuable insights into the characteristics of the activated sludge.
12. Bioflocculation forms the microstructure of activated sludge floc by aggregating individual bacteria into dense, settleable particles. It occurs as a result of exocellular polymers produced by the biomass. A minimum SRT (typically 1 day for domestic wastewaters and 3 days for industrial wastewaters) is required to obtain adequate bioflocculation.
13. Filamentous bacteria provide the macrostructure for activated sludge floc, thereby forming the "backbone" that allows floc to resist mechanical shear. A proper balance between floc-forming and filamentous bacteria results in strong, compact floc that resists shear in the bioreactor, while settling quickly and compacting well in the clarifier.
14. Excessive quantities of filamentous bacteria cause activated sludge bulking. About 30 types of filamentous bacteria can be found in activated sludge systems and the type present provides valuable clues about the situation that needs to be corrected to eliminate the bulking problem.

15. The SRT required for many activated sludge systems is determined by the need for bioflocculation. In other cases, longer SRTs are needed to nitrify, to treat certain industrial wastes containing less biodegradable organic matter, and to stabilize the biomass produced. The reliability and capacity of the solids processing system must also be considered in the selection of the SRT.
16. Two factors limit the bioreactor MLSS concentration. One is solids thickening, which limits the maximum economical MLSS concentration to about 5,000 mg/L as total suspended solids (TSS). The other is bioflocculation, which typically requires a minimum MLSS concentration of 500 to 1,000 mg/L as TSS.
17. The primary effect of the dissolved oxygen (DO) concentration is on the growth of filamentous bacteria, although it will also affect the occurrence of nitrification. A DO concentration of 2 mg/L is a reasonable benchmark, but in some situations successful treatment can be obtained with lower values, whereas in others, higher values will be required.
18. Use of the oxygen transfer equipment both to transfer oxygen and to maintain solids in suspension places constraints on the size of the bioreactor. If it is too small, then the volumetric power input required to transfer the needed oxygen will cause floc shear. If it is too large, then mixing will control the volumetric power input, resulting in increased power requirements.
19. Adequate nutrients are required to allow balanced growth of biomass in activated sludge systems. Nutrient limitations can result in the growth of undesirable quantities of filamentous bacteria and/or the production of exocellular slime. Both interfere with activated sludge settling and compaction.
20. The temperature of an activated sludge system must be maintained in either the mesophilic (35° to 40°C) or the thermophilic (45° to 60°C) range; it should not fluctuate between the two. The oxidation of organic matter and ammonia-N results in the liberation of heat, whereas the physical configuration of the bioreactor and the nature of the oxygen transfer device influence the loss of heat.
21. The design of an activated sludge process generally consists of the following six steps:
    a. Select the activated sludge option and SRT based on wastewater characteristics, effluent quality goals, facility capital and operating costs, and operational objectives.
    b. Calculate the mass of MLSS in the system, $(X_{M,T} \cdot V)_{System}$, the quantity of waste sludge, $W_{M,T}$, and the oxygen requirement, RO, by using the modified stoichiometric model of Chapter 5.
    c. Distribute the oxygen requirement as required by the system configuration.
    d. Calculate the upper and lower limits on the bioreactor volume based on mixing, floc shear, and oxygen transfer.
    e. Using $(X_{M,T} \cdot V)_{System}$ and the output from Step d, calculate the upper and lower limits on the MLSS concentration and choose a MLSS concentration within those limits based on consideration of final set-

tler design. Calculate the bioreactor volume associated with the chosen MLSS concentration.

f. Optimize the system using tools such as activated sludge model (ASM) No. 1.

22. The mass of MLSS and the quantity of waste sludge for any activated sludge system can be calculated once the SRT is established. They are calculated for cold weather conditions because they produce the largest values, thereby determining the required size of the bioreactor.
23. The system oxygen requirement can also be calculated once the SRT is fixed. It will be greatest at the highest temperature, and thus summer conditions are used in design. This requirement must be apportioned among the different vessels in a multitank system. Transient requirements must also be considered when sizing the oxygen transfer system.
24. The size of the bioreactor is determined from $(X_{M,T} \cdot V)_{System}$ based on the design MLSS concentration. The choices of the MLSS concentration and the bioreactor volume must be consistent with the constraints given in Key Points 16 and 18.
25. Aerobic selectors prevent filamentous sludge bulking by allowing removal of the readily biodegradable organic matter in an environment with a sufficiently high specific growth rate for kinetic selection of floc-forming bacteria. To ensure that this occurs under a variety of loading conditions, selectors are staged and are sized to give a desired average process loading factor.
26. The RAS flow rate and the distribution of the influent wastewater must be considered when sizing the bioreactor for systems with nonuniform MLSS concentrations, such as SFAS and CSAS, because both affect the distribution of MLSS within the system.
27. In the absence of simulation, the spatial distribution of the oxygen requirement in multitank systems can be approximated by partitioning it into its component parts and assigning each to the appropriate portion of the bioreactor system. Convenient divisions are the oxidation of readily biodegradable substrate, the hydrolysis and subsequent oxidation of slowly biodegradable substrate, nitrification, and decay of heterotrophs.
28. Because of the analogy between continuous flow and batch activated sludge systems, SBRAS systems can be designed with the same basic procedures as the other activated sludge options. The primary difference is that the smallest allowable bioreactor size is governed by its role as the final settler.
29. Dynamic models such as ASM No. 1 can be used to refine designs developed using the procedures described above. They allow estimation of the impact of short-term loading variations on effluent quality, and provide better estimates of the spatial and temporal variations in oxygen requirements.
30. Three procedures are routinely used to determine the waste activated sludge (WAS) flow rate required to achieve a given SRT: MLSS analysis, centrifuge analysis, and hydraulic control. The WAS flow rate should not be changed by more than 20% each day, but solids wasting should be

frequent enough so that MLSS concentrations do not change by more than 10%.

31. Observations of the color and appearance of the biomass in an activated sludge bioreactor provide information about the system operating conditions.
32. Phenomena such as denitrification in the final clarifier and an imbalance between floc-forming and filamentous bacteria can be detected by visual observations of the clarifier.
33. Microscopic examination of the activated sludge biomass should be performed routinely. Observation of the Eucarya present provides visual confirmation of the SRT value. Rapid changes in the Eucarya present indicate an inadequate DO concentration or the presence of toxic materials. Identification of the types of filamentous bacteria present can be used to determine the conditions causing excessive filament growth and the associated sludge bulking problems.
34. Oxidants such as chlorine and hydrogen peroxide can be used to oxidize excessive quantities of filamentous bacteria and control biomass settling characteristics. Oxidant doses are expressed as g oxidant/(kg MLVSS·day). The dose point should be selected to avoid contact with the influent wastewater while achieving a desired dosing frequency; convenience and adequate mixing should also be considered.
35. The dynamic response of an activated sludge system is constrained by the mass of biomass present in the system and by the ability of the microorganisms to synthesize additional enzymes. Longer SRTs generally provide greater capability to metabolize added organic matter. In some cases, equalization is necessary to limit the variations in activated sludge process loadings.

10.6 STUDY QUESTIONS

1. Prepare a table comparing the following factors for the eight major types of activated sludge: SRT, HRT, MLSS concentration, recycle ratio, bioreactor configuration, and design approach.
2. List the main advantages and disadvantages of an activated sludge system in comparison to anaerobic wastewater treatment processes. When would an activated sludge process be used in comparison to an anaerobic process?
3. List the benefits and drawbacks of each of the eight major activated sludge variations. When is each typically applied?
4. Explain why the SRT is generally maintained between 3 and 15 days to obtain an activated sludge that settles well.
5. Three benefits have been attributed to HPOAS: (1) an increased rate of treatment; (2) increased density of biomass with an associated increased settling rate; and (3) reduced rate of excess biomass production. Discuss and evaluate these claims.

6. Discuss the role of both floc-forming and filamentous bacteria in the formation of a good settling activated sludge, explaining why an optimum balance exists.
7. Explain the kinetic selection mechanism as applied to SAS. Describe how the selector should be configured to take advantage of this mechanism, and explain why.
8. Discuss the types of filamentous microorganisms that often occur in activated sludge systems and relate the filament types to the environmental conditions favoring them.
9. Discuss the impacts of DO concentration on the performance of activated sludge systems. Identify all potential effects and their typical importance.
10. Discuss the relationship between filament growth and activated sludge floc structure.
11. Describe and contrast the various indices used to characterize activated sludge settleability.
12. Discuss the benefits of plug–flow conditions within an activated sludge system. Describe how the process loading factor for the initial contact zone should be selected to optimize sludge settleability.
13. Why should the MLSS concentration in an activated sludge system normally lie between 500 and 5,000 mg/L as TSS? What factors affect the choice of the value?
14. Why must both an upper and a lower limit be placed on the mixing energy supplied per unit volume to an activated sludge bioreactor? List appropriate limit values for diffused air and mechanical surface aeration systems.
15. Why are heat losses from diffused air activated sludge systems less than those from systems using mechanical surface aerators? Does the HRT affect heat loss, and if so how?
16. Using the wastewater characteristics in Table E8.4, the stoichiometric and kinetic parameters in Table E8.5, and the temperature correction factors in Table E10.1, design a CMAS system to treat an average wastewater flow rate of 30,000 m^3/day using an SRT of 7 days. Assume a constant loading. Use mechanical surface aeration and justify all assumptions and decisions. The lowest sustained winter temperature is 13°C and the highest sustained summer temperature is 24°C.
17. The diurnal peak loading for the CMAS system considered in Study Question 16 is twice the average loading. What is the peak oxygen requirement during both low and high temperature operating conditions? Does consideration of the peak oxygen requirement affect the selection of the bioreactor volume? What bioreactor volume and MLSS concentration would you recommend under this condition? Why?
18. Design a CAS system for the situation considered in Study Question 16. Use a diffused air oxygen transfer system and assume that the in-process oxygen transfer efficiency is 12%. Also assume that the hydraulic characteristics of the CAS bioreactor are equivalent to three tanks in series.
19. Use a computer code implementing ASM No. 1 or a similar model to evaluate the steady-state distribution of oxygen requirements in the CAS system sized in Study Question 18.

20. Assume that the influent flow and pollutant concentrations for the problem considered in Study Question 18 vary as indicated in Figure 6.2. Use a computer code implementing ASM No. 1 or a similar model to determine the effect of these diurnal variations on the system effluent quality and the oxygen requirement in each equivalent tank of the CAS system designed in Study Question 18. Do this for the summer conditions only. How does the flow-weighted average effluent quality compare with the steady-state value obtained in Study Question 18? Would it be possible to deliver oxygen rapidly enough to meet the peak oxygen requirement in the first tank of the system sized in Study Question 18? Why? If the system is incapable of meeting the peak oxygen requirement, what modifications to the design would be required? What other impacts would such changes have?
21. Repeat Study Question 16 for an SAS system. Assume that the main bioreactor is a single completely mixed vessel, and determine the required sizes for all tanks in the system. Assume that diffused aeration is to be used with a 12% oxygen transfer efficiency.
22. An EAAS system is to be designed to treat the wastewater described in Study Question 16. Select and justify the SRT that will be used for the design. Determine the mass of MLSS in the system, the waste solids quantity, and the oxygen requirement under both summer and winter conditions. Assume that the bioreactor will be configured as an oxidation ditch with vertical mechanical surface aerators. The in-process oxygen transfer capacity of the aerators is 1.2 kg O_2/(kW·hr), and the minimum volumetric power input required to maintain solids in suspension is 7 kW/1000 m^3. Determine the allowable range in bioreactor volumes.
23. A four pass SFAS system with an SRT of 7 days is to be used to treat the wastewater considered in Study Question 16. The total bioreactor volume is 5,625 m^3 and the RAS flow rate is 15,000 m^3/day. Influent wastewater is to be distributed uniformly to each of the four passes. Determine the MLSS concentration in each of the four passes, as well as the oxygen requirement. Use a diffused air oxygen transfer system and assume that the in-process oxygen transfer efficiency is 12 percent. Is the bioreactor volume acceptable as far as the constraints on mixing energy and oxygen transfer are concerned? Why? How does the winter effluent soluble substrate concentration compare to that from the CMAS system in Study Question 16?
24. For the problem considered in Study Question 16, design a CSAS system to produce an effluent with a concentration of readily biodegradable substrate of less than 5 mg/L as COD, while maintaining an SRT of 7 days. For this design assume that the contact tank MLSS concentration will be 2,500 mg/L as TSS and that the stabilization basin MLSS concentration will be 8,000 mg/L as TSS. Use a diffused air oxygen transfer system and assume that the in-process oxygen transfer efficiency is 12%. Size the contact and stabilization basins and check the oxygen transfer, mixing, and floc shear requirements.
25. Use a computer code implementing ASM No. 1 or a similar model to determine the distribution of the steady-state oxygen requirement between

the contact and stabilization basins for Study Question 24 and compare the requirements to the estimates made in that study question.

26. An SBRAS process is to be designed to treat the wastewater defined in Study Question 16. An effective SRT of 15 days is to be used, and the SVI is expected to be less than or equal to 120 mL/g. Develop a plot that demonstrates the effect on the total SBR volume of the number of operating cycles per day and the fraction of the total bioreactor cycle devoted to fill plus react.
27. Discuss the factors that must be considered when selecting a wasting frequency for an activated sludge system.
28. Define each of the following terms and tell what each indicates about the operation of an activated sludge system: clumping, ashing, pin floc, straggler floc, and dispersed growth.
29. Discuss the factors that must be considered during selection of the dosing rate and dose point for application of chlorine to control filamentous sludge bulking.
30. Why is SRT control considered to be a long-term operational strategy?

REFERENCES

1. Albertson, O. and P. Hendricks, Bulking and foaming organisms control at Phoenix, AZ WWTP. *Water Science and Technology* **26**(3/4):461–472, 1992.
2. Amalan, S., Analysis of factors affecting peaking phenomena in activated sludge oxygen requirements due to diurnal load variations. M.S. Thesis, Clemson University, Clemson, South Carolina, 1992.
3. Andrews, J. F., *Dynamics and Control of the Activated Sludge Process*, Technomic Publishing, Lancaster, Pennsylvania, 1992.
4. Arden, E. and W. T. Lockett, Experiments on the oxidation of sewage without the aid of filters. *Journal of the Society of Chemical Industries* **33**:523, 1914.
5. Argaman, Y. and C. F. Adams, Jr., Comprehensive temperature model for aerated biological systems. *Progress in Water Technology* **9**(1/2):397–409, 1977.
6. Arora, M. L., E. F. Barth, and M. B. Umphres, Technology evaluation of sequencing batch reactors. *Journal, Water Pollution Control Federation* **57**:867–875, 1985.
7. Baillod, C. R., Oxygen utilization in activated sludge plants: Simulation and model calibration. EPA/600/S2-88/065, U. S. Environmental Protection Agency, 1989.
8. Bisogni, J. J. and A. W. Lawrence, Relationships between biological solids retention time and settling characteristics of activated sludge. *Water Research* **5**:753–763, 1971.
9. Blackbeard, J. R., G. A. Ekama, and G. v. R. Marais, A survey of bulking and foaming in activated sludge plants in South Africa. *Water Pollution Control* **85**:90–100, 1986.
10. Bossier, P. and W. Verstraete, Triggers for microbial aggregation in activated sludge. *Applied Microbiology and Biotechnology* **45**:1–6, 1996.
11. Boyle, J. D., Biological treatment process in cold weather. *Water and Sewage Works Reference Number* **123**:R28–R50, 1976.
12. Campbell, H. J. and R. F. Rocheleau, Waste treatment at a complex plastics manufacturing plant. *Journal, Water Pollution Control Federation* **48**:256–273, 1976.
13. Cha, D. K., D. Jenkins, W. P. Lewis, and W. H. Kido, Process control factors influencing *Nocardia* populations in activated sludge. *Water Environment Research* **64**:37–43, 1992.
14. Chao, A. C. and T. M. Keinath, Influence of process loading intensity on sludge clarification and thickening characteristics. *Water Research* **13**:1213–1223, 1979.

15. Chapman, T. D., L. C. Match, and E. H. Zander, Effect of high dissolved oxygen concentration in activated sludge systems. *Journal, Water Pollution Control Federation* **48**: 2486–2510, 1976.
16. Chudoba, J., P. Grau, and V. Ottova, Control of activated sludge filamentous bulking. II. Selection of microorganisms by means of a selector. *Water Research* **7**:1389–1406, 1973.
17. Costerton, J. W., T. R. Irvin, and K.-J. Cheng, The bacterial glycocalyx in nature and disease. *Annual Review of Microbiology* **32**:299–324, 1981.
18. Curds, C. R., The flocculation of suspended matter by *Paramecium caudatum*. *Journal of General Microbiology* **33**:357–363, 1963.
19. Daigger, G. T., Development of refined clarifier operating diagrams using an updated settling characteristics database. *Water Environment Research* **67**:95–110, 1995.
20. Daigger, G. T. and J. A. Buttz, *Upgrading Wastewater Treatment Plants*, Technomic Publishing, Lancaster, Pennsylvania, 1992.
21. Daigger, G. T. and C. P. L. Grady, Jr., A model for the biooxidation process based upon product formation concepts. *Water Research* **11**:1049–1057, 1977.
22. Daigger, G. T. and G. A. Nicholson, Performance of four full-scale nitrifying wastewater treatment plants incorporating selectors. *Research Journal, Water Pollution Control Federation* **62**:676–683, 1990.
23. Daigger, G. T., M. H. Robbins, Jr., and B. R. Marshall, The design of a selector to control low F/M filamentous bulking. *Journal, Water Pollution Control Federation* **57**: 220–226, 1985.
24. Das, D., T. M. Keinath, D. S. Parker, and E. J. Wahlberg, Floc breakup in activated sludge plants. *Water Environment Research* **65**:138–145, 1993.
25. Eckenfelder, W. W. and P. Grau, *Activated Sludge Process Design and Control: Theory and Practice*, Technomic Publishing, Lancaster, Pennsylvania, 1992.
26. Eikelboom, D. H., Filamentous organisms observed in bulking activated sludge. *Water Research* **9**:365–388, 1975.
27. Eikelboom, D. H., Identification of filamentous organisms in bulking activated sludge. *Progress in Water Technology* **8**(6):153–161, 1977.
28. Eikelboom, D. H. Biological characteristics of oxidation ditch sludge. *Oxidation Ditch Technology*, CEP Consultants Ltd, Edinburgh, Scotland, 1982.
29. Garrett, M. T., Jr., Hydraulic control of activated sludge growth rate. *Sewage and Industrial Wastes* **30**:253–261, 1958.
30. Garrett, M. T. and C. N. Sawyer, Kinetics of removal of soluble BOD by activated sludge. *Proceedings of the 7th Industrial Waste Conference, Purdue University Engineering Extension Series No. 79*, 51–77, 1952.
31. Gould, R. H., Operating experiences in New York City. *Sewage Works Journal* **14**:70–80, 1942.
32. Grady, C. P. L., Jr., A theoretical study of activated sludge transient response. *Proceedings of the 26th Industrial Waste Conference, Purdue University Engineering Extension Series No. 140*, 318–335, 1971.
33. Greenberg, A. E., L. S. Clesceri, and A. D. Eaton, *Standard Methods for the Examination of Water and Wastewater*, 18th ed., American Public Health Association, Washington, D.C., 1992.
34. Gujer, W. and D. Jenkins, A nitrification model for the contact stabilization activated sludge process. *Water Research* **9**:561–566, 1975.
35. Henze, M., R. Dupont, P. Grau, and A. De La Sota, Rising sludge in secondary settlers due to denitrification. *Water Research* **27**:231–236, 1993.
36. Higgins, M. J. and J. T. Novak, The effect of cations on the settling and dewatering of activated sludges: laboratory results. *Water Environment Research* **69**:215–224, 1997.

37. Irvine, R. L., L. H. Ketchum, R. Breyfogle, and E. F. Barth, Municipal application of sequencing batch treatment. *Journal, Water Pollution Control Federation* **55**:484–488, 1983.
38. Jenkins, D. and W. E. Garrison, Control of activated sludge by mean cell residence time. *Journal, Water Pollution Control Federation* **40**:1905–1919, 1968.
39. Jenkins, D., M. G. Richards, and G. T. Daigger, *The Causes and Cures of Activated Sludge Bulking and Foaming*, 2nd ed., Lewis Publishers, Ann Arbor, Michigan, 1993.
40. Jewell, W. J. and R. M Kabrick, Autoheated aerobic thermophilic digestion with aeration. *Journal, Water Pollution Control Federation* **52**:512–523, 1980.
41. Kessler, L. H., et al., Tapered aeration of activated sludges. *Municipal Sanitation* **7**:268, 1936.
42. Koopman, B. and K. Cadee, Prediction of thickening capacity using diluted sludge volume index. *Water Research* **17**:1427–1431, 1983.
43. McCarty, P. L., Thermodynamics of biological synthesis and growth. *Proceedings of the Second International Conference on Water Pollution Research*, Pergamon, New York, pp. 169–199, 1965.
44. McKinney, R. E., J. M. Symons, W. G. Shifrin, and M. Vezina, Design and operation of a complete mixing activated sludge system. *Sewage and Industrial Wastes* **30**:287–295, 1958.
45. McWhirter, J. R., *The Use of High-Purity Oxygen in the Activated Sludge Process*, Vols. I and II, CRC Press, West Palm Beach, Florida, 1978.
46. Metcalf & Eddy, Inc., *Wastewater Engineering: Treatment, Disposal, and Reuse*, 3rd ed., McGraw-Hill, New York, 1991.
47. Municipality of Metropolitan Seattle, *West Point Treatment Plant Secondary Treatment Facilities: High Purity Oxygen Design Test Facility, Final Report*, Seattle, Washington, 1989.
48. Murphy, K. L. and B. I. Boyko, Longitudinal mixing in spiral flow aeration tanks. *Journal of the Sanitary Engineering Division, ASCE* **96**:211–221, 1970.
49. Pagilla, K. R., D. Jenkins, and W. H. Kido, *Nocardia* control in activated sludge by classifying selectors. *Water Environment Research* **68**:235–239, 1996.
50. Palm, J. C., D. Jenkins, and D. S. Parker, Relationship between organic loading, dissolved oxygen concentration and sludge settleability in the completely-mixed activated sludge process. *Journal, Water Pollution Control Federation* **52**:2484–2506, 1980.
51. Parker, D. S. and Merrill, M. S., Oxygen and air activated sludge: Another view. *Journal, Water Pollution Control Federation* **48**:2511–2528, 1976.
52. Patry, G. G. and D. T. Chapman, *Dynamic Modelling and Expert Systems in Wastewater Engineering*, Lewis Publishers, Chelsea, Michigan, 1989.
53. Pavoni, J. L., M. W. Tenney, and W. F. Eckelberger, Jr., Bacterial exocellular polymers and biological flocculation. *Journal, Water Pollution Control Federation* **44**:414–431, 1972.
54. Randall, C. W., J. L. Barnard, and H. D. Stensel, *Design and Retrofit of Wastewater Treatment Plants for Biological Nutrient Removal*, Technomic Publishing, Lancaster, Pennsylvania, 1992.
55. Rittmann, B. E., W. Bae, E. Namkung, and C.-H. Lu, A critical evaluation of microbial product formation in biological processes. *Water Science and Technology* **19**(7):517–528, 1987.
56. Roper, R. E., Jr. and C. P. L. Grady, Jr., Activated sludge hydraulic control techniques evaluated by computer simulation. *Journal, Water Pollution Control Federation* **46**: 2565–2578, 1974.
57. Roper, R. E., Jr. and C. P. L. Grady, Jr., A simple effective technique for controlling solids retention time in activated sludge plants. *Journal, Water Pollution Control Federation* **50**:702–708, 1978.

58. Roper, R. E., Jr., R. O. Dickey, S. Marman, S. W. Kim, and R. W. Yandt, Design effluent quality. *Journal of the Environmental Engineering Division, ASCE* **105**:309–321, 1979.
59. Sanin, F. D. and P. A. Vesilind, Synthetic sludge: a physical/chemical model in understanding bioflocculation. *Water Environment Research* **68**:927–933, 1996.
60. Sedory, P. E. and M. K. Stenstrom, Dynamic prediction of wastewater aeration basin temperature. *Journal of Environmental Engineering* **121**:609–618, 1995.
61. Sezgin, M., D. Jenkins, and D. S. Parker, A unified theory of filamentous activated sludge bulking. *Journal, Water Pollution Control Federation* **50**:362–381, 1978.
62. Shao, Y. J., J. Crosse, E. Keller, and D. Jenkins, High rate air activated sludge operation at the City of Los Angles Hyperion Wastewater Treatment Plant. *Water Science and Technology* **25**(4/5):75–87, 1992.
63. Sheintuch, M., O. Lev, P. Einav, and E. Rubin, Role of exocellular polymer in the design of activated sludge. *Biotechnology and Bioengineering* **28**:1564–1576, 1986.
64. Sherrard, J. H. and A. W. Lawrence, Response of activated sludge to step increase in loading. *Journal, Water Pollution Control Federation* **47**:1848–1856, 1975.
65. Stall, T. R., and J. H. Sherrard, Evaluation of control parameters for the activated sludge process. *Journal, Water Pollution Control Federation* **50**:450–457, 1978.
66. Talati, S. N. and M. K. Stenstrom, Aeration basin heat loss. *Journal of Environmental Engineering* **116**:70–86, 1990.
67. Thompson, D. J., D. T. Chapman, and K. L. Murphy, Step feed control to minimize solids loss during storm flows. *Research Journal, Water Pollution Control Federation* **61**:1658–1665, 1989.
68. Toerber, E. D., W. L. Paulson, and H. S. Smith, Comparison of completely mixed and plug flow biological systems. *Journal, Water Pollution Control Federation* **46**:1995–2014, 1974.
69. Tomlinson, E. J., Bulking—A survey of activated sludge plants. Technical Report TR35, Water Research Centre, Stevenage, England, 1976.
70. Tomlinson, E. J. and B. Chambers, The effect of longitudinal mixing on the settleability of activated sludge. Technical Report TR 122, Water Research Centre, Stevenage, England, 1978.
71. U. S. Environmental Protection Agency, *Design Manual-Fine Pore Aeration Systems*, EPA-623/189023, U. S. Environmental Protection Agency, CERI, Risk Reduction Engineering Laboratory, Cincinnati, Ohio, 1989.
72. U. S. Environmental Protection Agency, *Process Design Manual for Nitrogen Control*, EPA/625/R-93/010, U. S. Environmental Protection Agency, Cincinnati, Ohio, 1993.
73. Urbain, V., J. C. Block, and J. Manem, Bioflocculation in activated sludge: An analytical approach. *Water Research* **27**:829–838, 1993.
74. Vaccari, D. A., T. Fagedes, and J. Longtin, Calculation of mean cell residence time for unsteady-state activated sludge systems. *Biotechnology and Bioengineering* **27**:695–703, 1985.
75. van Loosdrecht, M. C. M., J. Lyklema, W. Norde, and A. J. B. Zehnder, Bacteria adhesion: A physicochemical approach. *Microbial Ecology* **17**:1–15, 1989.
76. Wahlberg, E. J., T. M. Keinath, and D. S. Parker, Influence of activated sludge flocculation time on secondary clarification. *Water Environment Research* **66**:779–786, 1994.
77. Walker, L. F., Hydraulically controlling solids retention time in the activated sludge process. *Journal, Water Pollution Control Federation* **43**:30–39, 1971.
78. Wanner, J., *Activated Sludge Bulking and Foaming Control*, Technomic Publishing, Lancaster, Pennsylvania, 1994.
79. Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992.
80. Water Pollution Control Federation, *Activated Sludge*, Manual of Practice No. OM-9, Water Pollution Control Federation, Alexandria, Virginia, 1987.

81. Water Pollution Control Federation, *Activated Sludge Microbiology*, Water Pollution Control Federation, Alexandria, Virginia, 1989.
82. Water Pollution Control Federation, *Aeration in Wastewater Treatment*, Manual of Practice No. FD-13, Water Pollution Control Federation, Alexandria, Virginia, 1988.
83. Water Pollution Control Federation, *Clarifier Design*, Manual of Practice No. FD-8, Water Pollution Control Federation, Alexandria, Virginia, 1985.
84. Water Pollution Control Federation, *Operation of Municipal Wastewater Treatment Plants*, Manual of Practice No. 11, Water Environment Federation, Alexandria, Virginia, 1990.
85. West, A. W., Operational control procedures for the activated sludge process, Part I, Observations. Report No. EPA-330/9-74-001-a, U. S. Environmental Protection Agency, 1973.
86. Zita, A. and M. Hermansson, Effects of ionic strength on bacterial adhesion and stability of flocs in a wastewater activated sludge system. *Applied and Environmental Microbiology* **60**:3041–3048, 1994.

This page intentionally left blank

11
Biological Nutrient Removal

Biological nutrient removal (BNR) processes are modifications of the activated sludge process that incorporate anoxic and/or anaerobic zones to provide nitrogen and/or phosphorus removal. Many BNR variants have been developed, representing a wide range of nutrient removal capabilities. This chapter presents the basic design and operational principles for several of them. It builds upon the theoretical concepts presented in Chapters 2, 3, 6, and 7.

11.1 PROCESS DESCRIPTION

11.1.1 Historical Overview

The system components (aerobic and anoxic zones) that form the basis for biological nitrogen removal were developed in the 1960s, resulting in a number of approaches.[54,66,68,72,73] One used a series of separate suspended growth systems to accomplish removal of organic matter and nitrogen in a step-wise fashion. Organic matter was removed in the first step, nitrification was accomplished in the second, and denitrification was achieved in a third. These three-stage nitrogen removal systems were discussed extensively in the literature, but received little full-scale use due to high capital and operating costs.[57] Another approach, referred to as single-sludge nitrogen removal, incorporated both aerobic zones for nitrification and anoxic zones for denitrification in a single system, with carbon oxidation occurring in both zones. Among the concepts incorporated into the single-sludge approach were the recirculation of nitrate-N to an initial anoxic zone to allow use of the readily biodegradable substrate for denitrification, as discussed in Section 7.5, and the use of a second anoxic zone for additional denitrification using slowly biodegradable substrate and biomass decay, as discussed in Section 7.6. These concepts are now widely used in biological nitrogen removal.

Concurrently with development of nitrogen removal systems, enhanced phosphorus removal was observed in certain full-scale activated sludge systems.[54,57,73] These systems generally used plug–flow bioreactors with uniform aeration along their length, resulting in very low dissolved oxygen (DO) concentrations in the initial sections. We now know that these inadequately aerated sections provided the anaerobic zone necessary for selection of the phosphorus accumulating organisms (PAOs) required for biological phosphorus removal, as discussed in Sections 2.4.6 and 3.7. Nevertheless, controversy existed for more than a decade concerning the phosphorus removal mechanisms operating in these plants.[48] While some chemical phosphorus removal will occur in systems with anaerobic and anoxic zones,[2] it is now recognized

that biological mechanisms are responsible for most of the phosphorus removal. In spite of the controversy concerning the removal mechanisms, research during the 1960s resulted in the first commercial biological phosphorus removal (BPR) process, the Phostrip® process.[40,41]

With this background, the stage was set for the integration and refinement of the basic concepts into the single sludge biological nitrogen and phosphorus removal processes that we know today. The initial major strides were provided by Barnard,[9,10] who made two significant conceptual leaps forward. The first was the integration of aerobic and anoxic zones, along with nitrate recirculation, to create the effective and cost competitive single-sludge nitrogen removal system now known as the four-stage Bardenpho process. The second was the observation that biological phosphorus removal would occur in these systems if nitrate was sufficiently depleted in the initial anoxic zone. Comparing this observation with the operating conditions in the full-scale plants that achieved enhanced phosphorus removal, Barnard added an initial anaerobic zone to his nitrogen removal system to obtain the five-stage Bardenpho process, which removes both nitrogen and phosphorus. Since that time a great deal has been discovered about the mechanisms, microbiology, stoichiometry, and kinetics of BNR systems, and many process variants have been developed. Consequently, our understanding is sufficient to allow the design and operation of facilities that achieve reliable and predictable results.

11.1.2 General Description

BNR systems are modifications of the basic activated sludge process described in Chapter 10 and incorporate the four features common to them: (1) a flocculent slurry of microorganisms, (2) quiescent sedimentation, (3) settled solids recycle, and (4) solids retention time (SRT) control. In addition, the bioreactor of a BNR system is divided into anaerobic (ANA), anoxic (ANX), and aerobic zones (AER), with provision for mixed liquor recirculation (MLR), as illustrated in Figure 11.1. These zones are distinguished by the terminal electron acceptor utilized. In aerobic zones, oxygen is the electron acceptor; in anoxic zones, nitrate-N is the electron acceptor; and in anaerobic zones, neither oxygen nor nitrate-N is present. The division of the bioreactor to provide these alternative biochemical environments is the distinguishing

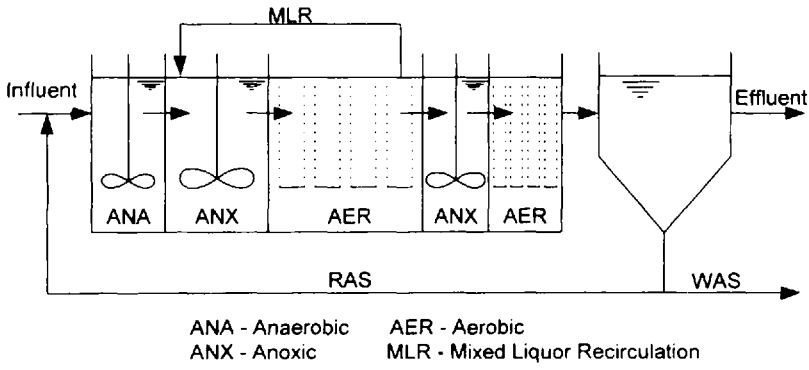


Figure 11.1 Single-sludge biological nutrient removal process.

feature of a BNR system. The aerobic zone is a necessary component of all BNR systems, while the anaerobic zone is necessary to accomplish phosphorus removal, and the anoxic zone is necessary for nitrogen removal.

Nitrogen removal occurs through the processes of nitrification and denitrification, as discussed in Chapters 2, 3, 6, and 7. Nitrification is an aerobic process and, consequently, will occur only in aerobic zones. Denitrification is the conversion of nitrate-N to nitrogen gas by heterotrophic bacteria that utilize nitrate-N as their terminal electron acceptor as they oxidize organic matter in the absence of dissolved oxygen, and thus it occurs in anoxic zones. The denitrification rate in the first anoxic zone is relatively rapid because the bacteria use readily biodegradable substrate added by the influent wastewater as the electron donor. Denitrification in the second anoxic zone is much slower because exogenous substrate concentrations are normally low due to their oxidation in the upstream anoxic and aerobic zones. Consequently, endogenous substrates must be used as electron donors, although some slowly biodegradable substrate may be available. The primary function of the final aerobic zone is stripping of nitrogen gas generated in the preceding anoxic zone and the addition of oxygen prior to passage of the mixed liquor suspended solids (MLSS) to the clarifier.

The incorporation of anoxic zones impacts the microbial ecology of the biomass. In nitrogen removal systems, the initial anoxic zone functions as an anoxic selector to minimize the growth of filamentous bacteria through metabolic selection, as mentioned in Section 10.2.1. Most filamentous bacteria are not capable of utilizing nitrate-N as an electron acceptor, whereas many floc-forming bacteria can.[37] Addition of the wastewater to an anoxic zone allows denitrifying bacteria to metabolize a portion of the readily biodegradable substrate and reduce the amount that passes into the aerobic zone, where it could be used by the filamentous bacteria. This limits the size of the filament population. In fact, anoxic zones have been incorporated into some nitrifying activated sludge systems simply to control filament growth. Denitrification also results in alkalinity production, which can partially offset the alkalinity consumed in nitrification.[57]

Biological phosphorus removal is accomplished by creating conditions favorable for the growth of PAOs, causing the activated sludge community to become enriched in them. As discussed in detail in Sections 2.4.6 and 3.7, and illustrated in Figure 11.2, the anaerobic zone provides the selective advantage for the PAOs by allowing them to grow at the expense of other heterotrophic bacteria. Because oxygen and nitrate-N are absent, oxidation of organic matter cannot occur in the time provided, making it impossible for most species of heterotrophic bacteria to transport and store or metabolize organic matter. Rather, they only carry out fermentation reactions, resulting in the formation of volatile fatty acids (VFAs). Phosphorus accumulating organisms are able to transport VFAs into the cell and store them as polyhydroxyalkanoic acids (PHAs) and other carbon storage polymers, using energy from the cleavage of intracellular polyphosphate, releasing inorganic phosphate. The VFAs are then unavailable to the other heterotrophic bacteria when the mixed liquor flows into the aerobic zone. Rather, the stored substrate is used exclusively by the PAOs for growth and to provide energy for reforming polyphosphate from inorganic phosphate in the wastewater. Only the slowly biodegradable substrate is available to the other heterotrophs. As a consequence, PAOs become a significant part of the community. Because of its role in microbial selection, the anaerobic zone is referred

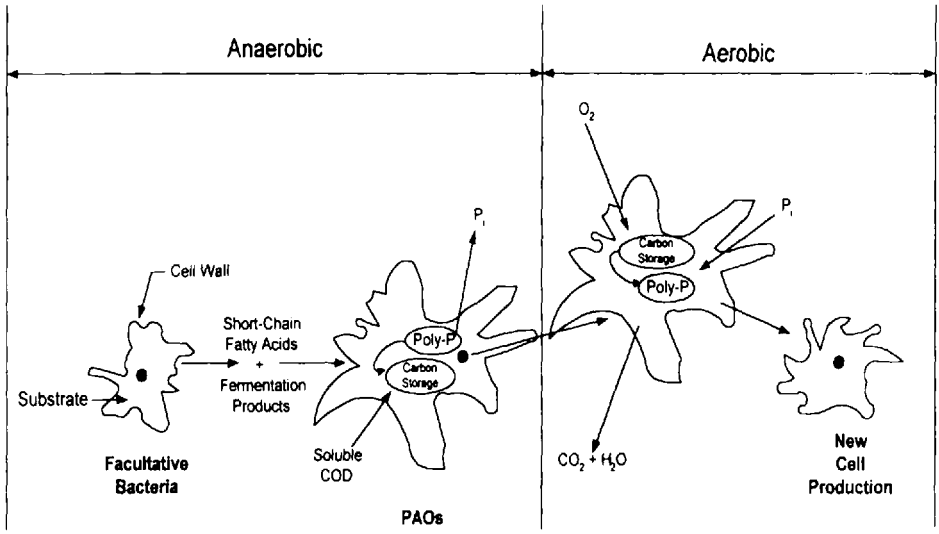


Figure 11.2 Relationship between phosphorus and organic matter metabolism in the anaerobic and aerobic zones of a BPR process.

to as an anaerobic selector.[37] Since the PAOs generally grow in a flocculent rather than a filamentous form, anaerobic selectors have also been used to control filamentous sludge bulking, providing another method of metabolic selection.

The enrichment of the biomass with PAOs, which contain a high concentration of polyphosphate at the end of the aerobic zone, provides the mechanism by which phosphorus is removed from the wastewater. The phosphorus content of a typical activated sludge is on the order of 1.5 to 2% (expressed on the basis of phosphorus to volatile suspended solids in the mixed liquor, or P/VSS), whereas when PAOs are present the P/VSS ratio will typically be increased to the 5 to 7% range, with values as high as 12 to 15% sometimes observed. Referring to Figure 11.1, phosphorus inputs to and outputs from the process include the influent, the treated effluent, and the waste activated sludge. By increasing the mass of phosphorus in the waste activated sludge (WAS), the mass in the treated effluent must be decreased in order to maintain a phosphorus mass balance.

The composition of the wastewater influences the reactions in the anaerobic zone, and hence its design. The readily biodegradable substrate in domestic wastewaters is a mixture of VFAs and other small biogenic organic compounds, with the proportions depending on the amount of fermentation that has occurred in the sewer. Generally, PAOs can only transport and store the short chain VFAs, acetic and propionic acid, in the anaerobic zone. However, the facultative bacteria will ferment the other readily biodegradable organic matter to produce VFAs, which can subsequently be utilized by the PAOs. Uptake of VFAs by the PAOs is relatively rapid, while fermentation is slower. Consequently, fermentation will be the rate limiting reaction in the anaerobic zone if only a small proportion of the readily biodegradable organic matter in the influent wastewater is present as VFAs. Thus, the size of the anaerobic zone in a phosphorus removal system will be influenced by the wastewater composition.

As demonstrated by the simulations of Section 7.7, nitrification adversely impacts biological phosphorus removal if nitrate-N is recycled to the anaerobic zone. There are two reasons for this: (1) denitrifying bacteria compete directly with the PAOs for readily biodegradable substrate, and (2) less formation of VFAs occurs by fermentation. Both reduce the selective advantage for PAOs. As a consequence, fewer PAOs are grown, the phosphorus content of the mixed liquor is reduced, and phosphorus removal is diminished.

Table 11.1 summarizes the biochemical transformations occurring in the various zones of a BNR process. It also presents the functions that these zones provide, as well as which zones are required to remove each nutrient. A key point is that an aerobic zone is required in all BNR systems. It is required for nitrogen removal because nitrifying bacteria are obligate aerobes. It is required for phosphorus removal because the stored and exogenous organic matter must be oxidized by the PAOs in an aerobic environment to generate the energy required for growth. This table may be used to understand the relative roles and the interactions between the various zones in BNR processes.

Because BNR processes are variations of the activated sludge process, they are constructed using the same materials and equipment components. The major differences are: division of the bioreactor into anaerobic, anoxic, and aerobic zones; provision of mixed liquor recirculation pumping facilities; and provision of mixing equipment in the anaerobic and anoxic zones to maintain solids in suspension while minimizing oxygen transfer. Figure 11.3 illustrates two types of mixers often used. Further discussion of the physical facilities is provided elsewhere.[54,57,68,72]

11.1.3 Process Options

Biological nutrient removal systems may be categorized according to their nutrient removal capabilities as nitrogen removal processes, phosphorus removal processes, and systems that remove both nitrogen and phosphorus.

Table 11.1 Summary of Biological Nutrient Removal Process Zones

Zone	Biochemical transformations	Functions	Zone required for
Anaerobic	• Uptake and storage of VFAs by PAOs • Fermentation of readily biodegradable organic matter by heterotrophic bacteria • Phosphorus release	• Selection of PAOs	• Phosphorus removal
Anoxic	• Denitrification • Alkalinity production	• Conversion of NO_x-N to N_2 • Selection of denitrifying bacteria	• Nitrogen removal
Aerobic	• Nitrification • Metabolism of stored and exogenous substrate by PAOs • Metabolism of exogenous substrate by heterotrophic bacteria • Phosphorus uptake • Alkalinity consumption	• Conversion of NH_3-N to NO_x-N • Nitrogen removal through gas stripping • Formation of polyphosphate	• Nitrogen removal • Phosphorus removal

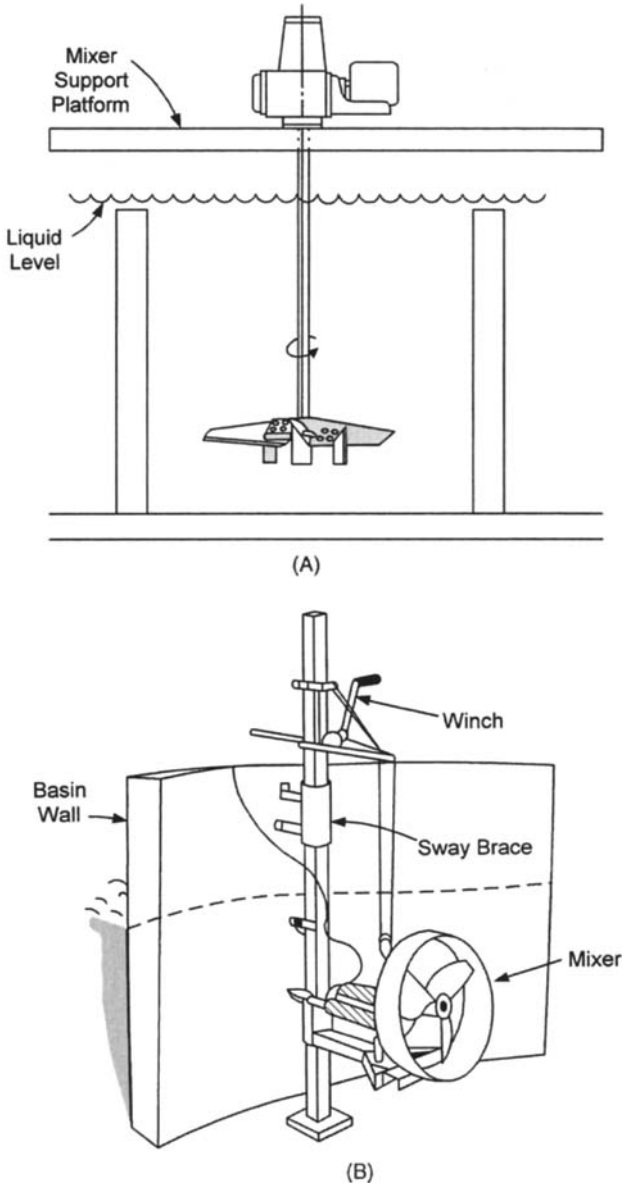


Figure 11.3 Typical vertical (A) and horizontal (B) mixers for anaerobic and anoxic zones.

Biological Nitrogen Removal Processes. Nitrogen removal processes incorporate aerobic zones for nitrification, anoxic zones for denitrification, and mixed liquor recirculation (MLR) to transfer the nitrate-N generated in the aerobic zone back to the initial anoxic zone. Many configurations are possible, resulting in a wide variety of performance capabilities and operational characteristics.

Figure 11.4 illustrates the modified Ludzak–Ettinger (MLE) process,[44] which is discussed in detail in Section 7.5. The denitrification rate in the anoxic zone is

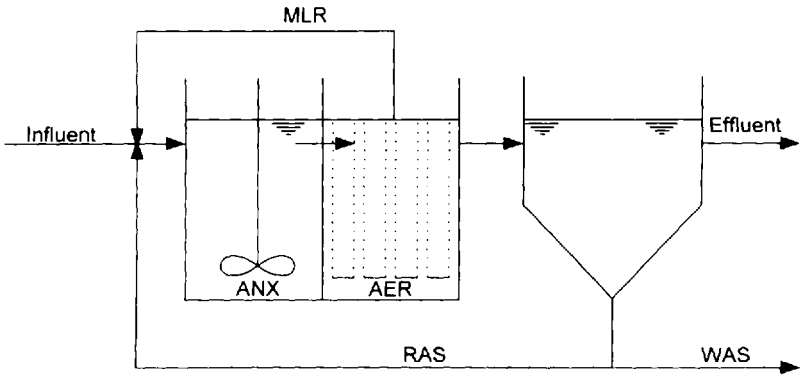


Figure 11.4 Modified Ludzak–Ettinger (MLE) process. A system with an anoxic selector has the same process flow diagram, but with a smaller anoxic zone.

relatively high because it is driven by the readily biodegradable substrate contained in the influent wastewater. Excellent nitrification and a good degree of denitrification (down to effluent nitrate-N concentrations of 4 to 8 mg/L as N) can be obtained with this process. However, further reductions in the effluent nitrate-N concentration are limited by the requirement to recirculate nitrate to the upstream anoxic zone, as illustrated in Figures 7.29 and 7.31. Mixed liquor recirculation flow rates typically range between one and four times the influent flow rate. The anoxic SRT ranges from about 1 to 4 days, with longer values being required when the wastewater contains smaller quantities of readily biodegradable substrate and when lower effluent nitrate-N concentrations are required. Aerobic SRTs are controlled by the need to maintain stable nitrification and range from about 4 to 12 days, depending on temperature, as discussed in Sections 9.3.2 and 10.2.2. Like all activated sludge systems, the system volume, and hence the hydraulic retention time (HRT), depends on the chosen MLSS concentration because the mass of biomass in the system is fixed once the SRT has been fixed. Typically, the HRT in the anoxic zone is in the 1 to 4 hr range, whereas the HRT in the aerobic zone may range from 4 to 12 hr. As defined by Eq. 4.15, HRTs for all BNR systems are calculated on the basis of influent flow alone; they do not include recycle and recirculation flows. The anoxic zone may be staged to increase denitrification efficiency and organism selection by provision of a substrate gradient.

Further nitrogen removal can be achieved by the addition of a second anoxic zone and a small final aerobic zone as provided in the four-stage Bardenpho[®] process discussed in Section 7.6 and illustrated in Figure 11.5. The second anoxic zone is used to remove the nitrate-N remaining in the effluent from the first aerobic zone. Although the denitrification rate is low in the second anoxic zone, the concentration of nitrate-N entering it is low and nearly complete nitrate-N removal can be obtained at anoxic SRTs of 2 to 4 days. HRTs are typically 2 to 4 hrs. The final aerobic zone is used only to strip nitrogen gas and to oxygenate the mixed liquor prior to entering the clarifier. The size of this zone is small to minimize decay reactions, which would liberate soluble nitrogen that would then pass into the effluent. Consequently, the HRT is typically around 30 min. The first anoxic and aerobic zones are similar in size to those in the MLE process, as is the MLR rate.

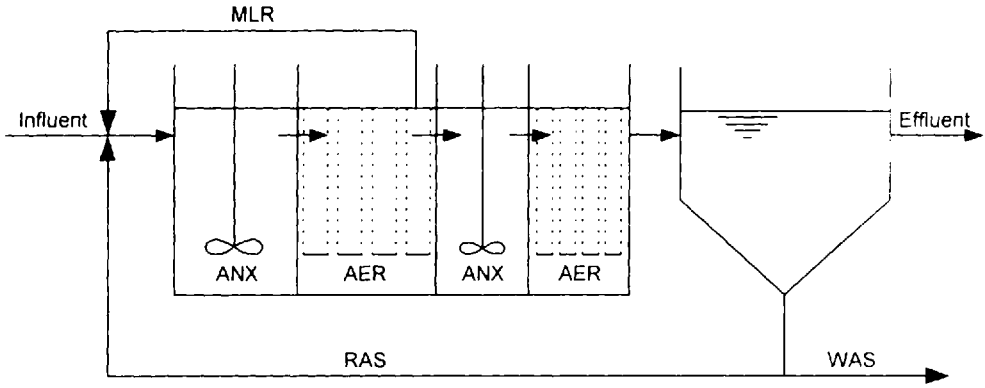


Figure 11.5 Four-stage Bardenpho process.

Significant nitrogen removal can also occur in SBRAS systems.[1,5,15,43] The performance of SBRAS systems and their analogy to continuous flow activated sludge systems, particularly the MLE process, is discussed in Section 7.8. The design and operation of SBRAS systems is discussed in Chapter 10. Many operating SBRAS facilities use an anoxic fill period or anoxic/aerobic cycling as an anoxic selector to control the growth of filamentous bacteria, also resulting in a significant degree of nitrogen removal. SFAS processes can also be configured to accomplish biological nitrogen removal by providing an anoxic zone at some or all of the step-feed addition points.[27,38,51] Significant nitrogen removal can be accomplished with this approach.

Significant denitrification is observed in some nitrifying activated sludge systems even though separate anoxic zones are not purposefully provided. Two circumstances are necessary for this to occur: (1) an operating SRT significantly greater than that required to achieve nitrification, and (2) regions of low DO concentration which allow denitrification to occur. If the SRT is considerably greater than that required for nitrification, the low DO regions can exist within the aerated bioreactor without negatively impacting its ability to nitrify. Low DO concentrations can occur in specific regions as a result of the bioreactor and aerator configurations, or they can occur inside the activated sludge floc particles throughout the bioreactor. The requirement to transport oxygen into the floc by diffusion will result in a DO concentration gradient within the floc. This gradient can produce anoxic zones inside the floc which can contribute to denitrification.

Extensive denitrification has been observed in some bioreactors oxygenated with point source aerators, such as mechanical surface aerators.[54,60] As illustrated in Figure 11.6, a high rate of oxygen transfer will occur at the surface of the basin as the mixed liquor is sprayed out from the aerator, allowing nitrification to occur. The mixed liquor will then circulate down into the basin and the rate of oxygen transfer will decrease as the mixed liquor is no longer in contact with air. As the DO concentration drops, nitrate-N generated through nitrification in the highly aerated surface zone can be denitrified. In fact, such systems provide the aerobic zone, anoxic zone, and MLR functions characteristic of the MLE process. It has been estimated that approximately 50% of the nitrate-N generated in such a basin can be denitrified if surface DO concentrations are maintained at approximately 1 mg/L. Even more

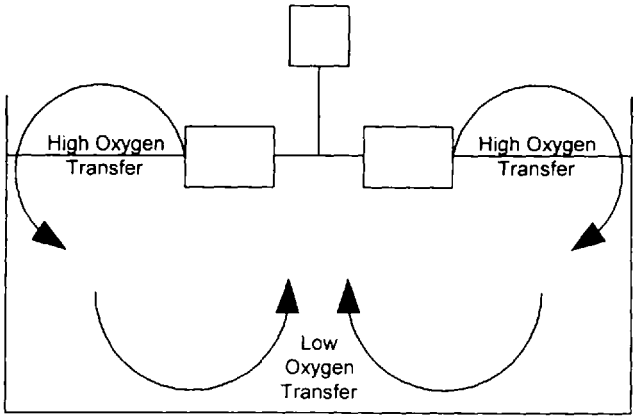


Figure 11.6 Variation in oxygen transfer rate in a biological reactor with a point source aerator.

complete denitrification, ranging up to 70 to 90%, has been observed in oxidation ditch activated sludge systems,[12,24,54,57] which incorporate the features of point source aeration and MLR on an even larger scale (see Figure 10.4). Poor mixing in the bioreactor can also encourage denitrification by creating low DO zones. Finally, the aeration input to continuous flow systems has also been cycled to create periods of high and low DO to encourage denitrification, as discussed in Section 6.5.2. While experience indicates significant denitrification potential for all of the above approaches, procedures to predict performance are, at best, approximate, and direct experience is the best guide to process design and operation. It is important to recognize, however, that operation with undefined anoxic zones may lead to growth of the filamentous bacteria classified in Table 10.5 as Group IV—aerobic, anoxic, and anaerobic zone growers.[29,37] Consequently, achieving nitrogen removal by encouraging denitrification in aerated basins may result in poorer sludge settleability. More work is needed on this subject to clarify the design and operational trade-offs.

Each of the systems previously considered in this section was developed to allow use of the biodegradable substrate in the wastewater as the electron donor for denitrification. In some cases, however, it may be necessary to denitrify the effluent from a nitrifying activated sludge system, which will be very low in organic matter. Such a stream can be readily denitrified in a separate stage suspended growth denitrification system[13,66,68,84] if a supplemental electron donor, such as methanol, is added, as illustrated in Figure 11.7. The electron donor is added in proportion to the nitrate-N content of the influent as it enters the anoxic zone, as discussed in Section 6.4. An aerobic zone is required to strip the entrained nitrogen gas and to prepare the biomass to settle in the clarifier. It will also allow metabolism of any excess electron donor that was not removed in the anoxic zone, thereby maintaining a high effluent quality. The SRT for such systems is typically between 3 and 5 days. Because of the low yields typically observed with substrates like methanol, the bioreactor HRT can be in the 2 to 4 hr range while maintaining typical MLSS concentrations. The aerobic HRT will typically be 45 to 60 min.

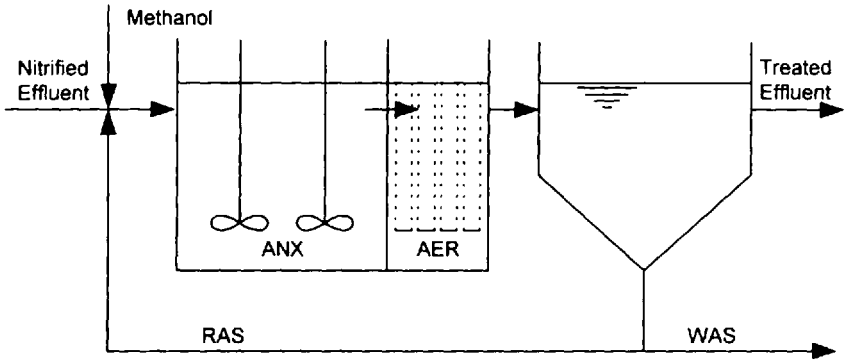


Figure 11.7 Separate stage suspended growth denitrification process.

Biological Phosphorus Removal Processes. Biological phosphorus removal processes utilize an anaerobic zone coupled with an aerobic zone to select for PAOs. Two systems will be considered: the A/O™ or Phoredox process, and the Phostrip® process.

Figure 11.8 provides a process schematic for the A/O™ process,[35,36,54,57] which is discussed in detail in Section 7.7. The term A/O™ stands for anaerobic/oxic (oxic = aerobic) and represents the sequence of environments provided in the bioreactor. The A/O™ process is referred to as a mainstream BPR process because the anaerobic zone is contained in the main process stream. Optimum phosphorus removal is obtained by high-rate operation, corresponding to an SRT in the 3 to 5 day range, with an anaerobic SRT of about 25 to 30% of the total. For typical MLSS concentrations, the resulting bioreactor HRT is often in the 3 to 6 hr range. High-rate operation maximizes phosphorus removal by minimizing nitrification and maximizing solids production. High solids production is desirable because it maximizes the wasting of high phosphorus content biomass. As indicated in Figure 9.2, aerobic SRT values as low as 2 to 3 days are required to prevent nitrification from occurring at temperatures of 20°C or above.

Figure 11.9 provides a simplified schematic diagram for the Phostrip® process.[41,42,57] It consists of a conventional activated sludge system that passes approximately 30 to 40% of the return activated sludge (RAS) flow through a stripper tank.

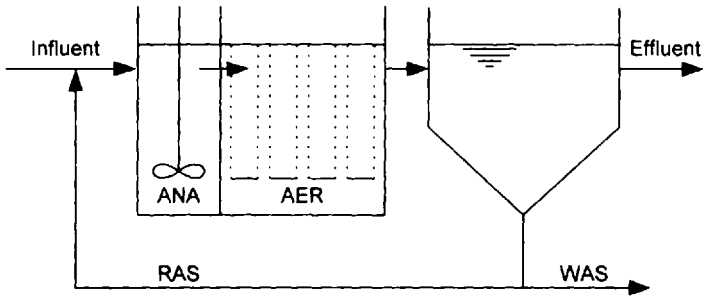


Figure 11.8 A/O™ or Phoredox process.

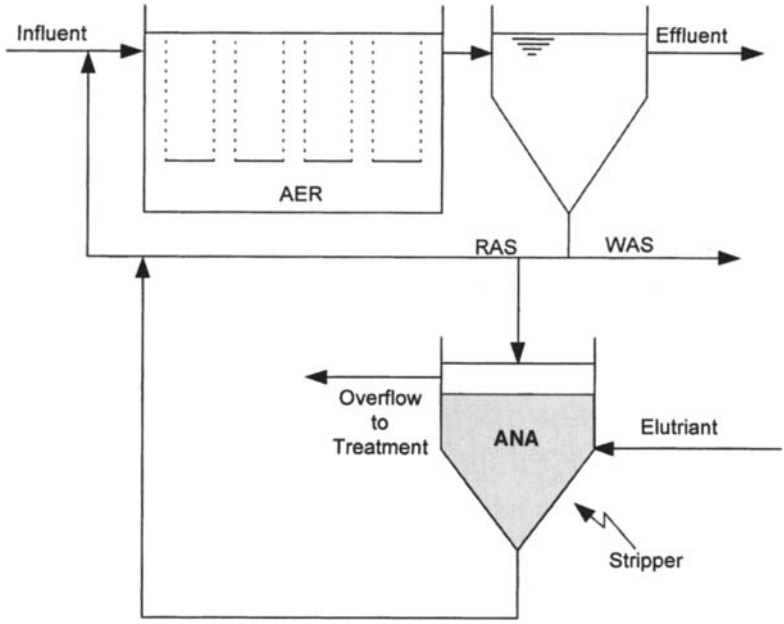


Figure 11.9 Phostrip® process.

The stripper is similar in configuration to a gravity thickener and maintains the MLSS in the RAS under anaerobic conditions in a sludge blanket, with a residence time of 8 to 12 hr. The stripper is the anaerobic zone where phosphorus release occurs, while the aerobic zone is provided by the main bioreactor where phosphorus uptake occurs. Since the anaerobic zone is provided in a sidestream bioreactor, the Phostrip® process is referred to as a side-stream process. In addition to a portion of the RAS, an external elutriant stream, such as a small portion of the influent wastewater or treated effluent, is added to the stripper to remove the released phosphorus. The loading of biodegradable organic matter on the stripper is relatively low. One source is particulate organic matter entrapped in the MLSS added to the stripper. The particulate matter can be solubilized and then fermented into VFAs in the stripper, thereby providing a VFA source for the PAOs. Another source is the organic matter contained in the elutriation stream. Indeed, experience indicates that process performance is enhanced when wastewater is used as the elutriant rather than treated effluent. The stripper overflow, which is enriched in phosphorus, is typically treated with lime to precipitate the phosphorus. Consequently, two phosphorus removal mechanisms operate in the Phostrip® process: through the WAS, due to the enriched phosphorus content of the MLSS, and in the stripper overflow.

Processes that Remove both Nitrogen and Phosphorus. Processes that remove both nitrogen and phosphorus incorporate anaerobic, anoxic, and aerobic zones along with MLR. Since nitrification occurs in these systems, the RAS will contain nitrate-N, which will adversely impact phosphorus removal if it is added directly to the anaerobic zone. Consequently, control of nitrate-N addition to the anaerobic zone is a key consideration in the selection and design of these processes.

Figure 11.10 provides a process schematic of the A^2/O^{TM} process.[3,50,54] The term A^2/O^{TM} stands for anaerobic/anoxic/oxic and, just as for the A/O^{TM} process, represents the sequence of environments provided in the bioreactor. It may be thought of as a combination of the MLE process for nitrogen removal and the A/O^{TM} process for phosphorus removal. Consequently, the SRT in the anaerobic zone is similar to that used in the A/O^{TM} process (0.75 to 1.5 days), and the SRTs in the anoxic and aerobic zones are similar to those used in the MLE process. The HRTs are also correspondingly similar, since the MLSS concentrations are in the same range. Mixed liquor recirculation rates of one to two times the influent flow rate are often used. The nitrogen removal capability of the A^2/O^{TM} process is similar to that of the MLE process, but the phosphorus removal capability is usually less than that of the A/O^{TM} process due to the recycle of nitrate-N to the anaerobic zone. Although the recirculation of mixed liquor to the anoxic zone results in substantial removal of nitrate-N, complete removal is not possible (just as it is not possible in the MLE process) and some nitrate-N is present in the RAS directed to the anaerobic zone. The impact of that nitrate-N depends on the organic content of the wastewater. If it is high relative to amount needed for both phosphorus and nitrogen removal, the nitrate-N recycle will not adversely impact effluent quality. On the other hand, if it is low, process performance will be adversely impacted. Maintenance of a solids blanket in the clarifier to allow denitrification of the RAS has been used with the A^2/O^{TM} process to reduce nitrate-N recycle. However, it requires careful clarifier operation to prevent clumping and floating sludge.

Another process for combined nitrogen and phosphorus removal is the five-stage Bardenpho[10] process illustrated in Figure 11.1. It may be viewed as a four-stage Bardenpho nitrogen removal process with an anaerobic zone added to achieve phosphorus removal. Nitrate-N recycle is controlled because of the extensive nitrogen removal obtained in the bioreactor. The overall process SRTs and HRTs in the anoxic and aerobic zones are similar to the corresponding SRTs and HRTs in the four-stage process, as are the MLR rates. The HRT in the anaerobic zone is typically 0.75 to 1.5 hr, making the SRT in the range of 0.75 to 1.5 days.

Two processes that eliminate nitrate-N recycle to the anaerobic zone are the University of Cape Town, South Africa (UCT)[25,54,57,75] and VIP[23,54,57] processes. Illustrated in Figure 11.11, elimination of nitrate-N recycle in the UCT process is accomplished by directing the RAS to the anoxic zone, where it is denitrified. Nitrified

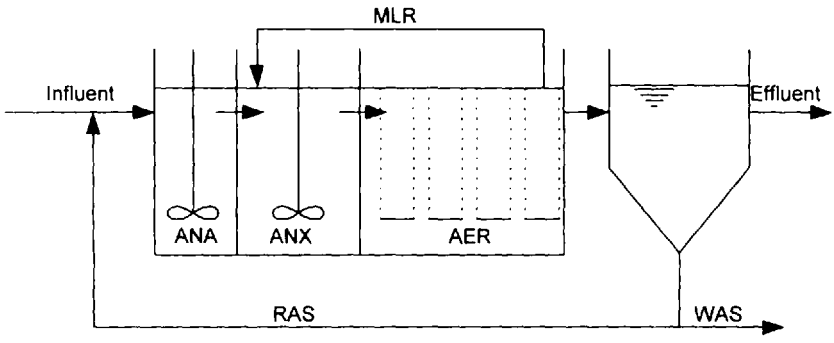


Figure 11.10 A^2/OT^{TM} process.

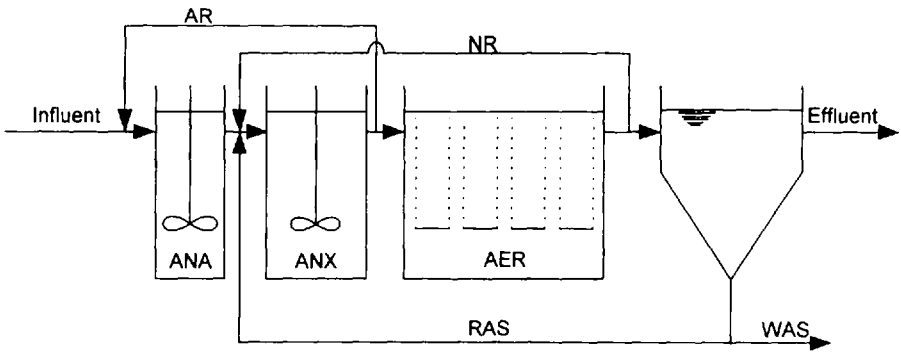


Figure 11.11 UCT process.

mixed liquor from the aerobic zone (referred to as nitrified mixed liquor recirculation, or NR) is also directed to the anoxic zone to increase the overall degree of nitrogen removal. A separate stream of denitrified mixed liquor from the end of the anoxic zone (referred to as anoxic mixed liquor recirculation, or AR) is recirculated to the anaerobic zone to provide the microorganisms needed there. Because mixed liquor, rather than RAS, is recycled to the anaerobic zone, the MLSS concentration in it is lower than the MLSS concentration in the remainder of the bioreactor. As a consequence, somewhat longer anaerobic zone HRTs (1 to 2 hr) are needed to achieve the desired SRTs. Anoxic and aerobic SRTs and HRTs are similar to those used in the MLE process. Both the anoxic and the nitrified mixed liquor recirculation rates are typically two times the process influent flow rate. One variation of the UCT process involves division of the anoxic zone into two compartments, one that receives and denitrifies the RAS and another that receives and denitrifies the NR. Referred to as the modified UCT (MUCT)[75] process, it is illustrated in Figure 11.12. The upstream anoxic zone, which receives RAS only, is oversized to ensure that complete denitrification occurs there and that essentially no nitrate-N is recirculated to the anaerobic zone. This makes control of the NR to the downstream anoxic zone less critical, and relatively high NR rates can be used to ensure that the full denitrification capability of this zone is used.

The term VIP stands for the Virginia Initiative Plant. As illustrated in Figure 11.13, it consists of anaerobic, anoxic, and aerobic zones and, like the UCT process,

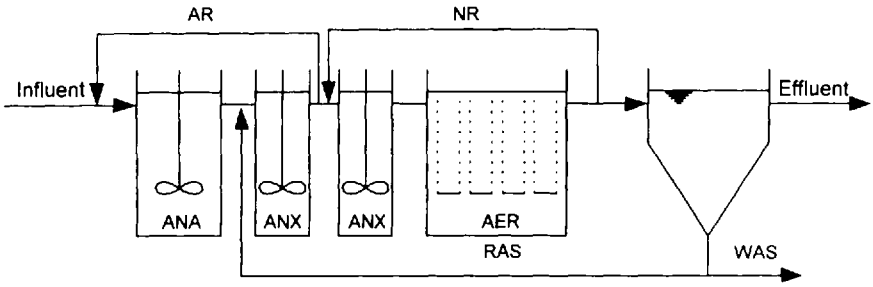


Figure 11.12 Modified UCT (MUCT) process.

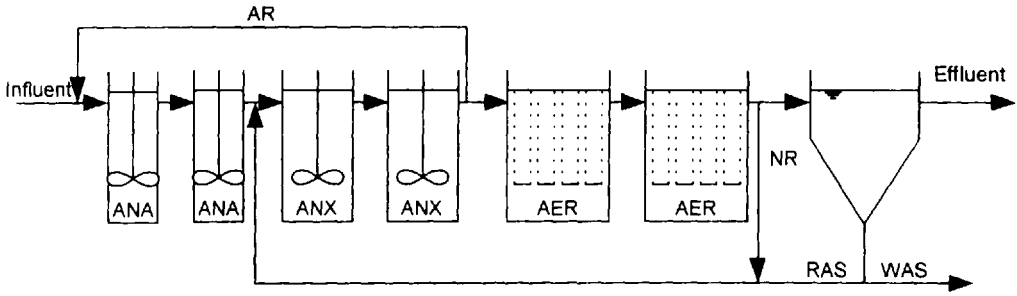


Figure 11.13 VIP process.

RAS is recycled to the anoxic zone while a denitrified AR stream is directed from the end of the anoxic zone to the anaerobic zone. However, several differences exist between the VIP and UCT processes. In the VIP process, all zones are staged, i.e., they consist of at least two (and as many as six) completely mixed cells in series; the process is designed as a high-rate, i.e., a short SRT, process to maximize phosphorus removal; and the NR is mixed with the RAS for recycle to the anoxic zone. Staging of the various zones provides several benefits. Increased efficiency is obtained with the use of a tanks-in-series hydraulic regime, just as it is for activated sludge systems. A tanks-in-series configuration also allows the respective environmental conditions to be more completely established in each zone. Finally, staging may allow the selection of microorganisms with increased reaction rates, since each zone functions more like a selector. High-rate operation is accomplished by minimizing the SRT, and hence the HRT, in each reactor zone. The combined SRT of the anaerobic and anoxic zones is generally 1.5 to 3 days, while the anaerobic and anoxic HRTs are typically 60 to 90 minutes each. The aerobic zone is sized just large enough to achieve a sufficient degree of nitrification to meet process objectives. The AR flow rate is typically 50 to 100% of the influent flow rate, while the NR flow rate is typically equal to the influent flow rate. Mixing of the NR with the RAS allows deoxygenation of the NR by the oxygen deficient RAS before it is added to the anoxic zone. This improves sludge settling characteristics by reducing the oxygen loading on the anoxic zone and minimizing the growth of low DO filamentous bacteria.[58]

Many other biological nitrogen and phosphorus removal processes have been developed and have received some full-scale use. Some minimize nitrate-N recycle to the anaerobic zone by allowing the RAS to denitrify under endogenous conditions. Increasing the residence time of the RAS in the clarifier, as described in connection with the A^2/O process, is one example. Two others are the Johannesburg[59] and the R-D-N[39] processes. Processes that use oxidation ditches include the VT2[54,60] and the Biodenitro[61] processes. The operation of SBRAS processes has also been modified to obtain both nitrogen and phosphorus removal.[40,46] Finally, full-scale facilities have been modified by simply turning off aerators and/or by other simple modifications to create the zones necessary to achieve nitrogen and phosphorus removal.[54,62] The potential to enhance the removal of nutrients by similar modifications of activated sludge systems appears to be limited only by the imagination and understanding of the process fundamentals by plant designers and operators.

Fermentation of primary sludge to generate an influent stream high in VFAs for use in systems that remove both nitrogen and phosphorus is a recent, and exciting, development.[11] It offers significant potential for enhancing the performance and improving the reliability of BNR systems.[52,53,64] The impact of fermentation on the performance of BNR facilities is described in Section 11.2.3, while the basic principles of sludge fermentation and the design of fermenters are discussed in Chapter 13.

11.1.4 Comparison of Process Options

Table 11.2 summarizes the primary benefits and drawbacks of several BNR systems. The MLE process offers good nitrogen removal, moderate bioreactor volume requirements, alkalinity recovery, good sludge settleability, reduced oxygen requirements compared to traditional activated sludge systems, and simple control. However, a high level of nitrogen removal cannot generally be achieved, as discussed previously. Practical MLR flow rates limit nitrate-N removal to between 60 and 85%. As illustrated in Figure 7.36, this constraint does not exist for the four-stage Bardenpho process, which includes a second anoxic zone. Performance data from full-scale wastewater treatment plants demonstrates this difference.[57] Processes with one anoxic zone typically produce effluents with total nitrogen concentrations ranging between 5 and 10 mg/L as N, while processes with two anoxic zones typically produce effluents with total nitrogen concentrations ranging between 1.5 and 4 mg/L as N.[57] However, this improved performance is at the expense of a larger bioreactor volume. Another benefit of the MLE and four-stage Bardenpho processes is alkalinity production by denitrification in the initial anoxic zone, which off-sets some of the alkalinity consumed by nitrification in the aerobic zone. Denitrification also reduces the oxygen requirement in the aerobic zone because nitrate-N serves as the electron acceptor during oxidation of some of the biodegradable organic matter, thereby removing the need for oxygen to do so. These effects are discussed in Sections 6.3, 6.4, 7.5, and 7.6, and illustrated in Figure 7.30. The reduced power requirements for oxygen transfer in the aerobic zone off-set some or all of the energy required to mix the anoxic zone and to pump the MLR. Good sludge settleability can be obtained with both the MLE and four-stage Bardenpho processes because the initial anoxic zone acts as a selector to control the growth of filamentous bacteria, as discussed previously. The incorporation of an anoxic zone in the bioreactor can also minimize denitrification problems in the secondary clarifier by reducing nitrate-N concentrations, making it impossible to generate sufficient nitrogen gas to cause sludge flotation.

Systems that encourage denitrification in an aerobic bioreactor provide the benefits of alkalinity recovery and oxygen requirement reduction associated with the MLE and four-stage Bardenpho processes. In fact, the total energy requirements in such systems are smaller since mixing and MLR facilities are generally not required. Some existing activated sludge facilities can easily be retrofitted. However, relatively large bioreactor volumes may be required since the microbial environment is not optimized, control can be more complex to restrict oxygen input to allow the anoxic regions to develop, and poor sludge settleability may result due to the growth of Group IV filamentous bacteria.

Table 11.2 Biological Nutrient Removal Process Comparison

Process	Benefits	Drawbacks
Nitrogen Removal		
MLE	• Good nitrogen removal • Moderate reactor volume • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• High level of nitrogen removal not generally possible
Four-stage Bardenpho	• Excellent nitrogen removal • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• Large reactor volume
Denitrification in aerobic reactor	• Alkalinity recovery • Reduced energy requirement • Easily applied to some existing facilities	• Large reactor volume • Complex control • May result in poor sludge settleability
Separate stage suspended growth denitrification	• Excellent nitrogen removal • Minimum reactor volume	• Requires upstream nitrification • Supplemental electron donor required • High energy requirement
Phosphorus Removal		
A/O™	• Minimum reactor volume • Good phosphorus removal • Good solids settleability • Simple operation	• Phosphorus removal adversely impacted if nitrification occurs
Phostrip®	• Excellent phosphorus removal	• Complex operation • Phosphorus removal adversely impacted if nitrification occurs
Nitrogen and Phosphorus Removal		
A²/O™	• Good nitrogen removal • Moderate reactor volume • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• High level of nitrogen removal not generally possible • Moderate phosphorus removal
VIP and UCT	• Good nitrogen removal • Good phosphorus removal • Moderate reactor volume • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• High level of nitrogen removal not generally possible • An additional MLR step is required
Five-stage Bardenpho	• Excellent nitrogen removal • Alkalinity recovery • Good solids settleability • Reduced oxygen requirement • Simple control	• Large reactor volumes • Moderate to poor phosphorus removal

Separate stage denitrification systems offer the benefits of excellent nitrogen removal and minimum bioreactor volumes. However, extensive upstream treatment is needed to convert ammonia-N to nitrate-N, a supplemental electron donor such as methanol is typically required, and energy requirements are higher due to the need to mix and aerate the bioreactor. Furthermore, the benefits of reduced oxygen requirements and alkalinity recovery associated with the other processes are not available since denitrification occurs after biodegradable organic matter has been removed from the wastewater and ammonia-N has been converted to nitrate-N.

When process objectives are limited to carbon oxidation and phosphorus removal, the A/O™ process offers the benefits of minimum bioreactor volume, good phosphorus removal, good solids settleability, and simple operation. Good solids settleability results from the anaerobic zone functioning as a selector to control the growth of filamentous bacteria. The primary drawback is that the phosphorus removal capability is adversely impacted if nitrification occurs because of the recycle of nitrate-N to the anaerobic zone via the RAS. The Phostrip® process achieves excellent phosphorus removal due to the two phosphorus removal mechanisms provided.[52] However, experience indicates that it is relatively complex and difficult for many to operate.[71] Like the A/O™ process, its phosphorus removal performance is adversely impacted if nitrification occurs due to the input of nitrate-N to the anaerobic stripper.

The A^2/O™ process offers the benefits and drawbacks of the MLE process, along with the capability to remove a significant degree of phosphorus. Like the A/O™ process, however, its phosphorus removal capability is adversely impacted by the presence of nitrate-N in the RAS. This limitation is eliminated in the VIP and UCT processes by directing the RAS to the anoxic zone. However, this added capability is obtained at the expense of another MLR pumping step, the AR. Several studies have demonstrated the superior phosphorus removal capabilities of the VIP and UCT processes in comparison to the A/O™ and A^2/O™ processes in nitrifying applications.[16,21,23,75] Similarly, the five-stage Bardenpho process provides the advantages of the four-stage Bardenpho process, along with the capability to remove phosphorus. Although nitrate-N recycle to the anaerobic zone is minimized in this process, experience indicates that its phosphorus removal capability may be only moderate to poor.[57] This results from the relatively long SRTs often used, which lead to relatively low waste sludge production rates, relatively low phosphorus uptake, and secondary release of phosphate in the second anoxic zone.

11.1.5 Typical Applications

Single-sludge nitrogen removal systems, such as the MLE and four-stage Bardenpho processes, have become the preferred alternatives for nitrogen removal during wastewater treatment. Physico–chemical technologies, such as ammonia stripping, ion exchange, and breakpoint chlorination, have been found to be expensive options that are difficult to operate and maintain.[57,68] Although separate stage biological nitrogen removal processes are technically feasible, they are rarely cost-effective relative to single-sludge systems. However, they may be the best alternative when nitrogen removal must be added to an existing wastewater treatment plant, particularly if little space is available to construct additional facilities. In such instances, attached growth bioreactors (Chapter 21), such as packed bed downflow denitrifying bioreactors or fluidized bed bioreactors (FBBRs), may be better than suspended growth denitrifi-

cation systems. Consequently, although separate stage suspended growth denitrification systems are capable of providing a high level of performance,[84] they have received little full-scale use.[57,68] Several examples of full-scale suspended growth single-sludge biological nitrogen removal facilities are presented in the literature, however.[57,68]

Biological phosphorus removal is also seeing increased use in comparison to chemical phosphorus removal. Biological phosphorus removal offers the advantages of reducing or eliminating chemical usage and decreasing the quantity of waste sludge that must be processed. These benefits must be evaluated relative to the capital and operating costs associated with the construction of the necessary facilities to accomplish biological phosphorus removal. Although the sidestream Phostrip® process was used extensively in the past, mainstream processes are being used much more frequently today. This is because of the mechanical and process complexity of the sidestream processes. Finally, solids fermentation is being used more frequently to improve the performance and reliability of BPR facilities. Sedlak[57] presents several examples of full-scale BPR facilities.

Biological nutrient removal capabilities have also been incorporated into the design of some activated sludge facilities due to the process benefits they provide. When properly designed, initial anoxic and anaerobic zones function as selectors to enhance solids settleability by controlling the growth of filamentous bacteria. The inclusion of denitrifying capabilities in nitrifying activated sludge systems can also reduce alkalinity and energy consumption, and can reduce solids flotation problems in secondary clarifiers by decreasing the nitrate-N concentration in them. In several instances, these benefits have proven to be cost-effective and have improved the performance and reliability of the activated sludge systems into which they were incorporated.[19,37,47]

11.2 FACTORS AFFECTING PERFORMANCE

Many of the factors that affect the performance of BNR processes are similar to those that affect activated sludge systems. This section will discuss those factors that are of particular concern in BNR systems.

11.2.1 Solids Retention Time

Solids retention time plays the same role in BNR processes as in activated sludge systems. In this case, however, the SRTs in the separate zones are of more interest than the total SRT, since they control what occurs in those zones,[43] as discussed in Chapters 6 and 7. In Section 6.5.2, the aerobic SRT ($\Theta_{c,AER}$) was defined as the fraction of the total system SRT that is aerobic. Because the SRT is defined in terms of the mass of biomass in a system (Eq. 5.1), the fraction of the system SRT that is aerobic is equivalent to the fraction of the biomass in the system that is maintained under aerobic conditions, $f_{XM,AER}$. Consequently:

$$\frac{\sum (X_{M,T} \cdot V)_{AER}}{(X_{M,T} \cdot V)_{System}} = f_{XM,AER} = \frac{\Theta_{c,AER}}{\Theta_c} \qquad (11.1)$$

Likewise, with the fractions that are anoxic:

$$\frac{\sum (X_{M,T} \cdot V)_{ANX}}{(X_{M,T} \cdot V)_{System}} = f_{XM,ANX} = \frac{\Theta_{c,ANX}}{\Theta_c} \tag{11.2}$$

and anaerobic:

$$\frac{\sum (X_{M,T} \cdot V)_{ANA}}{(X_{M,T} \cdot V)_{System}} = f_{XM,ANA} = \frac{\Theta_{c,ANA}}{\Theta_c} \tag{11.3}$$

The MLSS concentration in any given tank will depend on the positions of the solids recycle and MLR streams. Like conventional activated sludge (CAS) systems, MLE, four-stage Bardenpho, A/O™, and A^2/O™ systems can be considered to have uniform MLSS concentrations throughout. As a consequence, the fraction of the biomass held under any given environmental condition is equal to the fraction of the system volume maintained under that condition. Because of the AR in the UCT, modified UCT, and VIP processes, however, the MLSS concentration in their anaerobic zones will be smaller than the MLSS concentration in the rest of the system. Consequently, in those processes, the fraction of the system volume that is anaerobic is larger than the fraction of biomass held under anaerobic conditions and the fraction of the SRT that is anaerobic.

One way in which the aerobic SRT can dramatically affect a BNR system is through organism selection. If the system is to be designed and operated to achieve phosphorus removal, the aerobic SRT must be sufficiently long to allow PAOs to grow. Figure 9.4 provides guidance for selection of the required aerobic SRT as a function of temperature. Although the values are relatively low, full-scale activated sludge systems can be operated at aerobic SRTs below that required to grow PAOs.[61] Thus, care must be exercised to ensure that the aerobic SRT is sufficiently long. Just how long that SRT must be is determined in part by whether nitrification is required. If nitrification is an objective, it will control the choice of the aerobic SRT, because as Figure 9.4 illustrates, the minimum aerobic SRT for nitrification is longer than the minimum for growth of PAOs. If nitrification is not an objective, on the other hand, making the aerobic SRT short enough to preclude nitrification allows phosphorus removal to be maximized. This is particularly important for a system like the A/O™ process, in which nitrate-N will be recycled to the anaerobic zone should nitrification occur. Interestingly, Figure 9.4 suggests that at temperatures above about 25°C, it may be very difficult to operate at an aerobic SRT sufficiently high to allow PAOs to grow while also excluding nitrifying bacteria. Consequently, under those circumstances, systems like the VIP and UCT processes have distinct advantages. Appropriate safety factors must be added to the values presented in Figure 9.4 to derive design aerobic SRTs, as illustrated in Section 10.3.2.

Experience also suggests appropriate ranges for the SRT in the anaerobic and anoxic zones. If the removal of readily biodegradable substrate is the primary objective in these zones, then an SRT of about 1 day for temperatures above 20°C is appropriate, while it must be increased to about 1.5 days for colder temperatures.[47] These criteria would also be used if either anaerobic or anoxic zones were being incorporated into a system as selectors to control filament growth. As discussed in Section 10.2.1, readily biodegradable substrate must be removed upstream of the main, aerobic section of the bioreactor if the growth of filamentous bacteria is to be

controlled. Since this generally represents a baseline process objective, these values represent the minimum ones that should be used for the combination of the anaerobic and anoxic zones. The anaerobic plus anoxic SRT must be increased if significant quantities of slowly biodegradable substrate are to be metabolized in the zones. Anaerobic plus anoxic SRTs in the 2 to 3 day range have been used successfully in some pilot- and full-scale BNR systems.[23,56] In such circumstances, the portion of the total SRT allocated to the anaerobic zone may be as short as 0.5 days if the concentrations of VFA and readily biodegradable organic matter in the process influent are sufficiently high.[43] Increases in the anaerobic SRT will allow increased fermentation of biodegradable organic matter in the anaerobic zone, resulting in increased production of VFAs and increased biological phosphorus removal. Increased anoxic zone SRT will result in increased metabolism of slowly biodegradable substrate there and increased nitrogen removal capacity. Kinetic expressions such as those incorporated into activated sludge model (ASM) No. 2 must be used to specifically determine how much the anaerobic and anoxic SRTs must be increased to achieve a particular set of objectives.

Longer SRTs in BNR systems allow more complete metabolism of organic matter, which can increase nitrogen removal, but adversely impact phosphorus removal. Denitrification requires an electron donor. Readily biodegradable organic matter can be used rapidly as an electron donor, but slowly biodegradable organic matter must be hydrolyzed before it can be used. Hydrolysis is a relatively slow reaction under anoxic conditions and, consequently, long anoxic SRTs may be required when slowly biodegradable substrate must be used as the electron donor.[21,49] On the other hand, phosphorus removal can be adversely affected by the use of relatively long anoxic or aerobic SRTs.[43] This may occur for at least three reasons: (1) long SRTs result in reduced solids production so that less phosphorus is removed from the process in the WAS; (2) long aerobic SRTs result in relatively complete oxidation of organic storage products and a reduced rate of phosphorus uptake in the aerobic zone; and (3) decay reactions cause secondary release of phosphorus, i.e., the release of phosphorus without a corresponding uptake and storage of biodegradable organic matter. Thus, SRTs beyond that just required to meet treatment objectives should be avoided for BPR systems.

In summary, for BPR processes an anaerobic SRT of 1 to 1.5 days should be used, depending on temperature. The aerobic SRT for such systems should be selected from Figure 9.4, with an appropriate safety factor applied. A safety factor of about 1.5 is probably sufficient for most applications. For biological nitrogen removal processes, the initial anoxic SRT must be at least 1 to 1.5 days (depending on temperature), but a larger value may be used if significant hydrolysis of slowly biodegradable substrate is required. The aerobic SRT should be selected to achieve reliable nitrification, as discussed in Sections 10.2.2 and 10.3.2. For processes that remove both nitrogen and phosphorus, an anaerobic plus anoxic SRT of at least 2 to 3 days should generally be used, with higher values being used if significant hydrolysis of slowly biodegradable organic matter is necessary in the anaerobic and anoxic zones. The anaerobic SRT should be at least 0.5 days. The aerobic SRT should be selected in the same fashion as for nitrogen removal processes, although care should be exercised to use the minimum value that provides the necessary system performance.

11.2.2 Ratios of Wastewater Organic Matter to Nutrient

The concentration of biodegradable organic matter relative to the nutrient concentrations in an influent wastewater can dramatically affect the performance of a BNR system. This is because of the key role biodegradable organic matter plays in nutrient removal. Nitrogen removal is accomplished when biodegradable substrate is used as the electron donor by denitrifying bacteria under anoxic conditions. Phosphorus removal is accomplished when VFAs, which are either a part of the influent readily biodegradable substrate or are formed from it, are taken up and stored by PAOs in the anaerobic zone, thereby allowing them to increase the phosphorus content of the MLSS in the aerobic zone. Since only some PAOs can denitrify and their efficiency at forming polyphosphate under those conditions is less,[6,8,7] it would seem that the organic matter requirements for nitrogen and phosphorus removal could be additive when both nutrients are to be removed. As discussed in Section 11.3.3, however, evidence suggests that this is not the case and only the greater of the two requirements, i.e., for nitrogen or phosphorus removal, need be supplied.

Consider first the amount of organic matter required to remove nitrogen. As shown in Table 3.1, the chemical oxygen demand (COD) mass equivalent of nitrate-N as an electron acceptor is −2.86 g COD/g N. This does not mean, however, that an influent biodegradable COD to total Kjeldahl nitrogen (TKN) ratio of 2.86 will allow complete denitrification. There are several reasons for this. First, some of the biodegradable organic matter in the influent is incorporated into biomass, which is why the $\Delta S/\Delta N$ ratio for denitrification, as calculated with Eq. 6.4, is always greater than 2.86. Second, some of the influent nitrogen is incorporated into the biomass that grows in the process, which is why the nitrogen available for nitrification, S_{NH}, as calculated with Eq. 10.17, is always less than the influent TKN concentration. Third, some of the biodegradable organic matter is slowly biodegradable and may not be metabolized rapidly enough to be available in the anoxic zones. Thus, rather than being used for denitrification, it will be oxidized aerobically. Finally, differences in system configuration result in different utilization efficiencies for the biodegradable organic matter. Table 11.3 provides general guidance concerning the amenability of various wastewaters (characterized in terms of the amount of organic matter relative to the amount of nitrogen) to biological nitrogen removal. The values given can be used to screen candidate wastewaters to determine how difficult it may be to achieve good nitrogen removal. Alternatively, procedures such as those developed in South Africa are available to estimate the appropriate COD/TKN ratio for various nitrogen removal processes.[25,75] Experience has indicated that, in many instances,

Table 11.3 Relationship Between Expected Biological Nitrogen Removal Efficiency and Influent Organic Matter to Nitrogen Ratios

Nitrogen removal efficiency	COD/TKN	BOD_5/NH_3-N	BOD_5/TKN
Poor	<5	<4	<2.5
Moderate	5–7	4–6	2.5–3.5
Good	7–9	6–8	3.5–5
Excellent	>9	>8	>5

these procedures provide conservative estimates of the required ratio.[54] That is, adequate performance can be obtained at values of the COD/TKN ratio lower than the recommended value. They may be useful, however, in circumstances where "worst case" values need to be estimated.

Next consider the amount of organic matter required to remove phosphorus. The COD/total phosphorus (TP), the five-day biochemical oxygen demand (BOD_5)/TP, and the soluble BOD_5/soluble phosphorus (SP) ratios are often used to judge the phosphorus removal potential of a wastewater.[54,57,65,72] Figure 11.14[65] illustrates the impact of the BOD_5/TP ratio on the performance of some full-scale BPR processes. Results such as these demonstrate that a relationship exists, but that it varies with process type and operating conditions. The concept of a minimum organic matter requirement for biological phosphorus removal has resulted in a distinction between carbon limited and phosphorus limited wastewaters. A carbon limited wastewater is one in which insufficient organic matter is available to remove all of the phosphorus. As a consequence, phosphorus will be present in the process effluent at a concen-

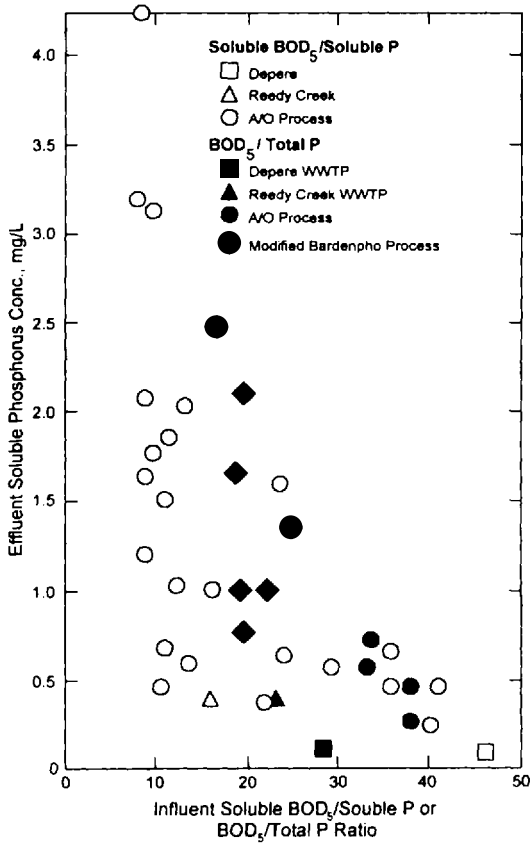


Figure 11.14 Example relationship between the influent BOD_5/P ratio and the effluent soluble phosphorus concentration. (From M. J. Tetrault et al., Biological phosphorus removal: a technology evaluation. *Journal, Water Pollution Control Federation* **58**:823–827, 1986. Copyright © Water Environment Federation; reprinted with permission.)

tration determined by the relative concentrations of phosphorus and organic matter in the influent. A phosphorus limited wastewater is one in which more than sufficient organic matter is available to remove the phosphorus. Consequently, the effluent phosphorus concentration will generally be low when it is treated in a BPR process. Thus, a phosphorus limited wastewater is desirable when a good quality effluent must be produced.

Recognition of the concept of carbon limited wastewaters has resulted in the development of benchmark ratios expressing the amount of organic matter required to remove a unit of phosphorus by various BPR processes. Such ratios have been determined from pilot- and full-scale BPR systems operating under carbon limited conditions and have been used to characterize the capabilities of those systems.[57-71] A commonly used ratio is the BOD_5 to phosphorus removal ratio ($BOD_5/\Delta P$), which is calculated as:

$$BOD_5/\Delta P = \frac{BOD_5 \text{ in biological process influent}}{\text{TP in biological process influent} - \text{SP in biological process effluent}} \quad (11.4)$$

Phosphorus removal is quantified as the total phosphorus in the biological process influent minus the soluble phosphorus in the biological process effluent. Influent total phosphorus is used because both soluble and particulate phosphorus are acted upon. Soluble phosphorus is in the form of, or rapidly converted to, inorganic phosphate, which is the form taken up and stored by the PAOs. Particulate phosphorus is either hydrolyzed and released as soluble phosphorus or it is entrapped in the MLSS and removed in the waste solids. In either case, the particulate phosphorus affects the overall phosphorus removal by the process. Effluent soluble phosphorus is used because effluent particulate phosphorus is generally associated with suspended solids that have escaped the clarifier and are a function of the efficiency of the clarifier, not the biological process.

Table 11.4 provides typical ranges for $BOD_5/\Delta P$ for a variety of BPR processes. For consistency in this text, values are also presented for $COD/\Delta P$ which were calculated from the BOD_5 values using Eq. 8.33. Note that a low value for the ratio

Table 11.4 BOD_5 and COD to Phosphorus Removal Ratios for Various BPR Processes

Type of BPR process	$BOD_5/\Delta P$ ratio (mg BOD_5/mg P)	$COD/\Delta P$ ratio (mg COD/mg P)
High efficiency (e.g., A/O™ without nitrification, VIP, UCT)	15–20	26–34
Moderate efficiency (e.g., A/O™ and A^2/O™ with nitrification)	20–25	34–43
Low efficiency (e.g., Bardenpho)	>25	>43

indicates an efficient process since little organic matter is required to remove a unit of phosphorus. Highly efficient BPR processes, such as the A/O™ process operating under nonnitrifying conditions or the VIP process, require only 15–20 mg BOD_5 (26–34 mg COD) to remove a mg of phosphorus. In these processes, essentially no nitrate-N is recycled to the anaerobic zone, either because it is not generated (for the nonnitrifying A/O™ process) or it is removed (for the VIP process). They are also both high-rate processes, which maximizes phosphorus uptake and waste solids production. A moderately efficient process, such as a nitrifying A/O™ or A^2/O™ process, will require 20–25 mg of BOD_5 (34–43 mg COD) to remove one mg of phosphorus. More organic matter is required for these systems because some will be consumed by non-PAO heterotrophs in the anaerobic zone due to the nitrate-N recycled there in the RAS. The ratio will be even higher for a low efficiency process, such as a five-stage Bardenpho process operating at long SRT, which require more than 25 mg of BOD_5 (43 mg COD) to remove a mg of phosphorus. Thus, it can be seen that the effects of many of the factors discussed in previous sections of this chapter are quantified in the organic matter to phosphorus removal ratio.

As discussed above, a BPR process will achieve good performance if it operates under phosphorus limited conditions. This occurs when the organic matter to phosphorus ratio of the influent wastewater is greater than the $BOD_5/\Delta P$ value for the BPR process being used, i.e., when more organic matter is available per unit of phosphorus than is required by the process to remove the phosphorus. Therefore, appropriate BPR processes can be identified for a particular application by comparing the organic matter to phosphorus ratio for the wastewater to the $BOD_5/\Delta P$ values for candidate BPR processes and selecting those processes with appropriate removal ratios. Thus, ratios such as those summarized in Table 11.4 can be quite useful in the initial stages of process evaluation and screening.

It must be emphasized that the organic matter to nutrient ratios discussed above are for the influent to the biological treatment system, not the ratios for the influent to the entire wastewater treatment plant. This is because the ratio can be significantly altered by treatment upstream of the biological process and by recycle streams from the solids handling system.

The potential for nutrient inputs to biological processes from solids handling systems is so great that it deserves emphasis. Some solids handling unit operations result in significant ammonification of organic nitrogen contained in the waste solids applied to them. Examples include anaerobic digesters, aerobic digesters, and heat treatment systems. A liquid stream with a high ammonia-N concentration is produced when the outflows from them are dewatered. If such streams are recycled to the liquid treatment process train, they can significantly increase the ammonia-N load without increasing the organic load. Moreover, such discharges are often periodic in nature, and the resulting ammonia-N shock load can overload the bioreactor system. Such solids handling systems can also result in solubilization of removed phosphorus and its recycle back to the liquid treatment process train, particularly if the solids are held under anaerobic conditions, which cause PAOs to release phosphorus. This includes, for example, wet wells and gravity sludge thickeners, as well as anaerobic digesters. Interestingly, full-scale experience indicates that of some of the phosphorus released during anaerobic digestion of BPR waste solids can precipitate and be retained with the solids.[54,57,59] Precipitates include struvite ($MgNH_4PO_4$), brushite

($CaHPO_4 \cdot 2H_2O$), and vivianite [$Fe_3(PO_4)_3 \cdot H_2O$]. The process designer must be aware of these potential impacts and balance the requirements of the liquid and solids processing trains to obtain an optimum treatment system.

11.2.3 Composition of Organic Matter in Wastewater

The composition of the organic matter present in a wastewater, particularly its biodegradability, also affects the performance of BNR processes. In the anaerobic zone, PAOs transport short chain VFAs into the cell and store them as PHAs.[79,82] There are two sources of VFAs for the PAOs; they are either present in the influent wastewater or they are produced by fermentation of other readily biodegradable substrate by facultative heterotrophs. The uptake of preformed VFAs is a rapid process, while fermentation is a slower process.[22,77,79] Ideally a wastewater that is to be treated in a BPR system will contain a high proportion of VFAs; this will result in their rapid uptake by the PAOs and a relatively small anaerobic SRT can be used. At a minimum, a sufficiently high concentration of fermentable organic matter must be present to generate VFAs for uptake by the PAOs. It has been estimated that a concentration of at least 25 mg/L as COD of readily biodegradable substrate must be available in the anaerobic zone to generate sufficient VFAs to allow adequate biological phosphorus removal.[62,75] Thus, the readily biodegradable substrate concentration in the influent wastewater, particularly the VFA concentration, will significantly affect the performance of a biological phosphorus removal system.

The readily biodegradable substrate concentration of the influent wastewater will also affect the denitrification rate in an initial anoxic zone. Denitrification is rapid when readily biodegradable substrate is available, but is much slower when only slowly biodegradable substrate is present. This is because the use of slowly biodegradable substrate is controlled by the rate of hydrolysis, which is relatively slow under anoxic conditions. Consequently, if the amount of readily biodegradable substrate entering an initial anoxic zone is insufficient to remove the nitrate-N added, the anoxic zone must be large enough to provide time for the hydrolysis of slowly biodegradable substrate. Hydrolysis and fermentation of slowly biodegradable substrate in an upstream anaerobic zone can produce readily biodegradable organic matter that can pass into an anoxic zone and produce a high rate of denitrification there.[18]

Significant fermentation will occur in some wastewater collection systems, resulting in a wastewater that contains sufficient quantities of readily biodegradable substrate (particularly VFAs) to allow efficient biological phosphorus removal and denitrification. Warm temperatures, low velocities, which minimize reaeration, and force main systems, which maintain the wastewater under anaerobic conditions and in contact with the fermentative bacteria that grow as slimes on the walls of the collection system, provide ideal conditions for fermentation. When this does not occur in the wastewater collection system, the influent wastewater can be treated to convert slowly biodegradable organic matter into a more readily biodegradable form. As discussed in Section 11.1.2, fermentation is a developing technology that can be used to accomplish this conversion. Either the raw wastewater itself can be fermented, or primary solids can be separated and fermented. Solids fermentation is discussed in Chapter 13.

11.2.4 Effluent Total Suspended Solid

The quality of the effluent from a biological wastewater treatment system is determined by the concentrations of soluble and particulate matter in it. Although this is the case for all wastewater treatment systems, it is particularly significant with BPR systems because of the elevated phosphorus content of the MLSS and their effect on the particulate phosphorus in the effluent. As discussed above, the phosphorus content of the MLSS in a BPR system will typically average about 6%, with values as high as 8 to 12% achievable in some cases. In contrast, conventional activated sludge will typically range from 1.5 to 2% on a P/VSS basis. Figure 11.15 illustrates the effect that increasing the phosphorus content of the MLSS can have on the particulate phosphorus concentration in the effluent from a BPR system. It indicates that significant quantities of phosphate can be contributed if effluent TSS concentrations exceed about 10 mg/L. Biological nutrient removal process mixed liquor also contains organic-N, with typical values in the 10 to 12% range on an N/VSS basis. This can amount to 1 to 2 mg/L of nitrogen for effluent TSS concentrations in the 10 to 30 mg/L range. These concentrations are significant in applications where an effluent low in total nitrogen must be produced.

Fortunately, BNR systems generally produce a well flocculated sludge that settles well in the clarifier and produces a clear effluent that is relatively low in suspended solids.[49] Nevertheless, the impact of effluent suspended solids on effluent nitrogen and phosphorus concentrations must be considered carefully when BNR systems are used.

11.2.5 Environmental and Other Factors

A number of environmental factors affect the performance of BNR systems. These factors also affect activated sludge systems, but their impacts can be more significant for BNR systems. The primary impact of temperature is on the kinetics of the various biochemical conversions, and its effect can be predicted quite well using the temperature correction factors described in Section 3.9. In general, temperature has the greatest impact on the nitrifying bacteria, as illustrated in Figure 9.4, but PAOs are also significantly affected.[45] Decreasing temperature will also reduce denitrification rates, resulting in the need for larger anoxic zones and/or in reduced nitrogen removal. Decreasing temperature in the collection system can reduce the rate of fermentation of organic matter and alter the composition of the wastewater entering the biological treatment system.[54,55] This impact is difficult to predict, but provides one of the major reasons that wastewater characterization or pilot plant studies should be conducted over an extended period of time.

Available data indicate that the activity of nitrifying bacteria is significantly reduced as the pH drops below 7.0, as illustrated in Figure 3.4, and that the impact of decreasing pH is much greater for them than for the PAOs or the denitrifying bacteria. Much of those data were collected in batch reactors using unacclimated cultures, but full-scale experience and some laboratory studies using acclimated cultures suggest that nitrifiers can acclimate to lower pHs, in the 6.5 to 7.0 range, with little decrease in activity.[54,68] Many nitrifying activated sludge systems operate quite successfully at pH values in this range, with efficiency dropping off only as the pH drops below 6.5. This potential should be considered in the design and operation of

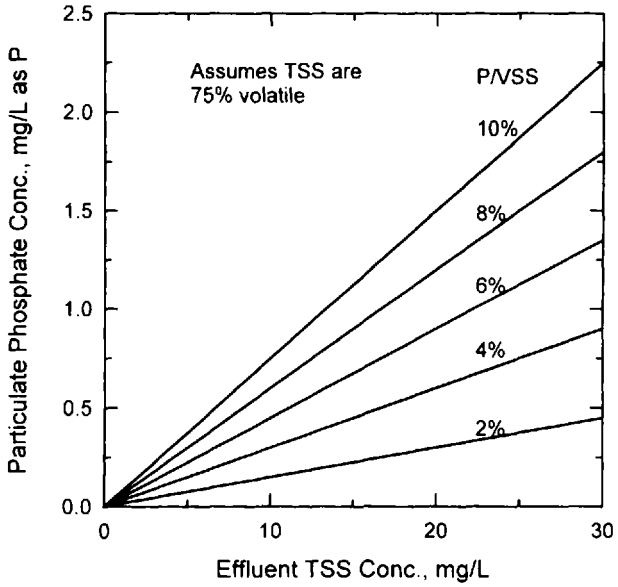


Figure 11.15 Effect of the effluent TSS concentration and the mixed liquor P/VSS ratio on the effluent particulate phosphorus concentration.

BNR systems, and site-specific data should be collected using acclimated cultures if this becomes a significant cost or operational issue.

Dissolved oxygen concentrations affect the rates of nitrification, denitrification, and phosphorus removal in a variety of ways. Enough DO must be present in the aerobic zone to allow growth of nitrifiers and PAOs at adequate rates. In general, the DO requirements for nitrifiers are controlling due to their high half-saturation coefficient for oxygen. Although a DO concentration of 2 mg/L is often specified to obtain efficient nitrification, effluent quality goals can be obtained at lower DO concentrations if aerobic SRTs are sufficiently long. This effect is illustrated in Figure 3.3 where it can be seen that many combinations of DO and ammonia-N concentrations can exist for a specified nitrifier specific growth rate, i.e., for a given system aerobic SRT. At a fixed aerobic SRT, a reduction in the DO concentration will result in an increase in the ammonia-N concentration, but the increase may be small if the aerobic SRT is sufficiently long. Consequently, DO concentrations must be evaluated on a relative basis and adjusted in accordance with system performance requirements. Operation at aerobic zone DO concentrations below 2 mg/L may result in adequate nitrification and phosphorus uptake, while also encouraging additional denitrification. Low DO operation, on the other hand, may encourage the growth of Group IV filamentous bacteria, as discussed previously. The addition of DO to anoxic and anaerobic zones should be minimized because it is used preferentially as a terminal electron acceptor, thereby reducing the amount of readily biodegradable substrate available for denitrification or for uptake by the PAOs. If the quantity of readily biodegradable organic matter in the influent wastewater is high, however, larger oxygen inputs can be tolerated while still producing acceptable process performance because sufficient quantities will still be available for the other needs. Care should

be taken not to introduce too much oxygen, however, because DO concentrations will be low in those situations and significant oxygen inputs can lead to excessive growths of low DO filamentous bacteria. In spite of the effects of oxygen entry into anoxic and anaerobic zones, both are often uncovered and simply rely on low surface turbulence to minimize oxygen transfer rates. As a consequence, care must be exercised in the selection and placement of mixers to minimize surface turbulence. In situations where interfacial oxygen transfer must be kept as low as possible, covers of various construction can be used. Since they need not exclude all air, but merely need to reduce surface transfer, the opportunity exists for the design engineer to be innovative in solving the problem.

The mixing energy provided to anoxic and anaerobic zones must be sufficient to keep the MLSS in suspension, but not so great as to cause significant surface turbulence, as discussed above. This means that the mixers in such zones should be selected with care. It is more difficult to generalize about the volumetric power input required to keep solids in suspension in anoxic and anaerobic zones than in aerobic systems because it depends very heavily on the type and placement of the mixer used. Consequently, mixer selection should be done in close cooperation with a qualified vendor. Nevertheless, as a rough approximation, vertical mixers often require around 12–16 kW/1000 m^3 whereas horizontal mixers require about 8 kW/1000 m^3.[3]

Bioreactor and aerator configurations also influence the environment produced in a BNR system. Both reaction kinetics and organism selection are improved by staging the reactor zones. This can be done either by building separate tanks or by using curtain walls in a single tank. Aerator configuration will influence localized DO concentrations, as discussed previously, which can influence organisms selection and process reaction rates in the aerobic zone.

11.3 PROCESS DESIGN

This section describes procedures and presents illustrative examples for the design of BNR systems. The design of these systems builds upon and utilizes all of the procedures and principles presented in Chapters 9 and 10. The major difference, however, lies in the fact that more interrelated processes are at work in BNR systems, making it impossible to derive accurate analytical expressions for them. As a consequence, the only way to accurately predict the performance of any proposed BNR system design is through simulation with a model like ASM No. 1 or No. 2. Nevertheless, it is very useful to have approximate procedures with which to estimate the required sizes of the various zones to use as starting points for simulations or to use with experience to initiate a design. The procedures and equations presented herein are provided for that purpose. The design process is organized in exactly the same way as activated sludge design, with all of the same constraints and limitations, such as on mixing energy input, oxygen transfer rates, MLSS concentrations, etc. Therefore, to save space and to allow the reader to focus on the unique aspects of BNR process design, that material will not be repeated here. This makes it is imperative for the reader to be thoroughly familiar with the material in Chapters 9 and 10 before proceeding. To provide continuity, the examples in this chapter are based on the same wastewater characteristics and parameter values used for the examples in Chap-

ter 10. The wastewater characteristics are presented in Table E8.4, the kinetic and stoichiometric parameters are presented in Table E8.5, and the temperature correction factors are presented in Table E10.1. In addition to the parameter values given in Table E8.5, for the design of denitrification facilities, information is needed about the values of the anoxic growth and hydrolysis coefficients, η_g and η_h, which were introduced in Section 6.1.2 and Table 6.3. Those values can be readily obtained during treatability studies by using the procedures presented in Section 8.5.3. For the examples in this chapter, η_g is assumed to have a value of 0.8 and η_h is assumed to have a value of 0.4, the default values in ASM No. 1.[33]

The section is organized according to objectives of the BNR process. First we consider processes for the removal of nitrogen, then those for the removal of phosphorus, and finally those that must remove both nitrogen and phosphorus.

11.3.1 Biological Nitrogen Removal Processes

Biological nitrogen removal processes require the oxidation of ammonia-N to nitrate-N through nitrification and the reduction of nitrate-N to nitrogen gas (N_2) through denitrification, thereby removing the nitrogen from the wastewater and transferring it to the atmosphere in an innocuous form. Because of the critical role that nitrification plays in nitrogen removal processes, a key decision during design is the selection of an aerobic SRT that will ensure reliable nitrification year-round. If carbon oxidation and nitrification are to occur in an ordinary activated sludge system, followed by separate stage denitrification, then the activated sludge system SRT will be equal to the aerobic SRT and design of that system proceeds exactly as outlined in Chapter 10. On the other hand, if carbon oxidation, nitrification, and denitrification are to be accomplished in a single-sludge system, then the system SRT will be larger than the aerobic SRT, with the difference being determined by the number of anoxic zones needed and the degree of nitrogen removal required. This increase in system SRT will decrease the excess solids production rate. In addition, the presence of denitrification in the system will decrease the oxygen requirement and the net destruction of alkalinity. In this section, we will first review the requirements for nitrification, and then we will examine the additional requirements for design of an anoxic selector, an MLE system, a four-stage Bardenpho system, and a separate-stage denitrification system.

Nitrification. The factors that must be considered in the selection of the aerobic SRT required to achieve stable nitrification are covered in Sections 10.2.2 and 10.3.2. For a fully aerobic system to be followed by separate-stage denitrification, the process design proceeds from there exactly as outlined in Section 10.3. For the other systems, the total system SRT must first be determined by adding the required anoxic SRT to the aerobic SRT, and then adjustments must be made for denitrification, etc. In this section, we consider design of a fully aerobic system.

Because nitrification, which destroys alkalinity, is required in all nitrogen removal systems and the activity of nitrifying bacteria is sensitive to pH, consideration must be given to the need for chemical addition for pH control. In general, a residual alkalinity of 1 mM (50 mg/L as $CaCO_3$) will result in an adequate pH value. Thus, the need for chemical addition will depend on the influent alkalinity and the amount of ammonia-N oxidized. As shown in Eq. 3.30, 6.71 g HCO_3^- are removed for each g of NH_4^+ removed. This corresponds to 7.07 g of alkalinity (expressed as $CaCO_3$)

per g of ammonia-N removed by nitrifying bacteria. Because some of the ammonia-N removed is incorporated into the nitrifying biomass (Eq. 3.30), only 0.98 g nitrate-N is formed for each g of ammonia-N removed. Consequently, the amount of alkalinity destroyed corresponds to 7.23 g of alkalinity (as $CaCO_3$) per g nitrate-N formed. This latter value is sometimes useful when determining the alkalinity destruction in a system performing both carbon oxidation and nitrification because ammonia-N is also removed for incorporation into heterotrophic biomass, making it simpler to determine the alkalinity requirement from the amount of nitrate-N formed. The concentration of nitrate-N formed, S_{NO}, is determined by the difference between the concentration of nitrogen available to the nitrifiers, $S_{N,a}$, as given by Eq. 10.17, and the effluent soluble nitrogen concentration:

$$S_{NO} = 0.98(S_{N,a} - S_{NH} - S_{NS}) \tag{11.5}$$

where the factor 0.98 is the mass of nitrate-N formed per mass of ammonia-N oxidized, as discussed above. The effluent ammonia-N concentration, S_{NH}, can be calculated with Eq. 10.18 for a CMAS system, but will generally be lower for other systems, as discussed in Section 10.3.2 and illustrated in Figure 10.17. If the bioreactor has plug–flow characteristics, S_{NH} will often be negligible. The effluent soluble biodegradable organic nitrogen concentration must either be measured directly or estimated by assuming that the ratio of soluble biodegradable organic nitrogen to soluble biodegradable organic matter in the effluent is the same as the ratio in the influent. For domestic wastewater, this value is often assumed to be negligible. It should be noted that the concentration of nitrogen available to the nitrifiers, $S_{N,a}$, will change with temperature because the heterotrophic nitrogen requirement, as given by Eq. 5.36, changes with temperature because of its influence on the decay coefficient, b_H. Consequently, the value of $S_{N,a}$ will be slightly larger in summer, making the effluent nitrate-N concentration, the amount of alkalinity destroyed, and the amount of oxygen required slightly higher then.

If a nitrifying system is to be configured as a CAS system, in which the oxygen requirement must be distributed along the bioreactor, special consideration must be given to the distribution of the autotrophic oxygen requirement. During the winter, when temperatures are low, the maximum nitrification rate will be low and nitrification will occur throughout a greater proportion of the bioreactor than in the summer, when the maximum nitrification rate is higher. The required distribution under each condition can be determined by applying the procedures of Section 10.3.4 as illustrated in Example 10.3.4.1. Because two different oxygen transfer systems would be needed to provide the oxygen according to both distributions, thereby increasing the cost of the facility, careful consideration must be given as to whether the expense is justified. During the summer, if the oxygen supply is not limiting, nitrification will generally be complete within the first half (or less) of the system, even during diurnal loads. However, if oxygen were supplied so as to meet the winter distribution (but in total amount sufficient to meet the higher summer requirement), the result would be a decreased DO concentration in the first part of the bioreactor. Because of the sensitivity of nitrification to the DO concentration, this would decrease the maximum nitrification rate in that part of the bioreactor, increasing the ammonia loading on the latter part, and forcing more of the oxygen requirement to it, where oxygen is available. As long as the DO concentration in the first part of the system is maintained sufficiently high to avoid problems with low DO filaments, such a procedure

is entirely acceptable. Consequently, many designers specify only a single oxygen transfer system.

The examples presented in Chapter 10 have considered all of the aspects of the design of a system for combined carbon oxidation and nitrification, except for the determination of the amount of nitrate-N produced and the need for pH control. The following example illustrates those procedures. In addition, it presents the determination of the required SRT and results from determination of the steady-state oxygen requirement, solids wastage rate, and required bioreactor volume because they will be used as baselines against which to make comparisons for the remainder of the examples in this section. As in the examples in Chapter 10, all organic substrate concentrations are given as biodegradable COD and all suspended matter concentrations are as TSS, unless noted otherwise.

Example 11.3.1.1

Consider the wastewater used for all examples in Chapter 10, with characteristics given in Table E8.4, kinetic and stoichiometric parameters in Table E8.5, and temperature correction factors in Table E10.1. As in Chapter 10, the average wastewater flow rate is 40,000 m^3/day and the diurnal peak pollutant loadings are 2.5 times the average. The characterization of the wastewater is considered to be sufficient to allow the safety factor for uncertainty, s_U, to be set to 1.0. Determine the major design characteristics of an activated sludge system to perform both carbon oxidation and nitrification year round, while maintaining an MLSS concentration of 3,000 mg/L and a DO concentration of 2.0 mg/L. A conventional bioreactor with hydraulic characteristics equivalent to four tanks in series is to be used.

a. What is the design SRT?
The design SRT should be chosen for winter operating conditions. The approach used is exactly the same as in Example 10.3.2.1, except that the peak load safety factor, s_{PL}, is 2.5. Since a CAS system is to be used, Eq. 10.11 is the appropriate expression with which to calculate the required SRT. The value of the minimum SRT can be obtained from Figure 9.4, which gives a value of 3.5 days at a temperature of 15°C. The safety factor for uncertainty, s_U, has a value of 1.0. Because the half-saturation coefficient for DO for nitrifiers is 0.75 mg/L (Table E8.5), the DO safety factor can be calculated with Eq. 10.9, giving a value of 1.375. Thus,

$$\Theta_{x,r} = (3.5)(1.0)(2.5)(1.375) = 12.0 \text{ days}$$

b. What concentration of nitrate-N will be formed?
This will be calculated for summer conditions since it will be slightly larger then because more nitrogen will be available to the nitrifiers. The concentration of nitrate-N formed can be calculated with Eq. 11.5. Since the bioreactor has hydraulic characteristics equivalent to four tanks in series, the concentrations of ammonia-N and soluble organic-N in the effluent can be considered to be negligibly small, even in winter. Thus, the concentration of nitrate-N formed will just be 98% of the concentration of nitrogen available to the nitrifiers, as calculated with Eq. 10.17. Use of Eq. 10.17 requires knowledge of the nitrogen requirement of the heterotrophs, which can be calculated with Eq. 5.36. This calculation was illustrated in Part c of Example 10.3.3.2 and will not be repeated here. However, it should be noted that the heterotrophic

nitrogen requirement for the SRT of 12 days is 0.022 mg N used/mg COD removed at 25°C, which is lower than the value in Example 10.3.3.2 because the longer SRT results in less net biomass synthesis. Therefore, the available nitrogen concentration is:

$$S_{NH} = 25 + 6.5 + 8.5 - 0.022(115 + 150)$$
$$= 34.2 \text{ mg/L as N}$$

Thus, the amount of nitrate-N formed in summer will be (0.98)(34.2) = 33.5 mg/L. Slightly less will be formed in winter because the heterotrophic nitrogen requirement will be slightly higher because of the decreased decay coefficient.

c. Will it be necessary to provide alkalinity to control pH in the process?
The wastewater contains 200 mg/L as $CaCO_3$ of alkalinity (Table E8.4) and the residual alkalinity must be 50 mg/L as $CaCO_3$ to provide a stable pH. Thus, the alkalinity available to neutralize the hydrogen ions formed during nitrification is 150 mg/L as $CaCO_3$. As noted above, nitrification destroys 7.23 g of alkalinity (as $CaCO_3$) per g nitrate-N formed. Since 33.5 mg/L of nitrate-N are formed:

$$\text{Alkalinity destroyed} = (7.23)(33.5) = 242 \text{ mg/L as } CaCO_3$$

This exceeds the alkalinity available by 92 mg/L. Thus, 92 mg/L of alkalinity as $CaCO_3$ must be provided to the process through chemical addition in the summer. The amount required in winter will be slightly less because slightly less nitrate-N will be formed.

d. What is the required bioreactor size?
As in other cases, the bioreactor size must be calculated for winter conditions because that is when solids production is highest. The first task is to calculate the mass of MLSS in the system at 15°C with Eq. 9.11, as illustrated in Part b of Example 10.3.3.3. The mass of autotrophic nitrifiers can be neglected because it is small, as seen in Chapter 10. The only difference in the calculation is the use of an SRT of 12 days. The resulting mass of MLSS is 51,800,000 g.

The MLSS concentration for this design is 3,000 mg/L, which is a higher value than was possible in the Examples in Chapter 10. This is permissible here because once nitrification is fully established, the mass of MLSS in the system increases faster than the oxygen requirement as the SRT is increased. Thus, the constraints due to oxygen transfer rates are encountered at higher MLSS concentrations. For an MLSS concentration of 3,000 mg/L (g/m^3), the bioreactor volume will be:

$$V = \frac{51{,}800{,}000}{3{,}000} = 17{,}300 \text{ m}^3$$

e. What is the average process oxygen requirement for design of the oxygen transfer system?
This is determined at the highest sustained wastewater temperature of 25°C because that is when the oxygen requirement is maximum. This requires calculation of the heterotrophic oxygen requirement with Eq. 9.13 and the autotrophic oxygen requirement with Eq. 10.16. The procedure was illustrated in Example 10.3.3.2, so it will not be repeated here. However, appli-

cation of the procedure results in a heterotrophic oxygen requirement (RO_H) of 7,930 kg O_2/day and an autotrophic oxygen requirement (RO_A) of 6,080 kg O_2/day, for a total requirement (RO) of 14,010 kg O_2/day.

h. What is the excess solids production rate?
Because the maximum solids production occurs in the winter, this is calculated at 15°C using Eq. 9.12, as illustrated in Example 10.3.3.4. Application of the equation for the conditions of this design gives a value for W_{SLE} of 4,310 kg TSS/day.

Design of an Anoxic Selector. An anoxic selector is an initial anoxic zone similar to that used in the MLE (Figure 11.4) and four-stage Bardenpho processes (Figure 11.5). As in those processes, it receives the influent wastewater and recirculated flow from a downstream aerobic zone where nitrification occurs. The influent wastewater contains biodegradable organic matter, which serves as the electron donor for the growth of heterotrophic denitrifying bacteria, whereas the recirculated flow from the aerobic zone provides nitrate-N, which serves as the electron acceptor for those bacteria. Unlike the initial anoxic zone in the MLE and four-stage Bardenpho processes, however, the primary purpose of an anoxic selector is the control of solids settleability. As discussed in Section 10.2.1, many filamentous bacteria use readily biodegradable organic matter very efficiently, but most are not able to use nitrate-N as an electron acceptor. Consequently, if the readily biodegradable organic matter in a wastewater is removed in an anoxic zone, the growth of many types of filamentous bacteria is prevented. Moreover, since readily biodegradable organic matter can be removed easily, a relatively small anoxic zone can be used. Initial anoxic zones designed to control the growth of filamentous bacteria are called anoxic selectors because their primary purpose is to select for floc forming bacteria and to selectively prevent the growth of filamentous bacteria.

Just as with aerobic selectors, the purpose of an anoxic selector is to remove the bulk of the readily biodegradable substrate so that little passes through to the aerobic zone where filamentous bacteria have a competitive advantage because of the long SRT and associated low substrate concentration. Few guidelines are available concerning the readily biodegradable substrate concentration that should leave the selector, so it is difficult to specify the specific growth rate that should be maintained there. Rather, most design experience has resulted from the application of a particular process loading factor, as has been done with aerobic selectors. Values that have been used successfully are 0.8 to 1.2 kg BOD_5/(kg MLSS·day) for operation at temperatures above about 20°C and 0.7 to 1.0 kg BOD_5/(kg MLSS·day) for operation at temperatures below about 17°C.[47] These correspond to biodegradable COD loadings of approximately 1.4 to 2.1 kg COD/(kg MLSS·day) and 1.2 to 1.7 kg COD/(kg MLSS·day), respectively. Thus, design of an anoxic selector could be approached in the same way as design of an aerobic selector, as outlined in Section 10.3.4, except that a single mixed basin is often used. However, since the total SRT is not known at the outset, an iterative procedure is required. This can be avoided if the anoxic SRT can be specified as the design criterion. Reexamination of the information leading to the process loading factors given above reveals that they correspond to anoxic SRTs of about 1.0 day at temperatures over about 20°C and of about 1.5 days at temperatures less than about 17°C. Thus, in the absence of site-specific data, these SRTs can be used to perform a preliminary process design.

The determination of the bioreactor size proceeds in the same manner as for activated sludge systems as discussed in Chapter 10. The anoxic SRT is added to the aerobic SRT, giving the total SRT to be used for the design. The mass of MLSS in the system is then calculated using the total SRT in Eq. 9.11, and it is apportioned to the anoxic and aerobic zones in proportion to the anoxic and aerobic zone SRTs. Since the MLSS concentration is uniform in this system, the total system volume can be calculated after deciding upon a MLSS concentration that will give a reasonable settler size, as discussed in Chapters 9 and 10. Finally, that volume can be allocated to the two zones in proportion to their respective SRTs, as suggested by Eqs. 11.1 and 11.2 for systems with uniform MLSS concentrations.

Once the anoxic SRT has been selected, the next task is to determine the amount of nitrate-N that must be recirculated to the anoxic zone by the solids recycle and MLR flows. The required quantity must be at least equal to the amount of electron acceptor that is needed for biomass synthesis and decay. Because this will be greatest at higher temperature, the computation should be made for summer conditions. As discussed in Section 6.4.2, the amount of electron donor required to reduce a given amount of nitrate-N to nitrogen gas ($\Delta S/\Delta N$) is a function of the SRT in the anoxic zone and it can be calculated using Eq. 6.4 for situations in which no oxygen enters the system. Since oxygen is used preferentially over nitrate-N as an electron acceptor, the oxygen equivalents of nitrate-N required will be decreased in direct proportion to the amount of oxygen entering the bioreactor. Even though all of the design equations assume no oxygen leakage, this effect should be kept in mind. During selection of the RAS and MLR flows we are interested in the amount of nitrate-N required, so the equation should be inverted. Making this modification and using only the anoxic SRT in the equation gives the expression for the amount of nitrate-N required to remove a given amount of COD:

$$\frac{\Delta N}{\Delta S} = \frac{1 + (b_H \cdot \Theta_{c,ANX}) - (Y_{H,T})(i_{O\,XB,T})(1 + f_D \cdot b_H \cdot \Theta_{c,ANX})}{2.86[1 + (b_H \cdot \Theta_{c,ANX})]} \tag{11.6}$$

The larger the amount of COD removed in the anoxic selector, the larger the amount of nitrate-N that must be provided, and the larger the MLR flow rate must be. Since the anoxic selector has been sized to remove the bulk of the readily biodegradable organic matter, the least acceptable MLR flow rate will result when ΔS is taken to be the mass flow rate of readily biodegradable substrate into the system, $F \cdot S_{SO}$. It is likely, however, that some of the slowly biodegradable substrate will also be oxidized, with the amount depending on the SRT of the system and the nitrate-N concentration, as indicated in process 7 of Table 6.1. Hydrolysis of slowly biodegradable organic matter under anoxic conditions is thought to be slower than aerobic hydrolysis, as reflected by the fact that a typical value of η_h, the anoxic hydrolysis factor, is 0.4, as given in Table 6.3. Because of this and because the kinetics of hydrolysis are still not well defined, it is difficult to predict with certainty the degree of hydrolysis that will occur in an anoxic selector. Furthermore, as discussed previously, it is difficult to know exactly how much oxygen will enter the anoxic zone. Since hydrolysis of slowly biodegradable substrate increases the nitrate-N requirement over the amount calculated with Eq. 11.6 and oxygen entry decreases it, and since both are hard to predict, the recommended procedure is to limit ΔS in Eq. 11.6 to readily biodegradable substrate and to consider the resulting MLR flow rate to be

the minimum allowable value. A value in slight excess of that can be adopted for the design, depending on the experiences and preferences of the design engineer.

Because the concentration of nitrate-N in the MLR stream depends on several factors, it is easiest to determine the MLR flow rate by considering the fraction of nitrate-N formed that must be denitrified. In the absence of denitrification, the concentration of nitrate-N in the process effluent would be given by Eq. 11.5. Multiplication of that concentration by the flow rate through the system, F, gives the mass formation rate of nitrate-N in the system. Therefore, since ΔN represents the mass of nitrate-N that is denitrified, the fraction of the nitrate-N denitrified in the initial anoxic zone, $f_{NO,D}$, is:

$$f_{NO,D} = \frac{\Delta N}{0.98F(S_{N,a} - S_{NH} - S_{NS})} \tag{11.7}$$

Furthermore, for the flow diagram depicted in Figure 11.4, the fraction of the nitrate-N produced in the aerobic zone that is recirculated to the anoxic zone ($f_{NO,R}$) can be calculated by simple mass balance. The result is:

$$f_{NO,R} = \frac{\alpha + \beta}{1 + \alpha + \beta} \tag{11.8}$$

where β is the MLR flow rate expressed as a ratio of the process influent flow rate and α is the RAS flow rate, also expressed as a ratio of the process influent flow rate. The minimum allowable MLR recirculation rate results when the fraction of nitrate-N recirculated is equal to the fraction denitrified. Higher recirculation rates will just result in higher nitrate-N concentrations in the anoxic zone. Substitution of $f_{NO,D}$ for $f_{NO,R}$ in Eq. 11.8 and rearrangement gives an expression that can be used to calculate the MLR and RAS flow rate ratios to achieve denitrification of a specified fraction of the nitrate-N formed:

$$\alpha + \beta = \frac{f_{NO,D}}{1 - f_{NO,D}} \tag{11.9}$$

While solids recycle from the final clarifier is done by pumping through a pipe, this may not be necessary for MLR, depending on the system configuration. If the exit from the aerobic zone is located some distance from the anoxic tank, then MLR is usually pumped through a pipe. On the other hand, if the anoxic and aerobic zones share a common wall, then it may be possible to recirculate mixed liquor by using a propeller pump in an opening in the common wall. Such pumps are similar to the horizontal mixer shown in Figure 11.3b. They are low head devices and work with lower power requirements than centrifugal pumps.

As discussed in Section 6.4.1, and shown in Table 6.1, approximately $(1 - Y_H)/14$ moles of alkalinity are produced for each gram of nitrate-N reduced to N_2. For a typical Y_H value, this amounts to about 3.5 g of alkalinity (as $CaCO_3$) per g of nitrate-N reduced. The production of this alkalinity in the anoxic zone helps to off-set somewhat the destruction of alkalinity due to nitrification in the aerobic zone. This can be another important benefit from using an anoxic selector. There will be a slight difference in the alkalinity recovery under summer and winter conditions, just as there will be for alkalinity destruction. Generally, however, the estimation is not precise enough to warrant recalculating it for both conditions. Either can be used to give a general idea.

The oxidation of biodegradable organic matter in the anoxic zone decreases the amount of substrate entering the aerobic zone, causing a corresponding decrease in the oxygen requirement. The decrease in the aerobic zone oxygen requirement can be calculated using one of two general procedures. In one, the oxygen requirement associated with the synthesis of biomass in the aerobic zone is reduced by the amount of substrate that is oxidized in the anoxic zone. It should be noted, however, that the oxygen requirement associated with decay of the biomass resulting from that substrate is not reduced. In the other, the nitrate-N reduced in the anoxic zone is converted to its oxygen equivalent (using the stoichiometric factor −2.86 g COD/g N for nitrate-N as shown in Table 3.1) and subtracted from the aerobic zone oxygen requirement calculated assuming that the anoxic zone is not present. In making this computation, the heterotrophic oxygen requirement should be calculated as if the total SRT were aerobic to account for the added decay, but the autotrophic requirement should be calculated only for the aerobic SRT. The two procedures will arrive at the same answer.

The procedure outlined above for the design of an anoxic selector for the removal of readily biodegradable substrate is approximate because of uncertainties about the amount of slowly biodegradable substrate used in the anoxic tank and the amount of oxygen leakage into the anoxic zone. In addition, it contains several approximations. Nevertheless, it should give sufficiently accurate estimates to allow a pilot study to be designed or to provide starting conditions for a series of simulations to refine the design. Its use is illustrated below.

Example 11.3.1.2

An anoxic selector is to be added to the nitrifying activated sludge system designed in Example 11.3.1.1. The purpose of this selector is to remove the readily biodegradable organic matter so that solids with good settling properties will be produced. Determine the approximate requirements for the system.

a. What size should the anoxic selector be?

As discussed above, experience suggests that the readily biodegradable organic matter can be removed from a typical domestic wastewater under cold weather conditions using an SRT for the anoxic zone of 1.5 days. Thus, since the aerobic SRT is 12.0 days, the total SRT will be 13.5 days. Using Eq. 9.11 to recompute the mass of MLSS in the system at 15°C for this slightly larger SRT gives a value of 56,700,000 g. At the design MLSS concentration of 3,000 mg/L (g/m^3), the total bioreactor volume will be

$$V = \frac{56{,}700{,}000}{3{,}000} = 18{,}900 \text{ m}^3$$

Because the MLSS concentration is uniform, the volume of the anoxic zone is equal to the total bioreactor volume times the fraction of the total SRT associated with the anoxic zone, as suggested by Eqs. 11.1 and 11.2. Thus:

$$V_{ANX} = 18{,}900 \left(\frac{1.5}{13.5}\right) = 2{,}100 \text{ m}^3$$

By difference

$$V_{AER} = 18{,}900 - 2{,}100 = 16{,}800 \text{ m}^3$$

In comparison to Example 11.3.1.1, the aerobic zone is slightly smaller due to the slightly larger total SRT, which results in additional decay.

b. What should the MLR flow rate be?
First, the mass of nitrate-N required to accept the electrons from use of the readily biodegradable substrate at the anoxic SRT is calculated using Eq. 11.6. This calculation is performed at 25°C as it represents the greatest value and will dictate the MLR pump sizing. Using the values for the various parameters from Table E10.2 and an anoxic SRT of 1.5 days gives:

$$\frac{\Delta N}{\Delta S} = \frac{1 + (0.22)(1.5) - (0.5)(1.2)[1 + (0.2)(0.22)(1.5)]}{(2.86)[1 + (0.22)(1.5)]}$$

$$= 0.182 \text{ g nitrate-N/g COD}$$

Since the concentration of readily biodegradable organic matter in the influent wastewater is 115 mg COD/L (g/m^3) and the influent flow is 40,000 m^3/day, the mass of nitrate-N required is

$$\Delta N = (0.182)(115)(40{,}000) = 837{,}000 \text{ g nitrate-N/day}$$

The fraction of nitrate-N denitrified in the anoxic selector can be calculated with Eq. 11.7. The value of available nitrogen, $S_{N,a}$, should be recalculated for 25°C and an SRT of 13.5 days because the longer SRT will reduce the nitrogen requirement for the heterotrophs. Following the procedure in Examples 10.3.3.2 and 11.3.1.1 gives a value of 0.021 mg N used/mg COD removed for the heterotrophic nitrogen requirement, which gives a value for $S_{N,a}$ of 34.4 mg/L as N. Substituting this into Eq. 11.7 and assuming that the effluent concentrations of ammonia-N and soluble organic-N are negligible, gives:

$$f_{NO,D} = \frac{837{,}000}{(0.98)(40{,}000)(34.4)} = 0.62$$

The sum of the recycle and recirculation ratios can now be calculated with Eq. 11.9:

$$\alpha + \beta = \frac{0.62}{1 - 0.62} = 1.63$$

Because some slowly biodegradable organic matter may be hydrolyzed and oxidized in the anoxic zone, the sum of the selected ratios should exceed 1.63, with higher values being used for systems that exclude more oxygen. Assuming an anoxic zone with little oxygen entry, use a value of 1.5 for β, the MLR ratio, which, when combined with typical values for α of 0.5 to 1.0, will provide a conservatively high recirculation of nitrate-N to the anoxic zone. The MLR flow rate will thus be (1.5) (40,000 m^3/day) or 60,000 m^3/day.

c. What will the effluent nitrate-N concentration be at 25°C?
Since the available nitrogen concentration is 34.4 mg/L as N, the concentration of nitrate-N formed is (0.98)(34.4), or 33.7 mg/L as N. At least 62% of this will be destroyed by denitrification, leaving an effluent concentration of less than 12.8 mg/L of nitrate-N. The exact value will depend on the amount of slowly biodegradable substrate used in the anoxic selector and the amount of oxygen entering it.

d. What is the effect of denitrification on the alkalinity requirement at 25°C?
Using the logic of Part c of Example 11.3.1.1, production of 33.7 mg/L of nitrate-N will destroy 244 mg/L as $CaCO_3$ of alkalinity. Since 150 mg/L of alkalinity are available in the wastewater, it would be necessary to add 94 mg/L in the absence of denitrification. However, destruction of 20.9 mg/L of nitrate-N by denitrification produces (3.5)(20.9) = 73 mg/L as $CaCO_3$ of alkalinity, thereby reducing the amount of alkalinity that must be provided by chemical addition to 21 mg/L.

e. By what percentage is the average oxygen requirement in the aerobic zone decreased by the presence of the anoxic selector at 25°C?
The reduction in the oxygen requirement in the aerobic zone can be approximated by subtracting the oxygen equivalents of the amount of nitrate reduced from the oxygen requirement for a totally aerobic system, as calculated with an SRT of 13.5 days for the heterotrophs and 12.0 days for the autotrophs. The autotrophic oxygen requirement, RO_A, will be slightly larger than calculated in Part e of Example 11.3.1.1 because the available nitrogen concentration is slightly higher due to the reduced nitrogen requirement for the heterotrophs associated with the additional SRT. Using a value of 34.4 mg/L for $S_{N,a}$ and an SRT of 12 days in Eq. 10.16 gives an autotrophic requirement of 6,120 kg O_2/day. Likewise, using an SRT of 13.5 days in Eq. 9.13 gives a value of 8,050 kg O_2/day for the heterotrophic requirement. Thus, the total requirement would be 14,170 kg O_2/day, from which would be subtracted the mass removal of nitrate-N, expressed as oxygen equivalents. From Part b above, the mass of nitrate-N reduced is 837 kg/day. This is equivalent to (2.86)(837) = 2,390 kg O_2/day. Thus, the total oxygen requirement is 11,780 kg O_2/day, which is 84% of the oxygen requirement in the totally aerobic system of Example 11.3.1.1.

f. What is the WAS production rate?
This is calculated at 15°C as illustrated in Part h of Example 11.3.1.1 using Eq. 9.12 and an SRT of 13.5 days. The result is 4,200 kg TSS/day, which is 97% of the quantity to be disposed of from the totally aerobic system of Example 11.3.1.1.

Design of a Modified Ludzack–Ettinger System to Achieve a Desired Effluent Nitrate-N Concentration. The design of an MLE system to achieve a particular effluent nitrate-N concentration usually involves degradation of some of the slowly biodegradable substrate in the anoxic zone. Because of the interactions between the reaction rates and various concentrations in the system, the only way to accurately predict the performance of such a system is by simulation with a model like ASM No. 1. Various anoxic SRTs and MLR rates can be tried in a systematic manner until a suitable design is developed. In the absence of complete kinetic data, or when a preliminary design is needed as input to a simulation program, an alternative approach is required, just as it was for an anoxic selector. Although the preceding discussion deals with the design of anoxic selectors for the removal of readily biodegradable substrate, the same general procedure can be used when a particular effluent nitrate-N concentration must be achieved. The primary difference is that the initial anoxic zone must be sized to remove a specified mass of nitrate-N rather than a specific mass of biodegradable organic matter.

The estimation of the organic substrate utilization and associated nitrate-N reduction for an MLE system requires two steps, one for readily biodegradable substrate and one for slowly biodegradable substrate. First, because the anoxic SRT usually exceeds 1.5 days, we can assume that all readily biodegradable substrate will be used, even in the winter. Thus, we can calculate the nitrate-N reduction associated with that activity by using a $\Delta N/\Delta S$ value. However, because the approach we will use to estimate the nitrate-N reduction associated with slowly biodegradable substrate utilization includes the effects of biomass decay, the reduction of nitrate-N associated with the removal of readily biodegradable substrate should consider only the electron acceptor requirement for biomass synthesis. Using Eq. 10.23 as a guide, the nitrate-N utilization for this case, ΔN_{SS}, is:

$$\Delta N_{SS} = \Delta S \left[\frac{1 - (Y_{H,1})(i_{O\,XB,1})}{2.86} \right] \tag{11.10}$$

in which ΔS is the mass rate of readily biodegradable substrate entry to the system. In other words, Eq. 11.10 came from eliminating all of the decay-associated terms in Eq. 11.6. Next, an empirical expression that has been widely used to size initial anoxic zones[5,7,68] can be used to estimate the nitrate-N utilization associated with biodegradation of slowly biodegradable substrate and biomass decay. This expression was developed with domestic wastewaters that contained high proportions of slowly biodegradable organic matter,[15] and is appropriate for estimating the nitrate-N utilization associated with hydrolysis and oxidation of that type of substrate in this context. It allows direct calculation of the specific nitrate-N utilization rate, $q_{NO\,XS}$, as a function of the process loading factor for the anoxic zone, U_{ANX}, when that loading factor is expressed in terms of slowly biodegradable substrate alone. In other words, since the effects of readily biodegradable substrate utilization are calculated with Eq. 11.10, readily biodegradable substrate should not be included in U_{ANX}. The expression is:

$$q_{NO\,XS} = 0.018U_{ANX} + 0.029 \tag{11.11}$$

The units of the specific nitrate-N utilization rate are mg NO_x-N/(mg MLSS · day). The units of the process loading factor in Eq. 11.11 are mg COD/(mg MLSS · day). In the original equation given by Burdick et al.,[15] the slope was 0.030 and the units of the process loading factor were mg BOD_5/(mg MLSS · day). For consistency within this text, Eq. 11.11 is expressed in biodegradable COD units by using Eq. 8.33 to convert BOD_5 to equivalent biodegradable COD. The relationship of Burdick et al.[15] was developed at 20°C and $q_{NO\,XS}$ can be adjusted to other temperatures by using Eq. 3.95 with an appropriate value of the temperature coefficient, θ, in the range of 1.04 to 1.08. It was developed using domestic wastewater and should not be used for a wastewater containing a high percentage of industrial wastes. In order to use Eq. 11.11, the mass of MLSS in the anoxic zone must be known. This requires use of an iterative approach to size the anoxic zone for an MLE system.

The approach to preparing a preliminary design of an MLE system is as follows. The selection of the anoxic SRT must be done for winter temperatures because that is when denitrification rates are slowest. First, an initial value for the anoxic SRT must be selected, and added to the aerobic SRT to give the total. The total SRT is then used to calculate the nitrogen available to the nitrifiers using Eq. 10.17, which in turn is used to calculate the mass rate of nitrate-N formation by multiplying Eq.

11.5 by the influent flow rate, F. The mass rate of nitrate-N reduction by biomass synthesis on readily biodegradable substrate, ΔN_{SS}, can be estimated with Eq. 11.10 from the known mass input rate of readily biodegradable substrate for ΔS. Using the total SRT, the mass of MLSS in the system, $(X_{M,1} \cdot V)_{System}$, is calculated with Eq. 9.11. The mass of MLSS in the anoxic zone can then be calculated with Eq. 11.2. Once it is known, the process loading factor on the anoxic zone due to slowly biodegradable substrate can be calculated from its definition, Eq. 5.37, in which slowly biodegradable substrate is the only substrate considered. This allows estimation of the specific denitrification rate in the anoxic zone with Eq. 11.11. Multiplication of it by the mass of MLSS in the anoxic zone gives the mass rate of nitrate-N reduction by slowly biodegradable substrate and decay, ΔN_{XS}. Summing ΔN_{SS} and ΔN_{XS} gives the total mass rate of nitrate-N reduction. Subtraction of this value from the mass production rate of nitrate-N gives the mass rate at which nitrate-N is leaving the process, from which the effluent nitrate-N concentration can be calculated by dividing by the flow rate. Repetition of these calculations for a number of anoxic SRTs provides information on the effluent nitrate-N concentration as a function of anoxic SRT, from which the anoxic SRT can be selected. Once that is fixed, the system volume is determined by considering the mass of MLSS in the system under winter temperature conditions and the design MLSS concentration, which is uniform throughout. The design MLR rate can be calculated with Eq. 11.9 for the summer temperature, since it will give the largest value. However, as mentioned previously, the need for chemical addition to control pH can be calculated for either winter or summer temperatures, because it is not precise enough to distinguish between the needs under the two conditions. The rest of the design can be accomplished exactly as in previous cases, with summer conditions determining oxygen requirements and winter conditions excess solids production. This procedure is illustrated in the following example.

Example 11.3.1.3

The process design developed in Example 11.3.1.2 is being revised to produce an effluent with a nitrate-N concentration of 6 mg/L, or less. Determine the approximate requirements for the system. Assume that the temperature coefficient for the specific denitrification rate is 1.05.

a. What anoxic zone SRT is required for this application?

The determination of the anoxic SRT must be made at 15°C since the effluent goals must be met under all conditions and low temperatures lead to the lowest rates of denitrification. Since the effluent nitrate-N concentration in Example 11.3.1.2 was 12.8 mg/L as N when the anoxic SRT was 1.5 days, and since anoxic hydrolysis of slowly biodegradable substrate is slow, it is likely to require a considerably longer anoxic SRT to achieve an effluent nitrate-N concentration of 6.0 mg/L as N or less. Thus, select an anoxic SRT of 4 days as an initial guess. This will give a total system SRT of 16.0 days since the aerobic SRT is 12.0 days.

The first task is to calculate the concentration of nitrogen available to the nitrifiers when the SRT is 16.0 days. This is done exactly as in Part b of Example 11.3.1.1, except that it must be done for a temperature of 15°C, giving a value of 34.0 mg/L.

The second task is to estimate the mass rate of nitrate-N production.

This is done by multiplying Eq. 11.5 by the influent flow rate. Because of the configuration of the aerobic zone, we can consider the effluent ammonia-N and organic-N concentrations to be negligible. Thus, in the absence of denitrification, the effluent nitrate-N concentration would be 33.3 mg/L as N and the mass rate of nitrate-N production would be 1,332,000 g/day.

The mass rate of nitrate-N reduction by heterotrophic biomass synthesis on readily biodegradable substrate, ΔN_{SS}, can be calculated with Eq. 11.10 by making use of the fact that ΔS is the mass input rate of readily biodegradable substrate, or (115 g COD/m^3)(40,000 m^3/day):

$$\Delta N_{SS} = (115)(40{,}000)\left[\frac{1 - (0.50)(1.20)}{2.86}\right] = 643{,}000 \text{ g/day}$$

Next, the mass of MLSS in the anoxic zone must be estimated to allow the specific denitrification rate in the zone to be approximated. The total mass of MLSS in the system at 15°C can be estimated with Eq. 9.11 for an SRT of 16 days, giving a value of 64,750,000 g. The mass of MLSS in the anoxic zone is in proportion to the fraction of the SRT in the anoxic zone, as given by Eq. 11.2:

$$(X_{M,t} \cdot V)_{ANX} = 64{,}750{,}000\left(\frac{4.0}{16.0}\right) = 16{,}190{,}000 \text{ g}$$

This can be used to calculate the process loading factor for addition of slowly biodegradable substrate to the anoxic zone. Since the concentration of slowly biodegradable substrate in the influent is 150 mg/L (g COD/m^3) and the flow rate is 40,000 m^3/day, the process loading factor is:

$$U_{ANX} = \frac{(150)(40{,}000)}{16{,}190{,}000} = 0.37 \text{ g COD/(g MLSS} \cdot \text{day)}$$

Substitution of the anoxic process loading factor into Eq. 11.11 allows calculation of the specific denitrification rate in the anoxic zone for 20°C:

$$q_{NO\,XS,20} = (0.018)(0.37) + 0.029 = 0.036 \text{ g } NO_3\text{-N/(g MLSS} \cdot \text{day)}$$

Since the design is being performed for 15°C, the specific denitrification rate must be corrected to that temperature using Eq. 3.95:

$$q_{NO\,XS,15} = (0.036)(1.05)^{15-20} = 0.028 \text{ g } NO_3\text{-N/(g MLSS} \cdot \text{day)}$$

The mass rate of denitrification associated with utilization of slowly biodegradable substrate and decay, ΔN_{XS}, is obtained by multiplying $q_{NO\,XS}$ by the mass of MLSS in the anoxic zone:

$$\Delta N_{XS} = (0.028)(16{,}190{,}000) = 453{,}000 \text{ g } NO_3\text{-N/day}$$

The total mass rate of denitrification is the sum of ΔN_{SS} and ΔN_{XS}, or

$$\Delta N = 643{,}000 + 453{,}000 = 1{,}096{,}000 \text{ g } NO_3\text{-N/day}$$

Since the mass rate of nitrate-N production is 1,332,000 g NO_3-N/day, the mass rate of nitrate release from the system is 1,322,000 − 1,096,000 = 226,000 g NO_3-N/day. Therefore, the effluent nitrate-N concentration is:

$$S_{NO} = \frac{226{,}000}{40{,}000} = 5.7 \text{ mg N/L}$$

This value is acceptable and can be used for preliminary design. If it had not

been adequate, it would have been necessary to adopt a new anoxic SRT and repeat the process until an acceptable value was found.

b. What are the sizes of the anoxic and aerobic zones?
The total mass of MLSS in the system at 15°C is 64,750,000 g. Since the design MLSS concentration is 3,000 mg/L (g/m^3) throughout the system, the total system volume is:

$$V = \frac{64{,}750{,}000}{3{,}000} = 21{,}600 \text{ m}^3$$

The mass of MLSS in the anoxic zone is 16,190,000 g, making its volume:

$$V_{ANX} = \frac{16{,}190{,}000}{3{,}000} = 5{,}400 \text{ m}^3$$

Therefore, the volume of the aerobic zone is:

$$V_{AER} = 21{,}600 - 5{,}400 = 16{,}200 \text{ m}^3$$

c. What MLR flow is needed at 15°C?
The specific denitrification rate is lowest in the winter, and thus the MLR flow for winter operation will be lower than that required for summer operation, which governs design of the MLR system. Nevertheless, because the degree of denitrification has only been calculated for the winter temperature, we will use it to demonstrate the procedure. The sum of the recycle and recirculation ratios can be calculated with Eq. 11.9. Since 1,322,000 g NO_3-N are produced per day and 1,096,000 g NO_3-N are denitrified per day, the fraction of the nitrate-N denitrified is 0.83. Therefore:

$$\alpha + \beta = \frac{0.83}{1 - 0.83} = 4.88$$

This is a relatively high value, but necessary to meet the process goals. Normally, for economic final settler performance, the solids recycle ratio, α, typically lies between 0.5 and 1.0. This means that the MLR ratio, β, should be at least 4.38. If a value of 4.5 is chosen, the MLR flow rate must be (4.5) (40,000 m^3/day) or 180,000 m^3/day. A final decision can be made only after the requirement for summer operation is determined. This will require recomputation of the mass of MLSS in the system in the summer, thereby giving the MLSS concentration associated with the system volume. This fixes the process loading factor for the anoxic zone, which determines the specific denitrification rate. The nitrogen available to the nitrifiers must also be recalculated for 25°C, allowing computation of the mass production rate of nitrate-N. Finally, the information can be combined to determine the fraction of the nitrate-N that is denitrified, allowing $\alpha + \beta$ to be calculated. Determination of the MLR rate needed for summer operation is left as an exercise for the reader.

d. What is the net alkalinity consumption at 15°C?
As calculated in Part a, the concentration of nitrate-N leaving the system in the absence of nitrification would be 33.3 mg/L as N. Its formation would lead to the destruction of (7.23)(33.3) = 241 mg/L as $CaCO_3$ of alkalinity. Since the effluent nitrate-N concentration is 5.7 mg/L as N, the amount of nitrate-N denitrified is equivalent to 27.6 mg/L as N. This would return

(3.5)(27.6) = 96.6 mg/L as $CaCO_3$ of alkalinity to the system. Thus, the net alkalinity destruction will be 241 − 96.6 = 144.4 mg/L as $CaCO_3$. Since 150 mg/L of the alkalinity in the influent could be destroyed while maintaining the needed residual of 50 mg/L, it would not be necessary to add chemicals in the winter. This is in contrast to the system with the anoxic selector, for which chemical addition was still required, although in lesser amount than the system without denitrification.

The remainder of the process design would proceed in exactly the same way as for the other systems. The oxygen requirement would have to be calculated for summer conditions and the savings due to denitrification determined in the same way as in Example 11.3.1.2. The heterotrophic oxygen requirement in the aerobic zone will be relatively low because the majority of the degradable organic matter will have been removed in the anoxic zone. This fact, coupled with the fact that the tank is large to accommodate the biomass needed to achieve a long SRT, means that the power requirements are associated primarily with solids suspension rather than with oxygen transfer. Investigation of these issues will be left as an exercise for the reader.

Examples 11.3.1.2 and 11.3.1.3 illustrate that initial anoxic zones of two quite different sizes may be used in systems of the same basic configuration, depending on process objectives. If an anoxic selector is used to remove readily biodegradable organic matter, then a relatively small anoxic zone results due to the rapid utilization of that substrate. Meaningful removal of nitrate-N will occur as a result of the presence of an anoxic selector, but the effluent nitrate-N concentration may still be significant. Relatively modest MLR rates may be adequate for these applications because the fraction of the nitrate-N that must be recirculated to the selector is fairly modest. Furthermore, a noticeable reduction in the oxygen requirement in the aerobic zone and in the net alkalinity consumption will occur. The reduction in power requirements associated with the decrease in the oxygen requirement will partially or completely off-set the power required to mix the selector and to recirculate mixed liquor. A larger initial anoxic zone may be required if the objective is to meet a specified effluent nitrate-N concentration through use of the MLE process. This occurs because both readily and slowly biodegradable organic matter may be required as electron donors to remove the larger mass of nitrate-N, and utilization of slowly biodegradable substrate results in slower denitrification rates. The greater degree of nitrate-N utilization in such applications results in increased MLR flow rates since a higher fraction of the nitrate-N formed must be recirculated to the initial anoxic zone. Greater reductions in the oxygen requirement and the associated power requirement in the aerobic zone occur, and net alkalinity consumption is reduced more. However, the further reduction in the power requirement for oxygen transfer may be small in comparison to the increase in power required to mix the larger anoxic zone and to pump the increased MLR flow.

Four-Stage Bardenpho Process—Addition of Second Anoxic and Aerobic Zones. A second anoxic zone, such as in the four-stage Bardenpho process (Figure 11.5), may be used to further reduce the nitrate-N concentration below that which can be economically achieved using an MLE system. However, because essentially all of the readily and slowly biodegradable organic matter will have been removed in the initial anoxic and aerobic zones, the primary source of electrons for the ad-

ditional denitrification is biomass decay. Since the decay of heterotrophs is a relatively slow process, particularly under anoxic conditions, and since not all of the heterotrophic bacteria are capable of denitrification, the resulting specific rate of denitrification will be low. This means that the size of a second anoxic zone can be significant. In addition, a small aerobic zone is required to prepare the MLSS for settling. Typically, an HRT of 30 min is used, which is sufficient to strip entrained gases from the denitrified mixed liquor exiting the second anoxic zone and to add dissolved oxygen to it before it enters the clarifier. The MLSS concentration is uniform throughout the four-stage Bardenpho system, making the SRT in each zone directly proportional to its volume.

As with the design of an MLE system, the best way to estimate the impact of adding second anoxic and aerobic zones is through the use of simulation with a model like ASM No. 1. Care should be exercised in the selection of the anoxic hydrolysis factor, η_h, because it will have a large impact on the rate of nitrate-N utilization associated with decay in the anoxic zones.

In the absence of a kinetic characterization sufficient to allow simulation, or as a prelude to simulation, it is possible to roughly estimate the nitrate-N requirement in the second anoxic zone by partitioning the electron acceptor requirement due to decay in much the same way that the oxygen requirement due to decay was partitioned to the various reactors in a CAS system, as illustrated in Section 10.3.4. The major uncertainty associated with this is the effect that the anoxic conditions have on the rate of decay.

An alternative to estimating the denitrification rate due to decay is to take an empirical approach. As they did for the specific denitrification rate in the first anoxic zone, Burdick, et al.[15] have reported an empirical relationship for the specific rate of denitrification due to decay in a second anoxic zone, $q_{NO\,XH}$. In this case, since no substrate enters the second anoxic zone, they were able to correlate the specific denitrification rate with the system SRT:

$$q_{NO\,XH} = 0.12\Theta_c^{0.706} \tag{11.12}$$

This relationship indicates that the specific denitrification rate will decrease as the SRT is increased, as expected. It was developed at a temperature of 20°C, but it can be corrected to other temperatures using Eq. 3.95 with an appropriate θ value; a value around 1.02 appears reasonable. This relationship has been widely reported and used,[5,7,8] and it is appropriate when other data are not available.

As with Eq. 11.11, use of Eq. 11.12 requires an iterative procedure. As with all previous nitrogen removal systems, selection of the system size should be done for winter operating conditions. After the first anoxic and aerobic zones have been sized using the procedures for the MLE system, an SRT is assumed for the second anoxic zone, thereby increasing the system SRT. Since the MLSS concentration is uniform throughout the system, its volume is calculated in proportion to the volume of the first anoxic and aerobic zones as suggested by Eqs. 11.1 and 11.2. Typically, the second aerobic zone is sized to give an HRT of around 30 min, as mentioned previously, thereby increasing the system volume. The SRT of this zone is also calculated by proportion, giving the new system SRT. The new system SRT allows computation of $q_{NO\,XH}$ with Eq. 11.12 and the mass of MLSS in the system with Eq. 9.11. The mass of MLSS in the second anoxic zone can then be calculated by proportioning with respect to its SRT as a fraction of the total SRT. Multiplication

of that mass by $q_{NO,XB}$ allows calculation of the mass denitrification rate due to biomass decay in the second anoxic zone, ΔN_{XB}. Division of that value by the flow rate gives the additional amount by which the effluent nitrate-N concentration is reduced. Subtraction of that value from the MLE effluent nitrate-N concentration will determine whether the required effluent nitrate-N concentration is attained. If it is not, then another SRT must be assumed for the second anoxic zone and the process repeated. Generally, the volume of the second aerobic zone is held constant at a value giving an HRT of about 30 min. The system size determined by this procedure will be approximate because each time the system SRT is increased the amount of nitrogen available to the nitrifiers and the rate of denitrification in the first anoxic zone will change. However, recalculation around the entire system is not justified because of the approximate nature of the empirical relationships used for denitrification in both anoxic zones. As stated earlier, the most accurate method of arriving at a final design is through simulation, provided the needed kinetic information is available.

The presence of a second anoxic zone will increase the amount of alkalinity recovery in the system, which will increase the alkalinity of the final effluent. However, the alkalinity produced in the second anoxic zone will have little impact on the alkalinity in the main aerobic zone because little of it will be recirculated through the system. Thus, it will not change the amount of chemical required to achieve stable nitrification from that required by an MLE system.

The inclusion of a second anoxic zone in a nitrogen removal process will increase the MLR requirements, which should be calculated for summer conditions. This occurs because the second anoxic zone will reduce the nitrate-N concentration entering the secondary clarifier, which reduces the nitrate-N concentration in the RAS flow to the initial anoxic zone, thereby decreasing the mass rate of nitrate-N return. If it is assumed that the nitrate-N concentration in the RAS is zero, Eq 11.8 can be modified to show the effect of the RAS and MLR ratios on the fraction of nitrate-N formed in the aerobic zone that is recirculated to the initial anoxic zone:

$$f_{NO,R} = \frac{\beta}{1 + \alpha + \beta} \tag{11.13}$$

Likewise, Eq. 11.9 can be modified to allow calculation of the MLR ratio required to allow denitrification of a specified fraction of the nitrate-N in the initial anoxic zone:

$$\beta = \frac{f_{NO,D}(1 + \alpha)}{1 - f_{NO,D}} \tag{11.14}$$

It is important to note that the value of $f_{NO,D}$ used in Eq. 11.14 is exactly the same as the value used in Eq. 11.9 because the fractional nitrate removal in the first anoxic zone is the same. As noted previously, the MLR flow rate should be calculated for summer conditions because it will be largest then.

The presence of the second anoxic zone will have little effect on the oxygen requirement in the first aerobic zone because the additional decay resulting from the increase in system SRT provides the electrons for denitrification in the second anoxic zone. The oxygen requirement in the second aerobic zone will be low, and thus the aeration rate in it will be governed primarily by the need to keep solids in suspension.

The impact of the additional SRT on the solids wastage rate can be calculated for winter conditions using the same procedures as used in all other designs.

The design of a second anoxic zone is illustrated in the following example.

Example 11.3.1.4

A second anoxic zone is to be added to the MLE process sized in Example 11.3.1.3. It is to reduce the nitrate-N concentration from 6 mg/L as N to 2 mg/L as N at 15°C. The second aerobic zone will have an HRT of 30 min. Assume that the temperature coefficient for the specific denitrification rate in the second anoxic zone is 1.02.

a. What is the required size of the second anoxic zone if the sizes and SRTs of the first anoxic and aerobic zones remain the same as in the MLE system? As a first guess, assume an SRT of 4 days for the second anoxic zone. From Example 11.3.1.3, the SRT of the MLE system is 16 days and its volume is 21,600 m^3. Thus, the volume of the second anoxic zone, $V_{ANX,2}$, is given by proportion:

$$V_{ANX,2} = \left(\frac{4}{16}\right) 21{,}600 = 5{,}400 \text{ m}^3$$

The second aerobic zone has an HRT of 30 min. Since the influent flow rate is 40,000 m^3/day, its volume is 833 m^3. The total system volume is:

$$V = 21{,}600 + 5{,}400 + 833 = 27{,}833 \text{ m}^3$$

By proportion, the total system SRT is:

$$\Theta_c = \left(\frac{27{,}833}{21{,}600}\right) 16 = 20.6 \text{ days}$$

Therefore, the SRT of the second aerobic zone is 0.6 day.

Substitution of the system SRT into Eq. 11.12 allows calculation of the specific denitrification rate in the second anoxic zone at 20°C:

$$q_{NO,NB,20} = (0.12)(20.6)^{0.706} = 0.0142 \text{ g NO}_3\text{-N/(g MLSS} \cdot \text{day)}$$

Since the design is being performed for 15°C, the specific denitrification rate must be corrected to that temperature using Eq. 3.95:

$$q_{NO,NB,15} = (0.0142)(1.02)^{15-20} = 0.013 \text{ g NO}_3\text{-N/(g MLSS} \cdot \text{day)}$$

This must now be multiplied by the mass of MLSS in the second anoxic zone to determine the mass removal rate of nitrate-N in the zone. The mass of MLSS in the system can be estimated with Eq. 9.11, as has been done several times, giving a value of 79,000,000 g. Since the SRT in the second anoxic zone is 4 days and the system SRT is 20.6 days, the mass of MLSS in the second anoxic zone is calculated by proportion to be 15,300,000 g. Therefore, the mass rate of denitrification in the second anoxic zone is:

$$\Delta N_{NB} = (0.013)(15{,}300{,}000) = 199{,}000 \text{ g NO}_3\text{-N/day}$$

The mass rate of NO_3-N release from the MLE system was calculated in Part a of Example 11.3.1.3 to be 226,000 g NO_3-N/day. Since the second anoxic zone can remove 199,000 g NO_3-N/day, the discharge into the second aerobic zone is 27,000 g NO_3-N/day. This corresponds to a concentration of 0.67

mg/L as N. Because the concentration is below the target value of 2.0 mg/L as N, it is possible that a smaller second anoxic zone could be used. However, several uncertainties exist in the design that suggest that it would be prudent to maintain an SRT of 4 days in the second anoxic zone. First, some additional nitrification will occur in the second aerobic zone on the ammonia-N released by the decay reactions in the second anoxic zone. Second, because of the longer SRT, the mass of MLSS in the first anoxic zone will be somewhat smaller than the amount calculated in Example 11.3.1.3. This means that a little less denitrification will occur there, increasing the mass flow rate of nitrate-N into the second anoxic zone. Consequently, it would be prudent to retain an SRT of 4 days in the second anoxic zone. Of course, if possible, simulations with a model like ASM No. 1 should be done to refine the design.

b. What MLR flow rate to the first anoxic zone is required for this application at 15°C?

Equation 11.14 is used to perform this calculation. From Part c of Example 11.3.1.3, 83% of the nitrate-N produced in the first aerobic zone must be recirculated to the first anoxic zone for denitrification. Typical solids recycle ratios for applications such as these range from 0.5 to 1.0. Calculate β for these two values of α. For $\alpha = 0.5$,

$$\beta = \frac{0.83(1 + 0.5)}{1 - 0.83} = 7.32$$

and for $\alpha = 1.0$

$$\beta = \frac{0.83(1 + 1)}{1 - 0.83} = 9.76$$

These are high values, which may not be practical. The alternative would be to reduce the size of the initial anoxic zone to reduce the fraction of the nitrate-N that would be reduced in that zone. This will necessitate an increase in the size of the second anoxic zone since more nitrate-N must be removed there. The entire computational procedure of this and the preceding example would have to be repeated to arrive at an estimate of the performance of an alternative system. Doing this several times would provide the information required to choose the optimal system design. Consideration of the effort involved in doing this demonstrates clearly the benefits associated with being able to use simulation to investigate alternative designs.

Simultaneous Nitrification and Denitrification. As discussed in Section 11.1.3, simultaneous nitrification and denitrification can be a significant nitrogen removal mechanism in a nitrifying activated sludge system oxygenated with a point source aerator or in a system with a uniformly low DO concentration, even though the bioreactor does not have a distinct and separate anoxic zone. Denitrification of as much as 50% of the nitrate-N produced has been reported in some applications.[54,69] The occurrence of simultaneous nitrification and denitrification requires three factors: (1) an oxygen transfer system that allows the development of zones of high and low DO concentration (either on a macroscopic or microscopic scale), (2) control of the oxygen input rate to the process, and (3) a sufficiently long SRT to allow full nitrification to occur even though parts of the bioreactor contain very low DO concentrations.

The impact of an oxygen transfer system that develops zones of high and low DO concentration was discussed in detail in Section 11.1.3. However, even if such a system is being used, regions of low DO concentration will not develop if the potential oxygen transfer rate to the system greatly exceeds the oxygen requirement. Rather, they will only develop when the potential oxygen transfer rate is less than the oxygen requirement. Furthermore, the mass rate of denitrification will be determined by the difference between the oxygen requirement for a totally aerobic system and the mass rate at which oxygen is actually being transferred to the liquid as an electron acceptor, TO:

$$\Delta N = \frac{RO_H + RO_A - TO}{2.86} \tag{11.15}$$

When denitrification is occurring, the effluent nitrate-N concentration is determined by the difference between the mass rate of nitrate-N formation and its utilization. The mass formation rate is just the flow rate times the concentration of nitrate-N that would be in the effluent from a totally aerobic system as given by Eq. 11.5. Therefore, the effluent concentration from a system experiencing simultaneous nitrification and denitrification is:

$$S_{NO} = \frac{0.98F(S_{N_0} - S_{NH} - S_{NS}) - \Delta N}{F} \tag{11.16}$$

Since S_{NO} cannot be negative, the ΔN from Eq. 11.15 cannot exceed the mass formation rate.

In order for simultaneous nitrification and denitrification to occur, the SRT must be sufficiently long for nitrification to occur even though the average DO concentration is low. Because the bioreactor contains regions (either microscopic or macroscopic) of high and low DO concentration and these regions are dynamic, it is difficult to estimate exactly the degree of nitrification. Thus, an approximation must be used. First, it is assumed that nitrification occurs at a rate consistent with the average DO concentration in the bioreactor. This allows computation of a DO safety factor for the system (Eq. 10.9) that can be used in Eq. 10.10 or 10.11 (depending on the bioreactor configuration) to estimate the system SRT required for stable nitrification. As long as that SRT is less than the equivalent aerobic SRT, $\Theta_{x,NR,eq}$, of the system, stable nitrification will occur, thereby allowing simultaneous nitrification and denitrification. The term equivalent aerobic SRT is used to signify that distinct aerobic and anoxic zones do not exist, preventing a purely aerobic SRT from being calculated as it was for the other nitrogen removal systems. Thus, some other approach must be used to approximate the equivalent aerobic SRT.

Estimation of the equivalent aerobic SRT requires the assumption that anoxic and aerobic regions (either microscopic or macroscopic) are distributed equally throughout the entire bioreactor. Under that condition, if nitrate-N were used as effectively as oxygen as an electron acceptor, the fraction of the system that was anoxic would be equivalent to the fraction of the heterotrophic oxygen requirement that was being met by nitrate-N. However, only a fraction of the heterotrophs are able to denitrify. If we take that fraction as being represented by the anoxic growth factor, η_g, then the fraction of the system volume that is anoxic would be given by:

$$f_{V,ANX} = \frac{2.86 \cdot \Delta N}{\eta_g \cdot RO_H} \tag{11.17}$$

Therefore, the equivalent aerobic SRT is:

$$\Theta_{c,AER,eq} = (1 - f_{V,ANX})\Theta_c \tag{11.18}$$

As stated above, as long as the required SRT as given by Eq. 10.10 or 10.11 (depending on the bioreactor configuration) is less than the equivalent aerobic SRT of the system, as given by Eq. 11.18, then stable nitrification will occur and the process will work. It should be recognized, however, that these computations are approximate and that it is difficult to predict with certainty the degree of nitrification that will occur.

Savings in the amount of oxygen that must be supplied and in the amount of alkalinity that will be destroyed can be calculated using the same procedures as those used for the other denitrification systems.

One other important point about simultaneous nitrification and denitrification in a single vessel should be emphasized. That is the potential for filamentous bulking from growth of Group IV bacteria (Table 10.5).[20,37] Currently, techniques are not available for preventing such growth without chemical control as discussed in Section 10.4.3. Thus, the potential impacts of poor settleability on overall process performance should be considered before implementing this strategy for nitrogen removal.

Example 11.3.1.5

Consider the nitrifying activated sludge system designed in Example 11.3.1.1. The design SRT, which was chosen for winter operation, was 12 days and it is maintained year round. This SRT provides a substantial safety factor during warm temperature operation. Consequently, the feasibility of operating the oxygen transfer system to achieve 50% denitrification during warm weather conditions is to be assessed. In making this assessment, assume that the reduction in oxygen input will result in an average DO concentration of 0.5 mg/L, that the value of the anoxic growth factor, η_g, is 0.8, and that the safety factor for uncertainty is 1.0.

a. What SRT will be required to achieve stable nitrification in summer?
The approach is the same as in Part a of Example 11.3.1.1. Since a CAS system is being used, Eq. 10.11 is the appropriate expression with which to calculate the required SRT. From Figure 9.4, the minimum aerobic SRT for growth of nitrifiers is 1.3 days. The peak load safety factor, s_{PL}, is 2.5. The safety factor for uncertainty, s_U, has a value of 1.0. Because the half-saturation coefficient for DO for nitrifiers is 0.75 mg/L (Table E8.5), and the average DO concentration is 0.5 mg/L, the DO safety factor (Eq. 10.9) has a value of 2.5. Thus:

$$\Theta_{c,r} = (1.3)(1.0)(2.5)(2.5) = 8.12 \text{ days}$$

Thus an SRT of at least 8.12 days is needed to ensure that stable nitrification will occur even during peak loads. The absolute minimum SRT for nitrification, however, is obtained by multiplying the DO safety factor by the minimum SRT, giving a value of 3.25 days. At that SRT, nitrification will be less stable during transients, with significant breakthrough of ammonia-N occur-

ring. It will become more stable, however, as the SRT is increased toward 8.12 days.

b. What is the equivalent aerobic SRT at 25°C?
From Example 11.3.1.1, the effluent nitrate-N concentration in the summer is 33.5 mg/L as N. Since the flow rate is 40,000 m^3/day, the nitrate production rate is 1,340,000 g NO_x-N/day. For 50% denitrification, ΔN would be half of that, or 670 kg NO_x-N/day. The heterotrophic oxygen requirement at 25°C was calculated in Part e of Example 11.3.1.1, where it was found to be 7,930 kg O_2/day. Substituting these into Eq. 11.17 gives the fraction of the system volume that is anoxic:

$$f_{V,ANX} = \frac{(2.86)(670)}{(0.8)(7,930)} = 0.30$$

The equivalent aerobic SRT can then be calculated with Eq. 11.18:

$$\Theta_{c,AER,eq} = (1 - 0.30)12.0 = 8.4 \text{ days}$$

The equivalent aerobic SRT of 8.4 days exceeds the required SRT of 8.12 days. Thus, stable nitrification will occur at the reduced DO concentration, allowing significant denitrification to occur. Thus, the proposed operating mode is feasible.

c. What oxygen transfer rate, TO, is required to achieve 50% denitrification at 25°C?
The summer oxygen requirement for a completely aerobic system was calculated in Part e of Example 11.3.1.1. The heterotrophic requirement, RO_H, was found to be 7,930 kg O_2/day, as noted above. The autotrophic requirement, RO_X, was found to be 6,080 kg O_2/day. The denitrification rate, ΔN, was found to be 670 kg NO_x-N/day in Part b above. The required oxygen transfer rate can be calculated with a rearranged form of Eq. 11.15:

$$TO = 7,930 + 6,080 - (2.86)(670) = 12,090 \text{ kg } O_2\text{/day}$$

This rate is 86.3% of the rate in the totally aerobic system.

Separate Stage Denitrification. Separate stage denitrification systems, such as illustrated in Figure 11.7, can be designed by the same procedures as those presented in Section 10.3 for the design of aerobic processes. In this case, however, the terminal electron acceptor (nitrate-N) is present in the influent wastewater and it is the electron donor that must be added. Since the required mass rate of nitrate-N removal, ΔN, is known, the mass rate at which the electron donor must be supplied, ΔS, can be calculated with Eq. 6.4 by setting the SRT equal to the anoxic SRT. Division of this rate by the influent flow rate gives the required electron donor concentration, expressed as COD, that must be provided. As discussed in Section 11.1.3, methanol is often used as the electron donor. An appropriate value for the anoxic SRT is selected, generally on the order of 3 to 5 days. The mass of MLSS in the system is then calculated for the selected anoxic SRT using Eq. 9.11 and the MLSS concentration can be chosen to give an economical settler, thereby fixing the volume of the anoxic zone. Since oxygen transfer is not required, the final settler size is the only factor that needs to be considered in choosing the MLSS concentration, unless the influent nitrate-N concentration is so high that evolution of N_2 gas would cause floc shear. For domestic wastewater treatment, mixing energy with a submerged mixer will

generally be required to keep the MLSS in suspension. An aerobic zone with an HRT of around 45 to 60 min is added to the process for the same reasons that the second aerobic zone was provided in a four-stage Bardenpho process. In addition, it must remove any residual electron donor, which is why it is slightly larger. The total SRT for the system can then be estimated from the total system volume by proportioning, as illustrated in the examples of this chapter. The mass of MLSS in the system can then be recalculated and the MLSS reconsidered if warranted. The quantity of waste solids is calculated as illustrated in Section 10.3.3. As indicated in Section 11.1.5, separate stage suspended growth denitrification systems are not used widely in practice. The interested reader is referred to the U. S. Environmental Protection Agency's, *Process Design Manual for Nitrogen Control*[18] for a detailed description of the design of such systems.

11.3.2 Biological Phosphorus Removal Processes

This section deals with those BNR processes that are designed to remove only organic matter and phosphorus. No nitrogen removal occurs. In fact, nitrification is undesirable because of the negative impact of nitrate-N recycle on the processes occurring in the anaerobic zone, as discussed in Sections 7.7.2 and 7.7.3. Section 11.3.3 will consider those BNR systems in which both nitrogen and phosphorus must be removed, or in which nitrification must occur in addition to phosphorus removal.

The mechanisms of biological phosphorus removal are discussed in Sections 2.4.6 and 3.7, while simulations showing the interactions between PAOs and the other bacteria commonly found in activated sludge systems are presented in Sections 7.7.2 and 7.7.3. It is apparent from all of this information that BPR is the most complicated suspended growth biochemical operation in use today, with many complex interactions possible between the different populations in the microbial community. Consideration of those interactions makes it clear that simulation is the only way to evaluate potential designs, short of actually testing them in pilot-scale facilities. However, models from which computer codes can be formulated have only recently become available,[7,34,78,80] and only a limited number of computer codes have been written for them, as indicated in Table 6.4. As a consequence, procedures for the design of BPR processes are not nearly as well developed as those for biological nitrogen removal systems, and are more empirically based. Nevertheless, the fundamental design approaches for activated sludge systems presented in Section 10.3 can be coupled with appropriate values of the SRT to establish the preliminary size and configuration of a BPR process. That approach will be presented here. Different approaches are used for mainstream and sidestream processes.

Mainstream Processes. Mainstream processes such as the A/O™ process are sized using the principles presented in Section 10.3. The SRTs in the anaerobic and aerobic zones are selected as described in Section 11.2.1 and added to give the total SRT. The mass of MLSS, the solids wastage rate, and the oxygen requirement are calculated for the selected total SRT and the bioreactor is sized using the procedures illustrated in Section 10.3.3.

Selection of the anaerobic SRT is influenced strongly by the amount and nature of the readily biodegradable substrate in the wastewater. PAOs sequester VFAs in the anaerobic zone, thereby providing themselves with the energy source that they will use to grow and store phosphorus in the aerobic zone. Approximately 7 to 10

mg of acetic acid are required to remove 1.0 mg of phosphorus.[78,81,82] This corresponds to 7.5 to 10.7 mg of COD due to VFAs for each mg of phosphorus that must be removed. Storage of VFAs as PHAs in the anaerobic zone is a very rapid process.[80] Consequently, if the needed amount of VFAs is present in the wastewater, the anaerobic SRT can be as short as 0.5 day at 20°C.[43] If no VFAs are present in the influent, but the amount of readily biodegradable substrate is sufficient to produce the needed quantity by fermentation, then the anaerobic SRT needs to be around 1.5 days at 20°C.[43] For situations in which some VFAs are present, but partial fermentation is still required, SRTs between 0.5 and 1.5 days should be sufficient, depending on the quantity of VFAs available. On the other hand, should the amount of readily biodegradable substrate be insufficient, then hydrolysis of slowly biodegradable substrate must occur before fermentation can form the VFAs and the anaerobic SRT must be much longer, on the order of 2.5 to 3 days. In that case, it may be more economic to carry out fermentation reactions off-stream, thereby allowing a VFA-rich stream to be added to the influent.

If nitrification is occurring in the aerobic zone, nitrate-N will be returned to the anaerobic zone through the RAS flow (making it anoxic) and the denitrifying heterotrophs will be able to outcompete the PAOs for VFAs. The return of nitrate-N to the anaerobic zone will also interfere with fermentation since the heterotrophs will simply use as much readily biodegradable substrate as is possible for the amount of nitrate-N returned, rather than fermenting it to VFAs. Equation 6.4 allows calculation of the amount of readily biodegradable COD (or VFAs) removed per unit of nitrate-N returned to the anaerobic zone. Using typical parameter values, it suggests that about 6 mg of COD will be lost per mg of nitrate-N recycled to the anaerobic zone. Consequently, if nitrification is occurring in the aerobic zone, this additional substrate requirement must be taken into consideration when making a judgement about the size of the anaerobic SRT. To do this, the fraction of nitrate-N formed that is returned to the anaerobic zone via the RAS flow must be considered. It can be calculated with Eq. 11.8 in which has been set equal to zero since no MLR is used.

Because of these negative impacts of nitrification, processes like the A/O™ process should be designed with short aerobic SRTs to minimize the likelihood of nitrification. This practice will also maximize the excess biomass production, thereby maximizing phosphorus removal. Selection of the aerobic SRT is complicated by the fact that the minimum SRTs required for growth of PAOs and nitrifiers are very similar at high temperatures, as illustrated in Figure 9.4. Therefore, to avoid nitrification at high temperature while providing sufficient aerobic SRT for growth of PAOs at low temperature, it may be necessary to use different aerobic SRTs in winter and summer. Generally, a safety factor of 1.5 to 2.0 is applied to the minimum SRT for PAOs to arrive at an aerobic SRT that will allow PAOs to grow while minimizing the chances for stable nitrification.

Since no substrate oxidation occurs in the anaerobic zone (VFAs are simply stored as PHAs), the oxygen requirement can be calculated with Eq. 9.13 using the total SRT. However, some evidence suggests that the oxygen requirement in a BPR process may be less than that calculated value.[5,70] The mechanism for such a reduction is not known, but the observed magnitude of the effect appears to be on the order of 10 to 30%. Until this issue is resolved, it is prudent to design the oxygen transfer system based on the results of Eq. 9.13, but to incorporate additional turn-

down capacity into the design to allow plant operators to save energy if actual oxygen requirements are less than calculated.

Since the MLSS concentration is uniform in this type of BPR process, the sizes of the aerobic and anaerobic zones will be proportional to their SRTs, as indicated by Eqs. 11.1 and 11.3. Likewise, the mass of MLSS in each zone will be proportional to its SRT. Consequently, once the mass of MLSS in the system has been calculated by using the total SRT in Eq. 9.11, the mass in the aerobic zone can be calculated by proportion. It can then be used in the selection of the aerobic zone volume, and the associated MLSS concentration, by considering floc shear, mixing, and oxygen transfer rates, just as was done in Section 10.3.3. The size of the anaerobic zone is then determined by proportionality. Benefits accrue from staging the anaerobic zone, with two tanks-in-series being a recommended configuration.[76] The need to cover the zone will be determined by the availability of readily biodegradable substrate relative to the amount of phosphorus to be removed. The mixing requirements in the anaerobic zone are similar to those in the anoxic zone of a nitrogen removal system, which are discussed in Section 11.3.1.

Because of the complexity of the events occurring in biological phosphorus removal, the only way to predict the effluent soluble phosphorus concentration with any certainty is to fully characterize the wastewater and use the resulting parameters in simulations. However, when that cannot be done, the organic matter to phosphorus removal ratio for the particular type of process (Table 11.4) can be used to approximate the value, or alternatively, the stoichiometry based on VFAs and readily biodegradable substrate reported earlier can be used. As long as more than enough organic matter is present in the wastewater to remove all of the phosphorus, i.e., the wastewater is phosphorus limited, then an effluent low in soluble phosphorus will be produced. On the other hand, if the process is carbon limited, the amount of phosphorus that can be removed will be determined primarily by the available organic matter, with the effluent soluble phosphorus concentration being determined by the difference between the amount entering the system and the amount removed. The total phosphorus concentration in the effluent will be the soluble phosphorus plus the phosphorus in the effluent suspended solids, which will depend on the effluent suspended solids concentration, as illustrated in Figure 11.15. The phosphorus content of the suspended solids can be determined on the basis of the phosphorus removed and the excess solids production. It is equal to the difference between the mass flow rates of phosphorus in and out of the process divided by the solids wastage rate.

The following example illustrates the design of a mainstream biological phosphorus removal process.

Example 11.3.2.1

Consider the design of a BPR process like the A/O™ process to remove phosphorus from the wastewater that was the subject of all of the examples throughout Sections 10.3 and 11.3.1. The phosphorus content of the wastewater is 7.5 mg/L as P but the VFA concentration is negligible. A safety factor of 1.5 is considered to be adequate for selection of the aerobic SRT to ensure stable PAO growth.

a. What SRTs are appropriate for cold and warm temperature operating conditions?

First consider cold temperature operation (15°C). Because the wastewater contains no VFAs, the anaerobic SRT will need to be at least 1.5 days to allow fermentation of the readily biodegradable organic matter. From Figure 9.4, the minimum aerobic SRT is 2 days at 15°C. Using a factor of safety of 1.5, the design aerobic SRT would be 3 days. Also from Figure 9.4, the minimum aerobic SRT for nitrification 3.5 days. Thus, for winter operation, an SRT of 3 days would allow the system to operate without nitrification. The total SRT would be 4.5 days.

Next, consider warm temperature operation (25°C). At 25°C it should be possible to reduce the anaerobic SRT to 1.0 day while still performing the necessary fermentation. From Figure 9.4, the minimum aerobic SRT is 1.3 days. Using a safety factor of 1.5, this gives a design aerobic SRT of 2.0 days. Unfortunately, it can be seen from Figure 9.4 that the minimum aerobic SRT for nitrifiers at 25°C is also around 1.3 days, so some nitrification will occur during the warmest operating condition. The extent of nitrification will depend on a number of factors, including the bioreactor configuration, and can only be accurately predicted from pilot studies. However, complete nitrification would definitely occur if the SRT were left at the winter value. Thus, in summer, the total SRT should be 3.0 days.

We now need to check to determine whether these two operating conditions are compatible with each other. Since the fraction of the SRT in each zone is equal to the fraction of the system volume in each zone, the fractions must be the same in summer and winter. Calculation of the anaerobic SRT as a fraction of total SRT for both conditions reveals that it is one-third both times. Thus, the choices are acceptable.

b. Selection of bioreactor volume and MLSS concentrations.
The mass of MLSS in the system should be calculated for winter conditions using Eq. 9.11 because winter conditions require the longer SRT and have the lower decay coefficient, both of which maximize the mass of MLSS. Once that is done, the bioreactor volume and the MLSS concentration must be selected in exactly the same way as in Example 10.3.3.2. Because this procedure has been demonstrated several times, it will not be repeated here. However, it should be recognized that the MLSS concentration will be lower in summer because of the shorter SRT and higher decay coefficient, and this should be considered in the choices.

c. What are the solids wastage rate and the oxygen requirement for heterotrophs?
Because of the use of different SRTs in summer and winter, we cannot assume that winter conditions will give the greater solids wastage rate, $W_{M,H}$, and summer the higher heterotrophic oxygen requirement, RO_H. Consequently, they must be calculated for each operating condition. As has been done several times, the solids wastage rate is calculated using Eq. 9.12, while the heterotrophic oxygen requirement is calculated using Eq. 9.13. The results are provided in Table E11.1. The differences are insignificant. It should be noted that such small differences will not always be the case.

d. What effluent soluble phosphorus concentration would be expected under winter operating conditions?
Nitrification will not occur in the winter, so no nitrate-N will be recycled to the anaerobic zone in the RAS. We can expect to use 7.5 to 10.7 mg of VFA

Table E11.1 Effect of Temperature on the Excess Solids Production Rate and the Oxygen Requirement in Example 11.3.2.1

	Warm weather	Cold weather
Temperature (°C)	25	15
SRT (days)	3.0	4.5
W_{VSS} (kg TSS/day)	5,350	5,330
RO_{tl} (kg O_2/day)	6,260	6,290

COD for each mg of phosphorus that must be removed. Little COD is lost during fermentation of readily biodegradable organic matter to VFAs, so we would expect to get a mg of VFA COD for each mg of readily biodegradable COD fermented. To be conservative, assume that 10.7 mg of readily biodegradable COD will be needed to remove a mg of phosphorus after fermentation. Since the wastewater contains 115 mg/L of readily biodegradable COD, the maximum phosphorus removal capability is 115 ÷ 10.7 = 10.7 mg/L. Furthermore, since the wastewater contains 7.5 mg/L of phosphorus, the system would be expected to remove essentially all of it in winter. Thus, we would expect the effluent soluble phosphorus concentration to be low.

Another way of addressing this question is with the $BOD_5/\Delta P$ ratios given in Table 11.4. Since the A/O™ process is considered to be a high efficiency process when nitrification is not occurring, the $BOD_5/\Delta P$ ratio will range from 15 to 20 mg BOD_5/mg P. Again, to be conservative, assume that the worst case applies, i.e, a ratio of 20. Since the wastewater contains 155 mg/L of BOD_5 (Table E8.4), the maximum phosphorus removal capability is 155 ÷ 20 = 7.8 mg/L. This, too, suggests that excellent phosphorus removal should occur, giving a very low effluent soluble phosphate concentration.

e. What effluent soluble phosphorus concentration would be expected under summer operating conditions?
During the warmest part of the summer, it is likely that nitrification will occur in spite of the reduction in the SRT to three days. In Example 10.3.3.2, the concentration of nitrogen available to the nitrifiers at an SRT of 3 days was calculated to be 30.5 mg/L as N. Thus, we can expect the maximum amount of nitrate-N formed to be 29.9 mg/L as N (Eq. 11.5). The amount of nitrate-N recycled to the anaerobic zone depends on the RAS flow rate. Typically, the RAS flow rate is one-half the influent flow rate, or α is 0.5. Using that in Eq. 11.8 with β = 0 tells us that one-third of the nitrate-N formed will be returned, or an effective concentration of 10 mg/L as N. Since about 6 mg of COD will be required to denitrify each mg of nitrate-N, the recycle of the nitrate-N will reduce the readily biodegradable COD by 60 mg/L, to a value of 55 mg/L. If only 7.5 mg of readily biodegradable COD were required to remove a mg of phosphorus, the process would be capable of removing 55 ÷ 7.5 = 7.3 mg/L of P, which would be sufficient to remove almost all of the phosphorus. However, if 10.7 mg of COD were required, the system could only remove 55 ÷ 10.7 = 5.1 mg/L of P. Although a small amount of phosphorus would be used in biomass synthesis, this suggests that a significant concentration of residual phosphorus would be left in the summer. The exact amount could only be determined with pilot studies. This

analysis did not consider the slowly biodegradable substrate because relatively little of it will be available at the short SRTs involved. Again, only pilot studies or simulations will reveal whether any slowly biodegradable substrate would become available.

As in Part d, another way of addressing this question is with the BOD:ΔP ratio. From Table 11.4 we see that the ratio is 20–25 mg BOD/mg P for an A/O™ process in which nitrification is occurring. A ratio of 20 would lead us to believe that almost all of the phosphorus could be removed, as we saw in Part d. If the ratio were 25, we would expect the removal to be 155 ÷ 25 = 6.2 mg/L as P, which would leave a significant residual. Thus this analysis, like the previous, gives mixed results, with one assumption suggesting full phosphorus removal while the other suggests incomplete removal. Therefore, pilot studies should be performed to more accurately define the capability.

f. What is the phosphorus content of the MLSS in the winter?
Since all of the phosphorus will be removed via the waste solids and the solids are homogeneous throughout the system, the phosphorus content of the MLSS can be determined by calculating the phosphorus content of the waste solids. The mass flow rate of phosphorus in the influent to the process is (7.5 g/m^3 as P) (40,000 m^3/day) ÷ 1,000 g/kg = 300 kg/day of P. As discussed in Part d above, this phosphorus will be almost completely removed. Therefore, to be conservative assume that the effluent soluble PO_4-P concentration is negligible. Therefore, the mass of phosphorus removed per day is 300 kg. From Table E11.1, the solids wastage rate in winter is 5,330 kg/day. Therefore, the phosphorus content of the mixed liquor is 300 kg P/day ÷ 5,330 kg TSS/day = 0.056 mg P/mg TSS.

g. What will be the concentration of particulate phosphorus in the effluent if the effluent suspended solids concentration is 10 mg/L as TSS?
Since the phosphorus content of the solids is 0.056 mg P/mg TSS and the effluent suspended solids concentration is 10 mg/L as TSS, the effluent particulate phosphorus concentration is 0.56 mg/L as P. The total phosphorus concentration will be the sum of this value plus any soluble phosphorus that is present.

Sidestream Processes. Sidestream processes such as Phostrip® can be designed in a similar fashion. The primary differences are that the anaerobic zone is created in a sidestream stripper and that two phosphorus removal mechanisms are provided, one through the waste solids like other BPR processes and the other in the stripper overflow. As discussed in Section 11.1.3, the low loading of biodegradable organic matter into the stripper in the Phostrip® process results in the need to retain the solids there for 8 to 12 hours to allow for the formation of sufficient quantities of VFAs by fermentation. The uptake of these VFAs results in substantial phosphate release, which is subsequently elutriated and removed in the stripper overflow stream. Mass balance procedures can be used to quantify the mass of phosphorus removed by stripping and elutriation. This is added to the phosphorus contained in the waste solids to give the total mass of phosphorus removed in the process. As discussed in Section 11.1.5, although the Phostrip® process is of historical importance, its use is relatively limited today, and this trend is expected to continue. Consequently, further

discussion of the design procedures will not be given here. They are available elsewhere.[57,67]

11.3.3 Processes that Remove both Nitrogen and Phosphorus

As discussed in Section 11.1.3, a large number of processes are available that remove both nitrogen and phosphorus. This would be of significant concern if unique design procedures were required for each. Fortunately this is not the case, and all processes that remove both nitrogen and phosphorus can be designed using the procedures already presented in Sections 11.3.1 and 11.3.2. Because those procedures are approximate, so is the procedure for designing systems to remove both nitrogen and phosphorus. The degree of interaction among the different components of the microbial community in a BNR system makes it impossible to develop exact analytical procedures. However, just as for systems that remove nitrogen or phosphorus, they are sufficiently exact to provide designs that can be verified through pilot studies and simulation. Just a few additional factors must be considered when designing a process that removes both nutrients.

Because complete nitrification must occur in the course of removing nitrogen, one consideration of particular importance is the minimization of nitrate-N recycle to the anaerobic zone. As discussed in Section 11.1.2, nitrate-N recycle adversely impacts phosphorus removal by allowing increased growth of non-PAO heterotrophs and by reducing fermentation in the anaerobic zone. Elimination of nitrate-N recycle is primarily a process selection issue, and several processes have been developed that minimize or eliminate it. The issues involved in process selection are discussed in Section 11.1.4 and the benefits and drawbacks of the alternative processes are presented in Table 11.2.

Another consideration in adapting the approaches of Sections 11.3.1 and 11.3.2 to the design of a process to remove both nutrients is the impact of the upstream anaerobic zone on denitrification in the downstream anoxic zone. It might be thought that the removal of readily biodegradable substrate in the anaerobic zone would result in reduced rates of denitrification in the anoxic zone because of the need to use slowly biodegradable substrate. Experience indicates that this is not the case.[18] It is hypothesized that fermentation of slowly biodegradable organic matter in the anaerobic zone results in the formation of readily biodegradable substrate, which produces a rapid rate of denitrification in the anoxic zone. While additional research is needed to elucidate the mechanism, it is clear that similar anoxic zone sizes can be used in processes that remove nitrogen alone and in processes that remove both nitrogen and phosphorus.

A final consideration is that some systems have different MLSS concentrations in the anaerobic zone than in the anoxic and aerobic zones. Examples are the UCT and VIP processes, shown in Figures 11.11 and 11.13, respectively, which add the RAS flow to the anoxic zone and provide biomass to the anaerobic zone by recirculating denitrified anoxic mixed liquor (AR) from the anoxic zone to the anaerobic zone. Although this is done to minimize nitrate-N recirculation to the anaerobic zone, it will result in lower MLSS concentrations in the anaerobic zone. The MLSS concentration in the anaerobic zone, X_{MLANA}, can be estimated from the MLSS concentration in the anoxic zone, X_{MLANX}, by performing a mass balance on the anaerobic

zone. Neglecting any change in the MLSS concentration that occurs in the zone, it can be shown that:

$$X_{MLANA} = X_{MLANX}\left(\frac{\delta}{1+\delta}\right) \tag{11.19}$$

where δ is the AR rate expressed as a fraction of the influent flow rate.

As with the other BNR processes, design begins with selection of the SRTs for the three environments. As discussed in Section 11.2.1, anaerobic SRTs can be reduced in processes that remove both nitrogen and phosphorus in comparison to the anaerobic SRT needed in a process that removes phosphorus alone. This is because readily biodegradable substrate is removed in both the anaerobic and anoxic zones of a process that removes both nutrients. Consequently, the anaerobic zone need not be relied upon to remove all of this material. However, reduction of the anaerobic SRT will result in reduced fermentation and reduced phosphorus removal capability for the process. Procedures are being developed to assess this trade-off, but at this time it must be done using judgment, experience, and simulation. Until more experience is gained, it is prudent to provide the flexibility to adjust the relative sizes of the anaerobic and anoxic zones, thereby adjusting their SRTs. Because it is the input of nitrate-N that distinguishes an anoxic zone from an anaerobic zone, one way to do this is to construct those zones as several completely mixed tanks in series, as in the VIP process (Figure 11.13), and to provide several possible discharge points for the RAS and NR. In this way, a tank can be changed from anaerobic to anoxic, and vice versa, simply by moving the RAS and NR input point. In addition, both mixing and oxygen transfer equipment can be installed in the initial sections of the aerobic zone to provide the flexibility to extend the anoxic SRT should the need exist. As discussed in Section 11.1.3, staging of the bioreactor provides several advantages including improved reaction rates and the selection of microorganisms with higher maximum specific growth rates. Thus, incorporation of staging not only provides operational flexibility, but also enhances overall process performance.

Just as with biological nitrogen removal processes, systems that remove both nitrogen and phosphorus can be designed for a widely varying degree of nitrogen removal. The principles presented in Section 11.3.1 can be directly applied here, depending on effluent total nitrogen objectives. The minimum degree of nitrogen removal required is that which will eliminate nitrate-N recycle to the anaerobic zone.

The design proceeds in exactly the same way that the other designs have proceeded. First, the SRTs of the three zones must be selected using the criteria presented above and in the preceding two sections, as well as in Section 11.2.1. This may require iteration, as discussed in Section 11.3.1. Using the total SRT, the mass of MLSS in the system is estimated with Eq. 9.11, giving the mass of MLSS in each zone by application of Eqs. 11.1–11.3. The RAS and NR flow rates are then selected to give the desired degree of nitrogen removal through the anoxic zone, using the principles articulated in Section 11.3.1 for an initial anoxic zone. The oxygen requirement in the absence of denitrification can be calculated with Eqs. 9.13 and 10.16, and corrected for denitrification to determine the net oxygen requirement as illustrated in Example 11.3.1.2. An MLSS concentration is then selected for the aerobic zone following consideration of the requirements for sedimentation, mixing, and oxygen transfer. This allows the total volume of the aerobic zone to be calculated. Since the MLSS concentration is the same in the anoxic zone as in the aerobic

zone, the chosen MLSS concentration can also be used to calculate the volume of the anoxic zone from the mass of MLSS present. A denitrified anoxic mixed liquor recirculation rate can then be selected, thereby setting the anaerobic MLSS concentration by Eq. 11.19. That, in turn is used to calculate the volume of the anaerobic zone from the known mass of MLSS present. Higher AR rates will result in higher anaerobic MLSS concentrations, thereby reducing the volume of the anaerobic zone. Thus, the opportunity exists for selecting a least-cost combination. Each zone can be subdivided into stages, if desired. If an effluent is desired with a lower nitrate-N concentration than can be accomplished with only an initial anoxic zone, then a second anoxic zone and a second aerobic zone can be added using the approach from Example 11.3.1.4.

Example 11.3.3.1

A facility to remove both nitrogen and phosphorus is to be designed to treat the wastewater considered throughout Sections 10.3 and 11.3, which contains 7.5 mg/L of phosphorus. Process objectives include oxidizing ammonia-N, maximizing phosphorus removal, and obtaining a moderate degree of nitrogen removal to reduce the net oxygen requirement and alkalinity consumption. The VIP process was selected based on these objectives.

a. What SRTs are appropriate for this application?
As discussed in Part a of Example 11.3.1.1, the required aerobic SRT is 12.0 days. Although an anaerobic SRT as low as 0.5 days could be used, a value of 1.0 is selected to allow some fermentation to occur. An anoxic SRT of 1.5 days is selected to ensure complete removal of readily biodegradable organic matter, which will result in good sludge settling characteristics. Thus, the total SRT will be 1.0 + 1.5 + 12.0 or 14.5 days.

b. If the design MLSS concentration in the anoxic and aerobic zones is 3,000 mg/L as TSS and the AR rate is equal to the influent flow rate, what are the sizes of the bioreactor and its individual components, as well as the HRTs?
First, calculate the mass of MLSS in the system at 15°C using Eq. 9.11 and an SRT of 14.5 days. Use of the procedure demonstrated in Example 10.3.3.3 gives a value of 60,000,000 g. Application of Eqs. 11.1–11.3 gives the mass in each zone:

$$\sum (X_{M,I} \cdot V)_{AER} = 49{,}700{,}000 \text{ g}$$

$$\sum (X_{M,I} \cdot V)_{ANX} = 6{,}200{,}000 \text{ g}$$

$$\sum (X_{M,I} \cdot V)_{ANA} = 4{,}100{,}000 \text{ g}$$

Since the MLSS concentration in the aerobic and anoxic zones is 3,000 mg/L, the volumes of those zones are 16,600 and 2,100 m^3, respectively.

The MLSS concentration in the anaerobic zone is obtained by application of Eq. 11.19. Since the AR flow rate is equal to the influent flow rate, δ has a value of 1.0. Therefore:

$$X_{M,I,ANA} = 3{,}000 \left(\frac{1.0}{1 + 1.0} \right) = 1{,}500 \text{ mg/L}$$

Using this value gives a volume for the anaerobic basin of 2,700 m^3.

The HRTs of each zone can be calculated using the design flow of

40,000 m^3/day. The volumes and resulting HRTs are summarized in Table E11.2. Note that the size and HRT of the anaerobic zone are larger than those of the anoxic zone, even though its SRT is smaller. This is a result of the lower MLSS concentration in the anaerobic zone.

Appropriate values for the anaerobic and anoxic zone SRTs are only approximately known, and therefore staging of these zones is desirable, as discussed in Section 11.1.3 and above. Consequently, use a total volume of 4,800 m^3 for the anaerobic and anoxic zones and configure the volume as 5 stages, each with a volume of 960 m^3. Provide flexibility in the NR and RAS discharge locations so that either two or three of the cells can function as the anaerobic zone, with the remainder functioning as the anoxic zone. Furthermore, because of the benefits of staging the aerobic zone with respect to nitrification, divide it into four cells, each with a volume of 4,150 m^3.

c. What total nitrogen removal can be expected, and what NR flow rate is appropriate for this application?
As shown in Part b of Example 11.3.1.2, an anoxic selector with an SRT of 1.5 days is expected to remove 837 kg NO_x-N/day based on the availability of readily biodegradable substrate. As discussed above, experience indicates that we should get about the same amount of denitrification in this application in spite of the presence of the anaerobic zone preceding the anoxic zone. It was also shown in Part b of Example 11.3.2.2 that the sum of the RAS and MLR rates should be 163% of the influent flow rate to provide sufficient nitrate-N to remove all of the readily biodegradable substrate. Thus, it is appropriate to provide an equivalent amount of RAS and NR flow in this case. However, because the goal is to fully denitrify in the anoxic zone to prevent recirculation of nitrate-N to the anaerobic zone, considerable flexibility should be provided in the possible pumping rates to ensure that this is accomplished.

d. How much phosphorus should this process be able to remove?
Because the PAOs are protected from nitrate by the process configuration, the system should be capable of removing equal quantities of phosphorus in winter and summer. Thus, the amount of phosphorus removal should be similar to that estimated for the A/O™ process in the winter in Part d of Example 11.3.2.1. Thus, essentially all of the influent phosphorus should be removed.

Table E11.2 Volumes and Hydraulic Residence Times for the Zones in the VIP Process of Example 11.3.3.1

Zone	Volume, m^3	HRT, day
Anaerobic	2,700	0.067
Anoxic	2,100	0.053
Aerobic	16,600	0.415
Total	21,400	0.535

11.3.4 Process Optimization by Dynamic Simulation

As discussed in Sections 9.4.3 and 10.3.7, process optimization using dynamic simulation is an extremely valuable tool for refining the design and operation of biological processes. This is particularly true for BNR systems because of the complex interactions among the components of the system. The most widely used models for BNR processes are ASMs No. 1[33] and 2,[34] as described in Section 6.1. ASM No. 1 is used in Chapter 6 to illustrate the impacts of nitrification and denitrification on the operational and performance characteristics of suspended growth biochemical operations, and both models are used in Chapter 7 to illustrate the performance of several BNR systems. ASM No. 1 has been widely used to model and optimize suspended growth nitrogen removal processes.[3,31,32,54,57,68] Experience with its use suggests that it can be used successfully if some limited wastewater characterization data are available.[3,31,32] Of particular importance is the fraction of the biodegradable organic matter in the influent wastewater that is readily biodegradable. Appropriate kinetic parameters for autotrophic biomass are also important as they have been observed to vary from one application to the next. If site-specific values of the kinetic parameters are not available for the nitrifying bacteria, standard values can be used with caution to provide approximate estimates of their performance in BNR processes. Due to the numerous interactions, the use of models such as ASM No. 1 is encouraged to optimize the design of nitrogen removal systems.

ASM No. 2 incorporates biological phosphorus removal. Although it has not been available as long as ASM No. 1, ASM No. 2 and a precursor model by Dold have received initial evaluation.[4,26,32] The results to date are very encouraging and suggest that ASM No. 2 can be used with confidence when properly calibrated.[4,17,20,26,83] Finally, Barker and Dold[7] have presented a model that includes some processes not incorporated into ASM No. 2. It is very complete and is capable of simulating a broad range of BNR systems.[8]

11.4 PROCESS OPERATION

The operation of BNR systems builds on the operating principles for the activated sludge process presented in Section 10.4. Long-term control of the process is achieved by control of the SRT; the techniques described in Section 10.4 for accomplishing this are equally applicable to BNR systems. Furthermore, the visual observations discussed in that chapter are also equally applicable to BNR systems.[74]

Complete nitrification is necessary if efficient nitrogen removal is to be achieved. Nitrification is controlled by adjustment of the overall SRT to attain the aerobic SRT necessary to maintain the nitrifying bacteria in the system at the given temperature. Adequate DO concentrations must be maintained in the aerobic zone. However, as discussed in Sections 11.1.3 and 11.3.1, if the SRT is long enough, the system can be operated at lower DO concentrations to encourage denitrification in the aerobic zone. The key is the maintenance of a sufficient nitrification safety factor to retain the nitrifiers in the process and to ensure an adequate level of activity.

Efficient denitrification is achieved by excluding DO from the anoxic zones and recirculating sufficient quantities of nitrate-N to initial anoxic zones. Because the MLR flow comes from the aerobic zone, avoiding high DO concentrations in

that zone helps to minimize the recirculation of DO with the nitrate-N. Entry of this alternative electron acceptor into the anoxic zone will allow some of the biodegradable organic matter to be consumed aerobically, resulting in less being available for denitrification. Moreover, since the DO concentration in an initial anoxic zone will be low, the recycle of significant quantities of DO to that zone may result in the growth of Group I filamentous bacteria (see Table 10.5), creating a poorly settling sludge. Mixed liquor recirculation provides the operator with considerable control over the mass of nitrate-N that can be denitrified in a BNR system, with increased mass flow rates resulting from increased MLR rates. Increased recirculation rates will increase the power requirements for recirculation pumping, but this must be balanced against the nitrogen removal requirements and the fact that increased denitrification will reduce both the oxygen and alkalinity requirements in the aerobic basin.

Biological phosphorus removal is optimized by excluding DO and nitrate-N from the anaerobic zone. Preventing the addition of these electron acceptors allows the maximum possible mass of organic matter to be stored by PAOs, thereby providing them with the maximum possible competitive advantage. Aeration of the wastewater upstream of the bioreactor should be avoided, both because it can add DO to the anaerobic zone, and because it can lead to stripping or aerobic biodegradation of VFAs. If fermentation facilities are available, their operation can significantly improve the performance of a BPR system.

In some instances, specialized operational techniques can be applied to BNR systems. Measurement of nutrient profiles through a staged bioreactor system can provide insight into the reactions occurring there. For example, phosphorus release in the anaerobic zone and phosphorus uptake in the aerobic zone can be monitored. These profiles can be used as an early indicator of process upsets, and they can be used to adjust process operating parameters. Nitrate-N profiles through the anoxic zone can be used to indicate incomplete denitrification that can result in nitrate-N recycle to the anaerobic zone, while nitrate-N and ammonia-N profiles through the aerobic zone will indicate the extent of nitrification. Furthermore, the ammonia-N profile can be used to adjust the aerobic SRT to obtain the necessary degree of nitrification. In some systems, alkalinity consumption and/or effluent turbidity are monitored and used to control aeration input to minimum values, allowing optimization of denitrification in the aerobic zone.[28,58]

As discussed above, nutrients contained in recycle streams from solids handling systems can adversely affect the performance of BNR processes. Consequently, the entire wastewater treatment plant must be viewed as an integrated system. The plant operator needs to monitor and understand the interaction between the liquid and solids treatment trains to optimize overall process performance and cost.

11.5 KEY POINTS

1. Segregation of the bioreactor into anaerobic, anoxic, and aerobic zones distinguishes biological nutrient removal (BNR) systems from other activated sludge systems. Anaerobic zones allow for selection of phosphorus accumulating organisms (PAOs), thereby increasing the phosphorus content of the mixed liquor suspended solid (MLSS) and allowing phospho-

rus removal in the waste solids. Anoxic zones allow for denitrification, thereby removing nitrogen as N_2. Aerobic zones are necessary for the growth of nitrifying bacteria and PAOs.

2. Phosphorus removal is adversely impacted by the recycle of nitrate-N to the anaerobic zone of a biological phosphorus removal (BPR) process. Nitrate-N allows consumption of volatile fatty acids (VFAs) by heterotrophic denitrifying bacteria, thereby reducing the competitive advantage of the PAOs. It also decreases the production of VFAs by fermentation of readily biodegradable substrate by non-PAO heterotrophic bacteria, thereby lowering the amount of VFAs available.
3. Incorporation of initial anoxic zones into nitrifying activated sludge systems provides several benefits: reduced alkalinity consumption, decreased oxygen requirement, diminished denitrification in the clarifier, and selection of floc-forming bacteria at the expense of filaments. The incorporation of anaerobic zones into nonnitrifying systems also offers the benefit of selection of floc forming bacteria.
4. Anoxic zones can be incorporated into nitrogen removal systems both upstream and downstream of the main aerobic zone. The rate of denitrification in an upstream anoxic zone is rapid due to the utilization of readily biodegradable substrate, while it is slow in a downstream anoxic zone because it must depend on decay reactions and utilization of slowly biodegradable substrate.
5. Biological phosphorus removal is maximized by operation at short solids retention times (SRTs). This occurs because: nitrification, and the resulting potential for recycle of nitrate-N to the anaerobic zone, is minimized; sludge production and the resulting wasting of phosphorus from the system is maximized; and uptake of phosphorus by the biomass is maximized.
6. The SRT of a BNR system impacts organism selection. The aerobic SRT must be long enough to allow PAOs to grow if phosphorus removal is desired, while it must be increased further to allow nitrifying bacteria to grow if nitrification is desired. Anaerobic and anoxic SRTs affect phosphorus and nitrogen removal, as well as control of filamentous bacteria.
7. Organic matter to nutrient ratios provide an indication of the suitability of wastewater for biological nutrient removal. The organic matter to nitrogen ratio indicates whether sufficient organic matter is available to denitrify the nitrate-N generated. The organic matter to phosphorus ratio indicates whether sufficient organic matter is available to grow an adequate quantity of PAOs to remove the influent phosphorus. If both nitrogen and phosphorus are to be removed in one system, the organic matter requirement for each should be considered and the larger amount determined. It will generally control overall performance.
8. The composition of the organic matter in a wastewater affects the performance of BNR systems. Volatile fatty acids are utilized by PAOs, and highly efficient phosphorus removal can be achieved with a small anaerobic zone when a high proportion of the influent organic matter is VFAs. If insufficient VFAs are present, readily biodegradable organic matter can be fermented to generate VFAs, but the reaction is slow; thus the anaer-

obic zone must be larger. Readily biodegradable organic matter also results in a high denitrification rate in initial anoxic zones.

9. Wastewaters can be fermented to increase their VFA content and improve the performance of BNR systems. Fermentation can occur in the wastewater collection systems and in specially designed facilities at the wastewater treatment plant.
10. Due to the high phosphorus content of the mixed liquor suspended solids (MLSS) in a BPR system, the effluent total suspended solids (TSS) can contribute a significant quantity of phosphorus to the process effluent. They also contribute some organic-N.
11. Biological nutrient removal processes are designed using the same procedures as for other activated sludge systems. SRTs must be chosen for each of the zones (aerobic, anoxic, and anaerobic) and summed to obtain the system SRT. Using the system SRT, the mass of MLSS, the solids wastage rate, and the total electron acceptor requirement are calculated using the equations from Chapter 9. The bioreactor is sized based on the mass of MLSS and the design MLSS concentration, and divided into the separate zones in proportion to their SRTs. Aerobic zone volumetric power inputs must be checked relative to the requirements for floc shear and mixing, and mixing energy inputs must be determined for anaerobic and anoxic zones. Mixed liquor recirculation (MLR) rates must also be calculated.
12. A key requirement for a biological nitrogen removal system is stable nitrification. Thus, selection of the aerobic SRT for such a system is governed by the maintenance of a stable population of nitrifiers.
13. Anoxic selectors are sized to remove readily biodegradable substrate, which typically requires an anoxic SRT of 1 day in summer and 1.5 days in winter for domestic wastewater. The MLR rate is set to return enough nitrate-N to the selector to accept the electrons removed during growth of heterotrophs on the readily biodegradable substrate.
14. The initial anoxic zone in the modified Ludzack–Ettinger (MLE) process is designed to remove sufficient nitrate-N to achieve a desired effluent concentration, which typically requires utilization of some slowly biodegradable substrate in addition to the readily biodegradable substrate. Because of theoretical uncertainties about the utilization rate of slowly biodegradable substrate, an empirical expression is used to select the anoxic SRT. The MLR rate is selected to return sufficient nitrate-N to the anoxic zone to meet the desired effluent concentration.
15. The second anoxic zone in a four-stage Bardenpho system is sized to remove additional nitrate-N by anoxic biomass decay. Selection of its SRT requires use of an empirical relationship. The introduction of the second anoxic zone essentially eliminates the recirculation of nitrate-N to the initial anoxic zone via the return activated sludge (RAS) flow, thereby requiring a larger MLR rate to achieve the same degree on nitrate-N removal in the initial anoxic zone.
16. Careful control of the oxygen transfer rate to a nitrifying activated sludge system provides the opportunity for simultaneous nitrification and denitrification in a single vessel, provided that the SRT is sufficiently long

for stable nitrification under reduced dissolved oxygen (DO) conditions. However, it is possible that such an activity will lead to filamentous bulking by Group IV bacteria (Table 10.5).

17. Selection of the aerobic SRT for a process to remove phosphorus but not nitrogen requires careful consideration of the minimum SRT required for both PAOs and nitrifiers. Under winter operating conditions it is usually possible to select an aerobic SRT that will allow growth of the PAOs while excluding the nitrifiers, but this will be difficult to do in summer.
18. Selection of the anaerobic SRT for a BPR process is influenced strongly by the nature of the readily biodegradable substrate and the recycle of nitrate-N to the anaerobic zone. If the readily biodegradable substrate is high in VFAs, then short anaerobic SRTs can be used. Fermentation of readily biodegradable substrate to form VFAs requires longer anaerobic SRTs, however.
19. The amount of phosphorus removed by a BPR process is determined by the amounts of readily biodegradable substrate and phosphorus in the influent, and the recycle of nitrate-N to the anaerobic zone by the RAS flow. Approximately 7.5 to 10.7 mg of VFA chemical oxygen demand (COD) are needed to remove 1.0 mg of phosphorus. In addition, around 6 mg of readily biodegradable COD are needed to remove 1.0 mg of recycled nitrate-N.
20. The design of BNR processes for the removal of both nitrogen and phosphorus is accomplished by combining the techniques used to design processes that remove either nitrogen or phosphorus. Because nitrification is necessary in systems removing both nutrients, a key task during the design is to minimize the entry of nitrate-N into the anaerobic zone.
21. Because of the complex interactions among the members of the microbial community in them, dynamic simulation is a valuable tool for optimizing the design of BNR systems. Activated sludge models No. 1 and 2 have been used for this purpose and have proven useful in a number of full-scale applications.
22. The operation of BNR systems is similar to that of other activated sludge processes, and the operational techniques applied to the activated sludge process are equally applicable. The SRT is controlled to provide long-term regulation of the process using the same approaches used with activated sludge systems. Visual observations of the bioreactor, the clarifier, the settleometer, and under the microscope can also be used to optimize the process. Specialized operational techniques used with BNR processes include measurement of nutrient profiles, alkalinity consumption, and effluent turbidity.
23. The operators of wastewater treatment plants containing a BNR process must recognize the interactions between the liquid and solids processing trains.

11.6 STUDY QUESTIONS

1. Prepare a table comparing the bioreactor configuration, SRT, HRT, recycle ratio, MLR ratio, and design approach for the following: MLE, four-

stage Bardenpho, five-stage Bardenpho, A/O™ A^2/O™, VIP, and UCT processes.

2. Describe the biochemical transformations occurring in a BNR process that achieves both nitrogen and phosphorus removal. Where do these transformations occur, and how do they interact to allow nutrient removal to occur?
3. Describe the benefits that biological nutrient removal can provide to the operation of an activated sludge system and list the circumstances under which nutrient removal capabilities would be incorporated into an activated sludge system design even though the effluent quality criteria did not require it.
4. Describe the mechanisms that allow denitrification to occur in a supposedly aerobic bioreactor. How can denitrification be increased? What are the impacts of this denitrification on the operational and performance characteristics of the system?
5. Prepare a process schematic of the Phostrip® process. Describe the phosphorus removal mechanisms that operate in this process, indicate how they operate, and how they differ from the mechanisms in other BPR processes. How would nitrification affect the phosphorus removal capability of this process? Why? How would the nature of the elutriant stream influence it? Why?
6. Make a list of all the process design and operational factors that maximize phosphorus removal in a BNR system. Make a similar list of factors that maximize nitrogen removal. Compare the two lists and identify those factors that are similar for the two systems and those that are different. Discuss the impacts of these similarities and differences on the design of a system that removes both nitrogen and phosphorus.
7. Make a list of factors that can affect the characteristics of a wastewater as it enters the first bioreactor in a BNR system and the resulting impact of those characteristics on the performance of a BNR system. Identify which factors are under the control of the process designer and which are under the control of the process operator. What does this analysis indicate about the design and operation of a BNR system?
8. Describe the concept of the organic matter to phosphorus removal ratio. How can this ratio be used as a screening tool to identify the BPR processes to be considered for a particular application?
9. List the factors that affect the size of the initial anoxic zone required for a particular nitrogen removal application. Under what circumstances would a relatively small zone be used? When would a relatively large zone be used?
10. List the factors that affect the size of the anaerobic zone required for a particular BPR application. Under what circumstances would a relatively small zone be used? When would a relatively large zone be used?
11. Derive the equation used to calculate the fraction of nitrate-N recirculated from the aerobic zone of a biological nitrogen removal system to an upstream anoxic zone. Do this for both the MLE and the four-stage Bardenpho systems and explain why the expressions are different.

12. Using the wastewater characteristics in Table E8.4, the stoichiometric and kinetic parameters in Table E8.5, and the temperature correction factors in Table E10.1, design a CAS system with an MLSS concentration of 2750 mg/L as TSS to produce a fully nitrified effluent year round while treating an average wastewater flow rate of 30,000 m^3/day. The lowest sustained winter temperature is 13°C and the highest sustained summer temperature is 24°C. Use a diffused air–oxygen transfer system and assume that the in-process oxygen transfer efficiency is 12%. Also assume that the hydraulic characteristics of the CAS bioreactor are equivalent to three tanks-in-series and that the safety factor for uncertainty is 1.0. The diurnal peak loading on the system is twice the average loading. Use this information in selecting the aerobic SRT for the system, but for simplicity, base all decisions about tank volumes and oxygen transfer rates on average loading conditions. Justify all assumptions and decisions.
13. Add an anoxic selector capable of removing all readily biodegradable substrate to the CAS system designed in Study Question 12. Determine its volume and the MLR rate. Justify all assumptions and decisions. Compare the following for the systems with and without the selector: total system volume, aerobic bioreactor volume, oxygen transfer rate and air flow rate to the aerobic zone, and alkalinity that must be added to maintain a residual concentration of 50 mg/L as $CaCO_3$.
14. Use a computer code implementing ASM No. 1 or a similar model to evaluate the effluent nitrate-N concentration and the oxygen requirement in the system designed in Study Question 13. Comment on any differences between the results from Study Question 13 and the simulation results, suggesting possible reasons for differences.
15. For the situation described in Study Question 12, design an MLE system to produce an effluent containing no more than 10 mg/L as N of nitrate-N, determining the volume of the anoxic zone and the MLR rate. Justify all assumptions and decisions. Compare the following to the system designed in Study Question 13: total system volume, aerobic bioreactor volume, oxygen transfer rate and air flow rate to the aerobic zone, and alkalinity that must be added to maintain a residual concentration of 50 mg/L as $CaCO_3$.
16. Use a computer code implementing ASM No. 1 or a similar model to evaluate the effluent nitrate-N concentration and the oxygen requirement in the system designed in Study Question 15. Comment on any differences between the results from Study Question 15 and the simulation results, suggesting possible reasons for differences.
17. Add a second anoxic zone to the process considered in Study Question 15, to lower the effluent nitrate-N concentration to 3 mg/L as N. Determine the size of the second anoxic and aerobic zones, and calculate the MLR required. Justify all assumptions and decisions. Compare the following to the system designed in Study Question 15: total system volume; aerobic bioreactor volume; oxygen transfer rate and air flow rate to the aerobic zone; and alkalinity that must be added to maintain a residual concentration of 50 mg/L as $CaCO_3$.

18. Use a computer code implementing ASM No. 1 or a similar model to evaluate the effluent nitrate-N concentration and the oxygen requirement in the system designed in Study Question 17. Comment on any differences between the results from Study Question 17 and the simulation results, suggesting possible reasons for differences.
19. Redo Study Question 17, but in this case, size the first anoxic zone to remove the amount of nitrate-N that can be returned with a MLR rate of four times the influent flow rate. Compare this system to the one designed in Study Question 17 and comment on the differences.
20. Evaluate the possibility of operating the oxygen transfer system in the CAS system designed in Study Question 12 in such a way that 45% of the nitrate-N formed during summer operation is denitrified. At what oxygen transfer rate would the system have to be operated to achieve the desired degree of denitrification? In making your assessment, assume that the reduction in oxygen input will result in an average DO concentration of 0.5 mg/L and that the anoxic growth factor, η_g, is 0.75.
21. For the situation described in Study Question 12, prepare the design of an A/O™ process operating at a temperature of 20°C at steady-state. Assume that the MLSS concentration is 2,500 mg/L as TSS. If the total phosphorus concentration of the influent wastewater is 8 mg/L as P and the effluent suspended solids concentration is 15 mg/L as TSS, what is the estimated effluent total phosphorus concentration for this process? What is the estimated effluent nitrate-N concentration? State and justify all assumptions.
22. For the situation described in Study Question 12, prepare the design of a VIP process to produce an effluent with minimal ammonia-N and phosphorus concentrations while operating at steady-state at 20°C. Assume that the MLSS concentration is 3,500 mg/L as TSS. If the total phosphorus concentration of the influent wastewater is 12 mg/L as P and the effluent suspended solids concentration is 10 mg/L as TSS, what is the estimated effluent total phosphorus concentration for this process? What is the estimated effluent nitrate-N concentration? State and justify all assumptions.
23. Using a computer code implementing ASM No. 1 or a similar model, systematically investigate the effects of the anoxic SRT and MLR rate on the performance of the MLE process developed in Study Question 15. Also determine the impact of staging the bioreactors on overall process performance. Discuss the implications of your findings to an optimal system design.
24. Using a computer code implementing ASM No. 1 or a similar model, systematically investigate the effects of the distribution of the anoxic SRT between the first and second anoxic zones on the performance of the four-stage Bardenpho process developed in Study Question 17. Adjust the MLR rate to match the nitrate-N need in the first anoxic zone for each size investigated. Also determine the impact of staging the bioreactors on overall process performance. Discuss the implications of your findings to an optimal system design.

25. Using a computer code implementing ASM No. 2 or a similar model, systematically investigate the effects of the anaerobic and aerobic SRTs on the performance of the A/O™ process developed in Study Question 21. Also determine the impact of staging the bioreactors on overall process performance. Discuss the implications of your findings to an optimal system design.
26. Using a computer code implementing ASM No. 2 or a similar model, systematically investigate the effects of the anaerobic and anoxic SRTs on the performance of the VIP process developed in Study Question 22. Adjust the MLR rate to match the nitrate-N need in the anoxic zone for each size investigated. Also determine the impact of staging the bioreactors on overall process performance. Discuss the implications of your findings to an optimal system design.
27. Discuss the impact of solids handling recycles on the performance of a BNR system. What steps can be taken to mitigate these impacts?

REFERENCES

1. Alleman, J. E. and R. L. Irvine, Storage-induced denitrification using sequencing batch reactor operation. *Water Research* **14**:1483–1488, 1980.
2. Arvin, E., Biological removal of phosphate from wastewater. *CRC Critical Reviews in Environmental Control* **15**:25–64, 1985.
3. Bailod, C. R., Oxygen utilization in activated sludge plants: simulation and model calibration. EPA-600/2–86/065, U.S. Environmental Protection Agency, Washington, DC, 1988.
4. Barbeau, D. S., S. B. Murphy, G. B. Sprouse, S. Reusser, W. C. Boyle, W. M. Karlovich, and T. F. Rubens, The use of pilot testing and EBPR modeling for alternative development and process selection of an EBPR retrofit. *Proceedings of the Water Environment Federation 68th Annual Conference & Exposition, Volume I, Wastewater Treatment Research and Municipal Wastewater Treatment*, Water Environment Federation, Alexandria, Virginia, pp. 535–546, 1995.
5. Barker, P. and P. L. Dold, COD and nitrogen mass balances in activated sludge systems. *Water Research* **29**:633–643, 1995.
6. Barker, P. and P. L. Dold, Denitrification behavior in biological excess phosphorus removal activated sludge systems. *Water Research* **30**:769–780, 1996.
7. Barker, P. S. and P. L. Dold, General model for biological nutrient removal activated-sludge systems: model presentation. *Water Environment Research* **69**:969–984, 1997.
8. Barker, P. S. and P. L. Dold, General model for biological nutrient removal activated-sludge systems: model application. *Water Environment Research* **69**:985–991, 1997.
9. Barnard, J. L., Biological nutrient removal without the addition of chemicals. *Water Research* **9**:485–490, 1975.
10. Barnard, J. L., A review of biological phosphorus removal in the activated sludge process. *Water SA* **2**:136–144, 1976.
11. Barnard, J. L., Activated primary tanks for phosphate removal. *Water SA* **10**:121– 126, 1984.
12. Barnes, D., C. F. Forster, and D. W. M. Johnstone, *Oxidation Ditches in Wastewater Treatment*, Pitman Publishing, Marshfield, Massachusetts, 1983.
13. Barth, E. F., R. C. Brenner, and R. F. Lewis, Chemical–biological control of nitrogen and phosphorus in wastewater effluents. *Journal, Water Pollution Control Federation* **40**:2040–2054, 1968.

14. Bundgaard, E., G. H. Kristensen, and E. Arvin, Full-scale experience with phosphorus removal in an alternating system. *Water Science and Technology* **15**(3/4):197–217, 1983.
15. Burdick, C. R., D. R. Refling, and H. D. Stensel, Advanced biological treatment to achieve nutrient removal. *Journal, Water Pollution Control Federation* **54**:1078–1086, 1982.
16. Burke, R. A., P. L. Dold, and G. v. R. Marais, *Biological Excess Phosphorus Removal in Short Sludge Age Activated Sludge Systems*, Research Report No. W58, University of Cape Town, South Africa, 1986.
17. Çinar, Ö., G. T. Daigger, and S. P. Graef, Evaluation of IAWQ Activated Sludge Model No. 2 Using Four Full Scale Wastewater Treatment Plants. *Proceedings of the Water Environment Federation 69th Annual Conference & Exposition, Volume I, Wastewater Treatment Research and Municipal Wastewater Treatment*, Water Environment Federation, Alexandria, Virginia, pp. 127–138, 1996.
18. Clayton, J. A., G. A. Ekama, M. C. Wentzel, and G. v. R. Marais, Denitrification kinetics in biological nitrogen and phosphorus removal activated sludge systems treating municipal wastewaters. *Water Science and Technology* **23**(3/4):1025–1035, 1991.
19. Daigger, G. T. and G. A. Nicholson, Performance of four full-scale nitrifying wastewater treatment plants incorporating selectors. *Journal, Water Pollution Control Federation* **62**:676–683, 1990.
20. Daigger, G. T. and D. Nolasco, Evaluation and design of full-scale wastewater treatment plants using biological process models. *Water Science and Technology* **31**(2):245–255, 1995.
21. Daigger, G. T., L. M. Morales, J. R. Borberg, and G. D. Waltrip, Full-scale and pilot-scale experience with the VIP process. *Proceedings of the First Australian Conference on Biological Nutrient Removal*, Bendigo, Australia, pp. 157–166, 1990.
22. Daigger, G. T., C. W. Randall, G. D. Waltrip, E. D. Romm, and L. M. Morales, Factors affecting biological phosphorus removal for the VIP process, a high-rate University of Capetown type process. *Proceedings of the IAWPRC Specialized Conference on Biological Phosphorus Removal from Wastewater*, Rome, pp. 185–200, 1985.
23. Daigger, G. T., G. D. Waltrip, E. D. Romm, and L. M. Morales, Enhanced secondary treatment incorporating biological nutrient removal. *Journal, Water Pollution Control Federation* **60**:1833–1842, 1988.
24. Drews, R. J. L. C. and A. M. Greef, Nitrogen elimination by rapid alternation of aerobic/anoxic conditions in orbal activated sludge plants. *Water Research* **7**:1183–1194, 1973.
25. Ekama, G. A., I. P. Siebritz, and G. v. R. Marais, Considerations in the process design of nutrient removal activated sludge processes. *Water Science and Technology* **15**(3/4): 283–318, 1983.
26. Elmendorf, H. A., M. Tetreault, and P. L. Dold, Process modeling of biological excess phosphorus removal: a basis for detailed design. *Proceedings of the Water Environment Federation 67th Annual Conference & Exposition, Volume I, Biological Treatment Systems and Biological Nutrient Removal*, Water Environment Federation, Alexandria, Virginia, pp. 471–482, 1994.
27. Fillos, J., V. Diyamandoglu, L.A. Carrio, and L. Robinson, Full-scale evaluation of biological nitrogen removal in the step-feed activated sludge process. *Water Environment Research* **68**:132–142, 1996.
28. Freed, A. J. and H. F. Davis, Biological process control using on-line alkalinity measurement. *Proceedings of the Water Environment Federation 65th Annual Conference and Exposition, Liquid Treatment Processes Symposia*, Water Environment Federation, Alexandria, Virginia, pp. 317–328, 1992.
29. Gabb, D. M. D., G. A. Ekama, D. Jenkins, and G. v. R. Marais, *Specific Control Measures for Filamentous Bulking in Long Sludge Age Activated Sludge Systems*, Research

Report No. W61, Department of Civil Engineering, University of Capetown, South Africa, 1987.

30. Goronszy, M. C. and D. Rigel, Biological phosphorus removal in a fed-batch reactor without anoxic mixing sequences. *Research Journal, Water Pollution Control Federation* **63**:248–258, 1990.
31. Henze, M. and W. Gujer, Editors, *Interactions of Wastewater, Biomass, and Reactor Configuration in Biological Treatment Plants*, Proceedings of the IAWPRC Specialized Seminar, Copenhagen, Denmark, August 21–23, 1991, *Water Science and Technology* **25**(6):1992.
32. Henze, M. and W. Gujer, Editors, *Modelling and Control of Activated Sludge Processes*, Selected Proceedings of the IAWQ International Specialized Seminar, Copenhagen, Denmark, 22–24 August, 1994, *Water Science and Technology* **31**(2):1995.
33. Henze, M., C. P. L. Grady Jr., W. Gujer, G. v. R. Marais, and T. Matsuo, Activated sludge model No. 1. *IAWPRC Scientific and Technical Report No. 1*, 1987.
34. Henze, M., W. Gujer, T. Mino, T. Matsuo, M. C. Wentzel, and G. v. R. Marais, Activated sludge model No. 2. *IAWQ Scientific and Technical Reports*, No. 3, 1995.
35. Hong, S., D. J. Krichten, K. S. Kisenbauer, and R. L. Sell, A biological wastewater treatment system for nutrient removal. Paper presented at the 54th Annual Conference of the Water Pollution Control Federation, Detroit, Michigan, 1981.
36. Hong, S., M. L. Spector, J. V. Galdieri, and R. P. Seebohm, Recent advances on biological nutrient control by the A/O™ process. Paper presented at the 56th Annual Conference of the Water Pollution Control Federation, Atlanta, Georgia, 1983.
37. Jenkins, D., M. G. Richards, and G. T. Daigger, *The Causes and Cures of Activated Sludge Bulking and Foaming*, 2nd ed., Lewis Publishers, Ann Arbor, Michigan, 1993.
38. Kayser, R., G. Stobbe, and M. Werner, Operational results of the Wolfsburg wastewater treatment plant. *Water Science and Technology* **25**(4/5):203–209, 1992.
39. Kos, M., J. Wanner, I. Sorm, and P. Grau, R-D-N activated sludge process. *Water Science and Technology* **25**(4/5):151–160, 1992.
40. Levin, G. V. and J. Shapiro, Metabolic uptake of phosphorus by wastewater organisms. *Journal, Water Pollution Control Federation* **37**:800–821, 1965.
41. Levin, G. V., G. J. Topol, and A. G. Tarnay, Operation of full scale biological phosphorus removal plant. *Journal, Water Pollution Control Federation* **47**:577–590, 1975.
42. Levin, G. V., G. J. Topol, A. G. Tarnay, and R. B. Samworth, Pilot-plant tests of a phosphate removal process. *Journal, Water Pollution Control Federation* **44**:1940–1954, 1972.
43. Liner, M. O. and C. P. L. Grady Jr., Development of design heuristics for biological excess phosphorus removal systems using BioWin. *Proceedings of the Water Environment Federation 70th Annual Conference & Exposition, Volume I, Wastewater Treatment Research and Municipal Wastewater Treatment*, Water Environment Federation, Alexandria, Virginia, 1997.
44. Ludzack, F. J. and M. B. Ettinger, Controlling operation to minimize activated sludge effluent nitrogen. *Journal, Water Pollution Control Federation* **34**:920–931, 1962.
45. Mamais, D. and D. Jenkins, The effects of SRT and temperature on enhanced biological phosphorus removal. *Water Science and Technology* **26**(5/6):955–965, 1992.
46. Manning, J. F. and R. L. Irvine, The biological removal of phosphorus in a sequencing batch reactor. *Journal, Water Pollution Control Federation* **57**:8794, 1985.
47. Marten, W. L. and G. T. Daigger, Full-scale evaluation of factors affecting the performance of anoxic selectors. *Proceedings of the Water Environment Federation 68th Annual Conference & Exposition, Volume I, Wastewater Treatment Research and Municipal Wastewater Treatment*, Water Environment Federation, Alexandria, Virginia, pp. 669–679, 1995.

48. Menar, A. B. and D. Jenkins, The fate of phosphorus in waste treatment processes: the enhanced removal of phosphate by activated sludge. *Proceedings of the 24th Purdue Industrial Waste Conference*, Purdue University Engineering Extension Series No. 135, pp. 655–674, 1969.
49. Morales, L. M., G. T. Daigger, and J. R. Borberg, Capability assessment of biological nutrient removal facilities. *Research Journal, Water Pollution Control Federation* **63**: 900–909, 1991.
50. Nicholls, H. A., D. W. Osborn, and A. R. Pitman, Improvement to the stability of the biological phosphate removal process at the Johannesburg Norther Works. *Proceedings of the IAWPRC Specialized Conference on Biological Phosphate Removal from Wastewaters*, Rome, pp. 261–272, 1987.
51. Nolasco, D., G. Daigger, J. Stephenson, D. Stafford, and G. Patry, An innovative BNR process using modified step feed configuration. *Proceedings of the Water Environment Federation 66th Annual Conference & Exposition, Volume I, Research*, Water Environment Federation, Alexandria, Virginia, pp. 239–248, 1993.
52. Pitman, A. R., L. H. Lotter, W. V. Alexander, and S. L. Deacon, Fermentation of raw sludge and elutriation of resultant fatty acids to promote excess biological phosphorus removal. *Water Science and Technology* **25**(4/5):185–194, 1992.
53. Rabinowitz, B. and W. K. Oldham, Excess biological phosphorus removal in the activated sludge process using primary sludge fermentation. *Canadian Journal of Civil Engineering* **13**:345–351, 1986.
54. Randall, C. W., J. L. Barnard, and H. D. Stensel, Editors, *Design and Retrofit of Wastewater Treatment Plants for Biological Nutrient Removal*, Technomics Publishing, Lancaster, Pennsylvania, 1992.
55. Randall, C. W., D. Waltrip, and M. V. Wable, Upgrading a municipal activated sludge plant for high-rate biological nutrient removal. *Water Science and Technology* **22**(7/8): 21–33, 1990.
56. Sadick, T. E., J. E. Semon, G. D. Waltrip, G. T. Daigger, and B. W. Newbry, Operational experience with BNR facilities in the Chesapeake Bay and Long Island Sound regions of the USA. *Proceedings of the European Conference on Nutrient Removal From Wastewater*, Leeds University, United Kingdom, 1992.
57. Sedlak, R. I., Editor, *Phosphorus and Nitrogen Removal from Municipal Wastewater: Principles and Practice*, 2nd ed., Lewis Publishers, Ann Arbor, Michigan, 1991.
58. Sen, D., C. W. Randall, and T. J. Grizzard, Alkalinity and turbidity measurements in feedback controls to improve performance of a biological nitrogen and phosphorus removal system. *Instrumentation, Control and Automation of Water and Wastewater Treatment and Transport Systems* (R. Briggs, ed.), Pergamon Press, Oxford, pp. 291–298, 1990.
59. Sen, D., C. W. Randall, T. J. Grizzard, and W. R. Knocke, Impact of solids handling units on the performance of biological phosphorus and nitrogen removal systems. *Residuals Symposium*, 64th Annual Conference of the Water Pollution Control Federation, Toronto, Canada, 1991.
60. Sen, D., C. W. Randall, T. J. Grizzard, and D. R. Rumke, Process design and operational modifications of oxidation ditches for biological nutrient removal. *Water Science and Technology* **25**(4/5):249–256, 1992.
61. Shao, Y. J., J. Crosse, E. Keller, and D. Jenkins., High rate air activated sludge operation at the City of Los Angles Hyperion Wastewater Treatment Plant. *Water Science and Technology* **25**(4/5):75–87, 1992.
62. Siebritz, I. P., G. A. Ekama, and G. v. R. Marais, A parametric model for biological excess phosphorus removal. *Water Science and Technology* **15**(3/4):127–152, 1983.
63. Silverstein, J. and E. D. Schroeder, Performance of SBR activated sludge processes with nitrification/denitrification. *Journal, Water Pollution Control Federation* **55**:377–384, 1983.

64. Skalsky, D. S. and G. T. Daigger, Wastewater solids fermentation for volatile acid production and enhanced biological phosphorus removal. *Water Environment Research* **67**: 230–237, 1995.
65. Tetreault, M. J., A. H. Benedict, C. Kaempfer, and E. D. Barth, Biological phosphorus removal: a technology evaluation. *Journal, Water Pollution Control Federation* **58**: 823–837, 1986.
66. U.S. Environmental Protection Agency, *Process Design Manual for Nitrogen Control*, U.S. Environmental Protection Agency, Washington, D.C., 1975.
67. U. S. Environmental Protection Agency, *Design Manual for Phosphorus Removal*, EPA-625/1–87/001, U. S. Environmental Protection Agency, Washington, D.C., 1987.
68. U.S. Environmental Protection Agency, *Process Design Manual for Nitrogen Control*, EPA/625/R-93/010, U.S. Environmental Protection Agency, Washington, D.C., 1993.
69. van Huyssteen, J. A., J. L. Barnard, and J. Hendriksz, The Olifantsfontein nutrient removal plant. *Water Science and Technology* **22**(7/8):1–8, 1990.
70. Wable, M. V. and C. W. Randall, Investigation of hypothesized anaerobic stabilization mechanisms in biological nutrient removal systems. *Water Environment Research* **66**: 161–167, 1994.
71. Walsh, T. K., B. W. Behrman, G. W. Weil, and E. R. Jones, A review of biological phosphorus removal technology. Paper Presented at the 56th Annual Conference of the Water Pollution Control Federation, Atlanta, Georgia, 1983.
72. Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992.
73. Water Pollution Control Federation, *Nutrient Control*, Manual of Practice No. FD-7, Water Pollution Control Federation, Alexandria, Virginia, 1983.
74. Water Pollution Control Federation, *Operation of Municipal Wastewater Treatment Plants*, Manual of Practice No. 11, Water Pollution Control Federation, Alexandria, Virginia, 1990.
75. Water Research Commission, *Theory, Design and Operation of Nutrient Removal Activated Sludge Processes*, Water Research Commission, Pretoria, South Africa, 1984.
76. Wentzel, M. C. and G. A. Ekama, Principles in the design of single sludge activated sludge systems for biological removal of carbon, nitrogen, and phosphorus. *Proceedings of the Water Environment Federation 69th Annual Conference and Exposition, Volume I, Wastewater Treatment Research and Municipal Wastewater Treatment*, Water Environment Federation, Alexandria, Virginia, pp. 681–694, 1996.
77. Wentzel, M. C., P. L. Dold, G. A. Ekama, and G. v. R. Marais, Kinetics of biological phosphorus release. *Water Science and Technology* **17**(11/12):57–71, 1985.
78. Wentzel, M. C., P. L. Dold, G. A. Ekama, and G. v. R. Marais, Enhanced polyphosphate organism cultures in activated sludge systems. Part III. Kinetic model. *Water SA* **15**: 89–102, 1990.
79. Wentzel, M. C., G. A. Ekama, and G. v. R. Marais, Processes and modelling of nitrification denitrification biological excess phosphorus removal systems—A review. *Water Science and Technology* **25**(6):59–82, 1992.
80. Wentzel, M. C., G. A. Ekama, P. L. Dold, and G. v. R. Marais, Biological excess phosphorus removal—Steady state process design. *Water SA* **16**:29–48, 1990.
81. Wentzel, M. C., G. A. Ekama, R. E. Loewenthal, P. L. Dold, and G. v. R. Marais, Enhanced polyphosphate organism cultures in activated sludge systems. Part II. Experimental behavior. *Water SA* **15**:71–88, 1989.
82. Wentzel, M. C., R. E. Loewenthal, G. A. Ekama, and G. v. R. Marais, Enhanced polyphosphate organism cultures in activated sludge systems. Part I. Enhanced culture development. *Water SA* **14**:81–92, 1988.

83. Wilson, A. W., T. Marstaller, P. L. Dold, m. J. Tetreault, and R. W. DeFore, Use of wastewater treatment process simulator as a tool for plant upgrading. *Proceedings of the Water Environment Federation 68th Annual Conference & Exposition, Volume I, Wastewater Treatment Research and Municipal Wastewater Treatment*, Water Environment Federation, Alexandria, Virginia, pp. 559–569, 1995.
84. Yoder, M. W., T. J. Simpkin, G. T. Daigger, and L. M. Morales, Denitrification Trio. *Water Environment Technology* 7(2):50–54, 1995.

12
Aerobic Digestion

The term aerobic digestion refers to the use of aerobic bioreactors to stabilize particulate organic matter arising from primary clarification (predominantly biodegradable organic matter) and biological treatment (predominantly biomass) of wastewaters. The solids are oxidized using either dissolved oxygen or nitrate-N as the terminal electron acceptor. The resulting residue consists primarily of a relatively inert, humus-like material that degrades quite slowly (months to years) in both aerobic and anaerobic environments. The destruction of pathogens is also an objective in municipal wastewater treatment facilities.

12.1 PROCESS DESCRIPTION

Aerobic digestion has been used for several decades to stabilize the waste solids produced at municipal and industrial wastewater treatment plants.[40] Its popularity increased throughout the 1960s and into the 1970s because of its simplicity and lower capital cost relative to anaerobic digestion. Although it had previously been used primarily in small wastewater treatment plants, during this period it was also used in medium to large facilities. This trend was halted in the mid 1970s as rapidly escalating energy costs adversely impacted its overall cost-effectiveness relative to other solids stabilization options. Then, in 1979, federal regulations governing the management of solids from municipal wastewater treatment plants were issued that set new requirements controlling pathogens when solids are to be reused.[24,39] This further decreased the attractiveness of aerobic digestion since its rates of pathogen inactivation are generally lower than anaerobic digestion. Nevertheless, aerobic digestion remained a popular option for small wastewater treatment plants because of its simplicity. In addition, thermophilic aerobic digestion processes, which have higher solids stabilization and pathogen inactivation rates, were developed in the late 1970s and early 1980s.[41,42] Consequently, aerobic digestion remains a viable option for the stabilization of waste solids at many wastewater treatment plants.

12.1.1 General Description

Figure 12.1 summarizes the biochemical transformations occurring in an aerobic digester. Biodegradable particulate organic matter is hydrolyzed and converted into biodegradable soluble organic matter, releasing nutrients such as ammonia-N and phosphate. The biodegradable soluble organic matter is then converted into carbon dioxide, water, and active biomass through the action of heterotrophic bacteria. The

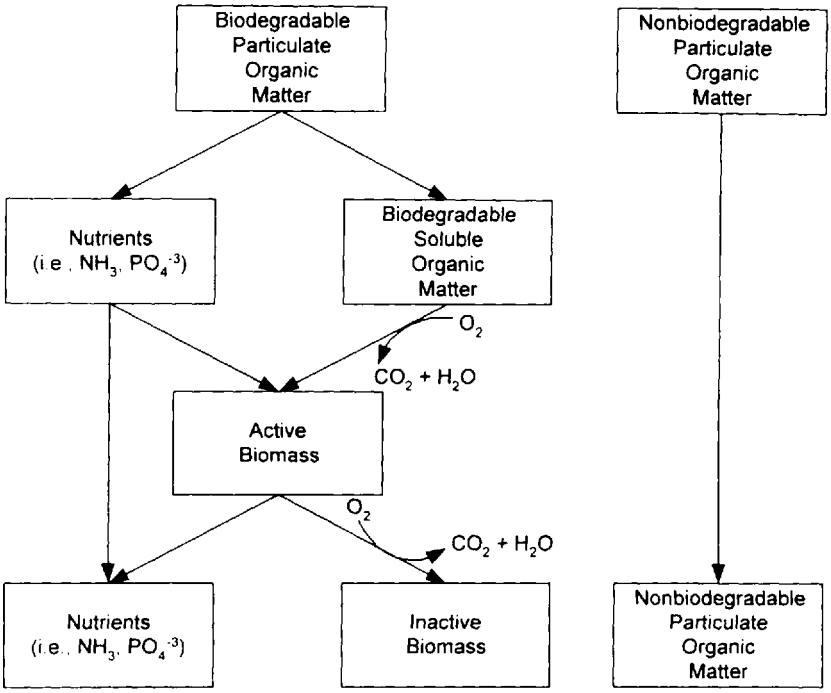


Figure 12.1 Schematic diagram of the events occurring during aerobic digestion.

active biomass, in turn, undergoes decay, resulting in the generation of additional carbon dioxide and water, along with inactive biomass, i.e., debris. Nonbiodegradable particulate organic matter in the influent is not affected by the digestion process and becomes a portion of the digested solids. Figure 12.1 is based on the traditional decay model for biomass destruction, as discussed in Section 3.3.1. The lysis:regrowth model, described in Section 3.3.2, is equally applicable and, in fact, International Association on Water Quality activated sludge model (IAWQ ASM) No. 1, with its explicit treatment of hydrolysis, nitrification, and denitrification, provides a more accurate description for some aerobic digestion process options. Nevertheless, the simplified models often used to design aerobic digesters are directly related to the traditional decay model, and thus it is emphasized herein.

Observations of aerobic digestion processes provide the following conceptual framework upon which design models are based:

- The suspended solids in the influent stream can be segregated into biodegradable and nonbiodegradable components.[14,35] The biodegradable components include particulate organic matter, X_S, and active biomass, both heterotrophic and autotrophic, $X_{B,H}$ and $X_{B,A}$. The nonbiodegradable component consists of particulate inert organic matter, X_I, and biomass debris, X_D.
- A nonbiodegradable residue will result from aerobic digestion, even if no nonbiodegradable particulate matter is present in the influent solids stream because biomass debris results from the decay of active biomass.[35,44]

- Aerobic digestion results in the destruction of both volatile suspended solids (VSS) and fixed suspended solids (FSS).[7,35] This occurs because both the organic and inorganic materials in the biodegradable suspended solids are solubilized and/or oxidized as the solids are digested. However, the volatile and fixed components of the biodegradable and nonbiodegradable suspended solids are not equal. Consequently, VSS and FSS will not generally be destroyed in the same proportion. However, in spite of the loss of fixed solids during aerobic digestion, most designers focus on loss of VSS.
- The biodegradable fraction of solids is a function of their source.[25,35,44] This is clearly illustrated by the models discussed throughout previous chapters. For example, both primary solids and waste activated sludge from a system with a short SRT will contain relatively high fractions of biodegradable material, whereas waste activated sludge from a system with a long SRT will contain a low fraction of biodegradable material and a high fraction of biomass debris.
- The destruction of biodegradable suspended solids can be characterized as a first order reaction.[14,22,35]. This occurs because the decay of active biomass is a first order reaction. Biodegradable particulate organic matter is rapidly converted to active biomass. Then that biomass, as well as any active biomass present in the influent, decays in a first order manner, resulting in an overall first order reaction for loss of biodegradable suspended solids. As a result of this relationship, the destruction of biodegradable suspended solids is often referred to as decay, and the first-order reaction rate coefficient is called a decay coefficient.
- For solids containing a relatively high proportion of active biomass, the value of the decay coefficient for biodegradable suspended solids is relatively independent of the SRT at which the waste solids were produced.[25,35,44] This is because the decay coefficient for the biodegradable suspended solids will be influenced strongly by the decay coefficient for heterotrophic bacteria, which is relatively constant.

Mathematically, these relationships can be summarized as follows. Because design of aerobic digesters is typically concerned with VSS destruction, the concentrations will be expressed as VSS, although it should be recognized that they could also be expressed as total suspended solid (TSS) or chemical oxygen demand (COD). The VSS undergoing aerobic digestion, $X_{M,V}$, can be subdivided into biodegradable and nonbiodegradable components, $X_{M,V,b}$ and $X_{M,V,n}$, respectively:

$$X_{M,V} = X_{M,V,b} + X_{M,V,n} \tag{12.1}$$

Furthermore, loss of biodegradable VSS occurs in a first-order manner:

$$r_{XMVb} = -b_{MV} \cdot X_{M,V,b} \tag{12.2}$$

where b_{MV} is the first-order decay coefficient based on the loss of VSS. Decay coefficients can also be determined on the basis of COD or TSS, depending upon how the solids concentration is expressed. Because the COD/VSS ratio can be considered to be constant, the decay coefficients based on VSS and COD loss have the same numerical values. However, because VSS and FSS are not generally destroyed in proportion to each other, the decay coefficient on a TSS basis will have a different

numerical value. The nonbiodegradable solids are considered to be totally inert so that nothing happens to them during aerobic digestion.

Sometimes it is desirable to consider the solids in terms of the constituents used in the simplified model of Chapter 5 and in ASM No. 1. In those terms, the VSS consists of:

$$X_{MV} = X_{S,V} + X_{B,H,V} + X_{B,A,V} + X_{D,V} + X_{I,V} \tag{12.3}$$

All symbols have been defined previously; the subscript V simply indicates that they are expressed on a VSS basis. The particulate substrate and the biomass are considered to be biodegradable, although decay of biomass leads to debris. The debris and the inert material are, of course, nonbiodegradable. The fate of these components can be modeled in ASM No. 1, just as the various activated sludge systems are modeled in Chapter 7. This has been done by Marais and coworkers[25,44] using the model upon which much of ASM No. 1 was based. The simplified model of Chapter 5 does not contain a term for particulate substrate and it is not as accurate as ASM No. 1. Nevertheless, it has been used to model aerobic digestion by lumping the particulate substrate with the biomass and considering the system to be a bioreactor receiving only biomass and nonbiodegradable solids. In that situation, the autotrophic biomass is generally neglected as being insignificant. This is done in Section 5.2.3 for COD units, with Eqs. 5.64 and 6.65 expressing the heterotrophic biomass and debris concentrations, respectively, in a single continuous stirred tank reactor (CSTR). Equation 5.68 gives the oxygen requirement for such a bioreactor when nitrification is not considered. Figure 5.10 shows the effects of solids retention time (SRT) on the theoretical performance of a single CSTR receiving only active heterotrophic biomass and debris. There it can be seen that a point of diminishing returns is reached at which further increases in the SRT have little effect. When that point is reached, most of the suspended solids will be nonbiodegradable, with only a small fraction of active biomass.

Since the purpose of aerobic digestion is to stabilize the biodegradable organic matter in an influent waste solids stream, criteria must be available for quantifying the degree of stabilization. Although several could be proposed, two frequently used ones are the VSS destruction efficiency (expressed as the percent VSS reduction) and the specific oxygen uptake rate (SOUR) of the digested solids (typically expressed as mg O_2/(g VSS · hr)).[12] A VSS reduction of 38% and a SOUR of 1.0 to 1.5 mg O_2/(g VSS · hr) are values typically used to represent stabilized solids.[24,30]

The simple first-order model presented above can be used to estimate the effect of the size of a completely mixed aerobic digester on the degree of solids stabilization achieved. The VSS destruction efficiency, E_{XMV}, in a CSTR can be calculated as a function of its SRT:

$$E_{XMV} = 100\left[\frac{X_{M,V,O} - X_{M,V,bO}}{X_{M,V,O}}\right]\left[\frac{b_{MV}\cdot\Theta_c}{1+b_{MV}\cdot\Theta_c}\right] \tag{12.4}$$

where the subscript O represents the influent concentrations. From Eq. 12.1:

$$X_{M,V,bO} = X_{M,V,O} - X_{M,V,bO} \tag{12.5}$$

Thus, it can be seen that the term $X_{M,V,O} - X_{M,V,bO}$ is just the concentration of biodegradable VSS entering the digester. Consideration of Eq. 12.4 in the limit as the SRT becomes very large reveals that the first bracketed term on the right side rep-

resents the highest possible VSS destruction efficiency since it is just the fraction of the influent VSS that can be degraded biologically. It is an important determinant of whether a target destruction efficiency of 38% can be economically achieved.

Calculation of the actual VSS destruction efficiency in an operating digester should be based on the mass flow rates of VSS entering and leaving it. While some sources[41] use the percent VSS content of the feed and effluent solids to make this calculation, such a procedure can give an inaccurate measure because of changes in the FSS as discussed above. Consequently, it should not be used.

The SOUR for a completely mixed aerobic digester can also be calculated as a function of the SRT:

$$\text{SOUR} = 1000 \left[\frac{b_{MV} \cdot i_{O\,XM,V}(X_{M,VO} - X_{M,V,nO})}{(X_{M,VO} - X_{M,V,nO}) + X_{M,V,nO}(1 + b_{MV} \cdot \Theta_c)} \right] \tag{12.6}$$

where $i_{O\,XM,V}$ is the conversion factor from VSS units to oxygen units. The numerical value for $i_{O\,XM,V}$ depends on whether the ammonia-N released through digestion of biodegradable VSS is nitrified in the aerobic digester. If the biodegradable VSS can be assumed to have the same elemental composition as biomass, i.e., $C_5H_7O_2N$, then the values of $i_{O\,XM,V}$ can be determined theoretically. If the released ammonia-N is not oxidized to nitrate-N by autotrophic biomass, the value will be that given in Table 3.1, or 1.42 g O_2/g VSS. If, on the other hand, the released ammonia-N is nitrified, the value of $i_{O\,XM,V}$ will be increased by 40%, to 1.98 g O_2/g VSS. Table 12.1 summarizes these values as well as the values the conversion factors would have if the solids concentrations were expressed in COD or TSS units. If the biodegradable solids cannot be assumed to have the same elemental composition as biomass, the values of the conversion factor must be determined experimentally.

Although Eqs. 12.4 and 12.6 are for a CSTR, many aerobic digester studies are done in batch reactors. The VSS concentration in a batch reactor at any time t can be estimated by writing the mass balance equation on biodegradable VSS with Eq. 12.2 as the loss term, integrating it, and adding the initial nonbiodegradable VSS concentration to give the total VSS concentration:

$$X_{M,V} = X_{M,V,nO} + (X_{M,VO} - X_{M,V,nO})\exp(-b_{MV} \cdot t) \tag{12.7}$$

This equation indicates that the VSS concentration will decline exponentially over time, but will approach a residual concentration equal to the nonbiodegradable solids.

The effects described in this section have been observed experimentally, as illustrated by the data of Reece, et al.[35] who performed aerobic digestion studies in batch reactors using solids produced in an activated sludge system. Since the wastewater was totally biodegradable, the waste solids consisted primarily of active biomass and biomass debris. The SRT of the activated sludge system in which the waste

Table 12.1 Oxygen Mass Equivalents for Biomass

Condition	$i_{O\,XM}$ g O_2/g COD	$i_{O\,XM,V}$ g O_2/g VSS	$i_{O\,XM,T}$ g O_2/g TSS
With nitrification	1.40	1.98	1.68
Without nitrification	1.00	1.42	1.20

solids were produced was varied to produce waste solids with varying proportions of active and inactive biomass. As indicated in Figure 12.2, both VSS and FSS (referred to in the figure as nonvolatile suspended solids [NVSS]) were destroyed during aerobic digestion. In fact, the proportions of VSS and FSS destroyed were similar. However, Figure 12.3 shows that the decay coefficient was relatively constant and independent of the SRT of the activated sludge system in which the waste solids were generated, although its numerical value depended on whether it was quantifying the loss of TSS or VSS. In contrast, the nonbiodegradable suspended solids content of the waste solids increased as the SRT of the activated sludge system was increased, as illustrated in Figure 12.4. Similar effects have been observed by others.[12,44]

12.1.2 Process Options

Three basic aerobic digestion process options exist. They are conventional aerobic digestion (CAD), anoxic/aerobic digestion (A/AD), and autothermal thermophilic aerobic digestion (ATAD).

Conventional Aerobic Digestion. Conventional aerobic digestion is quite simple. It consists of the addition of solids to an aerated vessel and their retention there for a period of time equal to the SRT. Figure 12.5 illustrates two CAD bioreactors, one with intermittent and one with continuous addition of feed solids.

In the intermittent process, Figure 12.5a, solids are added and removed from the digester periodically, usually once per day. This process is used in conjunction

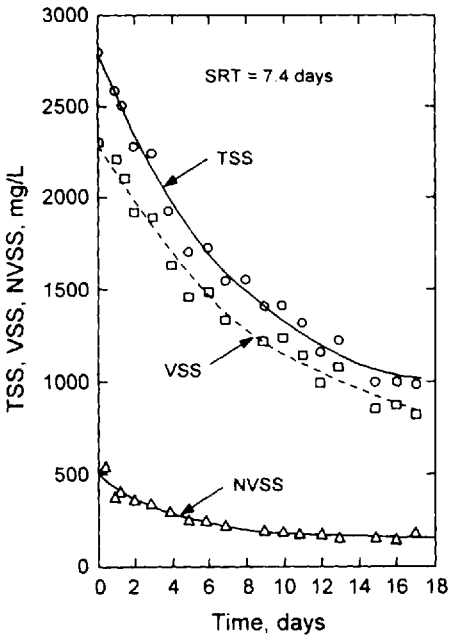


Figure 12.2 Destruction of TSS, VSS, and FSS (called NVSS) during batch aerobic digestion of waste activated sludge. (From C. S. Reece, et al., Aerobic digestion of waste activated sludge. *Journal of the Environmental Engineering Division, ASCE* **105**:261–272, 1979. Copyright © American Society of Civil Engineers; reprinted with permission.)

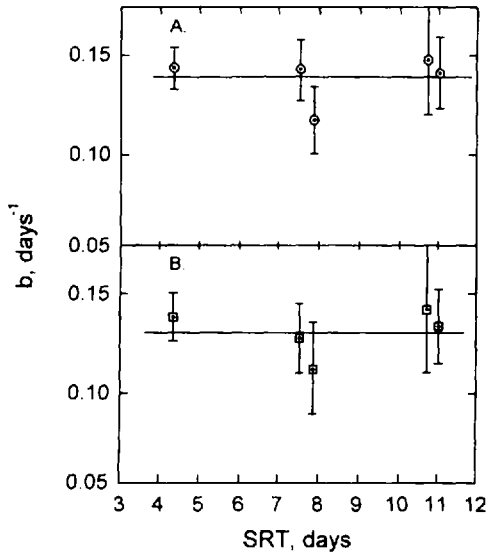


Figure 12.3 Effect of the SRT of an activated sludge system on the decay coefficient describing aerobic digestion of the resulting waste solids. A. TSS basis; B. VSS basis. (From C. S. Reece, et al., Aerobic digestion of waste activated sludge. *Journal of the Environmental Engineering Division, ASCE* **105**:261–272, 1979. Copyright © American Soceity of Civil Engineers; reprinted with permission.)

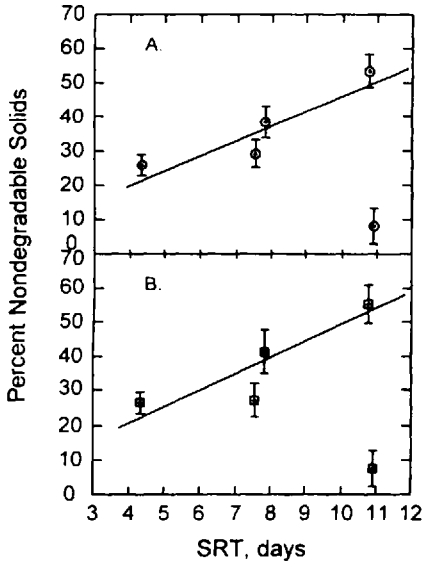


Figure 12.4 Effect of the SRT of an activated sludge system on the nonbiodegradable suspended solids content of the resulting waste solids. A. TSS basis; B. VSS basis. (From C. S. Reece, et al., Aerobic digestion of waste activated sludge. *Journal of the Environmental Engineering Division, ASCE* **105**:261–272, 1979. Copyright © American Society of Civil Engineers; reprinted with permission.)

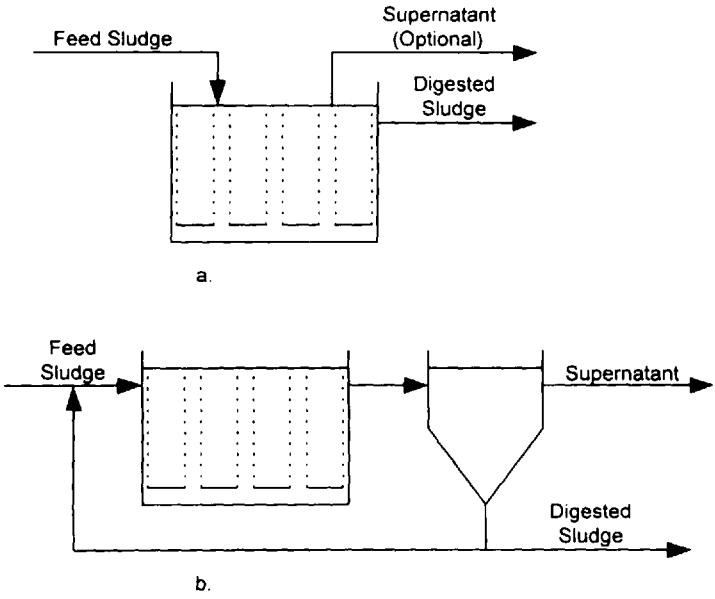


Figure 12.5 Conventional aerobic digestion: a. Intermittent feed; b. continuous feed.

with biological wastewater treatment systems in which solids are wasted on a daily basis, usually over a relatively short time period. Digested solids are removed from the digester as necessary, depending on the downstream solids handling system. In some moderate size wastewater treatment facilities, solids are thickened prior to addition to the digester. In those cases, the SRT of the digester will be equal to its hydraulic retention time (HRT). When thickening is not used, the aerobic digester is operated like a sequencing batch reactor (SBR) to provide both solids thickening and digestion. Consequently, the SRT is greater than the HRT. The typical steps for operation of an SBR are shown in Figure 7.42. In this application, digested solids are withdrawn for further processing at the end of the supernatant draw period. A suspended solids concentration in the range of 12,500 to 17,500 mg/L can typically be achieved if waste activated sludge (WAS) is being digested. Somewhat higher concentrations can be achieved if a mixture of primary solids and WAS (15,000 to 25,000 mg/L) or primary solids alone (30,000 to 40,000 mg/L) is being digested. Although feeding is intermittent, feed is added many times during one SRT, making the bioreactor perform like a CSTR rather than like an SBR.

Solids may also be wasted from a biological wastewater treatment system on a more continuous basis, a practice often used in larger plants. Figure 12.5b illustrates an aerobic digestion system that receives feed on a continuous basis. It looks like an activated sludge system, with feed solids displacing digesting solids to a gravity thickener. Supernatant overflows the thickener, while thickened solids are withdrawn from its bottom and returned to the digester. Thickened solids are also periodically directed to solids handling, with the rate of thickened solids removal being adjusted to maintain the desired SRT. Suspended solids concentrations in the thickened solids are similar to those achieved with the intermittent feed process. Consequently, suspended solids concentrations within the continuous feed digester are lower.

Other operating modes are possible for the intermittent and the continuous feed systems. For example, because VSS destruction is first order, arrangement of aerobic digesters in series can increase the efficiency of the process, as discussed in Section 12.2.5.

Conventional aerobic digesters are constructed using facilities and equipment similar to those used for activated sludge systems. In fact, the aerobic digester may simply be one or more of the aeration basins provided in an activated sludge system. The bioreactors can be concrete or steel, or they can be lined earthen basins. Submerged aeration systems, such as diffused air, may be more desirable in cold climates than mechanical surface aerators because they minimize heat losses, which can be quite significant because of the relatively long HRTs. However, surface aeration systems can be used, particularly in warm climates. Decant devices are required for intermittently fed digesters and several approaches have been successfully utilized. SRTs on the order of 20 days will usually produce a significant VSS destruction efficiency, although longer SRTs may be required to achieve a SOUR less than 1 mg O_2/(g VSS·hr). Relatively long SRTs may also be required to meet pathogen destruction requirements, depending on the operating temperature.

Anoxic/Aerobic Digestion. One difficulty often experienced with the CAD process is destruction of alkalinity by nitrification, as discussed in Section 6.3.3. The destruction of organic matter, particularly active biomass, results in liberation of organic nitrogen as ammonia-N. Because of the long SRTs required to accomplish solids stabilization, nitrifying bacteria typically grow in the digester even if they are not present in the feed solids. Furthermore, because of the relatively high feed solids concentrations typically used, the ammonia-N concentrations that develop are high, causing nitrification to deplete the alkalinity in the system, dropping the pH. Figure 12.6 illustrates typical pH profiles during digestion of a WAS.[30] Three operational modes and three temperatures were considered in this study. Two modes involved feeding of the digester, either on a continuous or a daily basis. The third was batch operation with only one feed addition. In all cases the pH dropped to relatively low values (4.5 to 5.5), which are typical of those that can be experienced in aerobic digesters without pH control. Although digestion will continue at these lower pH values, the rate will be reduced. The pH of the digester can be adjusted through the addition of bases such as lime, but this is an additional operating cost.

Just as is done with anoxic selectors (Section 11.3.1), A/AD incorporates an anoxic cycle to allow alkalinity production by denitrification to partially offset that consumed through nitrification.[44] An additional benefit is a small reduction in process energy requirements since some of the organic matter is oxidized by using nitrate-N generated through nitrification, rather than oxygen, as the electron acceptor.

As discussed above, digestion of cellular material using dissolved oxygen as the terminal electron acceptor, with concurrent oxidation of the ammonia-N to nitrate-N, leads to the net destruction of alkalinity. When the production of autotrophic biomass is neglected (since it is small), the molar stoichiometry is:

$$C_5H_7O_2N + 7\ O_2 + HCO_3^- \rightarrow 6\ CO_2 + NO_3^- + 4\ H_2O \qquad (12.8)$$

Consequently, one mole of bicarbonate alkalinity is destroyed for each mole of biomass destroyed. Furthermore, seven moles of oxygen are used for each mole of biomass destroyed, which is equivalent to 1.98 g O_2/g VSS destroyed as shown in Table 12.1. However, if a situation could be devised so that all of the nitrogen released was

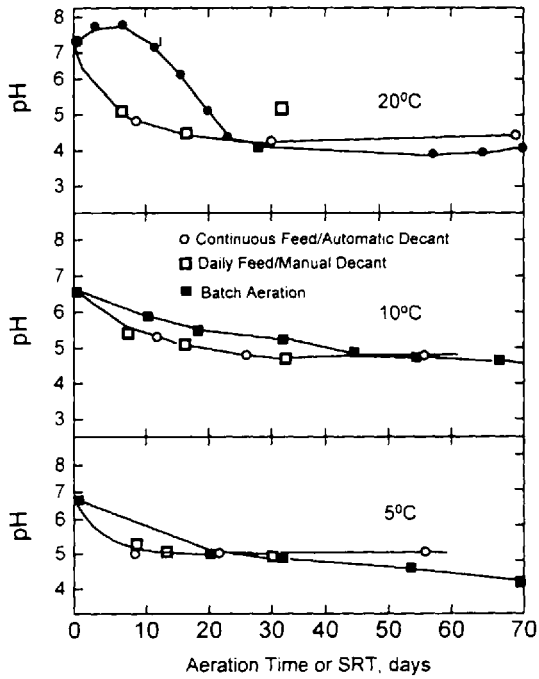


Figure 12.6 Depression of the pH during aerobic digestion. (From D. S. Mavinic and D. A. Koers, Fate of nitrogen in aerobic digesters. *Journal, Water Pollution Control Federation* **54**: 352–360, 1982. Copyright © Water Environment Federation; reprinted with permission.)

nitrified and then converted to nitrogen gas through denitrification with biomass as the electron donor, the overall molar stoichiometry would be:

$$4\ C_5H_7O_2N + 23\ O_2 \rightarrow 20\ CO_2 + 2\ N_2 + 14\ H_2O \qquad (12.9)$$

In that case there would be no alkalinity destruction and the oxygen requirement would be reduced to 5.75 moles per mole of biomass destroyed, or 1.63 g O_2/g VSS destroyed, which represents an 18% savings compared to the fully aerobic process. Considerations such as these led to the concept of the A/AD process. Evidence indicates that pH values near neutrality can be maintained with this process if a significant degree of denitrification is obtained.

Figure 12.7 illustrates A/AD process options. The option in Figure 12.7a is a modification of intermittent CAD in which the oxygen transfer system is cycled on and off to create aerobic and anoxic periods during the digester operational cycle. Mixers may be provided to maintain solids in suspension during the anoxic periods, but if a sufficiently high suspended solids concentration is maintained in the bioreactor, only limited settling will occur, making mixing unnecessary. Other options involve the use of separate anoxic and aerobic zones with recirculation of mixed liquor from the aerobic to the anoxic zone, just as in the modified Ludzak–Ettinger (MLE) process, as shown in Figures 12.7b and 12.7c. SRTs and suspended solids concentrations in A/AD systems are similar to those in CAD systems. Furthermore, the facilities required for A/AD systems are similar to those used with CAD systems,

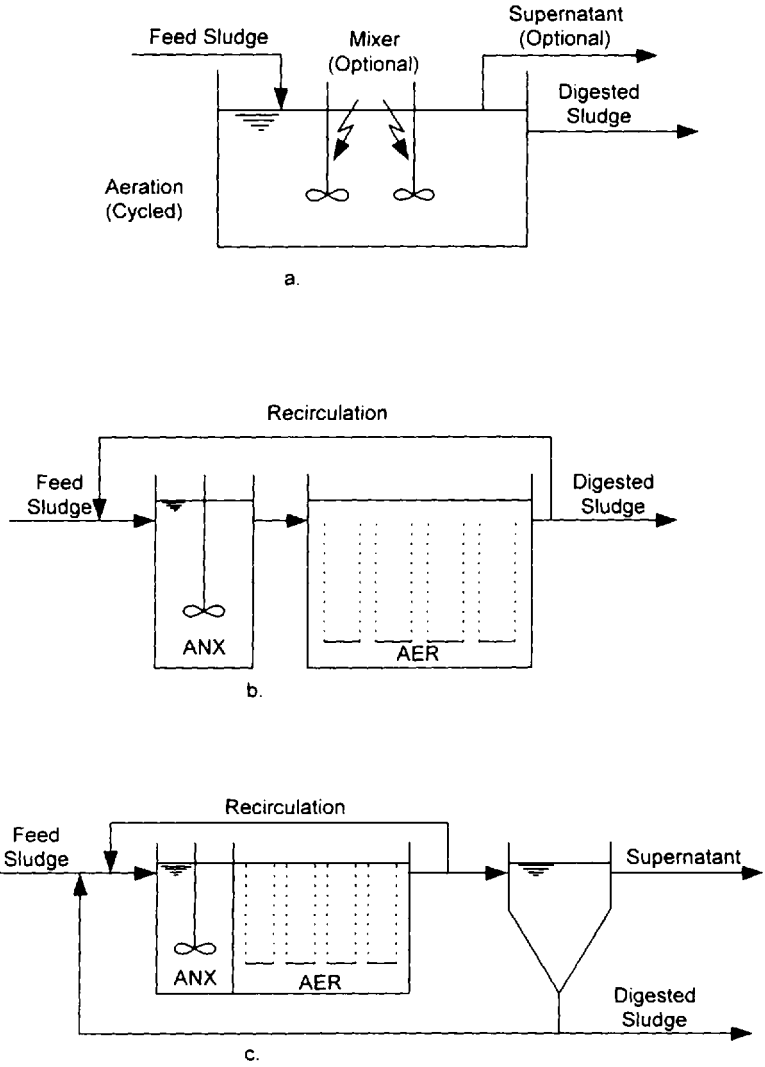


Figure 12.7 Anoxic/aerobic digestion: a. intermittent feed; b. continuous feed without thickening; c. continuous feed with thickening.

with the exception of the addition of mixing equipment where necessary for anoxic operation.

Because of the need to recirculate nitrified mixed liquor for denitrification, as shown in Figure 12.7, complete denitrification will not generally be achieved in continuous feed A/AD processes. Consequently, the reductions in alkalinity destruction and oxygen utilization will not be as great as suggested by Eq. 12.9. This means that chemical addition for pH control may still be required for locations with very low alkalinity water. Nevertheless, the implementation of A/AD is still worthwhile because of the savings involved.

Research has demonstrated the practicality of the A/AD process. It has also indicated that solids destruction rates under anoxic conditions are less than the rates under aerobic conditions. This phenomenon can be explained adequately using ASM No. 1 with its reduced rate of hydrolysis of particulate substrate under anoxic conditions. However, application of this model also indicates that the overall stabilization of organic matter in the A/AD process will be the same as achieved in a CAD system of equal SRT. Although a build-up of particulate substrate occurs in the anoxic zone of an A/AD process, it is rapidly oxidized in the subsequent aerobic zone.[19,20] The rate of degradation of active biomass to particulate substrate, which is rate limiting, is the same in the anoxic and aerobic zones.

Autothermal Thermophilic Aerobic Digestion. As discussed in Section 10.2.7, heat is released when organic matter is oxidized, and Table 12.2 summarizes the amount associated with biomass destruction.[42] Autothermal thermophilic aerobic digestion (ATAD) takes advantage of that heat to elevate the temperature of an aerobic digester. CAD does not experience a significant temperature increase because of the large mass of water flowing through the system, the conductive heat loss through the walls of the bioreactor, and the evaporative heat loss associated with operation of the oxygen transfer system. In ATAD elevated temperatures are achieved by thickening the feed solids to a concentration of 40,000–60,000 mg/L to minimize the mass of water that must be heated, covering and insulating the bioreactor to minimize conductive heat losses, and using a high efficiency oxygen transfer device to minimize evaporative heat losses. As a result, it is practical to achieve bioreactor temperatures in the thermophilic range (45° to 65°C) without external heating. Smaller bioreactors may be used for ATAD because of the smaller feed volumes resulting from the thickening of the feed solids and the higher digestion rates associated with the elevated temperature. The elevated temperatures also accelerate the inactivation rate of pathogens.[8,38,42]

Figure 12.8 presents a schematic of ATAD. Two tanks in series are often used; significant digestion and heating occur in the first tank, with further digestion and heating to a temperature of about 55°C in the second tank. Feeding is often intermittent, with removal of digested solids from the second tank, transfer of digesting solids from the first to the second tank, and addition of feed solids to the first tank. This promotes temperature elevation and minimizes short-circuiting of feed solids to the digested solids, thereby giving better pathogen inactivation.

The bioreactors may be constructed of steel or concrete, but covers and insulation are generally required to control conductive heat losses. Proprietary oxygen transfer devices achieve the necessary oxygen transfer while minimizing associated heat losses. Bioreactor geometry is typically constrained by the requirements of these devices. High-purity oxygen may also be used for oxygen transfer, resulting in even

Table 12.2 Energy Released by Oxidation of Biomass

Units	COD basis	VSS basis
kJ/kg	13,200	18,800
kcal/kg	3,500	5,000

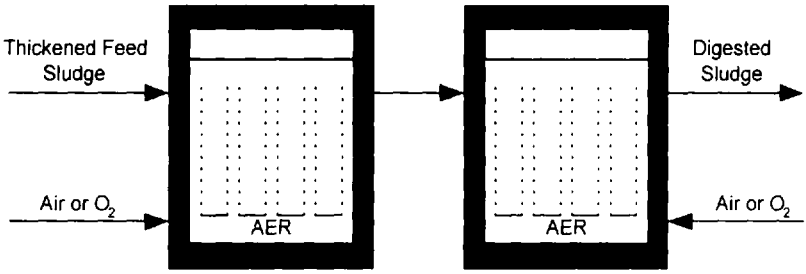


Figure 12.8 Autothermal aerobic digestion.

less heat loss.[27,38] Foaming can be severe because of the high feed-solids concentrations and high temperatures. Consequently, specialized foam control devices are provided with many designs. Nitrification of the released ammonia-N typically does not occur because the elevated temperature minimizes the growth of nitrifying bacteria, which are mesophilic. Thus, oxygen requirements are reduced compared to CAD and pH depression is generally not a problem. In fact, the destruction of biomass in the absence of nitrification produces alkalinity, which results in pH values in the 7.5 to 8.0 range. This increased pH, coupled with elevated ammonia-N concentrations, can result in increased inactivation of viruses.[37] Because of the accelerated destruction rates of VSS and pathogens, SRTs on the order of 5 to 6 days are often used. Although the concept of ATAD was originally developed in the U.S.,[18,19,27] most applications are in Europe.[11,42]

Autoheating can also occur in CAD systems if the feed-solids concentration exceeds 20,000 mg/L, the feed-solids are sufficiently biodegradable, and a high-efficiency oxygen transfer system is used.[18,26] In such cases the temperature in the digester will depend on the heat loss characteristics of the bioreactor and the oxygen transfer device, along with ambient temperatures. Significant variations in digester operating temperature will occur in locations with significant seasonal ambient temperature variations. Procedures for computing the heat balance necessary to estimate temperatures in aerobic digesters have been presented elsewhere.[31,43]

Another variation is the dual digestion process,[5,27,42] which uses a single-stage, high-rate ATAD system to heat solids for subsequent feeding to an anaerobic digester, as shown in Figure 12.9. Pure oxygen is typically used to provide oxygen to the ATAD unit, thereby minimizing heat losses from it. As a consequence, an HRT of about 1 day can be used to heat the solids to a temperature of 55 to 65°C. The short HRT minimizes the mass of biodegradable organic matter oxidized aerobically, thereby maximizing the mass fed to the downstream anaerobic digester. Oxygen feed is regulated to achieve the desired ATAD reactor temperature.[31] Significant solubilization of VSS occurs because of the high ATAD reactor loadings, and COD reduction, rather than VSS destruction, correlates best with heat generation. Excellent pathogen destruction occurs because of the elevated temperatures developed in the ATAD reactor.

12.1.3 Comparison of Process Options

Table 12.3 summarizes the benefits and drawbacks of the various aerobic digestion process options. CAD is a demonstrated, proven process. It is mechanically simple,

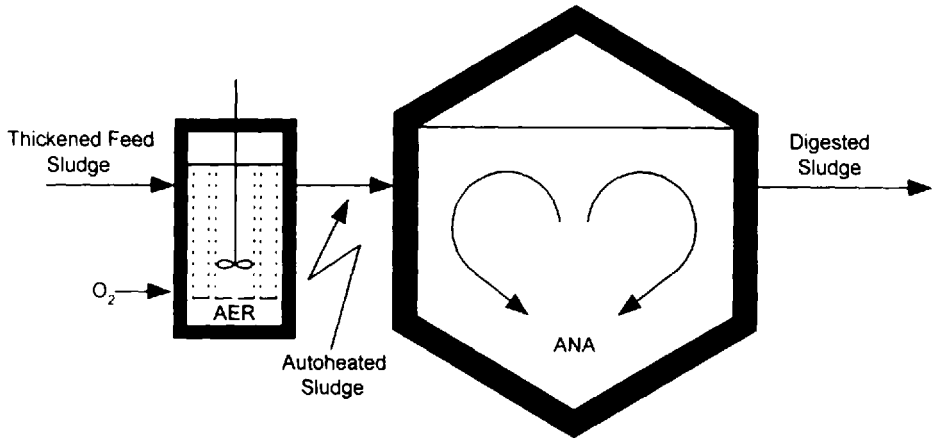


Figure 12.9 Dual digestion process.

and it is simple to operate. It is possible to incorporate both solids thickening and digestion into a single vessel, and the supernatant is of reasonably good quality. In contrast, the power costs for CAD are relatively high, and the rates of pathogen destruction are low. Relatively long SRT's are required, resulting in relatively large tank volumes. The pH will drop because of nitrification, and the solids that are produced will generally not dewater readily by mechanical means. The benefits and drawbacks of A/AD are similar to those of CAD, except that it provides pH control and affords a moderate reduction in power requirements.

ATAD offers significantly reduced bioreactor volumes because of the smaller feed flow rates associated with the thickened solids that must be used and the much lower SRTs that result from the higher operating temperature. Nitrification does not occur, so the pH is generally above neutral and the power requirement is less. The higher temperature, elevated pH, and increased ammonia-N concentration result in greater rates of pathogen inactivation. The dewatering characteristics of the solids may be somewhat better than those obtained with CAD and A/AD, although further work is needed to verify this. In exchange for these benefits, ATAD is mechanically more complex and subject to severe foaming. It is a newer process, and engineers have significantly less experience with it than with CAD. Consequently, its performance and operational characteristics are less predictable. Although power costs are reduced compared to other aerobic digestion processes, they may still be relatively high. Separate thickening of the feed solids is required to achieve a sufficiently high concentration to achieve autothermal conditions. Successful operation also requires a sufficient fraction of biodegradable solids in the feed.

12.1.4 Typical Applications

Conventional aerobic digestion is widely used to stabilize the solids at small- to medium-sized wastewater treatment plants (less than 20,000 to 40,000 m^3/day). Hundreds of examples exist. CAD is utilized in such circumstances because of its mechanical and process simplicity, and because solids thickening and stabilization can

Table 12.3 Aerobic Digestion Process Comparison

Process	Benefits	Drawbacks
Conventional aerobic digestion (CAD)	• Demonstrated process, proven • Mechanically simple • Simple to operate • Both thickening and stabilization can be provided in one vessel • Supernatant is of reasonable quality	• High power cost • Low rates of pathogen inactivation • Long SRTs required • Relatively large reactor volumes • pH drops due to nitrification • Digested solids dewater poorly
Anoxic/aerobic digestion (A/AD)	• pH control provided • Mechanically simple • Simple to operate • Power requirements less than CAD • Both thickening and stabilization can be provided in one vessel • Supernatant is of reasonable quality	• Relatively new process, less experience • Power costs still relatively high • Low rates of pathogen inactivation • Long SRTs required • Relatively large reactor volumes • Digested solids dewater poorly
Autothermal thermophilic aerobic digestion (ATAD)	• Low SRT • Small reactor volume • No pH drop • Excellent pathogen inactivation • Less power required than CAD or A/AD • Dewatering characteristics of digested solids may be better than CAD or A/AD	• Mechanically more complex • Foaming • Newer process, less experience than with CAD • Power costs still relatively high • Requires separate thickening • Requires adequate biodegradable solids in feed

be incorporated into a single vessel. Aerobically digested solids dewater quite poorly by mechanical means, but waste solids from small wastewater treatment plants are often handled in a liquid, not dewatered, form, thereby eliminating the problem. Even though power costs are relatively high, capital costs are relatively low, resulting in generally favorable economics. Pathogen destruction requirements may dictate the use of longer SRTs, thereby increasing capital and operating costs and reducing somewhat the use of CAD. Anaerobic digestion often offers cost advantages in larger wastewater treatment plants. Even though its capital costs are higher, power costs are lower, resulting in more favorable overall economics. Consequently, anaerobic digestion is frequently used in larger wastewater treatment plants.

Operational difficulties exist with CAD, particularly in colder climates where VSS reduction efficiencies are often low and operational difficulties such as freezing of equipment, and even of a portion of the bioreactor contents, can often occur. Design approaches are available to address these difficulties, such as covering and heat tracing of equipment, and thickening of solids prior to digestion to allow some autoheating. Other operational difficulties exist because of poor understanding of the process and the factors that affect its performance. In spite of these difficulties, successful operation and acceptable performance can be obtained through the application of proper design and operational principles, and many successful case histories exist (e.g., see Ref. 34).

Anoxic/aerobic digestion is a relatively new option that is receiving increased attention. Since it generally offers the benefits of CAD, along with minimal pH depression and reduced energy requirements, increased use of A/AD can be expected in the future. Existing CAD processes can easily be modified to operate in the A/AD manner by changing the operation of the oxygen transfer system to an intermittent mode. This can result in energy cost savings, along with improved control over bioreactor pH.

Increased concern over pathogen destruction and the desire for reduced bioreactor volumes is fueling interest in ATAD. Interest in ATAD is high in Europe where nearly 50 plants are in service, and the dual digestion process is used in more than 30 plants. Dual digestion has been used at three locations in the United States and ATAD has been applied in Canada. Additional ATAD applications are expected in North America.

The benefits, drawbacks, and costs of aerobic digestion must be compared to other solids stabilization options and the optimal one selected for a particular application. Nevertheless, it appears that use of aerobic digestion will continue, particularly at the small- to medium-sized wastewater treatment plants where it has traditionally been used. Design of CAD and A/AD processes to achieve autoheating is also likely, thereby providing the benefits of operation at elevated temperature (perhaps still in the mesophilic range) while retaining the benefits of CAD and A/AD.

12.2 FACTORS AFFECTING PERFORMANCE

Many of the factors that affect the performance of aerobic digestion have already been identified. This section summarizes the significant factors and quantifies their effects.

12.2.1 Solids Retention Time and Temperature

Figure 5.10 illustrates the effect of SRT on the performance of a conventional aerobic digester of the type shown in Figure 12.5b, as simulated with the simple model of Chapter 5. There it can be seen that the total solids concentration decreases as the SRT of the digester is increased, which is to be expected. Perhaps more importantly, however, the figure demonstrates that the buildup of biomass debris in the system limits the percent solids destruction that can be achieved. In this case about 53% solids destruction is achieved at an SRT of 1000 hrs, or about 42 days, as shown in Figure 12.10. At that SRT, debris accounts for about 83% of the remaining solids, suggesting that little additional solids destruction will occur, even if the SRT is extended greatly. This can also be seen from the slope of the VSS destruction curve in Figure 12.10. This characteristic should be kept in mind when selecting the SRT for an aerobic digester. A point may quickly be reached at which further increases in SRT will have minimal effect, making expenditures for additional tank volume questionable. That point will depend on the nature of the influent solids, their biodegradability, and the temperature of the system.

Figure 5.10 also shows the effect of SRT on the oxygen requirement of the digester and that curve has been converted into the SOUR, as shown in Figure 12.10. Because the oxygen requirement curve in Figure 5.10 does not consider nitrification of the released ammonia-N, neither does the SOUR curve in Figure 12.10. Should the pH and other factors allow nitrifiers to grow, the SOUR would be 40% greater than shown. Like the VSS destruction curve, the SOUR curve quickly reaches a point where further increases in SRT have little effect. Just how rapidly that occurs also depends on the nature of the influent solids, their biodegradability, and the temperature of the system. This, too, must be considered when an SRT is being

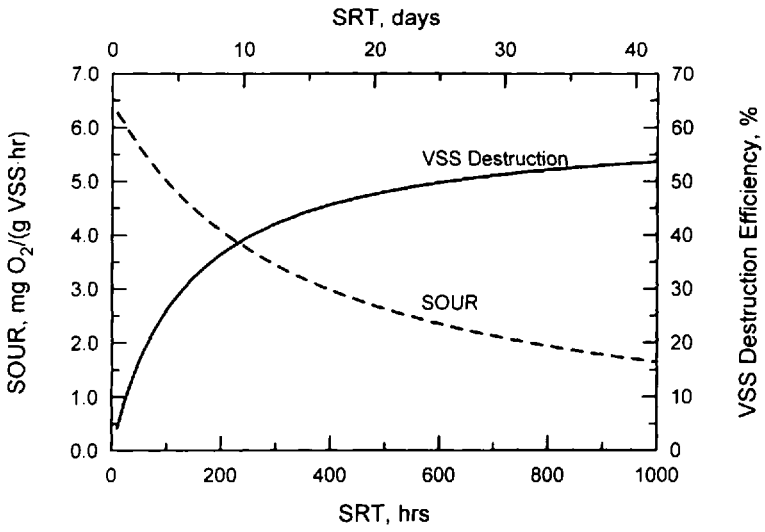


Figure 12.10 Effect of SRT on the percent VSS destruction and SOUR in a completely mixed aerobic digester. The curves are theoretical and were generated from the information in Figure 5.10.

chosen. These effects of aerobic digester SRT on the VSS destruction and the SOUR of the solids have been confirmed by the laboratory results of Novak, et al.[33]

Figure 12.10 illustrates a very important point: sometimes it is possible to meet one criterion of stabilized solids without meeting the other. In this particular case, 38% VSS destruction is achieved at a very reasonable SRT, but an SOUR of 1.0 mg O_2/(g VSS · hr) is not. This follows directly from the nature of the solids being digested, which were waste activated sludge from a completely mixed activated sludge (CMAS) system treating totally soluble substrate. As a consequence, it contained little nonbiodegradable VSS, which allowed a high degree of VSS destruction, but also made the SOUR high. Had the solids been high in nonbiodegradable VSS, just the opposite effect would have been seen, i.e., it would have been possible to reach an SOUR of less than 1.0 mg O_2/(g VSS · hr) at a relatively short SRT, but it would have been difficult to achieve 38% destruction of the VSS. Consequently, it can be seen that what constitutes stable solids depends to a large degree on their source and characteristics. This point is discussed further in Section 12.2.4 because it cannot be overemphasized.

Temperature has an important impact on the destruction of VSS during aerobic digestion because of its effect on the rate coefficient, b_{MV}. That effect is typically expressed with Eq. 3.95, using a θ value of 1.029 for mesophilic digestion. Less certainty is associated with the temperature coefficient for thermophilic conditions. Nevertheless, no matter what the temperature range, lower temperatures mean that longer SRTs are required, just as with activated sludge systems.

Simulations such as those presented above can be used for design when sufficient information is available. Otherwise, performance correlations from the literature must be used. Experience indicates that temperature exerts such an important effect that both temperature and SRT must be considered together in evaluating the performance of an aerobic digester.[17] Consequently, Koers and Mavinic[21,29] suggested that the VSS destruction efficiency be plotted as a function of the bioreactor temperature times the SRT. Figure 12.11 provides an example of such a plot that is widely used in aerobic digester design.[29,40] It suggests that operation of a mesophilic digester at a temperature-SRT product of 400 to 500°C-days will result in substantially complete VSS destruction. Further increases in the temperature-SRT product may result in additional VSS destruction, but the increase will be relatively small. Of course, as discussed above, the actual VSS destruction efficiency obtained will depend on the biodegradability of the solids being digested.

The effect of SRT and temperature on the SOUR is illustrated in Figure 12.12.[2] An SRT in excess of 60 days may be required at temperatures above 10°C to reduce the SOUR below 1 mg O_2/(g VSS · hr), whereas at temperatures below 10°C, SRTs in excess of 100 days may be required. These operating conditions correspond to temperature-SRT products ranging from 600 to well over 1,000°C-days. This again illustrates that it sometimes may be difficult to meet both criteria of stable solids. Similar results have been observed by others.[21,29,32]

Pathogen destruction occurs in aerobic digesters as a result of their natural die-off.[6,23,26,32] Consequently, digestion per se does not affect pathogen destruction; rather, it is the environment created in the aerobic digester and the retention of the pathogens in that environment that results in their destruction. Aeration is necessary primarily to maintain aerobic conditions and avoid nuisance conditions. Figure 12.13[32] illustrates that rates of pathogen die-off are significantly affected by digester temperature.

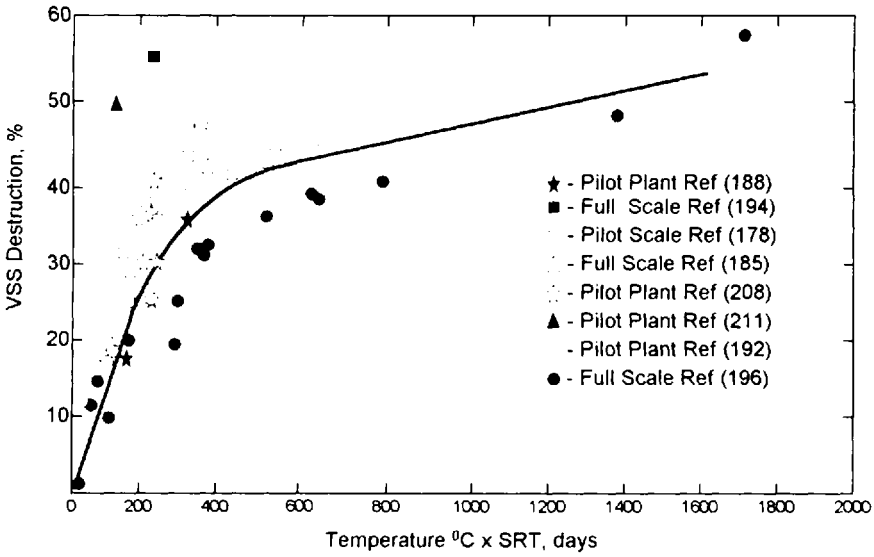


Figure 12.11 Effect of the temperature–SRT product on the VSS destruction efficiency during aerobic digestion. Reference numbers refer to original source. (From *Process Design Manual for Sludge Treatment and Disposal.*[10])

Relatively long aerobic digester SRTs are required for operation at lower temperatures (10°C), while much shorter SRTs are required at higher temperatures, particularly when they are in the thermophilic range. This observation provided one justification for development of ATAD.

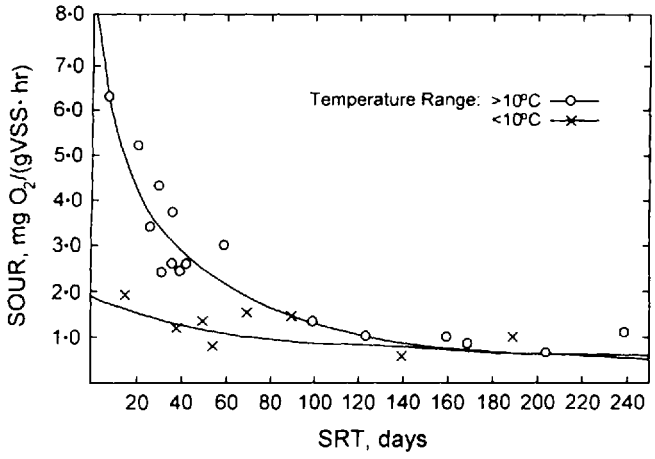


Figure 12.12 Effect of temperature and aerobic digester SRT on the SOUR of the digested solids. (From N. R. Ahlberg and B. I. Boyko, Evaluation and design of aerobic digesters. *Journal, Water Pollution Control Federation* **44**:634–643, 1972. Copyright © Water Environment Federation; reprinted with permission.)

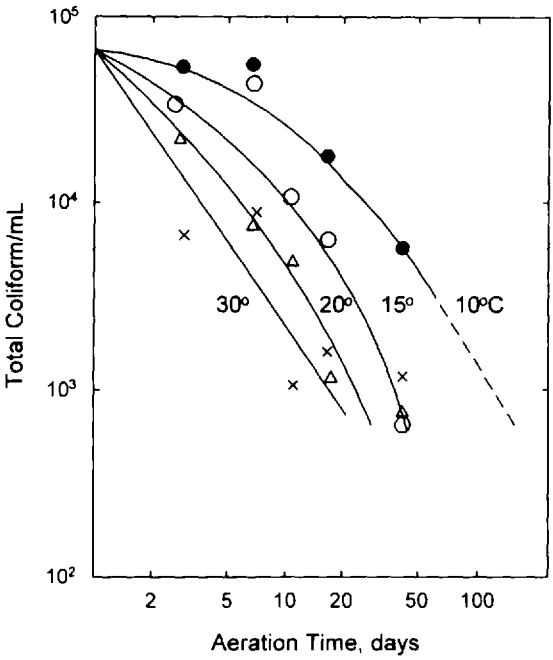


Figure 12.13 Effect of temperature on the inactivation of total coliforms during aerobic digestion in batch reactors. (From J. T. Novak et al., Stabilization of sludge from an oxidation ditch. *Journal, Water Pollution Control Federation* **56**:950–954, 1984. Copyright © Water Environment Federation; reprinted with permission.)

Since the inactivation of pathogens during aerobic digestion is a function of both temperature and SRT, regulatory approaches have specified the required digester SRT to achieve compliance with pathogen control requirements as a function of digester temperature.[24,39,41] The available data indicate that operating temperature has a stronger effect on pathogen destruction than the digester SRT. Consequently, the requirements for pathogen destruction cannot be expressed as a simple temperature–SRT product.

12.2.2 pH

As discussed in Section 11.3.1, a residual alkalinity of around 50 mg/L as $CaCO_3$ is required to maintain a stable pH near neutrality. However, alkalinity is destroyed if the ammonia-N released during aerobic digestion is nitrified. As shown in Eq. 12.8, the amount is one mole of HCO_3^- per mole of biomass destroyed. This corresponds to 0.44 g of alkalinity (expressed as $CaCO_3$) per g VSS destroyed. Consequently, as illustrated in Figure 12.6, unless pH control is practiced, the pH will decrease during CAD if the released ammonia-N is nitrified. pH values of 5 or below are routinely observed in operating aerobic digesters without pH control. Although the destruction of biodegradable organic matter proceeds at these low pH values, the rate is reduced compared to the rate at pH 7.[22] More rapid digestion has been demonstrated through

the addition of lime and other pH control chemicals to maintain the pH near neutrality.[3,4]

12.2.3 Mixing

Adequate mixing energy must be provided in aerobic digesters to maintain solids in suspension. Solids settlement will reduce the effective volume of the bioreactor and result in anaerobic conditions in the settled solids. The provision of adequate mixing can be a challenge because of the high suspended solids concentrations commonly maintained. Design and operational manuals typically recommend an air input rate to aerobic digesters of 20 to 40 $m^3/(min \cdot 1{,}000\ m^3)$ when diffused air systems are used.[45,46] Reynolds[36] has provided the following equation to calculate the volumetric power input (Π, expressed in units of kW/1000 m^3 of bioreactor volume) required to maintain solids in suspension during aerobic digestion:

$$\Pi = 0.935\ \mu_{wc}^{0.3}\ X_{M,T}^{0.298} \tag{12.10}$$

where μ_{wc} is the viscosity of water in centipoise and $X_{M,T}$ is the MLSS concentration in the digester, expressed as mg/L as TSS. At 25°C and a MLSS concentration of 10,000 mg/L, Eq. 12.10 gives a volumetric power input of 13.8 kW/1000 m^3, which is near the minimum recommended for mechanical aeration devices in activated sludge systems, as discussed in Section 10.2.5. While Eq. 12.10 can be used to estimate required mixing energy inputs for CAD applications, care should be exercised when applying it to the high suspended solids concentrations used in ATAD. For example, mixing energy inputs in the range of 85 to 105 kW/1.000 m^3 are typically used with ATAD,[42] whereas Eq. 12.10 would give lower values.

Evidence suggests the difficulty in maintaining aerobic conditions throughout an aerobic digester. For example, several researchers have indicated that the specific decay rate of a variety of waste solids declined as the suspended solids concentration was increased.[10,15,22,36] Since there is no biological basis for such an observation, the most likely explanation is the increasing difficulty of transferring oxygen and maintaining aerobic conditions throughout the digesting solids particles. Furthermore, a loss of nitrogen that can be attributed to denitrification has been observed in aerobic digesters, even though measurable dissolved oxygen concentrations were maintained on a continuous basis.[28,30] These observations suggest that it may not be possible to maintain fully aerobic conditions in many full-scale aerobic digesters.

12.2.4 Solids Type

Waste solids being sent to aerobic digestion vary in the proportions of their biodegradable and nonbiodegradable components. For example, if waste activated sludge is being digested, one factor that can influence this is the SRT of the activated sludge system from which it came. This can be seen in Figure 5.7, where the active fraction, which is numerically equivalent to the biodegradable fraction for the situation simulated, decreases as the activated sludge SRT is increased. It can also be seen in Figure 12.4, where experimental determinations of the nonbiodegradable fraction are shown as a function of the SRT of the activated sludge system from which the solids came. The implication of this is that it will be more difficult to achieve a given percent VSS destruction in solids from an activated sludge system with a long SRT,

simply because a smaller fraction of the solids is biodegradable. The nature of the influent to an activated sludge system will also influence the biodegradability of the solids wasted from it. This can also be seen in Figure 5.7, where two cases are compared, one with and one without inert, i.e., nonbiodegradable, suspended solids in the influent. Clearly, the MLSS in the system receiving inert solids has a lower active, i.e., biodegradable, fraction, which will make it more difficult to achieve a given percent VSS destruction. For other solids, the biodegradable content will depend on their source. On the order of 60 to 80% of the VSS in domestic primary solids will be biodegradable,[22] but no such generalization can be made for solids produced during the treatment of industrial wastewaters. Their biodegradable content must be measured.

Waste solids also differ with respect to the nature of their biodegradable component, depending on their source. This is important because the nature of that biodegradable component determines the rate at which it is stabilized. Theoretically the rate of degradation of waste biomass will be nearly independent of the solids source.[25,32,44] The data presented in Figure 12.11 suggest that this may be true, since they came from several sources. Likewise, the data in Figure 12.3 show that the rate coefficient for destruction of waste activated sludge is independent of the SRT of the system from which it came.[18] In general, the rate of destruction of domestic primary solids will be lower than the rate of destruction of waste biomass.[22] This occurs because of the need to first convert the particulate organic matter contained in the primary solids to active biomass that is subsequently oxidized.

12.2.5 Bioreactor Configuration

Because the destruction of biodegradable organic matter can be characterized as a first order reaction, the efficiency of an aerobic digester can be improved by configuring it as a series of CSTRs, provided that there is no solids recycle around the reactor chain. The impacts of bioreactor configuration on VSS destruction efficiency can be estimated with:

$$E_{XMV} = 100\left[\frac{X_{MVO} - X_{MV,effO}}{X_{MVO}}\right]\left[1 - \frac{1}{\left(1 + \frac{b_{MV}\cdot\Theta_c}{N}\right)^N}\right] \qquad (12.11)$$

The term Θ_c is the total SRT for the bioreactor system and N is the equivalent number of equal sized tanks in series. It is analogous to Eq. 12.4, which applies only to a single CSTR. If the tanks are not equal in size, the performance of the bioreactor system can be determined by sequential application of Eq 12.4. In a similar fashion, an expression equivalent to Eq. 12.6 can be developed to predict the SOUR of the treated solids from a tanks-in-series system:

$$SOUR = 1000\left[\frac{b_{MV}\cdot i_{O\,XMV}(X_{MVO} - X_{MV,effO})}{(X_{MVO} - X_{MV,effO}) + X_{MV,effO}\left(1 + \frac{b_{MV}\cdot\Theta_c}{N}\right)^N}\right] \qquad (12.12)$$

The impact of the number of CSTRs in series on the performance of CAD is illustrated in Figure 12.14. This figure was developed for the situation depicted in

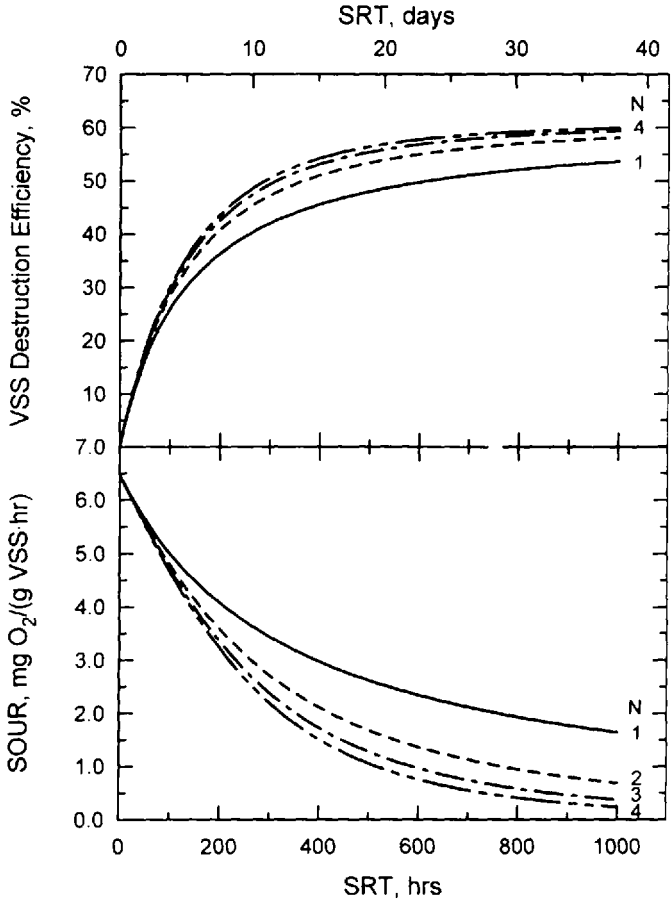


Figure 12.14 Effects of SRT and number of equivalent CSTRs in series on the percent VSS destruction and SOUR in a conventional aerobic digester. The curves are theoretical and were generated from Eqs. 12.11 and 12.12 using the same kinetics and solids characteristics used to generate Figures 5.10 and 12.10.

Figures 5.10 and 12.10, except that the bioreactor was divided into one to four equal sized compartments. Performance is significantly improved by configuring the bioreactor as two CSTRs in series rather than as a single CSTR, allowing a given degree of stabilization to be achieved at a lower SRT. However, compared to the two CSTR system, less improvement is obtained by going to three and four CSTR systems. Similarly, the pathogen destruction efficiency of an aerobic digester is improved by configuring it as a series of CSTRs. This has been clearly demonstrated for the ATAD process where two CSTRs in series are typically used, with significant improvements in bioreactor performance.

A CSTRs-in-series configuration can be obtained in several ways. Consider, for example, the intermittent feed CAD system illustrated in Figure 12.5a. If more than one bioreactor is available, they can be operated in an alternating fashion in which one is fed and decanted for a period of time while another is off-line to allow

digestion and pathogen inactivation to proceed. After digested solids are removed from the second bioreactor, feed is then directed to it while the first is taken off-line for further reaction. The advantages of CSTRs in series can be achieved with the continuous feed process by splitting the bioreactor into two compartments and directing the underflow from the settler to the second one. This approach is necessary because, as discussed in Section 7.2.2, the recycle of solids around the entire system would make it completely mixed with respect to biomass, which would make it behave like a single CSTR. Consequently, care must be taken to ensure that plug–flow type conditions are truly achieved with regard to the flow of solids through the bioreactor. Because the solids concentration in the first compartment will be less than that in the second, the volume of the first compartment should be larger to fully gain the benefits of the tanks-in-series configuration.

12.3 PROCESS DESIGN

12.3.1 Overview

The design of an aerobic digester can be accomplished by application of the principles presented in this and previous chapters. The decisions that must be made include:

- Selection of the process option, i.e., whether the system will be CAD, A/AD, or ATAD. In addition, a decision must be made as to whether operation will be intermittent or continuous.
- Selection of the bioreactor configuration, i.e, whether it will be a single CSTR or a series of CSTRs.
- Selection of the bioreactor feed solids concentration and physical reactor configuration. Both of these factors affect the heat balance for the bioreactor and determine whether significant autoheating will occur. If significant autoheating is expected, a heat balance should be performed to estimate the bioreactor operating temperature. Procedures for performing a heat balance are available elsewhere.[31,43]
- Selection of the bioreactor SRT. This is done on the basis of the desired percent VSS destruction or SOUR to be achieved, using equations like 12.4, 12.6, 12.11, or 12.12. Alternatively, graphical information like that in Figure 12.11 can be used.
- Determination of the required bioreactor volume. This determination is based on the selected SRT, influent flow rate, and desired bioreactor solids concentration.
- Calculation of the oxygen requirement. The power input required to meet the oxygen requirement can then be calculated using the procedures presented in Section 10.2.5.
- Determination of the power input required to achieve adequate mixing and to maintain solids in suspension. This can be done with Eq. 12.10 or with the procedures presented in Section 10.2.5. Just as in activated sludge design, the power required for mixing must be compared to the power required for oxygen transfer, and the larger of the two provided.
- Evaluation of the need for supplemental alkalinity for pH control As discussed in Section 12.1.2, nitrification of released ammonia-N will result in

destruction of alkalinity if CAD is used. If the amount of alkalinity available is insufficient, the pH in the digester will drop, reducing the rate at which digestion occurs. Consequently, the amount of alkalinity available should be compared to the amount of alkalinity likely to be destroyed to determine the need for supplementation.

Several procedures can be used to perform the necessary process calculations. Among them are those based on empirical correlations, those using batch data in simple models, and those using the simplified model of Chapter 5 or IAWQ ASM No. 1. The basic approaches are similar to those used to design activated sludge and biological nutrient removal (BNR) systems. Consequently, all of the procedures presented in those chapters will not be repeated here. Rather, just the unique points will be emphasized.

12.3.2 Design from Empirical Correlations

Empirical correlations such as Figure 12.11 can be used to select a design temperature-SRT product. Then estimation of the digester operating temperature allows direct calculation of the required SRT. If the solids are to be thickened prior to digestion, the SRT is equal to the HRT, allowing the bioreactor volume to be calculated from the influent solids flow rate using the definition of HRT, as given by Eq. 4.15. If solids are to be thickened during digestion, in either the intermittent or the continuous process, then the VSS concentration in the digester is given by:

$$X_{M,V} = \left(\frac{\Theta_c}{\tau}\right)\left[X_{M,VO}\left(\frac{100 - E_{XMV}}{100}\right)\right] \tag{12.13}$$

In that case, the required digester volume is given by:

$$V = \left(\frac{F \cdot \Theta_c}{X_{M,V}}\right)\left[X_{M,VO}\left(\frac{100 - E_{XMV}}{100}\right)\right] \tag{12.14}$$

In determining the degree of thickening that can be accomplished during digestion it is often desirable to know the suspended solids concentration on a TSS basis. The TSS is just the sum of the VSS and the FSS, $X_{M,F}$:

$$X_{M,T} = X_{M,V} + X_{M,F} \tag{12.15}$$

Although some FSS are solubilized during digestion, as discussed in Section 12.1.1, many designers assume that they remain unchanged, thereby giving a conservative estimate of the TSS concentration. In that case, the FSS act like inert suspended solids, so that;

$$X_{M,F} = \left(\frac{\Theta_c}{\tau}\right) X_{M,FO} \tag{12.16}$$

Thus, when solids are expressed on a TSS basis, the reactor volume is given by:

$$V = \left(\frac{F \cdot \Theta_c}{X_{M,T}}\right)\left[X_{M,FO} + X_{M,VO}\left(\frac{100 - E_{XMV}}{100}\right)\right] \tag{12.17}$$

The oxygen requirement (RO) can be calculated on the basis of the estimated VSS destruction efficiency:

$$RO = F \cdot i_{O\,XM,V} \cdot X_{M,VO} \left(\frac{E_{XMV}}{100} \right) \tag{12.18}$$

The value of $i_{O,XM,V}$ depends on whether nitrification is occurring, as expressed in Table 12.1. Once RO is known, the air flow rate or power requirement for oxygen transfer can be estimated as was done in Section 10.2.5. That can then be compared to the power required for mixing as given by Eq. 12.10 or other appropriate information.

Alkalinity destruction can be estimated by recognizing that approximately 0.44 g of alkalinity (expressed as $CaCO_3$) will be used for each g VSS destroyed if the released ammonia-N is nitrified. This can be compared with the available alkalinity to determine whether pH control will be needed. As discussed previously, a residual alkalinity of around 50 mg/L as $CaCO_3$ is required to maintain a stable pH near neutrality.

The following example illustrates the use of empirical correlations to accomplish a preliminary design.

Example 12.3.2.1

A preliminary design is needed for a conventional aerobic digester using intermittent feeding with in-basin thickening for stabilization of waste biomass from an aerobic biological wastewater treatment system. At least 40% VSS destruction is desired year round. The waste biomass concentration is 8,000 mg/L as TSS and the flow rate is 500 m^3/day. The solids are 75% volatile, but little else is known about their characteristics. The alkalinity of the carriage water is 150 mg/L as $CaCO_3$. The lowest operating temperature of the digester is expected to be 10°C based on similar installations in the region. Since those installations can achieve a TSS concentration of 20,000 in the digester, that will be used as a target in the design. Assume that an oxygen transfer device with a transfer efficiency of 1.2 kg O_2/(kW · hr) will be used.

a. What SRT might be appropriate for this application?
Figure 12.11 indicates that 40% stabilization can be achieved with a temperature–SRT product of 600°C-days. Since the lowest expected temperature is 10°C, the SRT should be 60 days.

b. What volume should the digester have?
The digester volume can be determined with Eq. 12.17. To use that equation the influent VSS and FSS concentrations must be known. Since the waste solids are 75% volatile:

$$X_{M,VO} = (0.75)(8{,}000) = 6{,}000 \text{ mg/L}$$

Therefore,

$$X_{M,IO} = 8{,}000 - 6{,}000 = 2{,}000 \text{ mg/L}$$

Recognition of the fact that 1.0 mg/L = 1.0 g/m^3 and substitution of these values into Eq. 12.17 gives:

$$V = \left(\frac{500 \cdot 60}{20{,}000}\right)\left[2{,}000 + 6{,}000\left(\frac{100 - 40}{100}\right)\right] = 8{,}400 \text{ m}^3$$

c. What is the oxygen requirement?
The oxygen requirement can be estimated with Eq. 12.18 after an appropriate value has been chosen for $i_{O\,VM,V}$. Provided that pH control is practiced, nitrification will occur even in the winter because of the long SRT. Consequently, from Table 12.1, the appropriate value for $i_{O\,VM,V}$ is 1.98 mg O_2/mg VSS destroyed. Consequently, the oxygen requirement is:

$$RO = (500)(1.98)(6{,}000)\left(\frac{40}{100}\right) = 2{,}376{,}000 \text{ g/day} = 99 \text{ kg/hr}$$

d. What is the power requirement for oxygen transfer?
Assuming the use of an oxygen transfer device with a transfer efficiency of 1.2 kg O_2/(kW · hr), the required power input is:

$$P = \frac{99}{1.2} = 82.5 \text{ kW}$$

e. What is the power requirement for mixing?
The volumetric power input for mixing can be calculated with Eq. 12.10. At 10°C, μ_w = 1.310 cp. Consequently,

$$H = (0.935)(1.31^{0.5})(20{,}000^{0.295}) = 19.4 \text{ kW}/1000 \text{ m}^3$$

Since the volume is 8,400 m^3, the power requirement for mixing is (19.4)(8.4) = 163 kW. Since this exceeds the power required for oxygen transfer, the larger power for mixing must be provided.

f. Will pH control be required to maintain a neutral pH?
The destruction rate of VSS in the digester is (500)(6,000)(0.40) = 1,200,000 g/day. Since 0.44 g of alkalinity is destroyed for each gram of VSS destroyed when nitrification occurs, the mass rate of alkalinity destruction will be 528,000 g/day. The waste solids contain 150 mg/L of alkalinity, but 50 mg/L must be retained as a residual. Therefore the mass of alkalinity available per day is (500)(100) = 50,000 g/day. This is inadequate, so pH control must be used.

12.3.3 Design from Batch Data

An alternative design approach involves the use of a batch reactor to characterize the solids to be digested, and it is commonly used (e.g., see Refs. 35 and 36). A sample of raw solids is placed in a well mixed vessel, aerobic conditions are maintained, and pH is controlled at 7. Data are then collected and used in Eq. 12.7 to determine the decay coefficient b_{MV} and the nonbiodegradable VSS concentration, $X_{M,V,nO}$. To obtain the most accurate assessment of b_{MV}, data should be collected on the oxygen uptake rate (OUR) and the TSS and VSS concentrations over time. Care

should be exercised to add distilled water to replace any evaporation losses and to scrape any solids that accumulate on the inside walls of the reactor back into the liquid to avoid changes in suspended solids concentrations not attributable to biological reaction. The batch reactor should be operated long enough to ensure that the majority of the biodegradable organic matter is destroyed, thereby allowing an accurate estimate of the nonbiodegradable VSS concentration to be made. This requires that the batch digestion time be greater than five times the reciprocal of the decay coefficient, b_{MV}. To determine b_{MV} the OUR data are analyzed in the same manner as described in Section 8.3.2 for the determination of b_H. Once b_{MV} is known, it can be substituted into Eq. 12.7 and the VSS data can be analyzed according to that equation for estimation of the nonbiodegradable VSS concentration. If OUR data cannot be collected, then b_{MV} and $X_{M,V,i0}$ can be determined simultaneously by fitting Eq. 12.7 to data on the VSS concentration over time. It should be recognized, however that an inaccurate estimate of the nonbiodegradable VSS concentration will result in an inaccurate estimate of b_{MV}.

While measured parameter values are frequently used for sizing aerobic digesters, care must be exercised for the following reasons:

- Significant variation in the measured decay coefficient and the nonbiodegradable fraction of the waste solids may occur with time. Consequently, it is recommended that several batch digestion tests be conducted over time. Then a statistical approach can be used to select the design values for the decay coefficient and the nonbiodegradable proportion of the waste solids.[1]
- Conditions in the batch tests can differ significantly from those anticipated for the full-scale digester. Factors that may differ include pH, temperature, and suspended solids concentrations that can lead to oxygen transfer limitations, as discussed previously. While the data from lab-scale batch reactors can be used successfully to predict the performance of full-scale continuous flow bioreactors, the batch results may significantly underestimate full-scale performance.[13] This can be due to acclimation and/or to the maintenance of more favorable conditions in the full-scale bioreactor.
- If the waste solids to be studied come from a nonnitrifying activated sludge system, they may give erroneous results when the OUR technique is used to determine the decay coefficient. When the waste solids come from a fully nitrifying activated sludge system, ammonia-N will be nitrified as it is released during the batch digestion test.[44] Consequently, the oxygen demand per unit of VSS destroyed will remain reasonably constant during the test and the decrease in OUR will be proportional to the destruction of biodegradable organic matter as assumed. On the other hand, if the waste solids are from a nonnitrifying system, the released ammonia-N will not be nitrified until a sufficient population of nitrifiers has developed.[28,30] In this situation, the oxygen demand per unit of VSS destroyed will not be constant and the change in OUR will not be proportional to the destruction of biodegradable organic matter. Ammonia-N concentrations should be monitored during the batch digestion test to detect whether consistent nitrification is occurring. If it is not, then either nitrification should be inhibited during the OUR measurements or nitrifiers should be added so that complete nitrification of ammonia-N occurs as it is released.

After the values of b_{MV} and $X_{M,V,nO}$ have been determined, the SRT required to achieve either a desired percent VSS destruction or a desired SOUR can be calculated. The SRT required for a given percent VSS destruction in a single-stage digester can be obtained with a rearranged form of Eq. 12.4:

$$\Theta_c = \frac{(E_{XMV}/100)X_{M,VO}}{b_{MV}[X_{M,VO} - X_{M,V,nO} - (E_{XMV}/100)X_{M,VO}]} \tag{12.19}$$

Similarly, the SRT required to achieve a given SOUR in a single-stage digester can be obtained with a rearranged form of Eq. 12.6:

$$\Theta_c = \frac{1000 \cdot i_{O\,XM,V}(X_{M,VO} - X_{M,V,nO})}{SOUR \cdot X_{M,V,nO}} - \frac{X_{M,VO}}{b_{MV} \cdot X_{M,V,nO}} \tag{12.20}$$

Consideration must be given to the SRT required to meet each criterion when deciding on the design SRT. If both are reasonable, then the larger of the two should be used. On the other hand, if one is inordinately high, then the SRT should be selected from consideration of both criteria.

Once the required SRT has been determined, the remainder of the design proceeds in exactly the same manner as described in Section 12.3.2.

Example 12.3.3.1

Batch tests with a waste activated sludge have revealed that the decay coefficient for its aerobic decomposition has a value of 0.216 day $^{-1}$ at 20°C. The solids to be digested are wasted from the bottom of the final settler at a concentration of 12,000 mg/L as VSS. Solids at that concentration were used to run the batch tests, revealing that the nonbiodegradable VSS concentration was 5,400 mg/L. If the lowest temperature expected in the digester is 12°C, what SRT would be required to achieve at least 38% VSS destruction and reduce the SOUR to 1.0 mg O_2/(g VSS · hr) or less? Is it realistic to meet both criteria? Assume that the temperature coefficient for b_{MV} has a value of 1.029.

a. What is the value of the decay coefficient at 12°C?
The decay coefficient can be corrected for temperature with Eq. 3.95:

$$b_{MV,12} = 0.216\ (1.029)^{12-20} = 0.172 \text{ day}^{-1}$$

b. What SRT is required to meet the percent VSS destruction criterion at 12°C?
This may be determined with Eq. 12.19:

$$\Theta_c = \frac{(38/100)(12{,}000)}{0.172[12{,}000 - 5{,}400 - (38/100)(12{,}000)]} = 13.0 \text{ days}$$

c. What SRT is required to meet the SOUR criterion at 12°C?
This may be determined with Eq. 12.20. Note that SOUR and b_{MV} must have consistent time units. To be conservative, assume that nitrification occurs, making the value of $i_{O\,XM,V}$ equal to 1.98.

$$\Theta_c = \frac{(1000)(1.98)(12{,}000 - 5{,}400)}{(1.0)(24)(5{,}400)} - \frac{12{,}000}{(0.172)(5{,}400)} = 87.9 \text{ days}$$

d. Is it realistic to meet both criteria?
An SRT of 88 days is very long and may not be realistic unless the waste activated sludge flow is very small. The reason that it is easy to meet the VSS destruction criterion but not the SOUR criterion is that the waste activated sludge has a fairly high percentage of biodegradable solids. A reasonable compromise would be to choose an SRT between the two values.

12.3.4 Design by Simulation

If an activated sludge system is being designed with the simple model of Chapter 5 or with IAWQ ASM No.1, the output may be used directly in the design of an aerobic digester by simulation. In either case, the characteristics of the waste solids are first determined from the simulations used in the activated sludge design. For the simple model of Chapter 5 the components will include active biomass, biomass debris, and inert organic matter. Only the active biomass will be degraded in the digester, leading to additional debris. The concentrations of active biomass and debris in a completely mixed digester can be calculated with Eqs. 5.64 and 5.65, respectively, while the MLSS concentration can be calculated with Eq. 5.66. The values of $X_{B,H,0}$ and $X_{D,0}$ in those equations are the concentrations in the waste activated sludge entering the digester. If the waste solids also contain inert organic matter (IOM) that originated in the influent to the activated sludge system, its concentration should be added within the bracket of Eq. 5.66 to reflect its presence in the digester. The concentrations of the various components as given by Eqs. 5.64–5.66 are in COD units. They can be converted to TSS or VSS units by using appropriate conversion coefficient, i.

When using ASM No. 1, the output from the activated sludge simulation will provide the concentrations of active biomass, biomass debris, slowly biodegradable substrate, and IOM in the waste activated sludge. These components can be used directly as inputs into a model for the aerobic digester. This is particularly useful when A/AD is being considered since ASM No. 1 can handle both nitrification and denitrification. Simulations conducted with different bioreactor configurations and different recirculation ratios will allow the designer to select a system capable of optimal performance.

Even if the activated sludge system has been designed by simulation, the aerobic digester can be designed with the simple first order model presented earlier in this chapter. In that case the total VSS concentration entering the digester would be calculated with Eq. 12.3, whereas the influent nonbiodegradable VSS concentration would be calculated as:

$$X_{M,V,i0} = f_D \cdot X_{B,H,V0} + X_{D,V0} + X_{I,V0} \tag{12.21}$$

The biodegradable solids can then be calculated with Eq. 12.5. Once those terms are known, everything can proceed exactly as presented in Sections 12.3.1–12.3.3.

12.4 PROCESS OPERATION

Historically aerobic digestion has been regarded as a simple process with very modest operational requirements. In fact, aerobic digesters have frequently been used

merely as solids holding and thickening tanks that are aerated to avoid nuisance conditions. In such situations, digestion occurs simply because of the aerobic conditions maintained. As a consequence, troubleshooting guides for aerobic digestion typically emphasize activities to keep equipment in working order and to avoid nuisance conditions, rather than activities aimed at process control.[45] This situation will change as more sophisticated performance requirements are imposed, including the need to achieve specified pathogen inactivation standards or adequate solids stabilization, as evidenced by either a stipulated VSS destruction efficiency or a specified SOUR. More stringent requirements will cause increased emphasis on the maintenance of desired values of the SRT, pH, and temperature. Many of these objectives can be accomplished at existing aerobic digestion facilities. For example, the operation of CAD systems, particularly intermittently fed ones, as A/AD systems, can avoid the precipitous drop in pH often associated with aerobic digestion. In addition, waste solids can be thickened prior to their addition to the digester so that autoheating will occur, resulting in improved digestion performance and pathogen inactivation. Minor physical modifications to reduce heat loss during the winter can also result in elevated digester temperatures. Such simple changes can significantly improve process performance. Furthermore, the use of oxidation–reduction potential as a technique for real-time control of the A/AD process should facilitate its operation and optimization, and encourage greater full-scale use.[33]

The supernatant from aerobic digestion is often of such poor quality that it cannot be discharged directly to the environment, requiring it to be recycled to the head of the liquid treatment train. This is due in part to the fact that the destruction of biomass results in the release of soluble cellular constituents including nitrogen, phosphorus, micronutrients, and nonbiodegradable organic matter.[9,30,40] In addition, the settleability of aerobically digested solids can be poor, making the suspended solids concentration in the supernatant high. Care must be exercised to minimize the suspended solids content of the supernatant and to control the timing of the return of supernatant to the liquid treatment train so as to minimize any adverse impacts on its performance. If the liquid treatment process is highly loaded, it may be desirable to return supernatant during low nighttime loading periods. In contrast, for nutrient removal systems it may be desirable to return the supernatant during the high loading period when an increased mass of organic matter is available to remove the recycled nutrients. Analysis of the entire treatment system will allow the operator to select the optimal approach.

12.5 KEY POINTS

1. Aerobic digestion has two primary objectives: (1) the destruction of biodegradable particulate organic matter, and (2) the inactivation of pathogens present in waste solids.
2. Aerobic digestion is most applicable to the stabilization of waste biological solids, such as those generated by activated sludge and trickling filter facilities. It can also be used to stabilize primary solids, but aerobic digestion of such solids is often less economic than anaerobic digestion.
3. The influent to an aerobic digester contains both biodegradable and nonbiodegradable particulate organic matter. The relative proportions of each

depends on the loading and operating characteristics of the process producing the solids.

4. The destruction of biodegradable particulate organic matter can be characterized as a first order reaction.
5. Both volatile and fixed suspended solids are destroyed during aerobic digestion, although the relative proportions destroyed may not be the same. Fixed suspended solids are lost as they are solubilized and released from the biodegradable particulate organic matter destroyed.
6. Solids stabilization is typically quantified as either the percent VSS destruction achieved during digestion or the specific oxygen uptake rate (SOUR) of the digested solids.
7. In conventional aerobic digestion (CAD) the solids are maintained under aerobic conditions at the ambient temperature for a period of time adequate to achieve the desired degree of solids stabilization and pathogen inactivation. Both intermittent and continuous feed options are available. Nitrification of released ammonia-N typically occurs, resulting in the destruction of alkalinity and depression of the pH.
8. Anoxic/aerobic digestion (A/AD) includes an anoxic and aerobic sequence in the digestion process. Alkalinity produced through denitrification can offset that consumed in the nitrification of the ammonia-N released. Oxygen requirements are also reduced in comparison to CAD.
9. Autoheating of the digester can be achieved if the solids are thickened prior to digestion and the vessel is designed to minimize heat loss. In autothermal thermophilic aerobic digestion (ATAD), such approaches are used to achieve bioreactor temperatures in the 45 to 65°C range. This results in increased rates of solids stabilization and pathogen inactivation. Because nitrification does not occur under thermophilic conditions, pH depression is avoided and oxygen requirements are reduced.
10. The destruction of biodegradable organic matter in an aerobic digester can be characterized using a variety of approaches. Mathematical approaches include a first-order decay model, the simplified model presented in Chapter 5, and IAWQ ASM No. 1. Another approach uses empirical correlations, such as those that relate percent VSS destruction to the operating temperature–SRT product.
11. Aerobic digestion is most efficient at neutral pH. Maintenance of a neutral pH can be accomplished by use of A/AD to denitrify any nitrate-N generated, ATAD to eliminate nitrification, or chemical pH control in CAD.
12. The mixing energy required to maintain solids in suspension increases as the suspended solids concentration in the aerobic digester is increased.
13. The performance of an aerobic digester can be improved by designing and operating it as a series of CSTRs rather than as a single CSTR.
14. Data collected using batch tests can provide the basis for the design of aerobic digesters. Because of the variability associated with the solids' characteristics, a series of tests should be run and a statistical approach used as the basis for the design. The batch tests must be conducted under conditions reflective of those anticipated in the full-scale system.

15. The physical design of an aerobic digester can significantly influence its operation and performance. Heat loss can be a particularly significant problem in colder climates.

12.6 STUDY QUESTIONS

1. Prepare a table that summarizes the typical SRT values, bioreactor suspended solids concentrations, operating temperatures, and pH values for the three types of aerobic digestion. Describe how the differences in operating conditions result in differences in performance.
2. List the benefits and drawbacks of aerobic digestion and describe where it is typically applied.
3. Discuss how ASM No. 1 can be used to evaluate A/AD. Describe the steps required to apply the model, including procedures to calibrate it.
4. What factors determine the biodegradable fraction of waste solids leaving an activated sludge system? Express the results in terms of the simplified model of Chapter 5.
5. Describe the batch technique used to determine the decay coefficient and nonbiodegradable fraction of waste solids, including the data analysis. Why must constant temperature and pH values be maintained during the test?
6. A batch aerobic digestion test was performed on waste activated sludge from a pilot plant treating a soluble wastewater. The temperature was 20°C and the pH was maintained at a value above 6.5. The results are presented in Table SQ12.1. (a) Determine the concentrations of the volatile and fixed nonbiodegradable suspended solids. (b) Determine the decay coefficients for both volatile and fixed suspended solids.
7. A completely mixed aerobic digester is to be designed to treat the waste solids characterized in Study Question 6. The solids will be thickened to 10,000 mg/L prior to digestion but no additional thickening will be practiced in the digester. The winter operating temperature will be 10°C, and the temperature correction factor, θ, is 1.04. The flow rate of the thickened solids will be 500 m^3/day. (a) What SRT is required to provide 38% VSS destruction during winter operating conditions? (b) The temperature of the bioreactor can be elevated to 25°C by insulation and selection of the oxygen transfer device. What SRT is required to achieve a 38% VSS destruction at this temperature? (c) What is the SOUR of the digested solids at 25°C? (d) What is the oxygen requirement for the aerobic digester? (e) What volume must the digester have at 25°C? (f) Assuming that FSS are lost during digestion in a first order manner, what is the percent reduction in TSS at 25°C? (g) How much power is required to mix the bioreactor at 25°C?
8. Use the simple model of Chapter 5 to define the characteristics of the waste solids produced by an activated sludge system operating at 20°C and an SRT of 5 days. For these calculations use the wastewater characteristics presented in Table E8.4 and the stoichiometric and kinetic parameters in Table E8.5. Assume that all slowly biodegradable substrate is

Table SQ12.1 Batch Aerobic Digestion Data

Time days	TSS, mg/L	VSS, %
0.00	3,080	87.0
0.34	3,000	86.6
0.84	2,890	85.2
1.9	2,630	85.7
2.9	2,320	85.9
3.9	2,140	87.3
4.9	1,926	87.5
5.9	1,710	88.0
6.9	1,690	87.9
7.9	1,590	88.5
8.9	1,520	87.7
9.9	1,490	88.2
10.9	1,320	88.0
11.9	1,260	87.3
12.9	1,280	87.8
13.9	1,140	88.6
14.9	980	86.6
15.9	1,060	87.7
16.9	1,030	87.3

solubilized and converted to readily biodegradable substrate, as was done in using the simple model in Chapter 10. Clearly state all other assumptions. The wastewater flow rate to the activated sludge system is 12,000 m^3/day. How many kg/day of waste solids are produced? Express your answer both as VSS and TSS. What fraction of the VSS is biodegradable? What is the SOUR of the waste solids? Does the estimated SOUR indicate that the solids are stabilized?

9. Using the information developed in Study Question 8, size an aerobic digester to reduce the SOUR to 1.0 mg O_2/(g VSS · hr). Assume that b_{MV} is numerically equal to b_H. Also assume that FSS are conserved. Compare the SRTs required for configurations consisting of one and two tanks in series. Size the bioreactors, calculate the oxygen requirements, and compare the power required for oxygen transfer to that required for mixing for each configuration. Assume that the digesters will be operated without solids recycle and that feed solids are thickened to 15,000 mg/L as TSS. Also assume that the efficiency of the oxygen transfer device is 1.2 kg O_2/(kW · hr). What percentage VSS destruction will each digester achieve?
10. Use ASM No. 1 to define the characteristics of the waste solids produced by an activated sludge system operating at 20°C and an SRT of 5 days. For these calculations use the wastewater characteristics presented in Table E8.4 and the stoichiometric and kinetic parameters in Table 6.3. Clearly state all assumptions. The wastewater flow rate to the activated sludge system is 12,000 m^3/day. How many kg/day of waste solids are produced? Express your answer both as COD and VSS. What fraction of

the waste solids is biodegradable? What is the SOUR of the waste solids? Does the estimated SOUR indicate that the solids are stabilized?

11. Use ASM No. 1 to evaluate the effect of SRT on the performance of a single CSTR CAD system receiving the waste solids characterized in Study Question 10. Model it as a continuous feed process with solids recycle. After preparing a graph of percent solids destruction versus SRT, choose an SRT to give 38% solids destruction and size the bioreactor to maintain a solids concentration of 15,000 mg/L on a VSS basis. How much alkalinity would have to be supplied to maintain a residual alkalinity of 50 mg/L as $CaCO_3$?
12. Reconsider the CAD system sized in Study Question 11. Maintaining the same SRT and total bioreactor volume, reconfigure the system as an A/AD system like that shown in Figure 12.7c. Then use ASM No. 1 to investigate the effects of the recirculation flow rate and the relative sizes of the anoxic and aerobic zones on the performance of the system. Specifically, investigate the effects of those variables on the percent solids destruction, the effluent nitrate-N concentration, the oxygen requirement, and the alkalinity required to maintain a residual alkalinity of 50 mg/L as $CaCO_3$.

REFERENCES

1. Adams, C. E., W. W. Eckenfelder, Jr., and R. M. Stein, Modification to aerobic digester design. *Water Research* **8**:213–218, 1974.
2. Ahlberg, N. R. and B. I. Boyko, Evaluation and design of aerobic digesters. *Journal, Water Pollution Control Federation* **44**:634–643, 1972.
3. Anderson, B. C. and D. S. Mavinic, Aerobic sludge digestion with pH control—Preliminary investigation. *Journal, Water Pollution Control Federation* **56**:889–897, 1984.
4. Anderson, B. C. and D. S. Mavinic, Improvement in aerobic sludge digestion through pH control: Initial assessment of pilot-scale studies. *Canadian Journal of Civil Engineering* **14**:477–484, 1987.
5. Appleton, A. R. and A. D. Venosa, Technology evaluation of the dual digestion system. *Journal, Water Pollution Control Federation* **58**:764–773, 1986.
6. Appleton, A. R., Jr., C. J. Leong, and A. D. Venosa, Pathogen and indicator organism destruction by the dual digestion system. *Journal, Water Pollution Control Federation* **58**:992–999, 1986.
7. Benefield, L. D. and C. W. Randall, Design relationships for aerobic digestion. *Journal, Water Pollution Control Federation* **50**:518–523, 1978.
8. Carrington, E. G., E. B. Pike, D. Auty, and R. Morris, Destruction of faecal bacteria, enteroviruses and ova of parasites in wastewater sludge by aerobic thermophilic and anaerobic mesophilic digestion. *Water Science and Technology* **24(2)**:377–380, 1991.
9. Chudoba, J., Quantitative estimation in COD units of refractory organic compounds produced by activated sludge microorganisms. *Water Research* **19**:37–43, 1985.
10. d'Antonio, G. Aerobic digestion of thickened activated sludge. *Water Research* **17**: 1525–1531, 1983.
11. Deeny, K., J. Heidman, and J. Smith, Autothermal thermophilic aerobic digestion in the Federal Republic of Germany. *Proceedings of the 40th Industrial Waste Conference, 1985, Purdue University*, Butterworths, Boston, Massachusetts, pp. 959–968, 1985.
12. Droste, R. L. and W. A. Sanchez, Microbial activity in aerobic sludge digestion. *Water Research* **17**:975–983, 1983.

13. Droste, R. L. and W. A. Sanchez, Modeling active mass in aerobic sludge digestion. *Biotechnology and Bioengineering* **28**:1699–1706, 1986.
14. Eckenfelder, W. W., Jr., Studies on the oxidation kinetics of biological sludges. *Sewage and Industrial Wastes* **28**:983–990, 1956.
15. Ganczarczyk, J., M. F. Hamoda, and H. L. Wong, Performance of aerobic digestion at different sludge solids levels and operating patterns. *Water Research* **14**:627–633, 1980.
16. Hao, O. J. and M. H. Kim, Continuous pre-anoxic and aerobic digestion of waste activated sludge. *Journal of Environmental Engineering* **116**:863–879, 1990.
17. Hartman, R. B., D. G. Smith, E. R. Bennett, and K. D. Linstedt, Sludge stabilization through aerobic digestion. *Journal, Water Pollution Control Federation* **51**:2353–2365, 1979.
18. Jewell, W. J. and R. M. Kabrick, Autoheated aerobic thermophilic digestion with aeration. *Journal, Water Pollution Control Federation* **52**:512–523, 1980.
19. Kambhu, K. and J. F. Andrews, Aerobic thermophilic process for the biological treatment of wastes—Simulation studies. *Journal, Water Pollution Control Federation* **41**:R127–R141, 1969.
20. Kim, M. H. and O. J. Hao, Comparison of activated sludge stabilization under aerobic or anoxic conditions. *Research Journal, Water Pollution Control Federation* **62**:160–168, 1990.
21. Koers, D. A. and Mavinic, D. S., Aerobic digestion of waste activated sludge at low temperature. *Journal, Water Pollution Control Federation* **49**:460–468, 1977.
22. Krishnamoorthy, R. and R. C. Loehr, Aerobic sludge stabilization—Factors affecting kinetics. *Journal of Environmental Engineering* **115**:283–301, 1989.
23. Kuchenrither, R. D. and L. E. Benefield, Mortality patterns of indicator organisms during aerobic digestion. *Journal, Water Pollution Control Federation* **55**:76–80, 1983.
24. Lue-Hing, C., D. R. Zenz, and R. Kuchenrither, *Municipal Sewage Sludge Management: Processing, Utilization and Disposal*, Technomics Publishing, Lancaster, Pennsylvania, 1992.
25. Marais, G. v. R. and G. A. Ekama, The activated sludge process. I. Steady state behavior. *Water SA* **2**:164–200, 1976.
26. Martin, M. H., Jr., H. E. Bostain, and G. Stern, Reductions of enteric microorganisms during aerobic sludge digestion. *Water Research* **24**:1377–1385, 1990.
27. Matsch, L. C. and R. F. Drnevich, Auto thermal aerobic digestion. *Journal, Water Pollution Control Federation* **49**:296–310, 1977.
28. Matsuda, A., T. Ide, and S. Fujii, Behavior of nitrogen and phosphorus during batch aerobic digestion of waste activated sludge—Continuous aeration and intermittent aeration by control of DO. *Water Research* **22**:1495–1501, 1988.
29. Mavinic, D. S. and D. A. Koers, Performance and kinetics of low-temperature aerobic sludge digestion. *Journal, Water Pollution Control Federation* **51**:2088–2097, 1979.
30. Mavinic, D. S. and D. A. Koers, Fate of nitrogen in aerobic sludge digestion. *Journal, Water Pollution Control Federation* **54**:352–360, 1982.
31. Messenger, J. R., H. A. de Villiers, and G. A. Ekama, Oxygen utilization rate as a control parameter for the aerobic stage in dual digestion. *Water Science and Technology* **22**(12):217–227, 1990.
32. Novak, J. T., M. P. Eichelberger, S. K. Banerji, and J. Yaun, Stabilization of sludge from an oxidation ditch. *Journal, Water Pollution Control Federation* **56**:950–954, 1984.
33. Peddie, C. C., D. S. Mavinic, and C. J. Jenkins, Use of ORP for monitoring and control of aerobic sludge digestion. *Journal of Environmental Engineering* **116**:461–471, 1990.
34. Pizarro, D. R., Alternative sludge handling and disposal at Kent County, Delaware. *Journal, Water Pollution Control Federation* **57**:278–284, 1985.
35. Reece, C. S., R. E. Roper, and C. P. L. Grady, Jr., Aerobic digestion of waste activated sludge. *Journal of the Environmental Engineering Division, ASCE* **105**:261–272, 1979.

36. Reynolds, T. D. Aerobic digestion of thickened waste activated sludge. *Proceedings of the 28th Industrial Waste Conference*, Purdue University Engineering Extension Series No. 142, pp. 12–37, 1973.
37. Scheuerman, P. R., S. R. Farrah, and G. Bitton, Laboratory studies of virus survival during aerobic and anaerobic digestion of sewage sludge. *Water Research* **25**:241–245, 1991.
38. Trim, B. C. and J. E. McGlashan, Sludge stabilization and disinfection by means of autothermal aerobic digestion with oxygen. *Water Science and Technology* **17**(4/5):563–573, 1985.
39. U. S. Environmental Protection Agency, Environmental Protection Agency, 40 CFR Part 257, Criteria for Classification of Solid Waste Disposal Facilities and Practices. *Federal Register* **44**(179):53438–53464, 1979.
40. U. S. Environmental Protection Agency, *Process Design Manual for Sludge Treatment and Disposal*, EPA/625/1–79/011, U. S. Environmental Protection Agency, Cincinnati, Ohio, 1979.
41. U. S. Environmental Protection Agency, *Control of Pathogens in Municipal Wastewater Sludge*, EPA/625/10–89/006, U. S. Environmental Protection Agency, Cincinnati, Ohio, 1989.
42. U. S. Environmental Protection Agency, *Autothermal Thermophilic Aerobic Digestion of Municipal Wastewater Sludge*, EPA/625/10–90/007, U. S. Environmental Protection Agency, Washington, D.C., 1990.
43. Vismara, R., A model for autothermic aerobic digestion. Effects of scale depending on aeration efficiency and sludge concentration. *Water Research* **19**:441–447, 1985.
44. Warner, A. P. C., G. A. Ekama, and G. v. R. Marais, The activated sludge process. IV. Application of the general kinetic model to anoxic-aerobic digestion of waste activated sludge. *Water Research* **20**:943–958, 1986.
45. Water Environment Federation, *Operation of Municipal Wastewater Treatment Plants*, Manual of Practice No. 11, Water Environment Federation, Alexandria, Virginia, 1990.
46. Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992.

This page intentionally left blank

13
Anaerobic Processes

The term anaerobic process refers to a diverse array of biological wastewater treatment systems from which dissolved oxygen and nitrate-N are excluded. In most instances they are operated to convert biodegradable organic matter, both soluble and particulate, to methane and carbon dioxide. Since methane is a sparingly soluble gas, most is evolved and recovered, thereby removing organic matter from the liquid phase and stabilizing any solids present in the influent or produced in the process. Anaerobic digestion of municipal wastewater solids also results in inactivation of pathogens, a step that is usually required prior to ultimate solids disposal. In some cases, anaerobic processes are operated to convert biodegradable particulate organic matter into volatile fatty acids (VFAs), which are subsequently separated from the particulate matter and fed to biological nutrient removal (BNR) systems to enhance their performance.

13.1 PROCESS DESCRIPTION

Anaerobic processes have been used in wastewater treatment systems for more than a century, initially to stabilize the solids produced.[46,50] These bioreactors, called anaerobic digesters, were simple concrete tanks in which the solids were placed as a slurry and allowed to decompose anaerobically. Hydraulic retention times of 60 days or more were common. Gradually, it was discovered that the decomposition could be accelerated by heating the digester to a consistent temperature of about 35°C and mixing it to provide uniform reaction conditions. These discoveries led to the current high rate anaerobic digestion process, which uses HRTs of 15 to 20 days. Anaerobic digestion remains an extremely popular and widely used solids stabilization process, particularly in municipal wastewater treatment.[45,46,72,75]

Development of high rate anaerobic digestion fostered interest in the use of anaerobic processes to treat high strength industrial wastewaters, leading to the development and use of a wide variety of innovative systems.[23,32] Some can be classified as either suspended growth or attached growth systems, but many are hybrid systems, incorporating elements of both. All anaerobic processes, regardless of the type of biomass employed, are described in this chapter because of the similarities of the design approaches employed. Additional details on attached growth systems are provided in Chapters 18 and 21.

The purposeful use of anaerobic digestion to inactivate pathogens in municipal wastewater solids is a relatively new and evolving application.[45,70] Just as in aerobic digestion, pathogen inactivation does not occur as a direct consequence of the di-

gestion process per se; rather it is a result of the environmental conditions in the digester. Pathogen inactivation in anaerobic digesters is relatively efficient because of the elevated temperatures that are typically maintained.

As mentioned above, anaerobic processes are beginning to be used to hydrolyze and ferment a portion of the biodegradable organic matter in wastewater solids, producing VFAs.[70] The VFAs are then removed from the solids by elutriation and used to enhance BNR processes, as discussed in Chapter 11. The solids are then concentrated prior to subsequent treatment.

13.1.1 General Description

A general description of the microbiology and biochemistry of anaerobic processes is presented in Chapters 2 and 3, while the kinetics of the transformations are summarized in Section 9.3.2. Although the chemistry, biochemistry, and microbiology of anaerobic decomposition are quite complex, it can be conceptualized as comprising three steps, as summarized in Figure 2.3: (1) hydrolysis of particulate organic matter to soluble substrates; (2) fermentation of those soluble substrates to produce acetic acid, carbon dioxide, and H_2; and (3) conversion of the acetic acid, the H_2, and a portion of the carbon dioxide to methane.[48,58,62] Methane is a sparingly soluble gas, which is evolved from solution and collected for subsequent use. The evolution of methane decreases the chemical oxygen demand (COD) of the waste stream and provides the mechanism for stabilization of the biodegradable organic matter contained in it. Only minimal COD reduction occurs without methane production, and it is associated with the formation and evolution of H_2. As discussed in Sections 2.3.2 and 9.3.2, the H_2-oxidizing methanogens are fast growing organisms and are present in most anaerobic treatment systems, resulting in conversion of most of the H_2 produced to methane.[58,62,65] However, since the greatest proportion of the methane produced comes from acetic acid, growth of aceticlastic methanogens is required to achieve significant waste stabilization.

Since COD stabilization in anaerobic processes is directly related to methane evolution, methane production can be calculated from the COD removed in the process, just as the oxygen requirement in an aerobic system can be calculated from a COD balance. As discussed in Section 2.3.2, two moles of oxygen are required to oxidize one mole of methane to carbon dioxide and water. Thus, the COD equivalent of methane is 4 kg COD/kg methane. At standard temperature and pressure (0°C and one atmosphere) this corresponds to 0.35 m^3 of methane produced per kg of COD converted to methane.[48,58] For municipal primary solids, the methane equivalent is 0.7 m^3 of methane produced per kg of volatile solids (VS) destroyed.[58] The carbon dioxide content of the gas produced in anaerobic processes ranges between about 30 and 50% and varies depending on the nature of the substrate. For example, the carbon dioxide content is higher when carbohydrates are being treated than when proteins are treated.[58]

Lema, et al.[40] have summarized those aspects of anaerobic processes that particularly affect their design. They are:

- The very low growth rates that the microorganisms have during methane fermentation.
- The low microbial specific activity, especially at the final step of the process.

- The very low values of the half-saturation coefficients, which means an extraordinary affinity of the microorganisms for their substrates.
- The importance of internal and external resistances to mass transfer.
- The inhibition produced by chemicals present in the wastewater or produced in the process.
- The necessity of keeping the physico–chemical parameters within relatively limited ranges to maximize the activity of the microorganisms.
- The need to design and operate a system that can handle fluctuations in wastewater flow and composition.

These challenges are addressed in the design of anaerobic bioreactors by providing a uniform reactor environment and an SRT that is sufficiently long to ensure the growth of all the necessary microorganisms. The mechanisms by which these objectives are achieved are discussed below.

Figure 13.1 provides a schematic of an anaerobic bioreactor that illustrates its four major components: (1) a closed vessel, (2) a mixing system, (3) a heating system, and (4) a gas–liquid–solids separation system. Table 13.1 relates those components to the aspects identified by Lema et al.[40] Anaerobic bioreactors are typically constructed of either concrete or steel, although earthen basins are used for some low-rate processes. An enclosed vessel is used to exclude dissolved oxygen and ensure the development of anaerobic conditions. The bioreactor is often insulated to minimize heat loss. Mixing is provided to increase the homogeneity of the reaction environment and to reduce the resistance to mass transfer. Uniform bioreactor conditions minimize the impacts of the inhibitory materials produced as metabolic intermediates, keep bioreactor physico–chemical parameters within limited ranges, and minimize the impacts of influent flow and composition fluctuations. Due to the high affinity of the reactions for their substrates, performance is not severely impacted by the uniform bioreactor environment. Several methods are used to mix the bioreactor, including devices such as gas recirculation or mechanical mixers, recirculation of bioreactor effluent to the influent, or bioreactor configurations that use the influent and recirculation flows to mix the contents. Gas evolution during treatment results in a degree of mixing that can be significant in certain bioreactor configurations. The

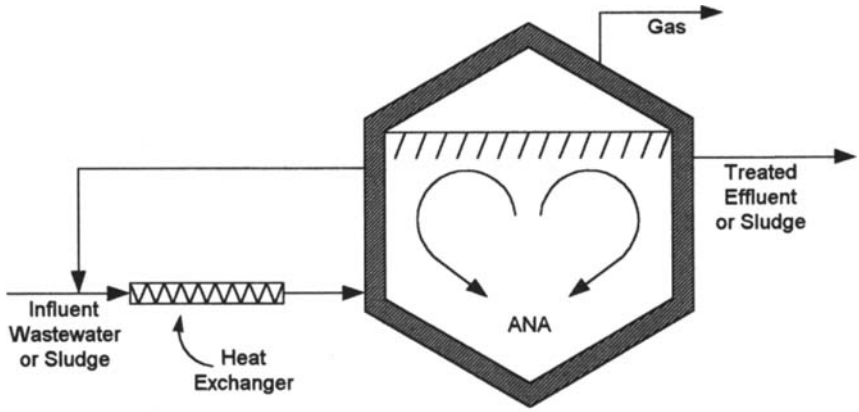


Figure 13.1 Anaerobic bioreactor.

Table 13.1 Relationships Between the Components of Anaerobic Bioreactors and the Aspects of Anaerobic Reactions that Affect Process Design

	Anaerobic reactor component			
Anaerobic reaction aspect of Lema et al.[40]	Closed vessel	Mixing system	Heating system	Gas–liquid–solids separation system
1. The very low specific growth rates.	Provides optimum reaction conditions by excluding dissolved oxygen.	Provides intimate contact between micro-organisms and their substrates to maximize achievable specific growth rates.	Allows maximum microorganism specific growth rate by maintaining temperature for optimal growth.	Accumulation of active microorganisms allows operation at increased SRT.
2. The low specific activities.	Provides optimum reaction conditions by excluding dissolved oxygen.	Provides intimate contact between micro-organisms and their substrates to maximize acheivable specific growth rate.	Allows maximum microorganism activity by optimizing temperature.	Accumulation of active microorganisms allows high total activity, in spite of low specific activity.
3. The very low values of the half-saturation coefficients.		High reaction conversion efficiency possible even though well mixed conditions are typically utilized.		
4. The importance of internal and external resistances to mass transfer.	Provides optimum reaction conditions by excluding dissolved oxygen	Mixing helps to overcome the adverse impacts of internal and external mass transfer resistance.	Allows optimal microorganism activity even though rate is reduced by mass transfer resistance.	Accumulation of active microorganisms and increased SRT can increase total reaction rate and compensate for effects of mass transfer resistances.

5. The inhibition produced by chemicals.	Excludes one reaction inhibitor, oxygen.	Minimizes buildup of reaction intermediates by providing uniform reactor environment.	Minimizes production of reaction intermediates by maintaining temperature for maximum biological activity.	Accumulation of active microorganisms allows increased SRT to be maintained, thereby limiting accumulation of reaction intermediates.
6. The necessity of keeping the physico-chemical parameters within relatively limited ranges.	Excludes one reaction inhibitor, oxygen.	Minimizes variations in reactor environment.	Reduces variation in one environmental factor, temperature.	Accumulation of active microorganisms increases reaction rates in spite of adverse environmental conditions.
7. The need to design and operate a system that can handle fluctuations in wastewater flow and composition.	Excluding oxygen allows optimal reaction rates in spite of fluctuating loading conditions.	Reduces the variation in reactor environmental conditions in spite of variations caused by fluctuating loading conditions.	Allows optimal reactor temperature to be maintained in spite of fluctuating loading conditions.	Accumulation of active microorganisms provides increased reactor biomass needed to treat peak process loadings.

configuration of the feed distribution system can also encourage mixing. Heating is typically provided to maintain temperatures that are constant and near the optimum values for the biomass. Methane gas produced by the system is generally used to fire boilers that provide the necessary heat.

Relatively long SRTs are required in anaerobic processes because of the low maximum specific growth rates of methanogens. Long SRTs also minimize the build-up of inhibitory reaction intermediates and allow the process to respond better to fluctuations in wastewater flow and composition. In some instances, the necessary SRT is achieved by providing a sufficiently long HRT.[46,58] In other cases, the necessary SRT is provided by separating solids from the treated effluent and retaining them in the bioreactor, thereby achieving an SRT that is significantly longer than the HRT.[16,23,68] The gas–liquid–solids separation device is critical to the performance of such systems because the efficiency of liquid–solids separation determines the extent to which active biomass can be accumulated. Gas separation from the solids is necessary to facilitate liquid–solids separation. Several approaches are used to retain active biomass in anaerobic treatment systems; they are described in Section 13.1.4.

A wide range of bioreactor configurations exists, depending on the type of waste, the type of gas–liquid–solids separation provided, and the treatment objectives. Four different process types are considered: (1) anaerobic digesters, (2) low-rate anaerobic processes, (3) high-rate anaerobic processes, and (4) solids fermentation processes. The first three are used to stabilize organic matter by converting it to methane and carbon dioxide. Solids fermentation processes are used to produce VFAs to enhance the performance of BNR systems.

13.1.2 Anaerobic Digestion

Anaerobic digestion (AD) is used for the stabilization of particulate organic matter and Figure 13.2 provides a schematic of the process. An anaerobic digester is well mixed with no liquid–solids separation.[46,72] Consequently, the bioreactor can be treated as a continuous stirred tank reactor (CSTR) in which the HRT and SRT are identical. An SRT of 15 to 20 days is typically used, although SRTs as low as 10 days have been used successfully and longer SRTs are employed when greater waste

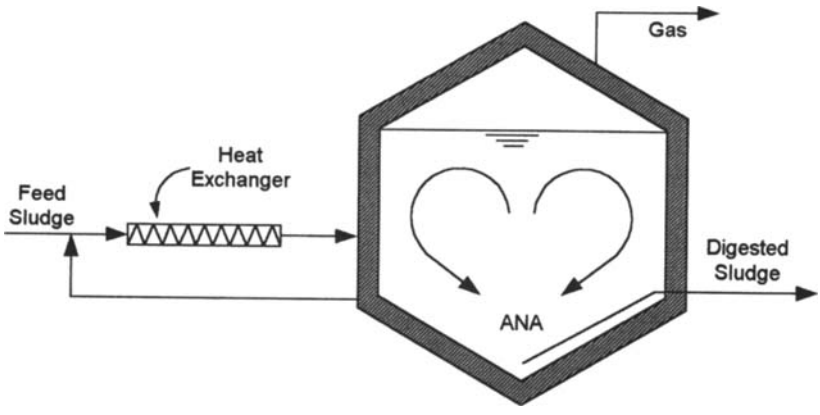


Figure 13.2 Anaerobic digestion.

stabilization is required.[46,72,75] Many anaerobic digesters are cylindrical concrete tanks with a cone-shaped bottom and steel or concrete covers, although other materials and configurations can be used. Diameters range from 10 to 40 m, and sidewall depths from 5 to 10 m. Mixing is required and is provided by internal mechanical mixers, external mechanical mixers that recirculate the tank contents, gas recirculation systems of various types, or pumped recirculation of the tank contents. Historically, relatively low volumetric power inputs have been used to mix anaerobic digesters. More recent experience suggests, however, that such practices may cause a significant portion of the bioreactor volume to be inactive, as well as in significant short-circuiting of feed to the effluent.[55] In contrast, tracer testing has demonstrated that newer approaches can produce essentially completely mixed conditions, thereby minimizing inactive volume and short-circuiting.[15,55,82]

Methane produced by the process is combusted and used to heat the feed stream and digester contents. Bioreactor temperatures in the mesophilic range (~35°C) are typically maintained,[46,58,72,75] although numerous investigations of the use of thermophilic operating temperatures (~55°C) have been conducted.[5,58] Gas storage is typically provided to accommodate variations in gas production rates, thereby facilitating the operation of boilers and other equipment using the gas as a fuel source. External pressurized storage is sometimes used, but more frequently gas is stored in the digester under a cover that floats on the digester contents, as illustrated in Figure 13.3.[46,52,72,75]

Historically, anaerobic digesters treating municipal wastewater solids have experienced operating problems associated with the accumulation of grit in the bottom and floating scum on the surface.[46,72,75] Consequently, bioreactor configurations have been developed that have improved mixing characteristics and reduced potential for grit and scum accumulation. One is the egg-shaped digester, illustrated in Figure 13.4.[46,75] Developed in Germany, it is receiving increasing interest in the United States, where several full-scale installations currently exist. The large height-to-diameter ratio and the steeply sloped lower and upper sections of the vessel result in improved mixing, reduced grit and scum accumulation, and easier removal of any

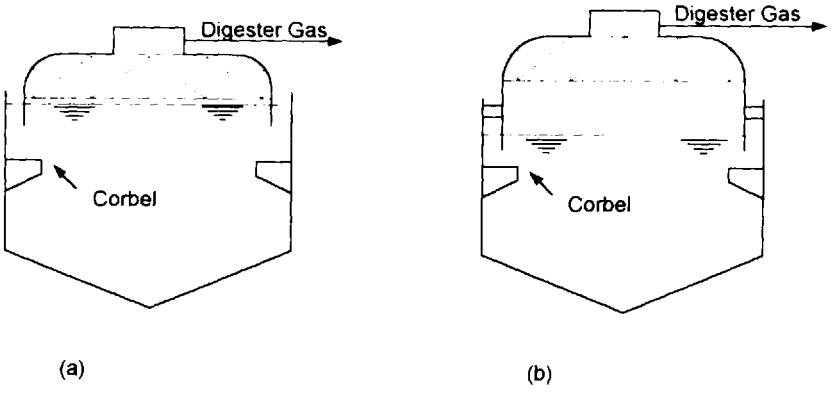


Figure 13.3 Gas storage covers for anaerobic digesters: (a) floating cover; (b) gas holder cover.

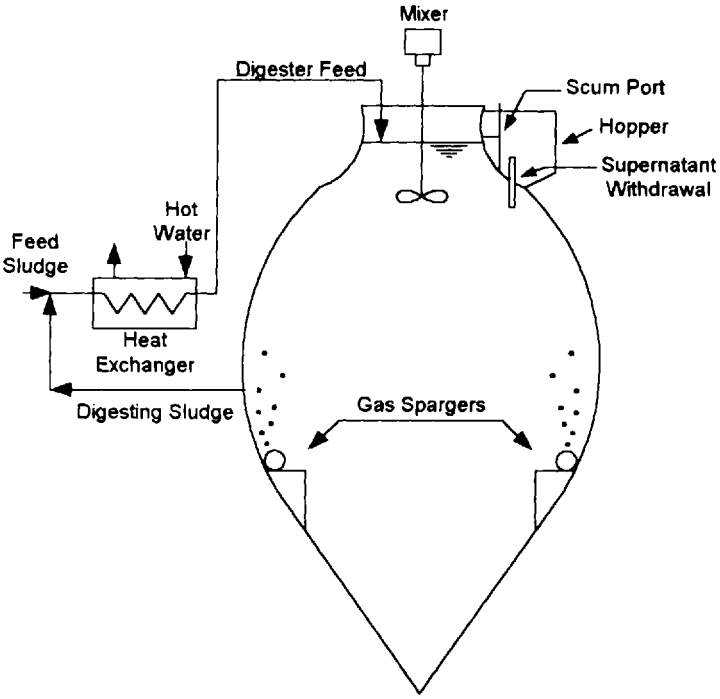


Figure 13.4 Egg shaped anaerobic digester.

that does accumulate. The waffle bottom digester is another configuration that facilitates grit and heavy solids removal.[72]

The purpose of anaerobic digestion is the stabilization of biodegradable particulate organic matter. Consequently, its performance can be quantified by the percent VS destruction. At an SRT of 15 to 20 days, 80 to 90% of the influent biodegradable particulate organic matter will be converted to methane gas.[58] This corresponds to destruction of about 60% of the VS contained in primary solids and 30 to 50% of the VS contained in waste activated sludges, as described further in Section 13.2.9.[21,46,58,72]

Many reference works and textbooks discuss two-stage anaerobic digestion, in which two digesters are operated in series.[46,52,72,75] Heating and mixing are provided in the first stage, where active digestion occurs, while quiescent conditions are provided in the second stage for liquid–solids separation. Supernatant from the second-stage is recycled to the liquid process train while thickened, settled solids are directed to further processing or ultimate disposal. Although of historical interest, use of the two-stage process has declined significantly in recent years for the following reasons:

- Experience indicates that while efficient liquid–solids separation will occur when treating primary solids or a mixture of primary solids and attached growth biomass, it can be quite poor when suspended growth biomass, either alone or mixed with primary solids, is digested. When suspended growth biomass is digested, the supernatant may be of poor quality, resulting

in the recycle of significant quantities of suspended solids to the liquid treatment process train, thereby causing adverse impacts on its performance.

- Significant advances have been made in recent years in solids thickening technology, particularly for suspended growth biomass. This technology is mechanically reliable, allows the consistent production of a thickened solids with a concentration of 50 g/L or more, and is cost-effective.
- Thickening the feed solids prior to anaerobic digestion results in a significant reduction in required tank volume and associated capital cost. Operating costs are also reduced since the volume of feed that must be heated is significantly reduced.
- The recycle of poor quality digester supernatant is eliminated.

Consequently, current practice is to thicken the feed solids prior to single-stage, high-rate anaerobic digestion, which is the process illustrated in Figure 13.2.

13.1.3 Low-Rate Anaerobic Processes

Low-rate anaerobic processes are slurry bioreactors that utilize a combination of solids sedimentation and accumulation to increase the SRT relative to the HRT. They often use earthen basins (Figure 13.5), although rectangular concrete vessels have also been used[23] (Figure 13.6). Mixing is typically provided simply by the addition of influent wastewater and by gas evolution. As a consequence, well mixed conditions are not generally provided and suspended solids settle and accumulate in the bioreactor. Some systems have incorporated settled solids recycle from a downstream settling zone to an upstream reaction zone, as indicated in Figure 13.6. Historically, materials in the wastewater were allowed to float to the surface and form a scum mat that provided some insulation and odor control, although gas would pass through it and escape to the atmosphere. More recently, membranes and similar materials have been used to trap and collect the gas for use elsewhere. Heating is not typically provided. Consequently, either the process is used to treat wastewaters that are already warm or a sufficiently long SRT is maintained to allow treatment to occur at ambient temperature.

Environmental conditions within low-rate processes are not well regulated and, even though active biomass accumulates, accurate control of the SRT is not generally

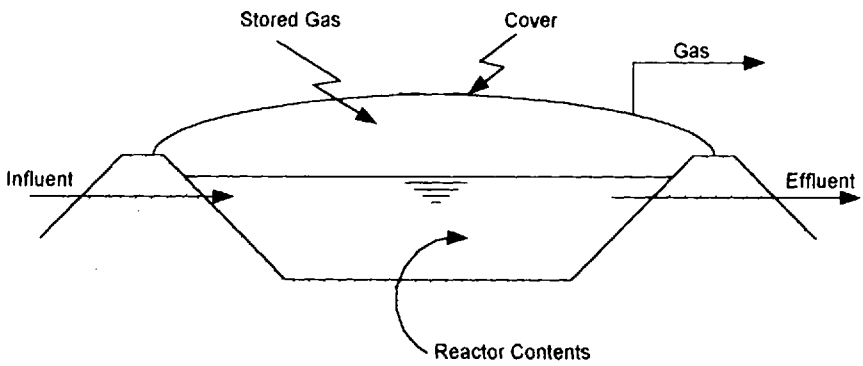


Figure 13.5 Low rate anaerobic process using an earthen basin.

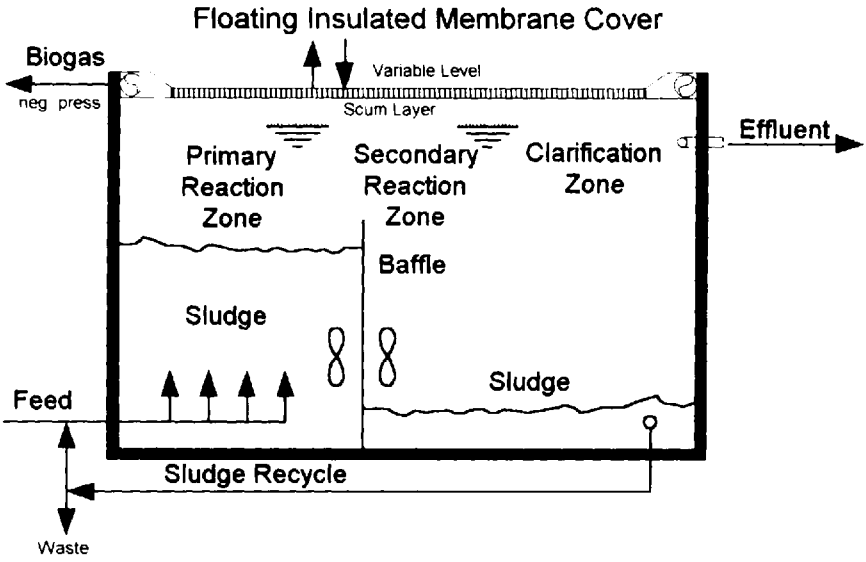


Figure 13.6 Low rate anaerobic process using a rectangular concrete structure.

possible. As a result, designs rely on experience with similar systems and wastewaters, and on the replication of HRTs and organic loading rates (expressed as kg BOD_5 or $COD/(m^3 \cdot day)$) found to be successful. HRTs in the 5 to 15 day range are often appropriate, with longer values required in some cases. Organic loading rates of 1 to 2 kg $COD/(m^3 \cdot day)$ are often found to be appropriate. Process performance varies, but approaches that achieved by high rate anaerobic processes, discussed below. Further discussion of anaerobic lagoons is presented in Chapter 14.

13.1.4 High-Rate Anaerobic Processes

High-rate anaerobic processes utilize bioreactor configurations that provide significant retention of active biomass, resulting in large differences between the SRT and the HRT.[16,23,68] Three mechanisms are used to retain biomass: (1) the formation of settleable particles that are retained by sedimentation, (2) the use of reactor configurations that retain suspended solids, and (3) the growth of biofilms on surfaces within the bioreactor. In many instances, more than one mechanism is operating in a bioreactor. Consequently, high-rate anaerobic processes represent a spectrum of bioreactor types ranging from suspended growth to attached growth, with hybrid bioreactors, which contain significant quantities of both suspended and attached biomass, in between. Six bioreactor types that span this range are described in this section: (1) anaerobic contact (AC), (2) upflow anaerobic sludge blanket (UASB), (3) anaerobic filters (AF), (4) hybrid UASB and anaerobic filters (UASB/AF), (5) downflow stationary fixed film (DSFF), and (6) fluidized bed/expanded bed (FB/EB).

Hall[23] has summarized the typical performance of high rate anaerobic processes, as presented in Table 13.2. A relatively high level of biodegradable organic matter removal can be achieved, as indicated by typical BOD_5 removal efficiencies of 80 to 90%. Biogas production is about 0.5 m^3/kg COD removed, corresponding to a

Table 13.2 Typical High-Rate Anaerobic Process Performance[a]

Parameter	Value
BOD_5 removal, percent	80 to 90%
COD removal, mass	1.5 × BOD_5 removed
Biogas production	0.5 m^3/kg COD removed
Methane production	0.35 m^3/kg COD removed
Biomass production	0.05-0.10 g VSS/g COD removed

[a]Adapted from Hall.[23]

methane production of 0.35 m^3/kg COD removed. Solids production is low, typically ranging from 0.05 to 0.10 kg VSS/kg COD removed. These performance levels can be achieved by all of the processes discussed in this section if appropriate organic loading rates are used.

Anaerobic Contact. Anaerobic contact systems, illustrated in Figure 13.7, consist of a completely mixed suspended growth bioreactor, a degassifier, and a liquid–solids separation device where the bioreactor effluent is separated into a relatively clear process effluent and a concentrated slurry of biosolids that is recycled to the bioreactor.[23,57] Therefore, AC is essentially an anaerobic activated sludge system. The degassifier is a device that facilitates removal of carbon dioxide and methane to allow settling of the biosolids in the liquid–solids separator. If the gas is not removed, bubbles attach to the solids, preventing their settling and subsequent recycle to the bioreactor. A variety of devices can be used to degassify the bioreactor effluent,[57] as illustrated in Figure 13.8. The bioreactor is often configured like an anaerobic digester, and the heating and gas handling systems provided are similar. Completely mixed conditions are achieved by mechanical mixing systems similar to those used in anaerobic digestion. Conventional clarifiers or plate settlers are often used as the liquid–solids separation device.

The AC process is designed and operated to maintain a desired SRT, which is accomplished by adjusting the solids wastage rate. As indicated in Section 9.3.2, SRTs in the 10 to 20 day range are required. Just as with the activated sludge process,

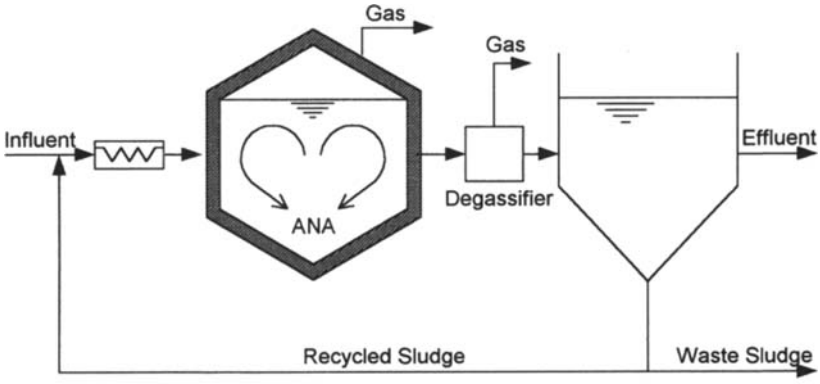


Figure 13.7 Anaerobic contact process.

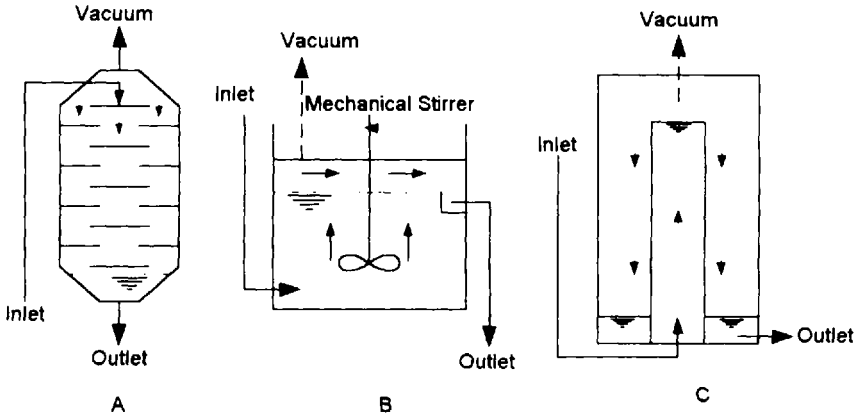


Figure 13.8 Systems for degassifying anaerobic mixed liquor before sedimentation: A cascade; B. flow-through tank; C. thin-film/trickle-film.

the range of HRTs associated with this range of SRTs is dependent on the strength of the wastewater and the concentration of active biosolids that can be attained in the bioreactor. Bioreactor suspended solids concentrations may range from 4 to 6 g/L as VSS to as high as 25 to 30 g/L as VSS, depending on the settleability of the solids that develop. The lower range of concentrations represents typical operation. Clarifier hydraulic loading rates on the order of 5 to 6 m/day are often used, with a solids recycle rate equal to the influent flow rate. Volumetric organic loading (VOL) rates often range between 0.5 and 10 kg COD/(m^3·day).

Upflow Anaerobic Sludge Blanket. The UASB process uses suspended growth biomass, but the gas–liquid–solids separation system is integral with the bioreactor. More importantly, the environmental conditions created in the bioreactor can result in the development of large, dense, readily settleable particles called granules, which allow very high concentrations of suspended solids, on the order of 20 to 30 g/L as VSS, to be accumulated.[23,42,43] These high suspended solids concentrations allow significant separation between the SRT and HRT, and operation at relatively short HRTs, often on the order of two days or less. Figure 13.9 provides a schematic of the process.

Influent wastewater enters the bottom of the bioreactor through a distribution system that is designed to provide relatively uniform flow across its cross section. A dense slurry of granules forms in the lower portion of the bioreactor, and the combined effects of the influent wastewater distribution and gas production result in mixing of the influent wastewater with the granules. Treatment occurs within the dense blanket of granules. For some wastewaters, a much less dense flocculent sludge also develops, and this accumulates on top of the blanket of granules. Other wastewaters contain suspended solids that are not trapped in the granular sludge, and these solids also accumulate as a flocculent sludge blanket overlying the granules. Treated effluent exits the granular and flocculent sludge zones and flows upward into the gas–liquid–solids separator. A variety of configurations can be used for this device, and the one illustrated in Figure 13.9 is only meant to represent the basic concepts used by several manufacturers. The device often consists of a gas collection hood

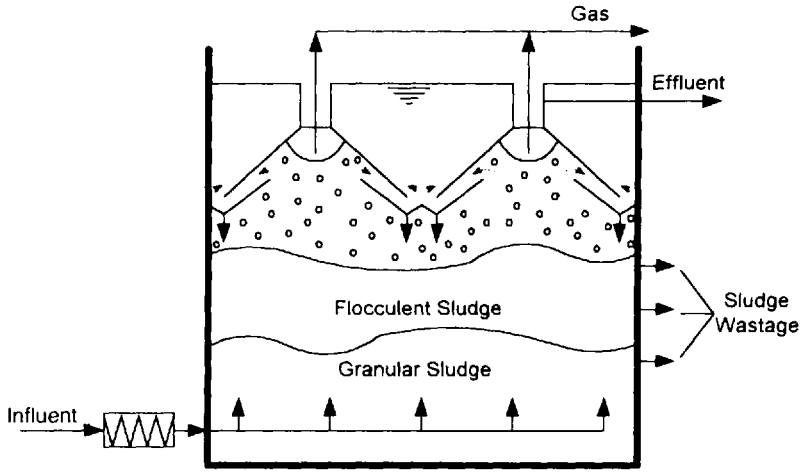


Figure 13.9 Upflow anaerobic sludge blanket bioreactor.

with a settler section above it. Gas bubbles cause some granular and flocculent solids (particularly small granules) to rise through the bioreactor and enter the gas–liquid–solids separator. Gas separation occurs in the hood area, thereby allowing some of this suspended material to return directly to the solids blanket. Gas collects in the upper inverted V section of the hood and is removed from the bioreactor. Liquid with some entrained solids flows out of the hood into the settler section where liquid–solids separation occurs. Clarifier effluent overflows the weirs and is discharged while separated solids settle back into the reaction zone. Design of the gas–liquid–solids separation device requires insight into the physical processes occurring there and experience with specific devices in a variety of applications.

Bioreactor dimensions are affected by process loadings, constraints on maximum upflow velocities, wastewater type, and the settling characteristics of the solids that develop in the process.[23,42,43] The solids inventory increases as treatment occurs and new biomass is grown. Consequently, provisions must be made for solids wastage, as illustrated in Figure 13.9. The relative proportions of flocculent and granular sludge can be controlled by the wasting locations used. Bioreactor HRTs in the 0.2 to 2 day range are typical, along with VOL rates of 2 to 25 kg COD/(m^3·day), depending on wastewater characteristics and whether granular or flocculent solids develop.

Anaerobic Filter. Anaerobic filter systems use upflow bioreactors that are filled with media. The packing is the same as that used with aerobic plastic media trickling filters (TFs), discussed in Chapter 19. Example media are illustrated in Figure 13.10; the specific surface area is typically 100 m^2/m^3 with a void volume of 90 to 95%. The presence of packing allows for the growth of some attached biomass, but the primary role of the media is to retain suspended growth.[23,67,80] The media may be thought of as performing like a set of tube settlers, which provide enhanced liquid–solids separation and retention of suspended biomass within the bioreactor. Gas–solids separation is also facilitated within the packed section. Although several types have been used successfully in AF systems, direct comparisons indicate ad-

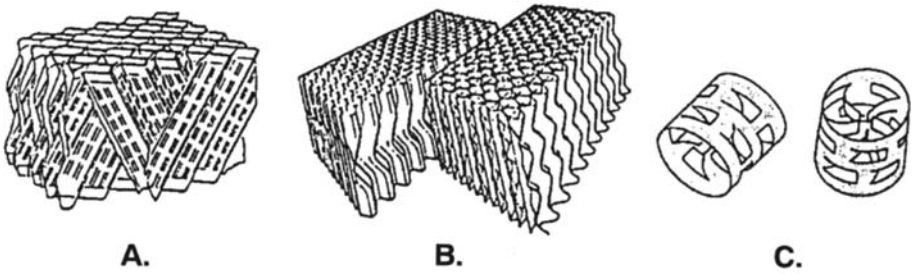


Figure 13.10 Typical media used in anaerobic filters: A. crossflow; B. tubular; C. pall rings. (From J. C. Young, Factors affecting the design and performance of upflow anaerobic filters. *Water Science and Technology* **24**(8):133–156, 1991. Copyright © Elsevier Science Ltd.; reprinted with permission.)

vantages for the crossflow modular media because of its superior gas–liquid–solids separation capabilities.[67,80]

Figure 13.11 provides a schematic of the overall AF process. Influent wastewater and recirculated effluent are distributed across the bioreactor cross-section and flow upward through the media. Treatment occurs as a result of the suspended and fixed biomass retained by the media. Effluent exits the top of the media section and is collected for discharge. Gas is collected under the bioreactor cover and is conveyed to subsequent use. Effluent is typically recirculated to maintain a reasonably uniform hydraulic loading on the bioreactor in spite of varying influent flow rates, thereby maintaining uniform bioreactor hydrodynamic conditions. Although performance is determined by the SRT maintained, accurate assessment of the bioreactor suspended solids inventory is not generally possible. Consequently, bioreactor designs are based on the HRTs and VOLs used successfully in other applications. Hydraulic retention times between 0.5 and 4 days are typical, along with VOLs in the 5 to 15 kg COD/(m^3·day) range. The biomass inventory is typically controlled by the hydrodynamic conditions that develop in the media as a result of the influent (wastewater plus recirculation) flow applied. Excess biomass is washed out of the system as it develops

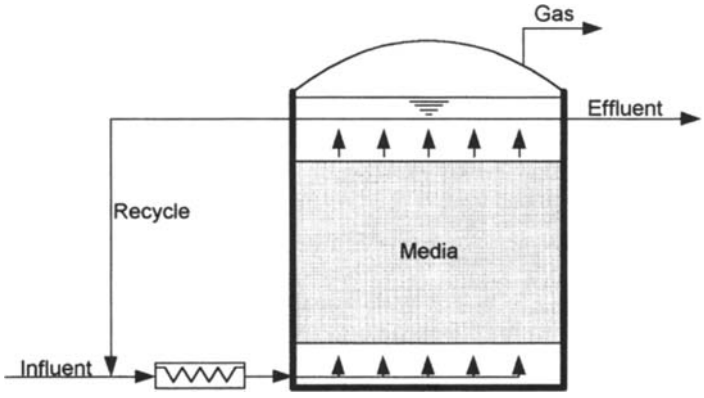


Figure 13.11 Anaerobic filter.

and becomes a part of the effluent. In some instances the capability to remove settled solids from the bioreactor bottom may be provided since heavy solids and precipitates can accumulate there. Solids removal from this location does not constitute an SRT control mechanism, however, because most of the active biomass is retained within the media section.

Hybrid Upflow Anaerobic Sludge Blanket/Anaerobic Filter. Hybrid UASB/AF systems combine aspects of the UASB process with aspects of the AF process.[23] As illustrated in Figure 13.12, influent wastewater and recirculated effluent are distributed across the bioreactor cross-section and flow upward through granular and flocculent sludge blankets where anaerobic treatment occurs. The effluent from the sludge blanket zone then enters a section of media identical to that used in AF systems where gas–liquid–solids separation occurs. Treated effluent then exits the media section and is collected for discharge from the bioreactor. Gas collects under the bioreactor cover and is transported to storage and/or use. The hybrid UASB/AF process primarily uses suspended biomass, and process loadings are similar to those used with the UASB process. The solids removal system is similar to that used with the UASB process.

Downflow Stationary Fixed Film. Downflow stationary fixed film systems use media just like an AF system, but flow is in the downward rather than the upward direction.[23,34] As a consequence, suspended biomass tends to be conveyed through the media rather than retained by it, and the process depends to a large extent on attached rather than suspended biomass. As illustrated in Figure 13.13, influent wastewater and recirculated effluent are distributed evenly across the bioreactor and flow downward through the media. Biomass growing on the media surface and that portion of the suspended biomass that is trapped within the media accomplish treatment of the applied wastewater. Solids removal systems are not generally provided because of the reduced importance of suspended biomass. However, the capability to remove heavy solids from the bottom of the DSFF bioreactor may be provided, just as with the AF process, but such provisions may be less useful as effluent is already being removed from this section of the bioreactor. Treated effluent exits the media zone and is collected for recirculation and discharge. Gas produced in the media flows upward and is collected under the bioreactor cover for transport to

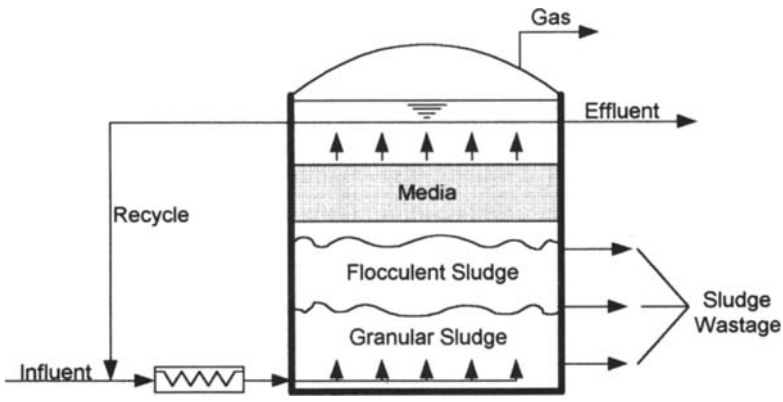


Figure 13.12 Hybrid USAB/AF process.

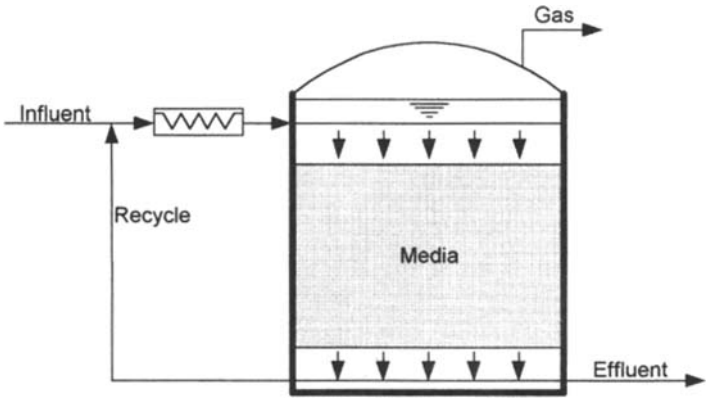


Figure 13.13 Downflow stationary fixed film process.

storage and/or use. The media used in DSFF systems is similar to that used in AF systems, although higher specific surface area media (on the order of 140 m^2/m^3) may be advantageous because of the importance of attached growth. As with many other high-rate anaerobic treatment processes, loadings are generally expressed in terms of both HRT and VOL. Typical HRTs for the DSFF process range from 0.5 to 4 days and VOLs range from 5 to 15 kg COD/(m^3 · day), both of which are similar to the AF process.

Fluidized Bed and Expanded Bed. Fluidized bed and expanded bed systems differ from those previously considered in that they are essentially attached growth systems with little or no suspended growth.[23,31] As illustrated in Figure 13.14, FB/EB systems use upflow bioreactors, just like the UASB, AF, and hybrid UASB/AF processes, but the upflow velocities are much higher, resulting in minimal retention of suspended biomass. Instead, the biomass grows attached to granular carrier particles that are fluidized by the upflow of influent wastewater and recirculated effluent. Fluidization is discussed in Section 18.2. The carrier particles are often silica sand with a diameter in the 0.2 to 0.5 mm range and a specific gravity of 2.65 or granular

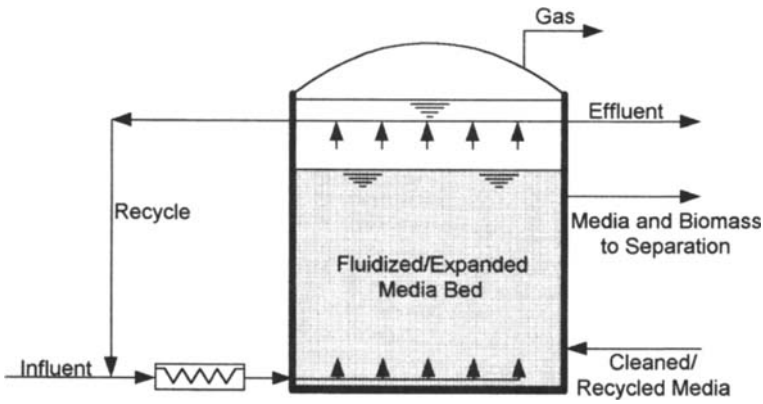


Figure 13.14 Fluidized bed and expanded bed process.

activated carbon. In FB/EB systems upward flow rates are sufficient to support the individual carrier particles with their attached biomass (bioparticles), resulting in expansion of the bed volume in comparison to its resting volume. In expanded beds the upflow velocity is sufficient to expand the bed by 15 to 30%. At this degree of expansion the bioparticles are supported partly by the fluid and partly by contact with adjacent bioparticles and, consequently, they tend to remain in the same relative positions within the bed. In fluidized beds a higher upflow velocity is used, resulting in further expansion of the bed to between 25 and 300% of its resting volume. Under these conditions the bioparticles are fully supported by the upward flowing fluid and move freely in the bed. Gas production also results in turbulence that tends to mix the bioreactor contents. Gas–solids separation devices are provided in some cases to allow the bioparticles to be retained in the bioreactor.

The turbulence created by the upward fluid flow and the gas production allows small carrier particles to be used without bioreactor plugging. It also encourages high mass transfer rates. The use of small carrier particles results in a high specific surface area and a high active biomass concentration. For expanded bed bioreactors, the specific surface area of the carrier particles is in the 9,000 to 11,000 m^2/m^3 range, with a void volume of 45 to 55%.[23,31] For fluidized bed systems the specific surface area is in the 4,000 to 10,000 m^2/m^3 range, with void volumes of 50 to 90%, depending on the degree of expansion. These specific surface areas are approximately two orders of magnitude greater than those provided in AF or DSFF bioreactors. The high specific surface areas allow high biomass concentrations to develop, on the order of 15 to 35 g/L as VSS, which are similar to those achieved with the UASB process.[23,31] The high biomass concentrations allow operation at relatively low HRTs and high VOLs while maintaining adequate SRTs for efficient treatment. HRTs in the 0.2 to 2 day range are used, depending on the concentration of the wastewater. Volumetric organic loadings over 20 kg COD/(m^3·day) are common with FB/EB systems.

As with any other biological process, excess biomass must be removed from the bioreactor to control the biomass inventory. As discussed in Section 18.2.2, accumulation of biomass on the carrier particles increases the bioparticle diameter and decreases its density. The result of these two contrasting factors is a decrease in the settling velocity of the bioparticles and the tendency for bioparticles with more biomass to accumulate in the upper portion of the bioreactor. As a consequence, solids wasting is generally from there. The bioparticles are conveyed to a device where biomass is sheared from the carrier particles. The carrier particles are then returned to the bioreactor and the biomass is taken to further processing or ultimate disposal. Several devices are available for this purpose, but they often consist of a rubber lined pump where the biomass is sheared from the carrier particles and a centrifugal separation device where the lighter biomass is separated from the denser media particles. Many of the commercially available devices are proprietary in nature. Fluidized beds are discussed in detail in Chapters 18 and 21.

The basic process options described in this section can also be combined in a variety of ways to produce a wide range of additional anaerobic treatment systems. For example, interest currently exists in the use of membranes as a means of further separating the SRT and the HRT, thereby producing an even more compact anaerobic process.[20,25]

13.1.5 Solids Fermentation Processes

Solids fermentation processes are used to solubilize particulate organic matter in primary solids and ferment the soluble products to VFAs, particularly acetic and propionic acid, for use in BNR processes.[66,76] The objectives of solids fermentation processes are different from those of the anaerobic stabilization processes discussed previously. They are to maximize the production of VFAs and recover them in a stream that can be delivered to a BNR system. The first objective is achieved by controlling the SRT to a value that allows the growth of hydrolytic and fermentative bacteria but prevents the growth of aceticlastic methanogens, which would consume the VFAs.[18,19,66] As indicated in Figure 9.5, at 35°C this requires an SRT in the 2 to 3 day range. In general, the feed solids and bioreactor contents are not heated, so the SRT must be increased to compensate for the lower temperature. Some methane will be produced as a result of the growth of H_2-utilizing methanogens, but the amount will be small. The second objective is achieved when the VFAs are separated from the residual primary solids by passing the bioreactor effluent through a liquid–solids separation step.

Figure 13.15 illustrates schematically the concepts of fermentation systems. Feed-solids are fed to a mechanically mixed bioreactor where fermentation occurs. The SRT is controlled by adding dilution water in sufficient quantities so that the HRT, which equals the SRT, is maintained at the desired value. The use of gravity sedimentation to achieve liquid–solids separation is illustrated in Figure 13.15. The option of adding elutriation flow to the bioreactor effluent is provided to ensure sufficient supernatant to effectively recover the produced VFAs. Typically, the settled solids are removed from the settler and taken to further processing. However, the capability to recycle a portion of those solids to the bioreactor may be provided to increase its SRT above its HRT.

Figure 13.16 illustrates how the concepts in Figure 13.15 have been implemented at several full-scale wastewater treatment plants. In an activated primary clarifier, primary solids are accumulated in a sludge blanket where fermentation occurs. The settled solids are then recycled to an upstream mixing/elutriation tank where the soluble VFAs are washed from the fermented primary solids and into the

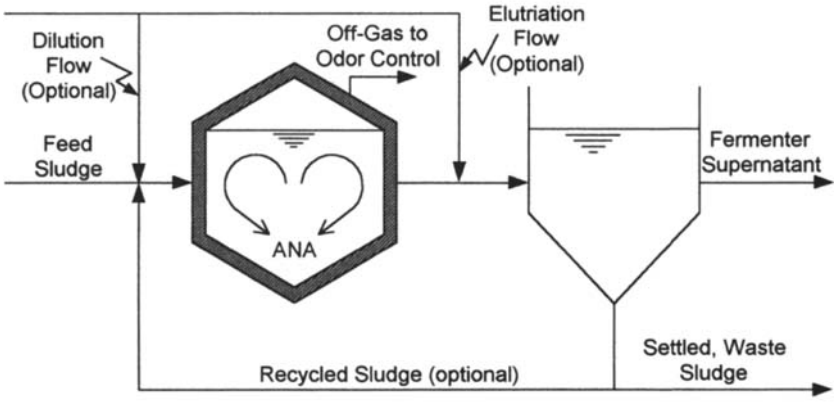


Figure 13.15 Solids fermentation process.

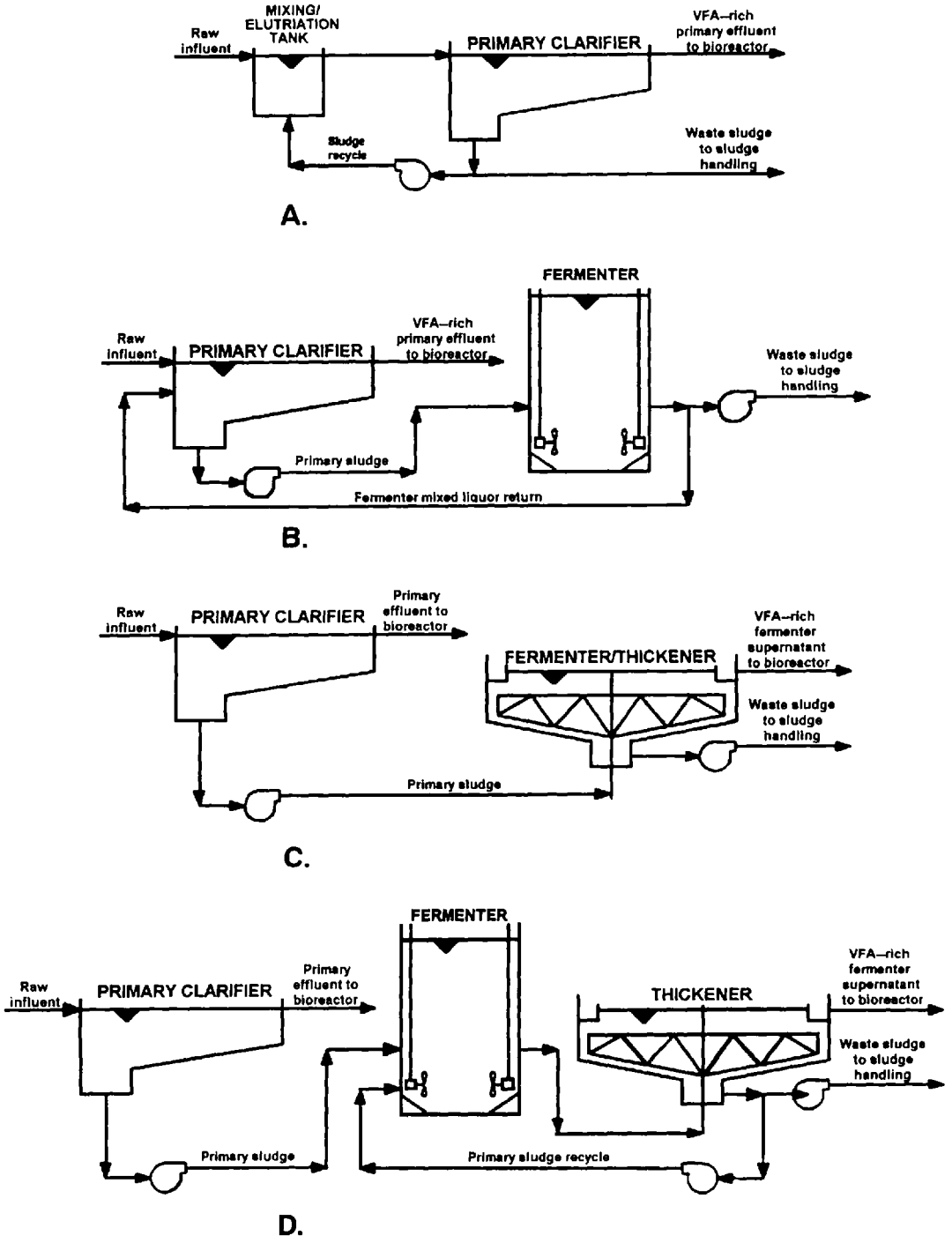


Figure 13.16 Alternative configurations for the solids fermentation process: A. activated primary tanks; B. complete mix fermenter; C. single-stage fermenter/thickener; D. 2-stage complete mix/thickener fermenter. (From Water Environment Federation, *Use of Fermentation to Enhance Biological Nutrient Removal*, Proceedings of the Conference Seminar, 67th Annual Conference and Exposition, Water Environment Federation, Alexandria, Virginia, 1994. Copyright © Water Environment Federation; reprinted with permission.)

primary clarifier effluent. The SRT is controlled by wasting settled solids from the process. In the completely mixed fermenter, solids are removed from the primary clarifier and fermented in a separate bioreactor. However, the fermented solids are then recycled to the primary clarifier where VFAs are removed by the wastewater flow and fermented solids are wasted to control the SRT. In the single-stage fermenter/thickener, primary solids are added to an oversized gravity thickener where both solids fermentation and liquid–solids separation occur. Solids wasting is controlled to achieve the desired SRT, and VFAs are removed in the overflow. Finally, in the two-stage completely mixed thickener/fermenter system, the fermentation and solids thickening steps are separated into two unit operations. Optional primary effluent addition points provide operational flexibility to control the SRT and VFA elutriation. This option is functionally identical to the prototype solids fermentation process illustrated in Figure 13.15.

All the options presented in Figure 13.16 use gravity liquid–solids separation. Relatively poor solids thickening has been experienced in some instances, probably as a result of the gases produced. Such operating problems have been controlled in some cases by operation at a reduced fermentation reactor solids concentration and/or by dilution of the bioreactor effluent prior to gravity separation. Alternatively, centrifuges have been used.[9,33] Other options include static fermenters or fermentation basins upstream of biological reactors.

Primary solids fermentation is a developing technology and many process configurations are being evaluated for full-scale application. However, all operate according to the basic principles described in this section. Details of the process configurations developed to date are presented elsewhere.[76]

13.1.6 Comparison of Process Options

Table 13.3 summarizes the primary benefits and drawbacks of the anaerobic treatment systems used to stabilize organic matter. Anaerobic digestion is suitable for a wide range of wastewaters, particularly those with high concentrations of suspended solids. Well mixed conditions are provided within the bioreactor, resulting in a uniform environment that produces predictable and stable performance. Process performance is not dependent on solids settleability since anaerobic digesters use completely mixed bioreactors with no biomass recycle. The long HRTs required to achieve adequate SRTs result in large bioreactor volumes which can effectively dilute toxic materials. However, they can cause high capital costs. Effluent quality can be poor if the influent contains high concentrations of nonbiodegradable organic matter. Process stability will be good if a sufficient SRT is provided, but it will be poor at shorter SRTs. The process requires a separate mixing system.

Low rate anaerobic processes are simple and economical to construct, and are suitable for a wide range of wastewaters, including those with high concentrations of suspended solids. The low VOLs used result in large bioreactors for the dilution of toxic inputs. These processes are also capable of accepting the waste solids from downstream post-treatment aerobic systems, but large bioreactors are required, along with relatively large land areas. Good performance is possible, and is not dependent on the development of a readily settleable sludge. However, because process control techniques are limited, the conditions within the bioreactor are poorly controlled, which can lead to reduced efficiency.

Table 13.3 Anaerobic Treatment Process Comparison—Organic Stabilization

Process	Benefits	Drawbacks
Anaerobic digestion (AD)	• Suitable for a wide range of wastewaters • Efficiently handles high suspended solids wastewaters • Easy to mix, thereby creating uniform reaction environment • Large bioreactor volume to dilute inhibitors • Performance not dependent on sludge settleability • Capable of accepting waste aerobic biomass	• Large bioreactor volumes required • Effluent quality can be poor if nondegradable organic matter is present or if a large concentration of anaerobic organisms is generated • Process stability and performance poor at short SRTs • Requires separate mechanical mixing
Low-rate anaerobic processes	• Simple and relatively economical construction • Suitable for a wide range of wastewaters • Efficiently handles high suspended solids wastewaters • Large bioreactor volume to dilute inhibitors • Good performance possible • Performance not highly dependent on solids settleability • Capable of accepting waste aerobic biomass	• Relatively large bioreactor volumes required • Large land area required • Poorly controlled conditions within bioreactor reduce efficiency • Limited process control capability
Anaerobic contact (AC)	• Suitable for concentrated wastewaters • Easy to mix, thereby creating uniform reaction environment • Relatively high effluent quality achievable • Reduced bioreactor volume compared to AD • Capable of accepting waste aerobic biomass • Significant process control capability available	• Biomass settleability critical to successful performance • Most suitable for wastes with low to moderate levels of suspended solids • System is relatively complex mechanically • Shorter bioreactor HRTs mean less equalization and dilution of inhibitors
Upflow anaerobic sludge blanket (UASB)	• High biomass concentrations and long SRTs achievable • Small bioreactor volumes due to high volumetric organic loading rates • High-quality effluent achievable • Mechanically simple • Compact system, relatively small land area • Well mixed conditions produced	• Performance dependent on development of dense, settleable solids • Much lower process loading required if wastewater contains suspended solids • Special bioreactor configuration required which is based on experience • Little process control possible • Shorter bioreactor HRTs mean less equalization and dilution of inhibitors

(*Table continues*)

Table 13.3 Continued

Process	Benefits	Drawbacks
Anaerobic filter (AF)	• High biomass concentrations and long SRTs achievable • Small bioreactor volumes due to high volumetric organic loading rates • High-quality effluent achievable • Mechanically simple • Compact system, relatively small land area • Performance not dependent on development of dense, settleable solids • Well mixed conditions produced in bioreactor	• Suspended solids accumulation may negatively impact performance • Not suitable for high suspended solids wastewaters • Little process control possible • High cost for media and support • Shorter bioreactor HRTs mean less equalization and dilution of inhibitors
Hybrid UASB/AF	• High biomass concentrations and long SRTs achievable • Small bioreactor volumes due to high volumetric organic loading rates • High-quality effluent achievable • Mechanically simple • Compact system, relatively small land area • Performance partially dependent on development of dense, settleable solids • Well mixed conditions generally produced in bioreactor • Reduced media cost	• Lower process loadings required if wastewater contains suspended solids • Little process control possible • Shorter bioreactor HRTs mean less equalization and dilution of inhibitors
Downflow stationary fixed film (DSFF)	• High biomass concentrations and long SRTs achievable • Small bioreactor volumes due to high volumetric organic loading rates • High-quality effluent achievable • Mechanically simple • Compact system, relatively small land area • Performance not significantly impacted by high suspended solids wastewaters • Performance not dependent on development of dense, settleable solids • Well mixed conditions generally produced in bioreactor	• Biodegradable suspended solids not generally degraded • High cost for media and support • Organic removal rate generally lower than other high-rate processes • Little process control possible • Shorter bioreactor HRTs mean less equalization and dilution of inhibitors

Table 13.3 Continued

Process	Benefits	Drawbacks
Fluidized bed/ expanded bed (FB/EB)	• High biomass concentrations and long SRTs achievable • Small bioreactor volumes due to high volumetric organic loading rates • Excellent mass transfer characteristics • High-quality effluent achievable, often better than other high-rate processes • Most compact of all high-rate processes; requires smallest land area • Performance not dependent on development of settleable solids • Very well mixed conditions generally produced in bioreactor • Increased process control capability relative to other high-rate processes	• Lengthy start-up period required • High power requirements for bed fluidization and expansion • Not suitable for high suspended solids wastewaters • Mechanically more complex than other high-rate processes • Increased process control required • Cost of carrier media is high • Shorter bioreactor HRTs mean less equalization and dilution of inhibitors

All high-rate anaerobic processes share certain characteristics. High biomass concentrations are maintained, thereby allowing long SRTs to be achieved while keeping the HRTs short. The high biomass concentrations allow high VOLs to be applied, resulting in relatively small bioreactors, and the long SRTs provide good process stability. Although the systems are compact and require relatively small land areas, a high-quality effluent can generally be achieved. Well mixed conditions are typically produced in the bioreactor, resulting in a uniform reaction environment. High-rate processes are best suited for the treatment of wastewaters containing soluble organic matter (SOM), and are adversely impacted by the presence of high concentrations of influent suspended solids. The relatively short HRTs possible with these systems mean that less equalization is provided and less dilution is available for toxic inputs. Important differences also exist among the various high rate processes.

Anaerobic contact processes are suitable for a wide range of wastewaters; are easy to mix, thereby creating a uniform reaction environment; can tolerate moderate suspended solids loadings, including the waste solids from coupled aerobic treatment systems; and provide the capability for significant process control. In contrast, their performance is dependent on the development of a settleable biomass, they best accomodate wastewaters with low to moderate concentrations of suspended solids, and they are mechanically complex.

Upflow anaerobic sludge blanket systems are mechanically simple and easy to operate, but their performance is dependent on the formation of dense, settleable

biomass granules. Furthermore, the allowable organic loading rate is adversely impacted by the presence of suspended solids in the influent wastewater, special bioreactor configurations are required, and little process control is possible.

The design of AF systems is quite straightforward. Furthermore, they are not dependent on the development of a dense, settleable biomass because the media provides the primary mechanism for biomass retention. However, excessive accumulations of suspended solids can lead to plugging, which negatively impacts process performance; therefore, they are not suitable for wastewaters containing high concentrations of suspended solids. Little process control is possible, and the cost of the media and its associated supports can be relatively high.

Hybrid UASB/AF systems combine the advantages of their parent systems. However, they are still adversely impacted by the presence of suspended solids in the influent wastewater, and little process control is possible.

The performance of DSFF systems is not dependent on the development of settleable biomass and is less influenced by high concentrations of suspended solids in the influent wastewater. The suspended solids are not retained by the system, so their presence causes a poor-quality effluent unless it is treated in a downstream clarifier. As with the anaerobic filter, the cost for the media and its associated support system is relatively high, and little process control is possible.

Fluidized bed and expanded bed systems share many of the benefits and drawbacks of the other high-rate anaerobic processes, but they also possess some unique characteristics. The high specific surface areas of the carrier particles allow the development of exceptionally high biomass concentrations, thereby allowing small bioreactors to be used. The uniformly mixed conditions and turbulence provide an extremely uniform reaction environment with excellent mass transfer characteristics. As a consequence, the effluent quality achievable with FB/EB systems is generally superior to that from other high-rate anaerobic processes. Process performance is not dependent on the development of a settleable biomass, and process control is superior to that of the other high-rate processes. In contrast, power requirements can be high because of the high flow rates required to achieve the necessary upflow velocities. Fluidized bed and expanded bed bioreactors are not suitable for wastewaters containing high concentrations of suspended solids, require more process control, are more expensive, and are mechanically more complex than some of the other processes.

Although high organic removal rates can be achieved with all of the high-rate anaerobic processes, differences exist for soluble materials. The highest rates of soluble substrate removal are generally achieved in FB/EB systems because of their high biomass concentrations and excellent mass transfer characteristics. High soluble substrate removal rates can also be achieved in UASB and hybrid UASB/AF systems, particularly when a dense, readily settleable, granular sludge develops. This is because of the high biomass concentrations in the granular sludge bed and the mixing caused by the introduction of influent wastewater and the evolution of gas. Soluble substrate removal rates are lower in AF systems because of their lower biomass concentrations and poorer mixing conditions. Biomass concentrations are even lower in DSFF systems because the downward flow pattern results in reduced accumulation of suspended biomass. This causes lower soluble substrate removal rates. The lowest biomass concentrations occur in AC systems, causing the lowest soluble substrate removal rates. Although a settleable biomass is developed, mechanical mixing gen-

erally prevents the development of a dense, readily settleable granular sludge such as in UASB and hybrid UASB/AF systems.

13.1.7 Typical Applications

As discussed in Chapter 9, anaerobic processes are typically used to stabilize the organic matter present in wastewaters with biodegradable COD concentrations greater than about 1,000 mg/L. Although anaerobic systems have been applied to the treatment of more dilute wastewaters,[41,74] aerobic systems are often more economical for them.[23] Compared to aerobic systems, the advantages of anaerobic processes include less solids production, lower nutrient requirements, lower energy requirements, and the production of a potentially useful product (methane). On the other hand, the effluent quality from anaerobic processes is generally not as good as from aerobic processes, and aerobic polishing may be required to achieve effluent quality goals. Anaerobic processes can be more sensitive than aerobic processes to shock loads and toxic materials, although the anaerobic process technology developed in the past ten years has demonstrated significant resistance to them. Finally, anaerobic processes are capable of metabolizing some organic compounds not readily biodegraded in aerobic systems. Examples include chlorinated organics, which can be dechlorinated in anaerobic treatment systems even though they are not readily biodegraded in aerobic systems.[68]

Several factors affect the choice between anaerobic and aerobic treatment systems for wastewaters with biodegradable COD concentrations in the 1,000 to 4,000 mg/L range. One is wastewater temperature. Anaerobic processes perform best when operated at temperatures near the optimum for either mesophilic (30° to 40°C) or thermophilic (50° to 60°C) microorganisms. Deviations from these ranges can result in significant reductions in microbial activity and increases in the required SRT. In general, the impact of temperature on the required SRT is greater for anaerobic than for aerobic processes. This is offset somewhat because the methane produced in the anaerobic process can be used to heat the influent wastewater. Because the quantity of methane produced is a function of the concentration of biodegradable organic matter in the influent wastewater, the potential heat rise depends on the wastewater strength, as illustrated in Figure 13.17. Two cases are considered. One incorporates recovery of the heat in the bioreactor effluent and uses it to heat the influent wastewater while the other does not. Sufficient energy is available to achieve a significant temperature increase only for wastewaters with biodegradable COD concentrations greater than about 2,000 mg/L if heat recovery is practiced and around 7,000 mg/L if heat recovery is not practiced. Wastewater flow rate also affects the choice between aerobic and anaerobic systems for wastewaters containing 1,000 to 4,000 mg/L of biodegradable COD. The simplicity of aerobic systems generally favors their use for smaller wastewater flows, while the significant energy and solids production savings available favors the use of anaerobic systems for larger wastewater flows.[23]

Wastewater composition also affects the choice between anaerobic and aerobic systems. High-rate anaerobic treatment technology was developed for the treatment of SOM, and the presence of significant quantities of suspended solids adversely impacts its efficiency. As indicated in Figure 9.5, either acidogenesis or methanogenesis can be the rate limiting step in the anaerobic stabilization of SOM, with the nature of the organic matter determining which is slower. If the organic matter is

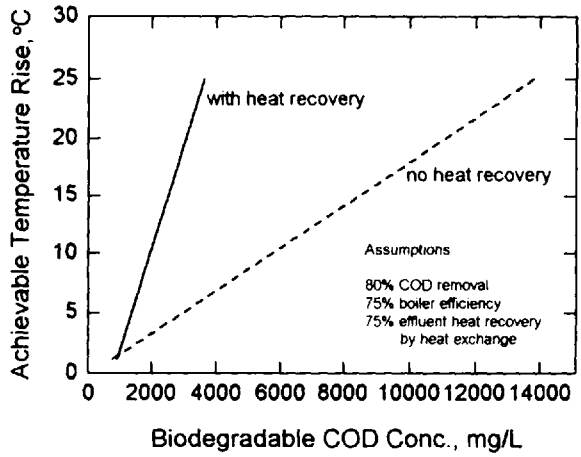


Figure 13.17 Relationship between wastewater strength and achievable temperature rise for anaerobic processes. (From E. R. Hall, Anaerobic treatment of wastewaters in suspended growth and fixed film processes. In *Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes*. J. F. Malina, Jr. and F. G. Pohland, eds. Technomics Publishing, Lancaster, Pennsylvania, pp. 41–118, 1992. Copyright © Technomics Publishing Co., Inc.; reprinted with permission.)

predominantly simple carbohydrates and proteins, methanogenesis will be slower, but can still be accomplished at short SRTs. In fact, some of the high-rate anaerobic systems were developed for food processing wastes containing such constituents. On the other hand, wastes high in lipids require much longer SRTs for acidogenesis, which can increase the SRT required in an anaerobic system. Hydrolysis and fermentation are generally the rate limiting steps in the anaerobic stabilization of particulate organic matter, and longer SRTs are required for them as well. Furthermore, some of the suspended solids are likely to be nonbiodegradable, and will accumulate in the bioreactor, thereby reducing the specific activity of the anaerobic biomass. Both of these factors significantly affect the VOLs that can be applied to an anaerobic bioreactor and negatively impact its economics. The presence of inhibitory or toxic materials also results in significant increases in the required SRT, which negatively impacts the economics of anaerobic processes. Finally, as illustrated in Figure 9.2, differences also exist among anaerobic processes with respect to the waste strengths for which they are suited.

Anaerobic digestion is generally applied to the treatment of high-strength wastewaters, particularly those with high suspended solids concentrations. In fact, historically it has been one of the most widely used processes for stabilizing organic solids produced in wastewater treatment plants. As a consequence, several thousand operating facilities exist around the world. The uniform reaction conditions and long SRTs used provide the conditions necessary for hydrolysis and stabilization of these materials. Volumetric organic loadings approaching those in low-rate anaerobic systems are achieved in anaerobic digestion because of the high wastewater strengths applied. Anaerobic digesters are capital intensive, but have low operating costs. Consequently, they are generally found in larger wastewater treatment plants where the

savings in operating cost more than offset the increased capital investment. For example, at current unit costs it is often found that anaerobic digestion is not cost-effective in wastewater treatment plants with capacities less than 40,000 to 100,000 m^3/day. Nevertheless, because cost relationships and available options were different in the past, they are often found in many older wastewater treatment plants with lower capacities. The increased availability of alternative approaches to solids management and stabilization has reduced the use of anaerobic digestion. However, it remains a viable solids stabilization technology, and continues to be widely used.

Low-rate anaerobic processes can be applied to wastewaters with a range of strengths, including those with high concentrations of suspended solids. They are most often applied to the treatment of wastewaters with biodegradable COD concentrations in the 20,000 to 30,000 mg/L range and to those with high concentrations of suspended solids. A main limitation, however, is the amount of land that must be available for their construction.

High-rate anaerobic processes are applied most often for the treatment of moderate to high strength wastewaters (those with biodegradable COD concentrations up to about 20,000 mg/L) containing mostly SOM. However, there are differences among them regarding their applications. Anaerobic contact and DSFF systems are not generally affected by influent suspended solids to the same extent as the others. Such solids can be accumulated to a certain degree in AC systems and sufficient SRT can be provided to allow their hydrolysis and stabilization to occur. While suspended solids will not adversely impact the performance of DSFF systems, they will generally pass through them with little stabilization.

If a wastewater contains solids that require stabilization, two options exist for handling them if a high-rate anaerobic process is to be used to treat the wastewater: they can be removed either before or after treament. An advantage of the latter option is that solids produced during anaerobic treatment will also be removed, but some systems are not amenable to that alternative. For example, prior removal is generally required for AF and FB/EB systems, in which case the removed solids can be processed in a separate anaerobic digester designed specifically for that purpose. Sedimentation is also often used prior to UASB and hybrid UASB/AF systems, but a larger bioreactor could be provided to allow operation at a lower organic loading rate to encourage hydrolysis and stabilization of the solids that accumulate. Stabilization of the accumulated solids can be further encouraged by periodically diverting wastewater away from the bioreactor to allow them to digest and/or by periodically increasing the bioreactor temperature to increase the rates of hydrolysis and acidogenesis. This may be particularly attractive if wastewater is not continuously applied to the treatment system.

Low-rate and high-rate anaerobic processes are widely used around the world, with several thousand installations in place. Their use to treat liquid waste streams is a newer application of anaerobic technology than anaerobic digestion of solids. Consequently, their development is ongoing. Each of the anaerobic technologies described above has a significant number of installations and a well developed experience base upon which to base process designs and evaluate potential applications. Some aspects of these technologies are proprietary, particularly the media and gas–liquid–solids separation devices. The suppliers of this equipment can be contacted to obtain installation lists and summaries of operating and performance experience.

Several recent symposia summarize current experience with the use of low- and high-rate anaerobic processes to treat municipal and industrial wastewaters.[7,17,24,40,63]

Fermentation of organic solids to produce VFAs for BNR systems is an evolving technology with relatively few installations. However, it is expected that the cost-effectiveness and popularity of BNR systems, coupled with their need for sufficient quantities of VFAs, will result in continued development of fermenation systems. Particular challenges include fermentation in smaller treatment plants where the wastewater is relatively fresh and may contain few VFAs, and where the absence of primary clarifiers means that no primary solids are available to feed to a fermenter. The proceedings of a workshop sponsored by WEF[76] provides a summary of current experience.

13.2 FACTORS AFFECTING PERFORMANCE

Many factors affect the performance of anaerobic treatment systems. These range from process loading factors such as the SRT, VOL, and hydraulic loading rate; to environmental factors such as temperature, pH, nutrient supply, and the presence of toxics; to operational factors such as mixing and the characteristics of the waste being treated. Historically, the stability and performance of anaerobic treatment systems have been considered to be poor in comparison to aerobic systems. However, with improved understanding of the factors that affect their performance, it has been possible to obtain stable and reliable performance. Consequently, a thorough understanding of these factors is critical to successful design and operation.

13.2.1 Solids Retention Time

The role of the SRT in controlling the performance of anaerobic processes is discussed briefly in Chapter 9 and has been referred to in previous sections of this chapter. Solids retention time controls the types of microorganisms that can grow in the process and the extent to which various reactions will occur. While SRT is the fundamental control parameter, it is difficult to routinely determine it in some anaerobic processes. Determination of the SRT is straightforward in flow-through systems such as anaerobic digesters, where it simply equals the HRT. Its calculation is also straightforward in anaerobic contact systems; the rate of solids wastage is controlled to achieve the desired SRT in the same fashion as in the activated sludge process. The SRT can also be measured and controlled in fluidized and expanded bed systems. The bioreactor bed is sampled to determine the VS inventory, and the biomass control device is adjusted to achieve the desired SRT. The SRT can also be measured and controlled in UASB and hybrid UASB/AF systems, but more often solids are simply wasted to maintain a set level for the granular and flocculent sludge layers. While it is possible to determine the solids inventory in low-rate anaerobic processes, solids wastage control is less certain. Likewise, while it is possible to determine the biomass concentration in pilot-scale AF and DSFF systems by removing sections of the media, this is not a practical approach for routine operation of full-scale systems. Consequently, it is not generally possible to control the operating SRT in low-rate, AF, and DSFF systems. Instead, process control is achieved by controlling the VOL, as discussed in the next section.

When the SRTs in pilot-scale anaerobic treatment systems are calculated, it is not unusual to find values of 30 to 40 days, with some systems ranging up to over 100 days.[23,32,68] Such values are significantly higher than required for wastewater treatment, and represent the accumulation of excess biomass. Experience indicates that very stable performance can be obtained from some anaerobic treatment systems, particularly if long SRTs are used. It also indicates that anaerobic systems can be shut down for extended periods of time (up to several months) and that good performance can be restored shortly after they are restarted.[23,32,68] In spite of these desirable features, it is possible that these long SRTs represent underloaded systems that could have been constructed more economically using shorter SRTs, while still achieving acceptable performance.

One benefit of increased SRTs is increased hydrolysis and stabilization of particulate organic matter. This can be particularly important for the stabilization of certain types of wastewater solids. More information will be provided on this topic in Section 13.2.9.

13.2.2 Volumetric Organic Loading Rate

Even though the VOL is not a fundamental parameter determining the performance of anaerobic treatment systems, it is related to the SRT through the active biomass concentration in the bioreactor. It is also a relatively easy parameter to calculate, and it has been used historically to characterize the loading on anaerobic treatment systems. Knowledge of the VOLs that can typically be achieved for a particular process quantifies how effectively the bioreactor volume is being utilized. Used in this fashion, the VOL provides useful information for the design and operation of anaerobic processes. The volumetric organic loading rate, $\Gamma_{V,S}$, can be calculated in units of kg COD/(m^3·day) as:

$$\Gamma_{V,S} = \frac{F(S_{SO} + X_{SO})}{V} \tag{13.1}$$

where $(S_{SO} + X_{SO})$ is the influent wastewater strength in g COD/L (kg COD/m^3), F is the influent wastewater flow rate in m^3/day, and V is the bioreactor volume in m^3. Substitution of Eq. 4.15 into Eq. 13.1 relates the VOL to the HRT:

$$\Gamma_{V,S} = \frac{S_{SO} + X_{SO}}{\tau} \tag{13.2}$$

This shows that the VOL is inversely proportional to the HRT, as illustrated in Figure 13.18. As discussed throughout Section 13.1, VOLs typically range from 1 to 2 kg COD/(m^3·day) for low-rate processes to between 5 and 40 kg COD/(m^3·day) for high-rate processes.

The SRT is defined by Eq. 5.1, just as for all other biochemical operations. However, we saw in Section 9.4.1 that it is often convenient to use the net process yield to relate the biomass inventory to the mass input rate of substrate and the SRT. Rearranging that equation and expressing the biomass concentration and net yield on a VSS basis gives:

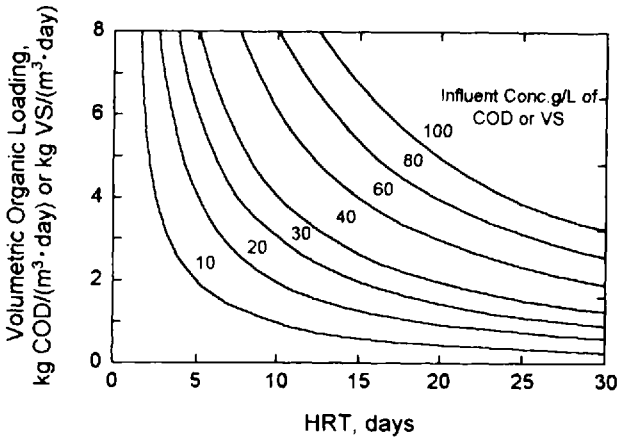


Figure 13.18 Effects of HRT and influent wastewater concentration on the volumetric organic loading of an anaerobic process.

$$\Theta_c = \frac{X_{M,V} \cdot V}{Y_{n,V} \cdot F(S_{SO} + X_{SO})} \tag{13.3}$$

Combining Eqs. 13.2 and 13.3 gives:

$$\Theta_c = \frac{X_{M,V}}{Y_{n,V} \cdot \Gamma_{V,S}} \tag{13.4}$$

Thus, it can be seen that the SRT and the VOL are inversely proportional to each other. Equation 13.4 also shows that for a fixed SRT, the VOL is increased as the biomass concentration is made larger, thereby allowing the bioreactor to be made smaller.

A similar approach is used for solids stabilization systems, such as anaerobic digesters, except that the VOL is expressed in terms of the mass of volatile solids applied, rather than COD, typically having units of kg VS/(m^3·day). Figure 13.18 also presents the relationship between the anaerobic process HRT, the influent volatile solids concentration, and the resulting volatile solids VOL. For single-stage, anaerobic digestion processes, i.e., systems operated without solids recycle, the HRT and the SRT are identical. In these instances, the volatile solids VOL simply indicates how effectively the digester volume is being utilized. For two-stage digestion systems, which incorporate solids recycle, the SRT will be greater than the HRT and the volatile solids VOL indicates both the anaerobic digester volume utilization efficiency and the overall process loading. In both instances volatile solids VOLs typically range from 2 to 6 kg VSS/(m^3·day).[46,58,72,75]

Interestingly, experience indicates that a maximum COD stabilization activity of 1 kg COD/(kg VSS·day) is achieved in a wide variety of anaerobic treatment processes.[23] Although higher values have been reported, especially in conjunction with the treatment of wastewaters rich in acetate, this value can be used to develop an initial estimate of the capability of a particular anaerobic process to stabilize organic matter.

13.2.3 Total Hydraulic Loading

In contrast to the suspended growth systems considered in Part II and the rest of Part III, some of the high-rate anaerobic processes are influenced by the total hydraulic loading (THL) applied to them. This is characteristic of the attached growth processes considered in Parts IV and V, and a detailed discussion of the effects of THL on them is presented in Chapters 16, 18, 19, and 21. This section presents the most important impacts of THL on UASB, AF, hybrid UASB/AF, DSFF, and FB/EB processes. With the exception of UASB systems, all of these anaerobic processes contain attached growth biomass. However, because of the physical similarity between UASB granules and the bioparticles in FB/EB systems, UASB systems also behave like attached growth systems.

The THL is simply the total flow applied to the bioreactor (including recirculation) divided by the bioreactor cross-sectional area perpendicular to the flow. It is calculated as:

$$\Lambda_H = \frac{F + F_R}{A_c} \tag{13.5}$$

where F_R is the recirculation flow rate and A_c is the cross-sectional area. The THL is a superficial velocity, i.e., a theoretical velocity based on the empty bed cross-sectional area.

The THL affects process performance in several ways. For upflow processes with sludge blankets, such as UASB, hybrid UASB/AF, and FB/EB systems, maximum allowable values of the THL correspond to the settling velocity of the particles to be retained in the bioreactor. If the THL exceeds these values, the particles will be washed out of the bioreactor. As a result, the desired biomass inventory and associated SRT cannot be maintained, and the process will fail. Procedures for calculating maximum THL values for FB/EB processes are presented in Section 18.2.2 and typical values are presented in Section 21.2.3.

For UASB and hybrid UASB/AF processes, the maximum allowable THL depends on the nature of the solids developing in the bioreactor.[42,43] For granular solids, the daily average THL should not exceed 72 m/day when treating fully soluble wastewater, and 24 to 30 m/day when treating partially soluble wastewater. The THL can be temporarily increased to 144 m/day for fully soluble wastewaters and 48 m/day for partially soluble wastewaters. For flocculent solids, the daily average THL should not exceed 12 m/day and the maximum THL should not exceed 48 m/day. The factors that lead to development of granular versus flocculent solids are discussed in Section 13.2.9. Knowledge concerning appropriate values of the THL for these bioreactor types continues to evolve, and the reader is urged to consult the literature for the most recent information.

For some bioreactors, a minimum THL must be maintained for various reasons. For AF and DSFF processes a minimum THL is needed to achieve uniform distribution of flow across the bioreactor cross section to minimize short circuiting. For AF processes, values in the range of 10 to 20 m/day appear to be appropriate.[79] For FB/EB processes, a minimum THL must be maintained to fluidize or expand the bed, as discussed in Chapters 18 and 21. As with the UASB and hybrid UASB/AF processes, THL criteria for these processes continue to evolve and the reader should consult the literature for further information.

THL constraints can affect the configuration of the anaerobic bioreactor. The bioreactor cross-sectional area must be adjusted to produce THL values within the necessary range. In some instances, recirculation must be initiated to maintain minimum required THLs. The impacts of the THL constraints on the design of anaerobic processes is discussed briefly in Section 13.3.2 and in substantially more detail in Chapters 19 and 21 for other attached growth systems.

13.2.4 Temperature

As with all biological processes, the performance of anaerobic processes is significantly affected by operating temperature. Best performance is typically obtained by operation in the optimal region of one of the two higher temperature ranges, i.e, 30°C to 40°C for mesophilic or 50°C to 60°C for thermophilic, and most anaerobic processes are designed to do so. These two regions generally represent the optima for growth of the methanogens. Nevertheless, it is possible to grow methanogens at lower temperatures, provided that longer SRTs are used to compensate for the lower maximum specific growth rates. Although anaerobic activity can be sustained at temperatures approaching 10°C, operating temperatures in the 20°C to 25°C range appear to be the lower limit from a practical perspective.[37,40,49,58,62,68]

Although the preceding paragraph focussed on methanogens, operating temperature affects hydrolytic and acidogenic reactions as well. For wastewaters consisting largely of simple, readily biodegradable organic matter, the effect of temperature on methanogenesis is the primary concern. However, for wastewaters consisting largely of complex organic compounds or particulate materials, the effects of temperature on hydrolysis and acidogenesis will be the primary concern. Table 13.4 presents $\hat{\mu}$ and K_s values for biodegradation of VFAs at temperatures of 25°C, 30°C, and 35°C. These data may be used to characterize the impact of temperature on the anaerobic biodegradation of simple organic compounds.

Figure 13.19 shows the combined effects of SRT and temperature on the anaerobic digestion of municipal primary solids. Essentially complete stabilization of biodegradable volatile solids is achieved at an SRT of 10 days when operating at a temperature of 35°C. A moderate increase in SRT to about 15 days is required when operating at a temperature of 25°C, but the stabilization is not complete as indicated by a residual VS concentration at SRT values as long as 60 days. The required SRT

Table 13.4 Average Values of Kinetic Parameters for Anaerobic Enrichment Cultures Grown on Various Volatile Fatty Acids[a]

	35°C		30°C		25°C	
Volatile fatty acid	$\hat{\mu}$ day^{-1}	K_s mg/L as COD	$\hat{\mu}$ day^{-1}	K_s mg/L as COD	$\hat{\mu}$ day^{-1}	K_s mg/L as COD
Acetic	0.36	165	0.26	356	0.24	930
Propionic	0.31	60	—	—	0.38	1145
Butyric	0.38	13	—	—	—	—

[a]Adapted from Lawrence.[37]

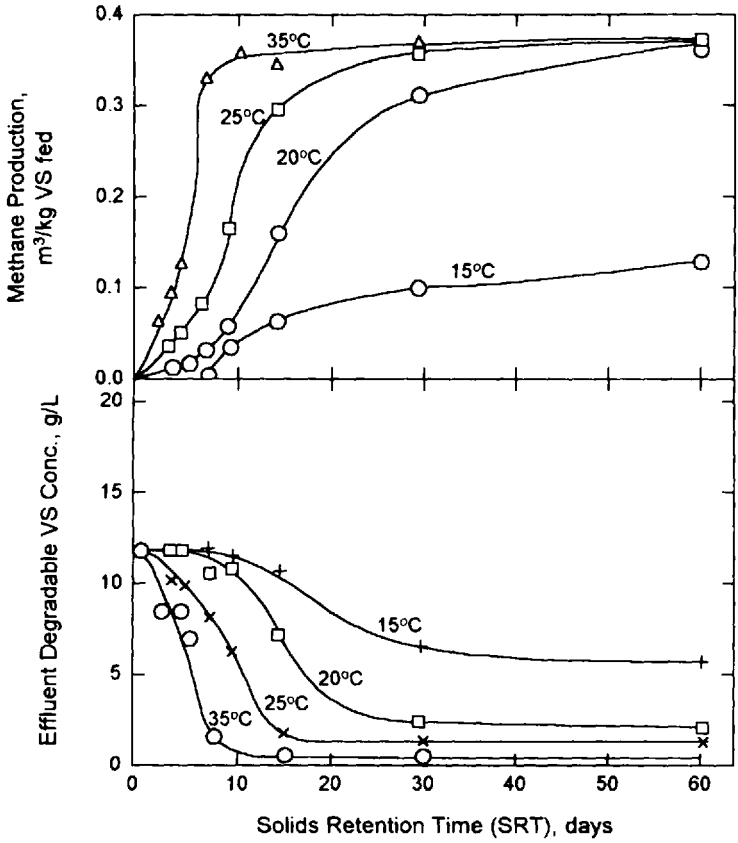


Figure 13.19 Effects of SRT and temperature on the anaerobic digestion of municipal primary solids. (From A. W. Lawrence, Application of process kinetics to design of anaerobic processes. In *Anaerobic Biological Treatment Processes*, ACS Advances in Chemistry Series **105**:163–189, 1971. Copyright © American Chemical Society; reprinted with permission.)

increases to about 25 days when operating at a temperature of 20°C, and a higher residual VS concentration is observed. At 15°C an SRT of about 30 days is required to obtain stable operation, and only about one-half of the biodegradable volatile solids are destroyed at SRTs as long as 60 days. The curves showing the correspondence between VS destruction and methane production suggest that hydrolysis of the solids is generally the rate limiting step at these temperatures. Taken together, these data suggest that a temperature of about 25°C is the practical minimum for the anaerobic stabilization of municipal primary solids.

Temperature variations are also of concern, and it is typically recommended that systems be designed and operated to achieve variations of less than ±1°C each day.[14,71,72,77] Some research indicates that anaerobic processes are capable of reacting successfully to temperature variations; although reaction rates decrease when the temperature is reduced, activity is restored quickly when the temperature returns to the optimum value. In contrast, experience with full-scale systems indicates that performance is adversely impacted by rapid temperature variations of as little as 2°C

to 3°C. This may be because of factors such as mixing and stratification within the bioreactor. Regardless of the mechanism, it appears prudent to adhere to recommended practice and to design and operate anaerobic processes to minimize short-term temperature variations.

Opinions vary concerning the benefits of operation under thermophilic conditions.[8,58,72,75] Potential benefits include increased stabilization rates, resulting in smaller bioreactors; improved solids dewatering properties, which benefits downstream processing; and increased inactivation of pathogenic organisms, which increases the options for disposing of treated solids. Potential drawbacks include the increased energy required to achieve thermophilic operating temperatures and decreased process stability. Increased stabilization rates and increased methane production sufficient to meet the increased heating requirements have been demonstrated by some workers, but not by others. Likewise, improved solids dewatering properties have been observed by some, but not by others. However, increased pathogen inactivation is certain to be observed as a result of thermophilic operation. Decreased process stability because of increased VFA concentrations, increased sensitivity to temperature variations, increased ammonia toxicity, increased foaming, and increased odor potential are all areas of concern. Because of its uncertain benefits and numerous drawbacks, designs based on thermophilic operation should be approached with caution unless site-specific pilot test results and/or full-scale experience are available.

13.2.5 pH

Like all biochemical operations, pH has a significant impact on the performance of anaerobic processes, with activity decreasing as the pH deviates from an optimum value. This effect is particularly significant for anaerobic processes because the methanogens are affected to a greater extent than the other microorganisms in the microbial community.[23,26,46,51,58,60,62] As a consequence, there is a greater decrease in methanogenic activity as the pH deviates from their optimum value. A pH range of 6.8 to 7.4 generally provides optimum conditions for the methanogens, whereas a pH between 6.4 and 7.8 is considered necessary to maintain adequate activity. pH will also affect the activity of the acidogenic bacteria; however, the effect is less significant and primarily influences the nature of their products. A decrease in pH increases the production of higher molecular weight VFAs, particularly propionic and butyric acid, at the expense of acetic acid. As discussed in Section 2.3.2, one mechanism causing this is the buildup of H_2 in the system. As its utilization by methanogens is slowed, it begins to accumulate, which then slows down the production of acetic acid by the acidogens and shifts their metabolism toward other VFAs. The activity of the hydrolytic microorganisms is affected the least by pH deviations from neutrality.

The pH sensitivity of the methanogens, coupled with the fact that VFAs are intermediates in the stabilization of organic matter, can result in an unstable response by anaerobic systems to a decrease in pH.[23,58,71,72] The unstable response may be triggered by a high VOL that results in an increase in the production of VFAs by the acidogenic bacteria. If the increased VFA production rate exceeds the maximum capacity of the methanogens to use acetic acid and H_2, excess VFAs will begin to accumulate, decreasing the pH. The decreased pH will reduce the activity of the methanogens, thereby decreasing their use of acetic acid and H_2, causing a further

accumulation of VFAs and a further decrease in the pH. If this situation is left uncorrected, the result is a precipitous decrease in the pH, the accumulation of higher molecular weight VFAs, and a near cessation of methanogenic activity. This condition is known as a "sour" or "stuck" anaerobic process. It can be corrected in its early stages by resolving the environmental factors causing the imbalance between the acidogenic bacteria and the methanogens. In the case considered above, this could be accomplished by reducing the VOL to the point where the VFA production rate is less than their maximum consumption rate. This will allow consumption of the excess VFAs in the system, thereby causing the pH to return to neutrality and the activity of the methanogens to increase. The VOL can then be increased as the process recovers until the full loading capability is utilized. In extreme cases, decreases in loading must be coupled with the addition of chemicals for pH adjustment, as discussed below.

For an anaerobic process functioning within the acceptable pH range, the pH is controlled primarily by the bicarbonate buffering system. Bicarbonate alkalinity is produced by the destruction of nitrogen-containing organic matter and the reaction of the released ammonia-N with the carbon dioxide produced in the reaction. This is illustrated by Eq. 13.6 for the conversion of primary solids (represented as $C_{10}H_{19}O_3N$) to methane, carbon dioxide, biomass, and ammonium bicarbonate[58]:

$$C_{10}H_{19}O_3N + 4.69\ H_2O \rightarrow 5.74\ CH_4 + 2.45\ CO_2 + 0.2\ C_5H_7O_2N + 0.8\ NH_4HCO_3 \quad (13.6)$$

As illustrated, bicarbonate alkalinity is produced in direct relation to the ammonia-N released. A strong base is needed to react with the carbon dioxide produced in the system to form the bicarbonate. In most instances, ammonia is the strong base, although the cations associated with soaps or the salts of organic acids can also serve to maintain electroneutrality in the reaction with carbon dioxide.

The concentration of bicarbonate alkalinity in solution is related to the carbon dioxide content of the gas space in the bioreactor and the bioreactor pH:

$$S_{BAlk} = 6.3 \times 10^{4} \left(\frac{\bar{p}_{CO_2}}{10^{\,pH}} \right) \quad (13.7)$$

where S_{BAlk} is the bicarbonate alkalinity expressed as mg/L as $CaCO_3$ and $\bar{p}_{CO_2}$ is the partial pressure of carbon dioxide in the gas space expressed in atmospheres.[10,48,58] This relationship is presented in Figure 13.20 and illustrates that typical anaerobic processes operate with bicarbonate alkalinities in the range of 1,000 to 5,000 mg/L as $CaCO_3$ and carbon dioxide partial pressures of 25 to 45%.

When VFAs begin to accumulate in an anaerobic process, they are neutralized by the bicarbonate alkalinity present. Consider for example, acetic acid. Acetic acid is released by the acidogenic bacteria in nonionized form, but exists as acetate ion at neutral pH. The reaction of acetic acid with bicarbonate alkalinity to convert it to acetate is:

$$HCO_3^- + HAc \rightleftharpoons H_2O + CO_2 + Ac^- \quad (13.8)$$

where HAc represents nonionized acetic acid and Ac^- represents acetate ion. When a pH end point of 4.0 is used in the alkalinity analysis, acetate will be partially converted to acetic acid and will, therefore, register as alkalinity. Thus, if VFAs are

present, the total alkalinity will represent the concentration of both bicarbonate ion and VFAs. If the concentration of VFAs is known and is expressed as acetic acid, the bicarbonate alkalinity can be calculated from the total alkalinity as:

$$S_{BAlk} = S_{TAlk} - 0.71(S_{VFA}) \tag{13.9}$$

where S_{TAlk} is the total alkalinity expressed as $CaCO_3$ and S_{VFA} is the concentration of VFAs expressed as acetic acid. The factor 0.71 converts the VFA concentration expressed as acetic acid to $CaCO_3$ and corrects for the fact that approximately 85% of the VFA anions are titrated to the acid form at a pH of 4.0.[48,58] Other organic and inorganic bases, such as sulfides, can also be titrated to their acid form and, consequently, measured as alkalinity. The concentrations of these anions are typically small relative to the bicarbonate concentration, but the potential for such interferences with bicarbonate alkalinity measurement should be recognized.

As discussed above, under stable operating conditions, bicarbonate is the primary form of alkalinity in anaerobic processes. However, under unstable operating conditions VFAs will react with bicarbonate alkalinity, both reducing its concentration and producing carbon dioxide (see Eq. 13.8), which increases the carbon dioxide content of the gas space. Reference to Figure 13.20 illustrates that both of these changes act to decrease the pH in the bioreactor. Stable operation of anaerobic processes is generally achieved by the maintenance of a relatively high concentration of bicarbonate alkalinity so that increased VFA production can be tolerated with a minimal decrease in bioreactor pH.

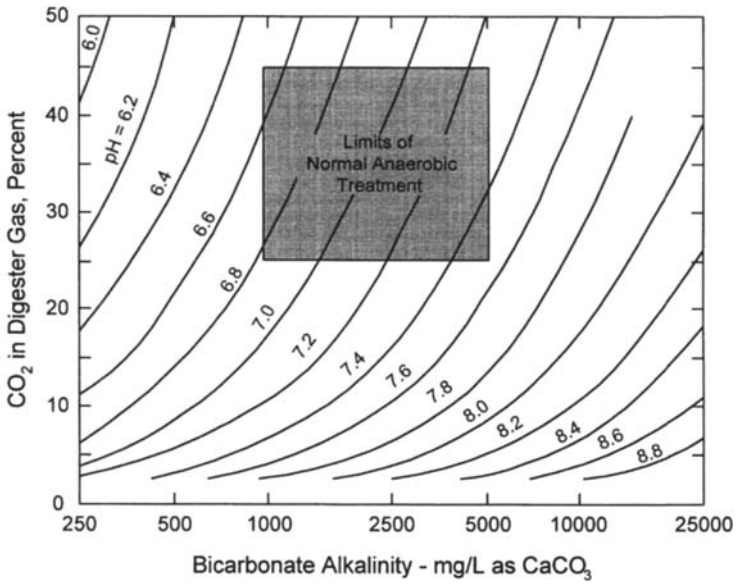


Figure 13.20 Effect of pH on the relationship between the bicarbonate alkalinity of the liquid phase and the carbon dioxide content of the gas phase in an anaerobic process. (From G. F. Parkin and W. F. Owen, Fundamentals of anaerobic digestion of wastewater sludges. *Journal of the Environmental Engineering Division, ASCE* **112**:867–920, 1986. Copyright © American Society of Civil Engineers; reprinted with permission.)

Adverse pH conditions can be corrected by the addition of appropriate chemicals, but, care must be exercised in their selection because of the complex interactions that can occur and the potential for adding toxicants. Commonly used chemicals include sodium bicarbonate, sodium carbonate, lime, sodium or potassium hydroxide, and ammonia.

Sodium bicarbonate is preferred for pH adjustment because its impact is longer lasting and its toxicity potential is low. It adjusts the pH by the direct addition of bicarbonate ions which, as illustrated in Figure 13.20, will result in a direct increase in the pH without affecting the carbon dioxide content of the gas space.

The addition of hydroxide ions by adding lime, sodium hydroxide, or potassium hydroxide, adjusts the pH because the hydroxide ion reacts with carbon dioxide to form bicarbonate alkalinity. Using lime as the example pH adjustment chemical gives:

$$Ca(OH)_2 + 2CO_2 \rightleftharpoons Ca^{+2} + 2HCO_3^- \qquad (13.10)$$

This reaction is accompanied by a decrease in the carbon dioxide content of the gas space, which further contributes to the rise in the bioreactor pH. Unfortunately, further production of carbon dioxide by the microorganisms in the process will restore the original gas space carbon dioxide content and reduce the pH.

The use of carbonate based chemicals reduces the magnitude of the pH variation, as follows:

$$Na_2CO_3 + H_2O + CO_2 \rightleftharpoons 2Na^+ + 2HCO_3^- \qquad (13.11)$$

Comparison to Eq. 13.10 illustrates that only one mole of carbon dioxide is required to produce two moles of bicarbonate from carbonate while two moles of carbon dioxide are required to produce two moles of bicarbonate from hydroxide. Thus, when carbonate-based chemicals are used for pH adjustment, the immediate consumption of carbon dioxide from the gas space is one-half of that when hydroxide-based chemicals are used.

These changes in pH are illustrated in Figure 13.21. Consider an initial condition represented by a gas phase carbon dioxide content of 40% and a bicarbonate alkalinity of 500 mg/L as $CaCO_3$ (point 1), which corresponds to a pH of about 6.3. The addition of sufficient sodium bicarbonate to elevate the bioreactor bicarbonate alkalinity to 2,100 mg/L as $CaCO_3$ would directly increase the bioreactor pH to about 6.9 (point 2). The addition of an equivalent amount of hydroxide-based chemical will result in not only an increase in the bicarbonate alkalinity to 2,100 mg/L as $CaCO_3$, but also in an immediate decrease in the carbon dioxide content of the gas space as it is removed to produce the bicarbonate (Eq. 13.10). The actual decrease in the carbon dioxide content of the gas will depend on the relative gas and liquid volumes in the bioreactor. If the requirement for carbon dioxide is large relative to the amount available, a negative pressure can be created, causing air to be drawn into the gas space, creating an explosive mixture of methane and oxygen. Furthermore, under extreme conditions, removal of carbon dioxide can cause a sufficiently strong negative pressure to collapse the structure. However, for the purposes of this example a decrease to 10% is assumed, and no other adverse consequences are experienced. This results in a pH of approximately 7.5 immediately after addition of the chemical (point 3). However, as additional carbon dioxide is produced by the biomass, the carbon dioxide content of the gas space will increase to its equilibrium

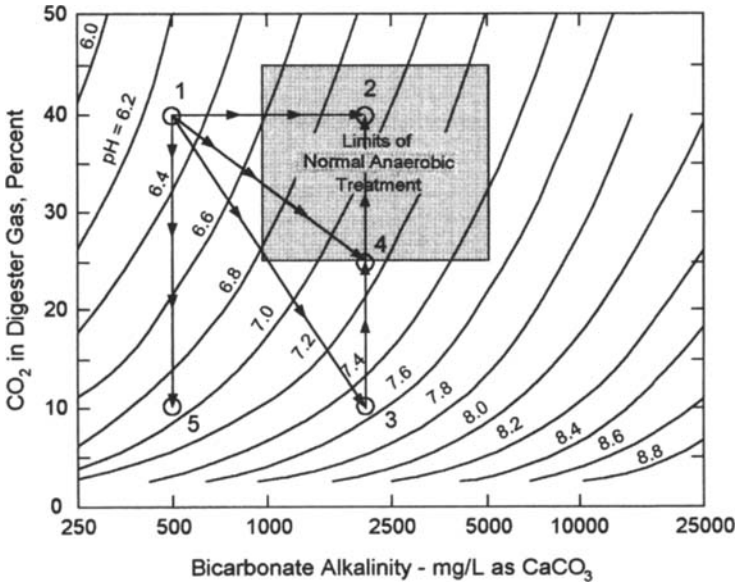


Figure 13.21 Illustration of the effects of changes in the bicarbonate alkalinity of the liquid phase and/or the carbon dioxide content of the gas phase on the pH in an anaerobic process.

value of 40% and the pH will decrease to 6.9, as illustrated by line 3-2 in Figure 13.21. This dramatic variation in pH, from 6.3 (point 1) to 7.5 (point 3) to 6.8 (point 2), can be detrimental to the process. Moreover, it makes pH control difficult from an operational perspective because the relationship between chemical addition and the resulting pH is not straightforward. The effect of adding a carbonate chemical, such as sodium carbonate, is illustrated by point 4 in Figure 13.21, where it is observed that the pH immediately after its addition will be about 7.1. The pH will decrease to 6.9 as carbon dioxide is produced and the carbon dioxide content of the gas is increased back to 40% (point 2). Thus, it can be seen that the addition of carbonate chemicals causes less drastic swings in pH than addition of hydroxide chemicals.

The cations associated with pH adjustment chemicals can also impact the anaerobic process. Use of lime increases the calcium concentration and if it becomes too high, calcium carbonate can precipitate and reduce the effective volume of the bioreactor. As discussed in Section 13.2.6, sodium, potassium, and calcium are toxic if concentrations become high enough. Ammonia can also be used to adjust the pH because it reacts with carbon dioxide to form ammonium bicarbonate:

$$NH_3 + H_2O + CO_2 \rightleftarrows NH_4HCO_3 \tag{13.12}$$

However, this results in the same variations in pH produced by the addition of hydroxide chemicals. Moreover, as discussed in Section 13.2.6, ammonia is also toxic at high concentrations. Thus, it can be seen that no ideal pH adjustment chemical exists and that some degree of care is required when any of them are used.

The control of pH by the removal of carbon dioxide from the gas space has been suggested.[22] As illustrated by point 5 in Figure 13.21, this would require re-

duction of the carbon dioxide content of the gas to about 10% on a sustained basis to achieve the same pH as achieved by increasing the bicarbonate alkalinity to 2,100 mg/L as $CaCO_3$. This would require continuous removal of carbon dioxide from the gas phase.

13.2.6 Inhibitory and Toxic Materials

As discussed in Section 13.1.1, one characteristic of anaerobic processes is their sensitivity to inhibition by chemicals present in the wastewater or produced as process intermediates. As discussed in Section 3.2.7, inhibition causes a reduction in the maximum specific growth rate of microorganisms, thereby requiring an increase in the SRT of a biochemical operation to produce the same effluent that would be produced in the absence of the inhibitor. However, if the inhibitor concentration is increased sufficiently, a toxic response is exhibited and the microorganisms are killed, causing total process failure. Unfortunately, the literature has not always made a clear distinction between inhibition and toxicity. Consequently, in the information that follows the two terms should not be interpreted strictly. However, it should be recognized that, in general, inhibition precedes toxicity as the concentration of a compound is increased. Several inorganic materials can cause an inhibitory response; the materials of greatest concern are light metal cations, ammonia, sulfide, and heavy metals. In addition, sulfate interferes with methane production by providing an alternate electron acceptor. Not only does the sulfide produce an offensive and dangerous gas, but soluble sulfide exerts an oxygen demand that reduces the amount of COD stabilized. Finally, many organic compounds are also inhibitory, particularly to methanogens.

Light Metal Cations. The light metal cations include sodium, potassium, calcium, and magnesium. They may be present in the influent, released by the breakdown of organic matter (such as biomass), or added as pH adjustment chemicals. They are required for microbial growth and, consequently, affect specific growth rate like any other nutrient. Consequently, they must be available if anaerobic treatment is to occur. Nevertheless, their inhibitory nature has been known for over three decades.[48,68] While moderate concentrations stimulate microbial growth, excessive amounts slow it, and even higher concentrations can cause severe inhibition or toxicity. Table 13.5 indicates the concentration ranges over which these various responses occur.

Table 13.5 Stimulatory and Inhibitory Concentrations of Light Metal Cations[a]

	Concentration, mg/L		
Cation	Stimulatory	Moderately inhibitory	Strongly inhibitory
Sodium	100–200	3,500–5,500	8,000
Potassium	200–400	2,500–4,500	12,000
Calcium	100–200	2,500–4,500	8,000
Magnesium	75–150	1,000–1,500	3,000

[a]Adapted from McCarty.[48]

The light metal cations exhibit complex interactions in their effects on microbial growth.[48,58,72] For example, inhibition can be increased when two light metal cations are present at their moderately inhibitory concentrations. This is known as a synergistic response because the combined effects of the two light metal cations exceeds that of either individually. Secondly, the inhibition caused by one light metal cation can be increased if the other light metal cations are present at concentrations below their stimulatory concentrations. Finally, the presence of one light metal cation at its stimulatory concentration can reduce the inhibition of another. This phenomenon is known as antagonism, since the effect is reduced. Table 13.6 summarizes antagonistic responses for the light metal cations and ammonia.

Ammonia. Ammonia-N is a required nutrient and stimulates bacterial growth at low concentrations. For anaerobic processes, ammonia concentrations between 50 and 200 mg/L as N are generally within the stimulatory range.[48,58] However, ammonia is inhibitory at higher concentrations, and toxic if the concentration is high enough. Ammonia may be present in the influent wastewater, or it may be formed as a result of the breakdown of organic materials that contain nitrogen, such as proteins. The production of ammonia by the breakdown of primary solids is illustrated in Eq. 13.6.

Ammonia is a weak base and dissociates in water:

$$NH_3 + H_2O \rightleftharpoons NH_4^+ + OH^- \tag{13.13}$$

Both species are inhibitory, but at significantly different concentrations. Free ammonia (NH_3) is more inhibitory and can cause a toxic response at concentrations of about 100 mg/L as N.[58] On the other hand, ammonium ion (NH_4^+) concentrations as high as 7,000 to 9,000 mg/L as N have been successfully treated without a toxic response with an acclimated culture,[58] although concentrations as low as 1,500 mg/L as N have been reported to be toxic.[48] This type of response has been observed for many materials that can ionize and the nonionized species is often the more inhibitory of the two. As noted in Table 13.6, ammonium ion is also an antagonist for inhibition by potassium.

The pK_a for the dissociation of ammonia is approximately 9.3, so ammonia is present primarily as the ionized species at the pH values typically occurring in anaerobic processes. However, if the total ammonia ($NH_3 + NH_4^+$) concentration is high enough, a sufficient concentration of free ammonia can be present to cause an inhibitory or toxic response. The proportion of total ammonia that is present as free ammonia increases with both pH and temperature. As illustrated in Figure 13.22, vastly different total ammonia concentrations can result in a toxic free ammonia

Table 13.6 Antagonistic Responses for Light Metal Cations and Ammonia[a]

Inhibitor	Antagonist
Na^+	K^+
K^+	Na^+, Ca^{++}, Mg^{++}, NH_4^+
Ca^{++}	Na^+, K^+
Mg^{++}	Na^+, K^+

[a]Adapted from Kugelman and Chin.[48]

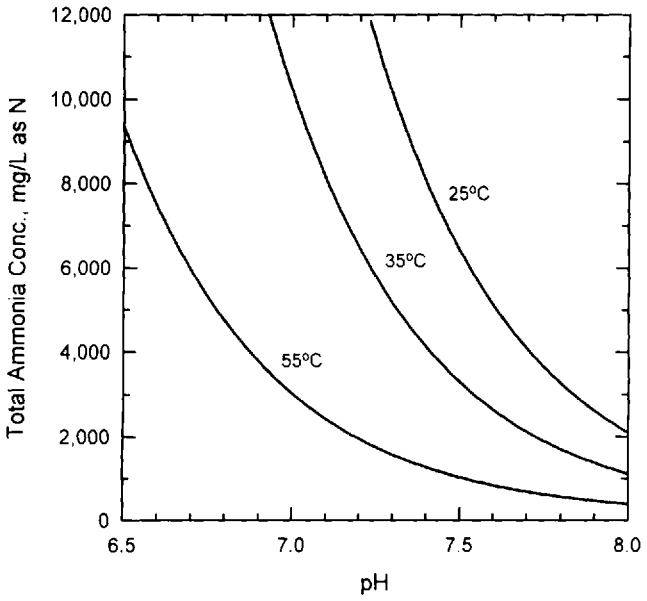


Figure 13.22 Effects of pH and temperature on the total ammonia-N concentration necessary to give a free ammonia concentration of 100 mg/L as N. The curves were generated from equilibrium and thermodynamic constants given in *Water Chemistry*. (V. L. Snoeyink and D. Jenkins, John Wiley & Sons, Inc., New York, 1980.)

concentration (100 mg/L as N), depending on the pH and temperature. For mesophilic conditions (25°C and 35°C), the total ammonia concentration can exceed 10,000 mg/L as N and the free ammonia concentration will still be below 100 mg/L as N at pH values of about 7. However, for thermophilic conditions (55°C) the total ammonia concentration must be maintained below 2,000 mg/L as N to keep free ammonia concentrations below toxic levels. Even for mesophilic operating conditions, total ammonia concentrations of about 2,000 mg/L can result in toxic free ammonia concentrations (100 mg/L as N) as the pH approaches 7.5 to 8.0.

Three strategies are available for reducing ammonia inhibition in anaerobic processes: (1) reduce the temperature, (2) reduce the pH, or (3) reduce the total ammonia concentration. As discussed in the preceding paragraph, Figure 13.22 illustrates the impact of mesophilic versus thermophilic operating temperatures on allowable total ammonia concentrations. This difference should be carefully considered when selecting the temperature range. Moreover, a reduction in operating temperature within a temperature range must also be carefully evaluated. For example, reducing the temperature from 35°C to 25°C causes a noticeable increase in the allowable total ammonia concentration. However, a temperature decrease of this magnitude would also result in a significant reduction in the maximum specific growth rate of the anaerobic biomass, which may be more detrimental to the process than the reduced inhibition associated with the reduction in free ammonia concentration. As further indicated in Figure 13.22, if pH values are relatively high, significant reductions in ammonia toxicity can result from their decrease. High pH values can occur when high-strength wastewaters or solids are treated because of the

high concentrations of bicarbonate alkalinity that result. If such wastewaters also contain high concentrations of ammonia or organic nitrogen, then the concentration of total ammonia will be elevated as well. The pH in such a bioreactor can be decreased by the addition of an acid. Hydrochloric acid is the ideal chemical for this purpose because chloride ion has little or no impact on anaerobic biomass. In addition, the total ammonia concentration can be reduced by dilution of the wastewater or solids with clean water. Care must be exercised if this is done because the larger flow rate may compromise the SRT of the system. However, if an adequate SRT can be maintained, this approach can be used quite successfully.

Sulfide. Sulfide is produced in an anaerobic process through the reduction of sulfate present in the influent and by the degradation of sulfur-containing organic matter (e.g., proteins). Only soluble sulfides are inhibitory and concentrations greater than 200 mg/L cause strong inhibition, while concentrations up to 100 mg/L can be tolerated with little or no acclimation. Concentrations between 100 and 200 mg/L may be tolerated after acclimation.[38,39,58] Sulfide reacts with heavy metal cations, including iron, forming highly insoluble precipitates. In fact, iron sulfide gives anaerobic processes their characteristic black color. Consequently, the concentration of soluble sulfide can be reduced by the addition of iron to the bioreactor, thereby reducing sulfide inhibition.

Hydrogen sulfide is a weak acid and, consequently, at neutral pH is present in equilibrium with the sulfide anion. Hydrogen sulfide is sparingly soluble in water, so it will partition between the liquid and gas phases. Hydrogen sulfide increases the corrosivity of anaerobic process gas and results in the formation of sulfur oxides when the gas is burned. Consequently, control of the hydrogen sulfide content of the product gas is desirable. This too can be done by adding iron to the bioreactor to precipitate the sulfide anion as iron sulfide.

Sulfate itself is not inhibitory to anaerobic bacteria, but it impacts anaerobic processes by providing an electron acceptor that can be used by sulfate reducing bacteria, allowing them to compete with methanogens for the electrons available in the organic matter. This has several effects. First, it produces sulfide, which is inhibitory, as discussed above. Second, it reduces the amount of methane produced because the electrons used to reduce the sulfate are not available for the reduction of carbon dioxide to methane. Third, it reduces the value of the product gas, as discussed above. Fourth, it decreases the removal of COD from the wastewater being treated. Although the organic matter is still oxidized to carbon dioxide, much of the sulfide produced remains in the process stream, where it represents an oxygen demand. Approximately two mg of carbonaceous COD are consumed by sulfate reducing bacteria for each mg of sulfate-S reduced to sulfide-S,[23,53] but any of that sulfide that is still present in the liquid phase exerts a COD. This can have a major impact when anaerobic processes are used to treat relatively dilute wastewaters.

The competition between methanogens and sulfate reducing bacteria is very complex and is influenced by many factors. Consequently, the conversion of sulfate to sulfide may not be complete.[11,13,53,59,68] Nevertheless, in the absence of other specific information, when judgements are being made about the potential for inhibitory sulfide levels being formed, it is generally prudent to assume that all of the influent sulfate will be reduced to sulfide. Some of the produced sulfide will be precipitated by heavy metals and some will partition into the gas phase. Both of these mechanisms will reduce the soluble sulfide concentration and, consequently, the potential

for the development of an inhibitory sulfide concentration. Nevertheless, when the influent COD is low relative to the influent sulfate concentration, insufficient methane gas may be produced to strip the sulfide produced from the liquid phase, resulting in soluble sulfide concentrations that are inhibitory or toxic. Experience suggests that inhibitory soluble sulfide concentrations may develop when treating wastewaters with a COD/SO_4 ratio less than about 7.5.[23] Hall[23] lists sulfide control strategies that can be applied in such cases, including, adding iron salts to precipitate sulfide from solution, purging hydrogen sulfide from the bioreactor liquid, scrubbing hydrogen sulfide from the biogas and recirculating it to the bioreactor to remove sulfide, and using biological sulfide oxidation and sulfur recovery.

Heavy Metals. As with other biochemical operations, heavy metals have strong effects on anaerobic processes, as indicated in Table 13.7 by the low concentrations causing 50% inhibition. Fortunately, only the soluble metal ions are inhibitory and the metal sulfides are extremely insoluble, giving residual heavy metal concentrations much less than the concentrations in Table 13.7. Consequently, heavy metal inhibition to anaerobic processes is often prevented by the sulfide produced in the process. In situations where inadequate sulfide is produced, sulfur can be added. Approximately 0.5 mg of sulfide is needed to precipitate one mg of heavy metal.[46,58] Ferrous sulfide is an ideal chemical to provide supplemental sulfide. Table 13.7 shows that ferrous iron is much less inhibitory than other heavy metals. In addition, the sulfide precipitates of the more inhibitory heavy metals are more insoluble than ferrous sulfide, and consequently the added sulfide will maintain the concentration of those heavy metals at low concentrations. Furthermore, the presence of residual iron will maintain soluble sulfide concentrations at low values. Finally, as long as the pH is 6.4 or above, any excess iron will precipitate as iron carbonate, thereby preventing any inhibition caused by soluble iron.

Volatile Acids. It is uncertain whether VFAs are inhibitory to methanogens.[46,58] Although early evidence suggested that VFA concentrations above 2,000 mg/L were inhibitory to methanogens, when the pH was held near neutral, neither acetic nor butyric acid inhibited methane formation at concentrations up to 10,000 mg/L.[29] Propionic acid was inhibitory at a concentration of 6,000 mg/L at neutral pH, but this is an extremely high concentration that is unlikely to be found in an anaerobic

Table 13.7 Soluble Heavy Metal Concentrations Exhibiting 50% Inhibition of Anaerobic Digesters[a]

Cation	Concentration mg/L
Fe^{+2}	1 to 10
Zn^{+2}	10^{-4}
Cd^{+2}	10^{-7}
Cu^{+}	10^{-12}
Cu^{+2}	10^{-16}

[a]Adapted from Mosey and Hughes.[56]

process at neutral pH.[36] Andrews and coworkers have suggested that it is the nonionized form of the VFAs that is actually inhibitory, with concentrations on the order of 30 to 60 mg/L having an effect.[2,3,4] Volatile fatty acids are weak acids that are largely dissociated at neutral pH. For example, a total acetic acid concentration of approximately 5,500 mg/L is required to produce a nonionized acetic acid concentration of 30 mg/L at pH 7. On the other hand, at pH 6.5 a total acetic acid concentration of only 1,800 mg/L produces the same nonionized acetic acid concentration,[58] showing that pH and VFA concentration are interrelated in their effects. Furthermore, some evidence suggests that the early observations of VFA inhibition were actually a result of the accumulation of H_2 during anaerobic process upsets.[58] While this controversy has not been fully resolved, it appears that inhibition caused by VFAs will be of little concern as long as the pH remains within the normal range for the growth of methanogens (6.8 to 7.4). For pH values below this range, pH impacts themselves will be significant and will be compounded by any inhibition caused by nonionized VFAs.

Other Organic Compounds. As with aerobic processes, a wide range of organic compounds can inhibit anaerobic processes. Also like aerobic processes, significant biodegradation of these chemicals can occur with sufficient acclimation.[6,22,58,68] Table 13.8 summarizes inhibitory concentrations of some typical organic compounds, while Table 13.9 compares the relative effects of several organic compounds on anaerobic processes. The concentration ranges presented in these tables represent the response of anaerobic cultures upon initial exposure to the compounds. However, it has been found that, with acclimation, anaerobic cultures can tolerate concentrations of 20 to 50 times those values while successfully metabolizing the compounds.[58] Table 13.10 demonstrates the biodegradative capability of anaerobic systems by summarizing petrochemical wastewater components that were inhibitory initially, but biodegradable following acclimation. During acclimation, the activity of a methanogenic community may nearly cease. However, even after long periods of inactivity (50 days or more), a community capable of degrading the target compound can develop. This suggests that some organisms survived and served as seed for the development of a healthy community capable of degrading the target compound. Procedures have been developed to assess the effects of compounds on anaerobic cultures, and they may be used to determine the concentration range over

Table 13.8 Concentrations of Organic Compounds Reported to be Inhibitory to Anaerobic Processes[a]

Compound	Inhibitory concentration mg/L
Formaldehyde	50–200
Chloroform	0.5
Ethyl benzene	200–1,000
Ethylene dibromide	5
Kerosene	500
Linear ABS (detergent)	1% of dry solids

[a]Adapted from Parkin and Owen.[58]

Table 13.9 Relative Inhibition of Selected Organic Compounds to Anaerobic Processes[a]

Compound	Concentration causing 50% inhibition, mM
1-Chloropropene	0.1
Nitrobenzene	0.1
Acrolein	0.2
1-Chloropropane	1.9
Formaldehyde	2.4
Lauric acid	2.6
Ethyl benzene	3.2
Acrylonitrile	4
3-Chlorol-1,2-propandiol	6
Crotonaldehyde	6.5
2-Chloropropionic acid	8
Vinyl acetate	8
Acetaldehyde	10
Ethyl acetate	11
Acrylic acid	12
Catechol	24
Phenol	26
Aniline	26
Resorcinol	29
Propanal	90

[a]Adapted from Parkin and Owen.[55]

Table 13.10 Petrochemicals Metabolized by Enriched Methanogenic Cultures[a]

Petrochemical		
Acetaldehyde	Formaldehyde	Phthalic acid
Acetone	Formic acid	Propanal
Adipic acid	Fumaric acid	Propanol
1-Amino-2-propanol	Glutaric acid	2-Propanol
4-Aminobutyric acid	Glycerol	Propionic acid
Benzoic acid	Hexanoic acid	Propylene glycol
Butanol	Hydroquinone	Resorcinol
Butyraldehyde	Isobutyric acid	Sec-butanol
Butyric acid	Maleic acid	Sec-butylamine
Catechol	Methanol	Sorbic acid
Crotonaldehyde	Methyl acetate	Succinic acid
Crotonic acid	Methyl ethyl ketone	Tert-butanol
Ethyl acetate	Nitrobenzene	Valeric acid
Ethyl acrylate	Phenol	Vinyl acetate

[a]Adapted from Parkin and Owen.[55]

which an inhibitory response may be observed.[35,68,81] They may also be used to develop cultures capable of biodegrading a target compound.

The response of both aerobic and anaerobic processes to inhibitory organic chemicals is an area of continued research, and the reader is urged to consult the literature for on-going developments. This topic is discussed further in Chapter 22.

13.2.7 Nutrients

Like all other biochemical operations, nutrients are required by anaerobic processes because they are essential components of the biomass produced. However, biomass yields are much lower in anaerobic processes than in aerobic ones, and this results in reduced nutrient requirements.[23,62] While the nutrient requirements in Table 9.3 are appropriate for anaerobic processes, only about 4 to 10% of the COD removed is converted into biomass, and thus the nutrient quantities required will be much lower. Consequently, adequate nutrients will generally be available when complex wastes are being treated. However, nutrient addition may be required when carbon rich industrial wastes are being treated. Such wastewaters may be deficient in the macronutrients nitrogen and phosphorus. The concentrations of micronutrients such as iron, nickel, cobalt, sulfur, and calcium may also be limiting.[58,62,68] Nickel and cobalt are particularly important for growth of methanogens.

13.2.8 Mixing

As discussed in Section 13.1 and indicated in Table 13.1, an effective mixing system is critical to the successful operation of an anaerobic process. It provides intimate contact between the microorganisms and their substrates, reduces resistance to mass transfer, minimizes the buildup of inhibitory reaction intermediates, and stabilizes environmental conditions. Mixing is an integral part of the design of many high-rate systems. For example, introduction of the influent wastewater directly into the sludge bed in a UASB bioreactor promotes intimate contact between the wastewater and the granules. Likewise, fluidization in a FB/EB bioreactor promotes intense mixing, which allows high process loadings. Mixing is less efficient in other high-rate anaerobic processes, such as AF and DSFF systems, and this is one of the factors restricting their loading. Likewise, poorer mixing, along with less effective mechanisms for solids retention, result in lower allowable loadings for low-rate anaerobic processes.

Mechanical or gas mixing is an integral component of some anaerobic processes, such as anaerobic digestion and anaerobic contact. Several systems have been developed to mix these processes, and the reader is referred to design references for a detailed discussion.[46,72,75] The contents of such processes are viscous, thixotropic slurries, and mixing criteria applied to other processes are not generally applicable. The solids and wastewaters treated may contain rags and hair, which can wrap around and damage mixing equipment, and inorganic solids such as grit, which can accumulate and reduce the effective volume of the bioreactor if mixing is inadequate. Floating material can accumulate in a scum layer, which also reduces effective volume. Given these challenges, it is interesting that anaerobic digester volumetric power inputs are often lower than those used in aerobic suspended growth processes, such as activated sludge and aerated lagoons. Volumetric power inputs in anaerobic

digesters are often in the range of 5 to 8 kW/1000 m^3, but successful performance has been obtained at inputs as low as 1 kW/1000 m^3. The importance of the configuration and efficiency of the mixing system is illustrated by the fact that power densities as high as 20 kW/1000 m^3 have been ineffective in some instances. Egg shaped digesters, shown in Figure 13.4, have superior mixing characteristics and can be properly mixed using lower than normal volumetric power inputs.

Specialized techniques are now routinely used to determine the mixing pattern within full-scale anaerobic digesters.[15,55,82] The most frequently used technique involves the pulse addition of lithium into the digester and the monitoring of its concentration in the effluent for at least 3 SRTs. The results are then analyzed as discussed in Section 4.3.2 to determine the residence time distribution, from which the effective volume of the bioreactor and the proportion of the feed that short-circuits can be estimated. Application of this technique allows the effectiveness of various mixing systems to be determined and compared. It has revealed that significant differences exist in the effectiveness of such systems.

13.2.9 Waste Type

The nature of the wastewater being treated significantly affects its performance in an anaerobic process. One consideration is the relative amounts of soluble and particulate organic matter. Some anaerobic processes are better suited to treat wastewaters containing primarily particulate matter, while others are ideally suited to remove soluble substrates. For example, anaerobic digesters and solids fermentation systems were developed specifically to handle particulate organic matter, while the low-rate processes and some of the high-rate ones, such as AC, AF, and DSFF, can tolerate high concentrations of it. They all effectively retain the particulate material and allow the slow hydrolysis reactions to proceed. On the other hand, FB/EB, UASB, and hybrid UASB/AF systems do not retain particulate organic matter as effectively, allowing it to pass through the bioreactor with little hydrolysis and stabilization. They are better for soluble wastes.

Soluble organic matter can be further subdivided into readily and slowly biodegradable components. Slowly biodegradable soluble substrate consists of high molecular weight and/or recalcitrant materials requiring significant metabolism to convert them to the simple VFAs that are the substrates of the acidogenic bacteria. Examples include polymers such as carbohydrates and proteins, as well as the complex organic compounds found in many industrial wastewaters. Long SRTs may be required to metabolize these materials.[5,6,43,69] One characteristic of high-rate processes is their ability to accumulate high concentrations of biomass, which allows maintenance of long SRTs even though their HRTs are short. Thus, effective metabolism of slowly biodegradable soluble organic matter can be achieved in them. Long HRTs are required to degrade such substrates in anaerobic process that are unable to achieve such an effective separation of SRT and HRT. Examples include the low-rate anaerobic processes as well as AC and DSFF systems. In contrast, a wide range of bioreactor types and process loadings can be used to treat wastewaters containing primarily soluble, readily biodegradable substrates.

The nature of the wastewater has a strong impact on the performance of UASB systems because it affects granule development.[42,43,68] Research is still under way to characterize all of the factors that affect the development of granules, but an inter-

esting model has been developed by researchers at the University of Capetown.[76] It suggests that separation of acidogenesis and methanogenesis, along with a deficiency of the amino acid cysteine, causes the formation of excessive quantities of proteins by the H_2-utilizing methanogen *Methanobrevibacter*. The protein provides a polymeric matrix that results in the formation of granules. Several other researchers have speculated on the role of the filamentous methanogen *Methanosaeta* (formerly *Methanothrix*) in determining the structure of UASB granules. While further research will undoubtedly define the conditions that facilitate granule formation, experience indicates that it is encouraged during the treatment of wastewaters consisting primarily of carbohydrates and retarded during the treatment of wastewaters consisting primarily of VFAs or proteins.[42,43] Granule formation may also be impeded when the wastewater contains a large proportion of particulate or slowly biodegradable organic matter.

The extent and rate of biodegradation of organic solids varies, and this can affect the performance of anaerobic digesters. Approximately 70% of the organic matter in municipal primary solids, measured as either COD or VS, is biodegradable in an anaerobic environment.[37,58,72] In contrast, the biodegradability of waste solids from aerobic biochemical operations depends on how much stabilization they have undergone in the operations from which they came. For example, Gossett and Belser[31] found that the biodegradable fraction of waste activated sludge under anaerobic conditions is equal to the active fraction, as defined in Section 5.1.3. Furthermore, the active biomass degraded under anaerobic conditions in a first order manner, with a rate coefficient of 0.22 day $^{-1}$ at 35°C. In short, the anaerobic stabilization of waste activated sludge is qualitatively and quantitatively similar to its aerobic stabilization, as described in Chapter 12. However, the rate coefficient for waste activated sludge is lower than the rate coefficient for primary solids. Moreover, because the active fraction of waste activated sludge is often on the order of 50%, and only about 80% of the active mass will be stabilized, i.e., $1 - f_p$, only a small fraction of the total organic matter in waste activated sludge will be stabilized during anaerobic digestion. These effects are illustrated in Figure 13.23 where the COD reduction efficiencies of municipal primary solids, waste activated sludge, and a mixture of primary solids and waste activated sludge are plotted as a function of anaerobic digester SRT.

13.3 PROCESS DESIGN

As we saw in Section 13.1, a wide range of anaerobic process options exists. Although the various processes operate according to a unified set of principles, they differ in many ways. In some, such as anaerobic digestion, the bioreactor functions as a CSTR without biomass recycle so that the SRT is equal to the HRT. In others, significant quantities of biomass are accumulated, allowing long SRTs to be maintained at relatively short HRTs. However, because of the mechanisms used to accumulate biomass in some anaerobic processes, it is impossible to calculate the resulting biomass concentration or SRT. In these instances, empirical correlations between the VOL and the performance must be used for design purposes. In short,

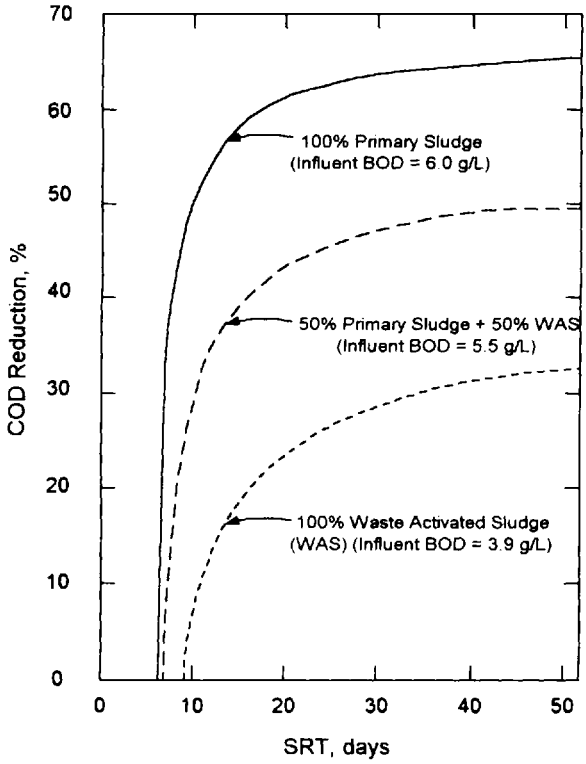


Figure 13.23 Effect of SRT on the stabilization of primary solids and waste activated sludge. (From G. F. Parkin and W. F. Owen, Fundamentals of anaerobic digestion of wastewater sludges. *Journal of the Environmental Engineering Division, ASCE* **112**:867–920, 1986. Copyright © American Society of Civil Engineers; reprinted with permission.)

a wide range of design procedures must be used to accommodate the wide range of anaerobic processes.

13.3.1 Anaerobic Digestion

The design of an anaerobic digester to stabilize solids is quite straightforward. Since anaerobic digesters are simple CSTRs, the SRT is equal to the HRT. Consequently, the process design consists simply of selecting an appropriate SRT and calculating the bioreactor volume directly from the solids flow rate and the definition of HRT, as given by Eq. 4.15. Principal concerns in choosing the SRT include the degree of stabilization and pathogen inactivation required, digester mixing efficiency, requirements for equipment and digester redundancy, and variations in solids flow rates.

Several factors must be considered when selecting the minimum acceptable SRT, including washout of methanogens, hydrolysis of particulate organic matter, and pathogen inactivation. As indicated in Figure 9.5, growth of aceticlastic methanogens can be maintained at SRTs as low as 5 days at 35°C, which is the most common digester operating temperature. While full-scale digesters have been suc-

cessfully operated at SRTs this low,[75] it really is a lower limit and operation at such an SRT places the digester at risk for rapid washout of methanogens and process failure. Furthermore, the hydrolysis of particulate organic matter and its conversion to acetic acid will generally be the rate limiting steps when treating complex organic material. Consequently, longer SRTs are usually used.

A distinction must be made between the design of anaerobic digesters for treatment of primary solids and waste activated sludge, as discussed in Section 13.2.9. Figure 13.19 demonstrates that for municipal primary solids, an SRT of 8 to 10 days is needed at 35°C to ensure reasonably complete stabilization. Figure 13.24 presents

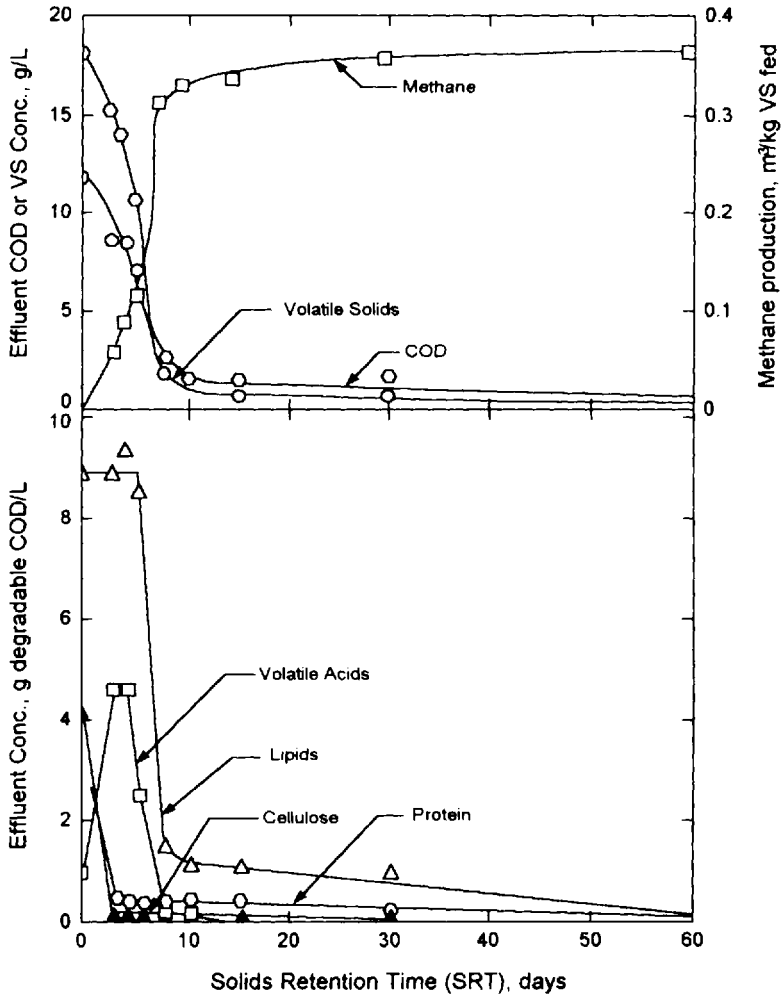


Figure 13.24 Fate of various components of municipal primary solids during anaerobic digestion at 35°C. (From A. W. Lawrence, Application of process kinetics to design of anaerobic processes. In *Anaerobic Biological Treatment Processes*, ACS Advances in Chemistry Series **105**:163–189, 1971. Copyright © American Chemical Society; reprinted with permission.)

information about the degradation of the various components of those solids and shows that the overall performance is limited by the degradation of lipids. This is consistent with Figure 9.5, which shows that anaerobic oxidation of long and short chain VFAs requires an SRT of about 10 days. The data presented in Figures 13.19 and 13.24 are also consistent with Figure 13.23, where it is observed that an SRT of 10 days results in reasonably complete stabilization of primary solids. Thus, an SRT of at least 10 days is needed to stabilize primary solids at 35°C. However, the hydrolysis of the biomass in waste activated sludge occurs at a slower rate than the hydrolysis of primary solids. As a consequence, a longer SRT is required if waste activated sludge is to be stabilized. Using Figure 13.23 as a guide, an SRT on the order of 15 to 20 days is required to achieve substantial stabilization of waste activated sludge. These conclusions are consistent with observations at full-scale plants.[72,75]

Pathogen control is a relatively new requirement for anaerobic digesters. It has been known for some time that digestion reduces the concentration of indicator organisms. In fact, that is one reason anaerobic digestion has been used. However, the purposeful design of digesters to achieve a specific degree of pathogen control is new, and few data exist upon which to base such a design. Currently, U.S. regulations require a minimum SRT of 15 days for anaerobic digesters operating at 35°C to ensure effective reduction of pathogens in municipal wastewater solids.[70] Operation of anaerobic digesters in series also increases pathogen destruction, just as it does in aerobic digesters. Continued evolution of procedures to design anaerobic digesters to control pathogens is expected.

Once the SRT, i.e., HRT, has been selected, the effective volume of the digester is calculated by multiplying the design solids flow rate by the SRT. The design solids flow rate should be for the month or week in which the highest volume of solids is produced to allow the digester to function properly under all reasonable operating conditions. Variations in both the mass of solids produced and the performance of upstream solids thickening devices should be considered in choosing that flow rate. The total volume is then calculated considering the relationship between the effective and total volumes. The effective volume is less than the total volume because of ineffective mixing, leading to the accumulation of grit in the bottom and scum at the top of the digester. In digesters with older style mixing systems, the effective volume can be less than 50% of the total volume, but in digesters with modern mixing systems the effective volume is generally at least 90% of the total volume. Volume should also be allocated to grit and scum accumulations. Typical designs allocate the volume of the floor cone (see Figures 13.2 and 13.3) to grit accumulation and the top 0.6 m of the digester to scum accumulation.[52,72,75] Minimal grit and scum accumulations occur in egg shaped digesters (Figure 13.4), so the total and effective volumes can generally be assumed to be the same.

Once the total bioreactor volume has been determined, the number of individual units and their dimensions must be selected. Provisions must be made for units to be removed from service for maintenance, so a minimum of two units should be provided. The impacts on performance of having a unit out of service must also be considered, and this may dictate the number of units provided and/or the total volume. In doing so, it may be assumed that the unit will be removed from service during average, rather than peak, solids production. The gas production rate is estimated based on the mass of volatile solids stabilized and the conversion factor of

0.7 standard m^3 of methane produced/kg of VS destroyed, as presented in Section 13.1.1. The process design for an anaerobic digester to stabilize the waste solids produced at a municipal wastewater treatment plant is illustrated in the following example.

Example 13.3.1.1

An anaerobic digestion system is to be designed to stabilize the solids produced by a municipal wastewater treatment plant. It must be capable of destroying pathogens, implying that the SRT must be at least 15 days at 35°C. The estimated masses of primary solids and waste activated sludge to be produced daily under various conditions are given in Table E13.1. After blending and thickening, the solids concentration entering the digestion system is expected to average 60 g/L (kg/m^3) and to range from 50 to 70 g/L. The volatile solids concentration is 75% of the total solids concentration. Design the system with multiple digesters, but assume that one will be taken out of service for cleaning only under average loading conditions.

a. What solids flow rates must be processed by the system?
The mass flow rates of dry solids under various conditions are given in Table E13.1. These may be converted to volumetric flow rates by assuming solids concentrations. It is likely that the thickener can maintain the average solids concentration under average and maximum month conditions, but that performance will deteriorate during the maximum solids production week. Consequently, the average solids concentration is used to calculate the average and maximum month volumetric flow rates but the minimum solids concentration is used to calculate the maximum week volumetric flow rate. The results are summarized in Table E13.2.

b. What SRT should be used in the design?
Because an excellent degree of solids stabilization is desired under average loading conditions, an SRT of 20 days is appropriate, based on Figure 13.23. This value should be attained even during the maximum solids production month, but it is unrealistic to maintain it during the maximum week. However, to ensure pathogen destruction under all conditions, an SRT of at least 15 days must be maintained even during the maximum week.

c. What effective total digester volume must be provided?
The required effective digester volume must be calculated in two steps. First the volume required for each flow rate must be calculated based on the assumption that all units are in service. Since the SRT is the same as the

Table E13.1 Solids Production Rates for Design of the Anaerobic Digester in Example 13.3.1.1

Type of solids	Mass of dry solids, kg/day		
	Average	Maximum month	Maximum week
Primary	18,000	22,500	27,000
WAS	16,000	20,000	24,000
Total	34,000	42,500	51,000

Table E13.2 Anticipated Volumetric Solids Flow Rates Under Various Conditions for Example 13.3.1.1

	Solids mass kg/day	Concentration kg/m^3	Flow rate m^3/day
Average	34,000	60	567
Maximum month	42,500	60	708
Maximum week	51,000	50	1,020

HRT, this is done by multiplying the volumetric flow rate by the SRT. The results are given in Table E13.3. If all units could be kept in service all of the time, then the maximum week would control the design and a total volume of 15,300 m^3 would be required. However, it must be possible to take a unit out of service for maintenance during average conditions, so this must also be considered. If two units were used, then one would have to have a volume of 11,340 m^3 under average conditions to maintain the 20 day SRT, making the total volume 22,680 m^3. This is larger than the volume required during the maximum month or maximum week since both units would be in service then, and would control. Similarly, if three units were used, two would have to have a total volume of 11,340 m^3 under average conditions, making the system volume 17,100 m^3. This, too, is larger than the volume required during the maximum month or maximum week, and would control. In this case, using three units reduces the total volume by 25%.

Some savings in digester volume could be achieved by allowing the SRT to decrease to 15 days during the period when one unit was out of service for maintenance. This would have only a minimal impact on performance, as seen by Figure 13.23, and would still ensure pathogen destruction. If this were done, the total effective volume for a two unit system under average conditions would be 17,010 m^3. This is larger than the volumes required for the maximum month and maximum week, which would remain unchanged from the values in Table E13.3, and thus would control. However, for a three unit system, the total effective volume under average conditions would be 12,760 m^3. This is smaller than the volume required for the maximum month and maximum week, so the maximum week would control. Thus, a three unit system would have to have a total volume of 15,300 m^3. Consequently, in this case, using three units only reduces the total volume by 10%.

The choice between these possible designs would have to be made on the basis of economics. However, given the small sacrifice in performance associated with short-term operation at a 15 day SRT, a reasonable decision would be to allow the SRT to drop to 15 days when one unit is out of service for maintenance and to use two units, with a total volume of 17,010 m^3.

d. What volatile solids destruction efficiency and methane production rate would be achieved under the three loading conditions with all units in service?

The volatile solids destruction efficiencies for the primary solids and waste activated sludge must be estimated separately and then combined to obtain the overall digester performance. From Figure 13.23, the COD destruction efficiency for primary solids, which equals the volatile solids destruction

Table E13.3 Required Total Digester Volumes Under Various Conditions with All Units in Service for Example 13.3.1.1

	Flow rate m^3/day	SRT days	Effective volume m^3
Average	567	20	11,340
Maximum month	708	20	14,160
Maximum week	1,020	15	15,300

efficiency, will be about 60% for SRTs of 15 to 20 days. The effect of SRT over that range is so small that it need not be considered. Although the volatile solids (or COD) destruction efficiency of waste activated sludge depends on its biodegradable fraction, which depends in turn on the operating conditions of the activated sludge system, we will assume that the curve in Figure 13.23 is applicable. It suggests that the volatile solids destruction efficiency of the waste activated sludge would be 20 to 25% at an SRT of 15 to 20 days. Consequently, to be conservative, we will use 20%. Using this information, the overall volatile solids destruction efficiency can be calculated as shown in Table E13.4. It would be around 41% for all three situations and would represent stable solids, as defined in Section 12.1.1.

Table E13.4 Estimation of the Volatile Solids Destruction Efficiency and Methane Production Rate for the Anaerobic Digester in Example 13.3.1.1

	Average	Maximum month	Maximum week
Primary solids			
Total solids, kg/day	18,000	22,500	27,000
Volatile solids,[a] kg/day	13,500	16,875	20,250
Volatile solids destroyed[b], kg/day	8,100	10,125	12,150
Methane[c], m^3/day	5,670	7,090	8,505
Waste activated sludge			
Total solids, kg/day	16,000	20,000	24,000
Volatile solids[d], kg/day	12,000	15,000	18,000
Volatile solids destroyed[d], kg/day	2,400	3,000	3,600
Methane[c], m^3/day	1,680	2,100	2,520
Total solids			
Total solids, kg/day	34,000	42,500	51,000
Volatile solids, kg/day	25,500	31,875	38,250
Volatile solids destroyed			
kg/day	10,500	13,125	15,750
Percent	41	41	41
Methane, m^3/day	7,350	9,190	11,025

[a] 0.75 × Total solids.
[b] 0.6 × Volatile solids.
[c] 0.7 m^3/kg VS destroyed.
[d] 0.2 × Volatile solids.

Furthermore, by assuming that destruction of 1 kg of VS results in the formation of 0.7 m^3 of methane under standard conditions, the methane production rate can also be estimated, as shown in the table. The average production rate would be 7,350 m^3/day. This information can be used to plan for use of the methane.

Another important design consideration for anaerobic digesters is the feeding frequency. Wastewater sludges are thixotropic, and they contain grit, rags, and other debris that can clog piping if adequate velocities are not maintained. Furthermore, the presence of debris requires the use of minimum pipe diameters of 10 cm. The need to maintain minimum velocities in sludge piping often precludes continuous feeding of digesters. Fortunately, because of the relatively long SRTs and HRTs used, periodic feeding will not adversely impact digester performance if the time between feedings is sufficiently short.[71,72,75,77] A feeding frequency of several times per day spaced relatively uniformly will provide acceptable performance.

13.3.2 Low- and High-Rate Anaerobic Processes

Low- and high-rate anaerobic processes are used primarily to treat industrial wastewaters and occasionally to treat municipal wastewaters. Because of the mechanisms that some use to accumulate active biomass, it is often impossible to precisely determine the degree of biomass accumulation and the resulting SRT. Consequently, process performance is commonly correlated with the volumetric organic loading rate. In many instances, this is done by operating a pilot plant for a particular bioreactor type treating a specific wastewater. In other instances, the correlation can be based on experience. Further information on the use of these procedures and detailed examples are presented in Chapter 21.

The general procedure used to design a low- or high-rate anaerobic process is as follows:

- Characterize the wastewater to be treated. Characterization involves determination of conventional wastewater parameters such as total and soluble BOD_5 and COD, total and volatile solids, total and volatile suspended solids, pH, alkalinity, temperature, and nutrient concentrations. It also involves assessment of the nature of the organic matter present, i.e., readily or slowly biodegradable; carbohydrate, protein, or synthetic organic compounds of a particular type, and the potential presence of inhibitory materials.
- Summarize experience with treatment of the particular type of wastewater in the low- or high-rate anaerobic process.
- Compare the subject wastewater with the characteristics of the low- or high-rate anaerobic process. Some preliminary cost analyses may be conducted to help discriminate among the various process options. Based on the available information, the option or options most appropriate for treating the subject wastewater can then be selected.
- Determine the need for bench or pilot-scale studies of treatment of the subject wastewater in the selected process(es). This will depend on the available knowledge base.
- Conduct bench or pilot-scale studies, as appropriate.

- Develop a correlation between process performance (generally COD removal efficiency) and the VOL, or some other measure of loading. Also characterize other pertinent design parameters such as HRTs, THLs, and bioreactor geometry.
- Use the process performance relationships developed above to size and configure the bioreactor.
- Calculate the methane production rate using the stoichiometric relationship 0.35 standard m^3 of methane per kg of COD stabilized.
- Perform a heat balance and determine the need to insulate the bioreactor and/or to provide supplemental heat. If supplemental heat is needed, size the system.
- Determine the need for any ancillary facilities such as nutrient addition, pH adjustment or alkalinity addition, iron addition to control sulfide, or sulfur addition to control heavy metal toxicity.
- Summarize the results of the process design in a succinct table of process loadings and required facilities.

Several procedures are used to size and configure low- and high-rate anaerobic processes. Many anaerobic processes are directly analogous to other suspended and attached growth systems discussed elsewhere in this book, and the procedures used to size and configure them are similar.

AC systems are analogous to the activated sludge process, as described in Chapters 9 and 10. Thus, the general design procedures described in Section 9.4 are applicable to AC systems. The design SRT is selected and used to calculate the necessary bioreactor suspended solids inventory and waste solids mass flow rate. The bioreactor MLSS concentration is then chosen considering the trade-off between the sizes of the bioreactor and clarifier, and the bioreactor is sized. Finally, the methane production rate is calculated with Eq 9.6.

Upflow anaerobic sludge blanket, hybrid UASB/AF, and FB/EB processes are designed using the procedures for submerged attached growth bioreactors described in Chapter 21. Upflow anaerobic sludge blanket and hybrid UASB/AF processes behave essentially as upflow packed bed bioreactors, and can be sized and configured using the procedures described in Section 21.3.2. FB/EB processes can be designed using the procedures outlined in Section 21.3.3. These procedures involve selection of appropriate VOLs and THLs and determination of the necessary bioreactor volume, cross-sectional area, and recirculation flow rates (for FB/EB processes) using these criteria. The waste solids mass flow rate and methane production rate can be calculated using the procedures described above for the AC process.

Anaerobic filter and DSFF processes are similar in configuration to trickling filters, and are designed using procedures like those described in Chapter 19. Volumetric organic loading and THL criteria are used most often, so the approach presented in Section 19.3.2 is appropriate. Anaerobic lagoons are generally sized and configured using VOL and HRT criteria, as discussed in Chapter 14. Again, the waste solids mass flow rate and methane production rate can be calculated using the procedures described above for the AC process.

Rather than repeat information presented in greater detail elsewhere, the reader is referred to the appropriate sections of this book for detailed descriptions of the process design procedures. Further information on the design of anaerobic processes is provided in recent books devoted entirely to anaerobic systems.[47,48]

13.3.3 Fermentation Systems

The primary objective of most anaerobic processes is stabilization of biodegradable organic matter through its conversion to methane. In contrast, the objective of fermentation processes is conversion of biodegradable organic matter to VFAs and the harvesting of those VFAs for addition to biological nutrient removal systems.

Several differences exist between fermentation systems and other anaerobic processes. The first is the SRT. Conversion of biodegradable organic matter to VFAs is accomplished by operation at SRTs that allow the growth of hydrolytic and acidogenic bacteria but preclude the growth of aceticlastic methanogens. The latter is necessary because acetic acid is the most desireable VFA for nutrient removal systems, and thus we do not want it to be converted to methane. Analysis of Figure 9.5 suggests that growth of H_2-utilizing methanogens is likely at the SRTs used, so that some methane will be produced. This results in the loss of some COD in the form of methane, but this loss is beneficial because the consumption of H_2 minimizes its partial pressure in the bioreactor and allows the fermentation reactions to proceed with acetic acid as the main product. Suppression of the growth of sulfate reducing bacteria is also desirable since they will consume acetic acid. Experience with full-scale fermentation systems indicates that their growth can also be controlled by appropriate selection of the SRT.

A second difference is that fermentation bioreactors are generally operated at ambient temperature without heating. Since only limited quantities of methane are produced, heating generally requires an external energy source. Because fermentative bacteria can grow at significantly lower SRTs than aceticlastic methanogens, the economic benefit of reduced bioreactor volume as a result of elevated temperature is much less for fermentation systems. Operation at reduced temperature also makes it easier to limit the growth of aceticlastic methanogens and maximize the production of VFAs.

Another difference is the operating pH. Since the primary objective of a fermenter is production of VFAs, the bioreactor pH will be significantly less than the pH in methanogenic anaerobic processes. pH values are typically less than 6, and may be less than 5. Reduced pH values also aid in controlling the growth of aceticlastic methanogens.

Although the production of methane is limited, other gases are produced. Significant quantities of carbon dioxide are generated in the hydrolytic and acidogenic reactions, along with some H_2, which will be converted to methane by H_2-utilizing methanogens. Limited quantities of hydrogen sulfide will be produced if sulfate reducing bacteria are able to grow, and nitrogen may also be present because of the entrance of air into the bioreactor or the reduction of any nitrate-N present. Thus, the gas produced will consist primarily of carbon dioxide with small quantities of methane and trace quantities of nitrogen, H_2, and hydrogen sulfide. Consequently, the gas will not generally be combustible.

The typical feed to a fermentation process is primary solids collected from the influent wastewater. At the SRTs and temperatures used, only a portion of the biodegradable organic matter in those solids is converted to VFAs, with yields on the order of 0.05 to 0.3 g VFA produced/g VS fed to the fermenter.[18,19,66,76] Consequently, the solids still contain significant quantities of biodegradable organic matter that must be stabilized prior to final disposal. Primary solids also contain inert suspended solids

(both volatile and fixed). Thus, the VFAs are normally separated from the solids stream for addition to the BNR process, while the remaining solids are sent for further treatment in the solids processing train.

As illustrated in Figure 13.16, gravity settling is often used for liquid–solids separation in fermentation systems, with the overflow carrying the VFAs for use in the downstream BNR system. Since the concentration of VFAs, which are soluble, is the same in the overflow and underflow from the settler, the recovery of VFAs will be equal to the fraction of flow leaving the settler via the overflow. However, because the solids in the bioreactor are highly concentrated, little additional concentration can occur in the settler, which means that, without dilution, the overflow rate will be a small fraction of the inflow rate, thereby limiting VFA recovery. This problem is overcome by adding an elutriation stream to increase the total flow and make the overflow a larger fraction of the total.

The design of a two-stage completely mixed thickener/fermenter, shown schematically in Figures 13.15 and 13.16, illustrates the solids fermentation processes. It is presented in the following example.

Example 13.3.3.1

A primary solids flow of 385 m^3/day at a solids concentration of 25 g/L (75% volatile) is to be fermented to produce VFAs to add to a BNR process. A two-stage completely mixed thickener fermenter is to be used, as illustrated in Figure 13.16. The process will be sized based on the following assumptions: primary solids are added directly to the fermenter, no overflow from the primary clarifier is added to the fermenter, no thickened solids are recycled from the thickener to the fermenter, and overflow from the primary clarifier is added to the fermenter effluent to dilute it prior to the thickener. Experience with fermentation of these solids indicates that a conversion efficiency of 0.12 g VFA/g VS fed can be achieved at an SRT of 5 days. In addition, the fermented solids will thicken to 40 g/L. Design the system for 80% recovery of the VFAs produced.

a. What is the volume of the fermenter?
At the design condition the completely mixed fermenter operates as a CSTR. For a CSTR the volume is just the flow rate times the SRT. Therefore:

$$V = (385)(5) = 1{,}925 \text{ m}^3$$

b. How many kg/day of VFAs will be produced?
The mass of volatile solids fed to the process is:

$$(25)(385)(0.75) = 7{,}220 \text{ kg VS/day}$$

$$\text{VFA production} = (0.12)(7{,}220) = 866 \text{ kg VFA/day}$$

c. What volume of overflow from the primary clarifier must be added to the fermenter effluent if 80% of the VFAs are to be recovered in the gravity thickener overflow?
The thickened solids concentration will be 40 g/L. Minimal destruction of solids will occur in the fermenter. Therefore, the mass of solids in the thickener underflow will be approximately equal to the solids fed to the process. From a mass balance, the thickened solids flow rate will be:

$$\text{Thickened solids flow} = \frac{(385)(25)}{40} = 241 \text{ m}^3\text{/day}$$

To achieve 80% VFA recovery, the thickener overflow must be 80% of the total flow leaving the thickener and the thickened solids flow must be the other 20%. Since the total flow out must equal the flow in, the flow to the thickener must be:

$$\text{Thickener influent flow} = \frac{241}{0.2} = 1{,}205 \text{ m}^3\text{/day}$$

The thickener influent flow consists of primary solids plus overflow from the primary clarifier. Thus, the primary overflow required is:

$$\text{Primary overflow} = 1{,}205 - 385 = 820 \text{ m}^3\text{/day}$$

13.3.4 Other Design Considerations

Once the process design is completed, the design of the other components of the system can commence. Mixing and recirculation systems must be selected and sized, with the type depending on the particular anaerobic process being designed. A heat balance must also be done, as discussed previously. If that balance shows that more methane will be produced than is needed to heat the process, then plans can be made for use of the gas. One potential use is in an engine-driven electric generator, with waste heat from the engine being used to heat the anaerobic process. Alternatively, excess gas can be used directly for various heating purposes at the facility, or it can be processed and sold as a fuel. Insulation of the bioreactor will reduce its heat requirements, and the cost of the insulation can be compared to the value of the extra gas made available to determine whether insulation is justified.

Materials selection is of particular concern in the design of anaerobic processes. Corrosion of process components is minimal as the long as the environment in which they are housed remains completely anaerobic. For example, concrete inside of anaerobic digesters that have been in service for several decades is generally in excellent condition because the environment has consistently been anaerobic. However, corrosion can be excessive at interfaces between anaerobic and aerobic environments because reduced anaerobic reaction products can be oxidized to acidic products as they come in contact with oxygen. One example is hydrogen sulfide, which can be oxidized to sulfuric acid that will attack metal and concrete bioreactor components, causing rapid deterioration. Care must also be exercised in the handling of anaerobic process gas because a combustible mixture can result if it mixes with air. Standard safety equipment is available to prevent atmospheric air from entering anaerobic bioreactors and to suppress an explosion if one begins. However, these devices are not foolproof, and care must be exercised in anaerobic bioreactor design and operation. Further details on the physical design of anaerobic processes are available elsewhere.[52,72,75]

13.4 PROCESS OPERATION

The great variety of anaerobic processes results in a corresponding multiplicity of process monitoring and control techniques, as well as numerous operating problems. Nevertheless, because all of the processes employ similar microbial communities, a

number of similarities exist between the monitoring and control techniques used and the operating problems encountered. These similarities are discussed below.

13.4.1 Process Monitoring and Control

Control of anaerobic processes is accomplished primarily by maintaining appropriate loadings and operating conditions. Loadings are controlled by controlling the rate at which biodegradable organic matter is added to the process. Operating conditions of particular concern include temperature and pH. As discussed in Section 13.2.4, temperature must be maintained in an optimum range, but, more importantly, changes in temperature must be held to less than 1°C per day. Optimum performance is generally obtained at pH values between about 6.8 and 7.4. Lower pH values lead to inhibition of methanogens, while higher pH values can lead to ammonia toxicity because of increased free ammonia concentrations. Fortunately, most anaerobic processes operate naturally within this pH range as a result of the carbonate/bicarbonate buffering system. The pH may deviate from this desirable range during process upsets, and pH adjustment chemicals must be added as discussed in Section 13.2.5. Process upsets can result from temporarily high loadings, deviations in the environment provided, or the presence of toxic or inhibitory materials in the bioreactor influent. While pH adjustment is necessary to prevent process failure, the root cause of the upset should be identified and corrected to ensure long-term process stability.

Several parameters can be monitored to assess anaerobic process performance. As indicated by the previous discussion, deviations in bioreactor pH are associated with process upsets. However, both experience and theoretical analysis indicate that pH is not a good indicator of process upsets.[71,77] Because of the buffering capacity inherent in the system, by the time that a noticeable decline in bioreactor pH occurs, the upset may be well under way. Other indicators, such as the relative proportions of VFAs and alkalinity, or the methane production rate, are better indicators of impending failure.

The ratio of VFAs to alkalinity indicates the relative proportion of compounds acting to lower the pH and of buffering capacity acting to maintain it. Any change that suggests an increase in acids or a decrease in buffering capacity indicates an imbalance between the acid forming and consuming microbial populations and an impending upset. Alkalinity concentrations are generally measured by titration, whereas VFA concentrations can be measured directly by gas chromatography or by titration.[1] The VFA to alkalinity ratio and the bioreactor pH should be plotted chronologically to allow detection of trends indicative of an impending upset and to assist in the identification of potential causes.

Methane production is another good indicator of process performance because it is proportional to the mass of biodegradable organic matter stabilized.[48,58,71,77] It is also a direct indicator of the activity of the methanogens. Various specific indicators have been used to quantify process performance. One is the methane content of the gas produced. A decrease in this parameter suggests a decrease in the activity of the methanogens. Another parameter is the volume of methane produced per unit of COD or VS fed to the bioreactor. As long as the composition of the bioreactor feed remains constant, then a fixed proportion of it should be converted into methane if operating conditions remain constant. Therefore, deviations in this ratio suggest deviations in operating conditions. Methane production will change more quickly than

bioreactor effluent COD or VS concentrations, and consequently, the ratio of methane produced to COD or VS fed provides an early indication of decreased bioreactor performance. This ratio can also be plotted chronologically and the trends used to identify the onset of upsets.

For many anaerobic processes, the collection of operating data and calculation of process performance indicators on a daily basis is adequate. For these, the HRT is on the order of several days, causing process changes to occur over the course of days. For the high-rate processes, however, HRTs are just a few days and significant changes can occur from one day to the next. These processes benefit from the increased data collection and analysis frequency provided by on-line analysis. Research is ongoing to evaluate the use of various on-line control strategies, such as those based on gas-flow rate and composition, and bicarbonate alkalinity measurements.[12,27,54]

13.4.2 Common Operating Problems

Two of the most common operating problems in anaerobic processes are foaming and the formation of precipitates. Two types of foaming can occur in anaerobic processes. One can occur in all of them. It is associated with incomplete metabolism of the influent organic matter, which leads to the production of intermediates with surfactant properties. Because of the surfactants, the gas produced by the process forms bubbles, to which particulate matter attaches, forming a foam. The foam can plug gas piping, interfering with proper operation of the bioreactor, and can escape from the bioreactor, resulting in unsightly and unsafe operating conditions. Moreover, the removal of active biomass by the foam reduces the SRT, thereby decreasing the treatment capacity. A downward spiral of performance can result as the decreased treatment capacity causes more surfactant production, which produces more foaming, thereby removing more biomass, causing more surfactant production, etc. The primary corrective measure in such instances is to reduce the organic loading to a value that allows complete treatment to occur so that intermediates with surfactant properties are no longer produced.

The second type of foaming occurs in anaerobic digesters that are stabilizing solids from activated sludge systems containing significant quantities of the nuisance microorganism *Nocardia* (see Section 10.4.2). *Nocardia* is a branched, filamentous microorganism that causes foaming in activated sludge systems. When the waste solids from such a system are added to an anaerobic digester, the *Nocardia* retains its physical integrity and characteristics, thereby causing foaming in the digester.[28,73] The effects are similar to those described in the preceding paragraph. In severe cases, significant disruption of the digestion process can occur. Correction of anaerobic digester foaming caused by the presence of *Nocardia* requires its elimination from the feed solids. This requires correction of the conditions causing the *Nocardia* to grow in the upstream activated sludge system.

The other major operational problem in anaerobic processes is the formation of precipitates resulting from the liberation of inorganic constituents during stabilization of complex organic matter like wastewater solids. This follows from the high feed concentrations to many anaerobic processes, which result in high concentrations of inorganic constituents. Precipitation of metal sulfides and their impact on heavy metal solubility were discussed in Section 13.2.6. Other precipitates can form in

large quantities, forming scale on surfaces and clogging pipes. Two precipitates of particular concern are struvite ($MgNH_4PO_4$) and calcium carbonate ($CaCO_3$).

Struvite precipitation occurs most frequently when waste biomass and vegetable matter are digested because their stabilization releases inorganic cell constituents, including magnesium, ammonia, and phosphate.[71,77] Struvite is moderately soluble (pK_{sp} of 12.6 at 25°C), but the high biomass concentrations in the feed to many anaerobic digesters result in a supersaturated solution with respect to struvite precipitation.[44] The kinetics of struvite precipitation are slow and precipitation does not occur immediately or uniformly. Instead, it begins at some location within the digester system, and further precipitation occurs rapidly at that location, resulting in the formation of a scale. Struvite precipitates frequently occur at points of turbulence, such as in overflow structures, in piping immediately adjacent to pumps, and in heat exchangers or heat exchanger piping. These points of turbulence strip carbon dioxide, resulting in a localized increase in pH. Struvite solubility decreases with increasing pH, thereby increasing the degree of supersaturation and the propensity for a precipitate to form.

Work is ongoing to characterize the precise conditions under which struvite precipitation occurs.[44] Research is also ongoing to determine how the recycle of nitrogen and phosphorus to anaerobic digesters from solids handling systems can be minimized.[61,64] Design features to minimize the impacts of struvite precipitation include the use of polyvinyl chloride (PVC) or glass-lined pipes to minimize the adherence of precipitates, the design of piping systems with long radius elbows and other features to minimize turbulence, and the incorporation of features to allow easy cleaning of piping.[77]

Calcium carbonate precipitates can form when treating high-strength wastewaters that also contain high concentrations of calcium (for example, dairy wastes).[43] Carbonate formed as a result of stabilization of the organic matter reacts with the calcium to form the precipitate. The precipitate may form on surfaces, but it may also form within biomass flocs or biofilms. For example, calcium carbonate precipitates have been observed in the granules in UASB systems. These precipitates may or may not cause operating problems. However, they introduce a nonreactive solid phase that reduces treatment capacity by reducing the unit activity of the biomass in the process. These effects must be accounted for in the process design.

13.5 KEY POINTS

1. Anaerobic processes stabilize biodegradable organic matter by converting it to methane gas. At standard temperature and pressure, 0.35 m^3 of methane are produced per kg of chemical oxygen demand (COD) removed. For primary solids this is equivalent to 0.7 m^3 of methane per kg of volatile solids (VS) removed.
2. Anaerobic processes consist of four major components: (1) a closed bioreactor, (2) a mixing system, (3) a heating system, and (4) a gas–liquid–solids separation system.
3. Anaerobic processes can be grouped into four principal types: (1) anaerobic digesters, (2) low-rate anaerobic processes, (3) high-rate anaerobic processes, and (4) solids fermentation processes.

4. Anaerobic digesters are used to stabilize biodegradable organic matter and inactivate pathogens in slurries that contain high concentrations of particulate matter. They are completely mixed bioreactors with no cell recycle and can be characterized as single continuous stirred tank reactors (CSTRs).
5. Low-rate anaerobic processes are generally mixed only by the gas produced in the bioreactor. Consequently, solids settle and accumulate within the bioreactor, resulting in some difference between the solids retention time (SRT) and the hydraulic retention time (HRT). Covers can form as a result of the accumulation of scum and other floating materials, or membrane covers can be provided to capture the methane produced.
6. High-rate anaerobic processes incorporate a variety of biomass retention mechanisms, including: the formation of readily settleable particles which are retained by sedimentation, the use of bioreactor configurations that retain suspended solids, and the growth of biofilms on surfaces. High-rate anaerobic processes are generally able to remove 80 to 90% of the 5-day biochemical oxygen demand (BOD_5) applied (COD removed is approximately 1.5 times the mass of BOD_5 removed), to produce 0.35 m^3 of methane per kg COD removed, and to produce 0.05 to 0.10 kg of biomass (as volatile suspended solid [VSS]) per kg COD removed.
7. The various high-rate anaerobic processes differ in their ability to successfully treat wastewaters with high concentrations of particulate matter and in the volumetric organic loadings (VOLs) applied. Anaerobic contact (AC) processes can effectively treat wastewaters with high concentrations of particulate matter. Anaerobic filter (AF) systems can treat wastewaters with moderate levels of suspended solids, while suspended solids will generally pass through downflow stationary fixed film (DSFF) bioreactors. Upflow anaerobic sludge blanket (UASB), hybrid UASB/AF, and fluidized bed/expanded bed (FB/EB) bioreactors do not generally respond well to wastewaters containing high concentrations of particulate matter. High VOLs can be applied to UASB, hybrid UASB/AF, and FB/EB bioreactors if the wastewater contains primarily soluble organic matter (SOM).
8. Solids fermentation systems differ from other anaerobic processes since their objective is the production of volatile fatty acids (VFAs) and their separation from the waste stream for feeding to biological nutrient removal processes.
9. Anaerobic processes are generally competitive with aerobic processes for the treatment of wastewaters with biodegradable COD concentrations greater than 1,000 mg/L, and are usually the process of choice when the biodegradable COD concentration exceeds 4,000 mg/L. Factors affecting the choice include waste strength, flow rate, and temperature. Combustion of the methane produced and recovery of heat from the effluent can be used to achieve the temperatures required for effective anaerobic treatment.
10. The SRT is the primary factor determining the performance of anaerobic processes. An SRT of 15 to 20 days is generally required to achieve stable, reliable performance at 35°C. Many high-rate processes have SRTs in

excess of 30 to 50 days, and sometimes even 100 days. These very high SRTs may partially account for their stability.

11. It is not possible to precisely determine the SRT for some bioreactors. In such cases, the VOL is used to characterize process performance. The VOL is related to the SRT through the process yield and the biomass concentration.
12. The performance of some high-rate anaerobic processes, such as AF, UASB, hybrid UASB/AF, DSFF, and FB/EB, is affected by the total hydraulic loading (THL). The THL is the total bioreactor influent flow rate, including recirculation, divided by the cross-sectional area perpendicular to the flow. Total hydraulic loading criteria include maximum values to prevent biomass washout and minimum criteria to fluidize/expand media or to ensure good flow distribution.
13. Optimum performance of anaerobic processes is generally achieved by operation at a temperature near the optimum for mesophilic (30°C to 40°C) or thermophilic (50°C to 60°C) microorganisms. Acceptable performance can be achieved at temperatures below these values if an increased SRT is provided and if sufficient time is allowed for acclimation. Short term temperature fluctuations must be avoided, with a typical goal of no more than 1°C/day.
14. The optimum pH range for growth of aceticlastic methanogens is 6.8 to 7.4, with their activity decreasing significantly at pH values below this range. In contrast, the growth of acidogenic bacteria is much less sensitive to low pH. This difference in pH sensitivity can result in a downward spiral in process performance in which retardation of the aceticlastic methanogens causes reduced consumption of VFAs relative to their formation, which results in further reductions in pH, etc. This condition can be resolved by pH adjustment and reduction of the loading.
15. The pH in anaerobic processes is determined by the bicarbonate buffering system. Excess carbon dioxide is generally produced, and the quantity of bicarbonate is determined by the concentration of strong base available to react with it. Ammonia-N present in the wastewater or produced through biodegradation of nitrogen-containing organic matter is the base carbon dioxide reacts with most often.
16. Sodium bicarbonate is the most desirable chemical for pH adjustment. Other options result in pH variations as the added chemical reacts with carbon dioxide, removing it from the gas space. When the carbon dioxide balance is restored through continued metabolic activity, a second pH shift occurs. Addition of calcium based chemicals can also cause the precipitation of calcium carbonate.
17. The light metal cations sodium, potassium, calcium, and magnesium are required nutrients in anaerobic processes and their presence at low concentrations causes a stimulatory effect on microbial growth. However, elevated concentrations can cause moderate inhibition, and even higher concentrations can cause severe inhibition or toxicity. Interactions between the light metal cations can cause either increased or decreased inhibition.

18. Total ammonia (the sum of the free plus ionized ammonia species) concentrations of 50 to 200 mg/L as N stimulate microbial growth in anaerobic processes. However, free ammonia (NH_3) can be inhibitory if it reaches concentrations of about 100 mg/L as N. The fraction of total ammonia present as free ammonia increases with increasing temperature and pH.
19. Three strategies are available for reducing ammonia toxicity, including: (1) reducing the temperature, (2) reducing the pH, and (3) reducing the total ammonia concentration. The pH can be reduced by the addition of hydrochloric acid.
20. Dissolved sulfide is toxic to anaerobic processes at a concentration of about 100 mg/L (200 mg/L with acclimation). Sulfide is formed by the destruction of sulfur-containing organic matter and by the reduction of sulfate. The possibility of sulfide inhibition must be considered for wastewaters with COD/SO_4 ratios less than about 7.5. Sulfide reacts with heavy metals, forming insoluble precipitates that are not inhibitory. The reduction of sulfate requires electrons from biodegradable organic matter, thereby decreasing the number available for methane production. Sulfide production also decreases the degree of waste stabilization because soluble sulfide exerts an oxygen demand.
21. Dissolved heavy metals can be quite toxic to anaerobic processes. However, the presence of dissolved sulfides minimizes their effect since the sulfide precipitates of heavy metals are quite insoluble.
22. Evidence concerning inhibition by volatile acids is mixed. Andrews and coworkers have suggested that it is the nonionized form of the VFAs that is actually inhibitory, with concentrations on the order of 30 to 60 mg/L having an effect.[2,3,4] At neutral pH relatively high total VFA concentrations are required to cause nonionized concentrations in that range.
23. A wide variety of organic compounds can inhibit anaerobic processes. However, biomass can become acclimated to many of these compounds and cultures can acquire the ability to biodegrade many of them.
24. Because the net process yield is low in anaerobic systems, nutrient limitations are seldom encountered when treating complex wastewaters, but they may occur when treating certain high-strength industrial wastewaters. Nutrients of concern include the macronutrients nitrogen and phosphorus, and the micronutrients iron, nickel, cobalt, sulfur, and calcium.
25. Several approaches are used to mix anaerobic processes, including effluent recirculation, gas recirculation, and mechanical mixing. Mixing in anaerobic digesters is particularly challenging because of the thixotropic nature of the solids processed.
26. The nature of the organic matter fed to an anaerobic process can dramatically affect its performance. For example, the biodegradable portion of the particulate organic matter in primary solids is typically about 70%, while the biodegradable portion of the particulate organic matter in waste activated sludge typically ranges from 30 to 50%, depending on the SRT of the activated sludge system from which it came.
27. Anaerobic digesters are typically designed with SRTs on the order of 15 to 20 days at 35°C to achieve good stabilization of biodegradable organic

matter and control of pathogens. Design procedures must also consider variations in influent flow rates and requirements to periodically remove units from service for maintenance.

28. Low- and high-rate anaerobic processes are often designed using VOLs and other parameters based on pilot-scale and full-scale experience. Pilot tests may be needed to design a specific installation, or previous experience with the subject wastewater in anaerobic processes may provide sufficient information for design. The procedures used to design many anaerobic process options are analogous to those used to design other processes considered in this book.
29. Several solids fermentation process configurations are available. In general, fermentation processes are operated at SRTs that are short enough to preclude the growth of aceticlastic methanogens.
30. Control procedures for anaerobic processes generally require monitoring the bioreactor pH, the volatile acids to alkalinity ratio, and the methane production rate.
31. Foaming removes active biomass from the liquid phase in the bioreactor, thereby interfering with anaerobic treatment. It can be caused by incomplete metabolism of influent organic matter or by the presence of *Nocardia*. Foaming caused by incomplete metabolism can be reduced by reducing the process loading. Foaming caused by *Nocardia* requires its elimination from the feed by appropriate control of the activated sludge system producing it.
32. Precipitates can form in anaerobic processes and cause scaling of surfaces and plugging of piping. One frequently encountered precipitate is struvite ($MgNH_4PO_4$). It is encountered when complex wastes are degraded, resulting in releases of high concentrations of magnesium, ammonia, and phosphate. Calcium carbonate can form when wastes that are high in calcium (such as dairy wastes) are treated.

13.6 STUDY QUESTIONS

1. Prepare a table summarizing the advantages and disadvantages of anaerobic digestion compared to aerobic digestion for the stabilization of waste solids.
2. Prepare a table summarizing the advantages and disadvantages of low-rate and high-rate anaerobic wastewater treatment processes relative to aerobic processes. When is each typically used? When might either be used?
3. Discuss the roles of H_2 utilizing and aceticlastic methanogens in anaerobic processes.
4. Prepare a table summarizing the typical design criteria for the various low-rate and high-rate anaerobic processes. Contrast those processes in terms of the fate of soluble and particulate organic matter within them.
5. List the principal biomass retention mechanisms used in each of the low-rate and high-rate anaerobic processes.

6. Prepare a table summarizing the advantages and disadvantages of mesophilic versus thermophilic anaerobic processes.
7. Discuss the impact of SRT on the reactions occurring in anaerobic processes. What SRT should be selected for various applications, and why?
8. A wastewater with a biodegradable COD concentration of 20 g/L is being treated in an anaerobic process operated at an HRT of 5 days. What is the VOL? If the bioreactor biomass concentration is 10 g VSS/L, what is the SRT? What net yield value did you use in the calculation of the SRT, and why?
9. An anaerobic digester is treating waste solids with a volatile solids concentration of 40 g/L. If the VOL is 3 kg VS/(m^3·day), what is the HRT?
10. Municipal primary solids with characteristics similar to those used to develop Figure 13.19 are to be treated in an anaerobic digester. A minimum of 70% of the biodegradable organic matter is to be converted to methane. What SRT is required to achieve this objective at temperatures of 35°C, 25°C, and 20°C? Should somewhat larger SRTs be used in some cases to increase process stability? If so, when and why?
11. An anaerobic digester treating municipal primary solids with a volatile solids concentration of 60 g/L has an HRT of 25 days. If it is operating at 35°C, what volatile solids destruction efficiency would be expected? Explain how you arrived at your answer.
12. Prepare a diagram demonstrating the downward spiral that occurs as a "stuck" or "sour" anaerobic process develops.
13. An anaerobic process is operating with a bicarbonate alkalinity concentration of 750 mg/L as $CaCO_3$ and a gas carbon dioxide content of 40%. What is the bioreactor pH?
14. For the anaerobic process described in Study Question 13, how much sodium bicarbonate, lime, sodium carbonate, or ammonia must be added to the bioreactor to adjust the pH to 7.0?
15. For the anaerobic process described in Study Question 13, to what value must the gas carbon dioxide content be adjusted to produce a bioreactor pH of 7.0?
16. Describe what is meant by "stimulatory" and "inhibitory" concentrations of a chemical. Describe what is meant by "synergistic" and "antagonistic" interactions of inhibitors.
17. Discuss the relationship between temperature and pH as it affects ammonia toxicity in anaerobic processes.
18. A wastewater with a flow rate of 1,000 m^3/day and a biodegradable COD concentration of 25 g/L is to be treated in an anaerobic process. Assuming typical performance for a high-rate anaerobic process, what will the methane production rate be?
19. A second waste stream with a flow rate of 300 m^3/day, a sulfate concentration of 5 g SO_4/L, and a biodegradable COD concentration of 1,000 mg/L is to be added to the anaerobic process described in Study Question 18. If all of the sulfate is reduced to sulfide and the sulfide is precipitated with iron, how will the addition of this waste stream affect the methane production rate from the anaerobic process?

20. What can be done to reduce the dissolved sulfide concentration in an anaerobic process?
21. Heavy metal toxicity is occurring in an anaerobic process. What chemicals can be added to eliminate this toxicity? What are the relative advantages and disadvantages of these chemicals?
22. A wastewater with a biodegradable COD concentration of 20 g/L is to be treated in an anaerobic process. What concentrations of ammonia-N and phosphorus are required to achieve efficient treatment?
23. The primary solids and waste activated sludge from a wastewater treatment plant are to be stabilized by anaerobic digestion. For the primary solids, 75% of the total solids are volatile and 70% of the volatile solids are biodegradable. For the waste activated sludge, 80% of the total solids are volatile and 40% of the volatile solids are biodegradable. The solids masses and thickened solids concentrations are given in Table SQ13.1. For this plant, do the following:
    a. Select an appropriate SRT to stabilize the biodegradable organic matter and inactivate pathogens, and calculate the total digester effective volume required under average, maximum month, and maximum week conditions. Assume that the operating temperature is 35°C.
    b. Evaluate options that provide two, three, or four digesters and determine the option that requires the minimum total bioreactor volume. Assume that digester cleaning occurs only under average loading conditions.
    c. Calculate the methane production rate under all loading conditions.
24. Reconsider Study Question 23. The SRT in the activated sludge system is to be reduced, resulting in a 25% increase in the mass of biodegradable volatile solids in the waste activated sludge (WAS) stream. How much more methane will be produced when the solids are anaerobically digested?
25. Discuss when treatability tests and a pilot study should be conduted prior to the design of an anaerobic process to treat an industrial waste. When would it not be necessary to conduct treatability tests or a pilot study?
26. Consider a wastewater with a flow of 125,000 m^3/day and a TSS concentration of 200 mg/L (75% volatile). Primary treatment of this wastewater results in removal of 60% of the TSS. Solids are removed from the primary clarifier at a concentration of 10 g/L. A completely mixed/thickener fermenter is to be designed to produce VFAs to add to a BNR

Table SQ13.1 Data for Study Question 23

	Average		Maximum month		Maximum week	
Type of solids	Mass kg/day	Conc. g/L	Mass kg/day	Conc. g/L	Mass kg/day	Conc. g/L
Primary	25,000	50	27,500	50	30,000	40
WAS	20,000	45	22,500	40	25,000	35

system. The fermented primary solids can be gravity thickened to 25 g/L. You wish to recover 85% of the VFAs produced. Do the following:

a. Size the completely mixed fermenter.
b. Determine the flow rate of any elutriation streams required.
c. Determine the mass of VFAs formed in the process and the mass elutriated for addition to the BNR system.

REFERENCES

1. Anderson, G. K. and G. Yang, Determination of bicarbonate and total volatile acid concentration in anaerobic digesters using a simple titration. *Water Environment Research* **64**:53–59, 1992.
2. Andrews, J. F., Dynamic model of the anaerobic digestion process. *Journal of the Sanitary Engineering Division, ASCE* **95**:95–116, 1969.
3. Andrews, J. F. and S. P. Graef, Dynamic modeling and simulation of the anaerobic digestion process. In: *Anaerobic Biological Treatment Processes*, American Chemical Society Advances in Chemistry Series **105**:126–163, 1971.
4. Andrews, J. F. and E. A. Pearson, Kinetics and characteristics of volatile acid production in anaerobic fermentation processes. *International Journal of Air and Water Pollution* **9**:439–461, 1965.
5. Azhar, N. G. and D. C. Stuckey, The influence of chemical structure on the anaerobic catabolism of refractory compounds: a case study of instant coffee waste. *Water Science and Technology* **30**(12):223–232, 1994.
6. Blum, D. J. W., R. Hergenroeder, G. F. Parkin, and R. E. Speece, Anaerobic treatment of coal conversion wastewater constituents: biodegradability and toxicity. *Journal, Water Pollution Control Federation* **58**:122–131, 1986.
7. Britz, T. J. and F. G. Pohland, eds., *Anaerobic Digestion VII, Water Science and Technology*, **30**(12), 1994.
8. Buhr, H. O. and J. R. Andrews, Review paper: the thermophilic anaerobic digestion process. *Water Research* **11**:129–143, 1977.
9. Bundgaard, E., P. P. Brinch, M. Henze, and K. Andersen, Process optimization by fermenter technology. *Proceedings of the Water Environment Federation 65th Annual Conference & Exposition, Volume III*, 343–354, 1992.
10. Capri, M. G. and G. v. R. Marais, pH adjustment in anaerobic digestion. *Water Research* **9**:307–313, 1975.
11. Choi, E., and J. M. Rim, Competition and inhibition of sulfate reduces and methane producers in anaerobic treatment. *Water Science and Technology* **23**(7/9):1259–1264, 1991.
12. Chynoweth, D. P., S. A. Svoronos, G. Lyberatos, J. L. Harman, P. Pullammanappallil, J. M. Owens, and M. J. Peck, Real-time expert system control of anaerobic digestion. *Water Science and Technology* **30**(12):21–29, 1994.
13. Colleran, E., S. Finnegan, and R. B. O'Keeffe, Anaerobic digestion of high-sulphate-content wastewater from the industrial production of citric acid. *Water Science and Technology* **30**(12):263–273, 1994.
14. Dague, R. R., Application of digestion theory to digester control. *Journal, Water Pollution Control Federation* **40**:2021–2032, 1968.
15. Daigger, G. T. and J. A. Buttz, *Upgrading Wastewater Treatment Plants*, Technomic Publishing, Lancaster, Pennsylvania, 1992.
16. Defour, D., D. Derycke, J. Liessens, and P. Pipyn, Field experience with different systems for biomass accumulation in anaerobic reactor technology. *Water Science and Technology* **30**(12):181–191, 1994.

17. De Souza, M. E. and F. G. Pohland, eds., *Anaerobic Treatment in Tropical Countries*, *Water Science and Technology* **18**(12):1986.
18. Eastman, J. A. and J. F. Ferguson, Solubilization of particulate organic carbon during the acid phase of anaerobic digestion. *Journal, Water Pollution Control Federation*, **53**: 352–366, 1981.
19. Elefsionitis, P. and W. K. Oldham, Anaerobic acidogenesis of primary sludge: the role of solids retention time. *Biotechnology and Bioengineering* **44**:7–13, 1994.
20. Fakhru'l-Razi, A., Ultrafiltration membrane separation for anaerobic wastewater treatment. *Water Science and Technology*, **30**(12):321–327, 1994.
21. Gossett, J. M. and R. L. Belser, Anaerobic digestion of waste activated sludge. *Journal of the Environmental Engineering Division, ASCE* **108**:1101–1121, 1982.
22. Graef, S. P. and J. F. Andrews, Stability and control of anaerobic digestion. *Journal, Water Pollution Control Federation* **46**:666–683, 1974.
23. Hall, E. R., Anaerobic treatment of wastewaters in suspended growth and fixed film processes. In: *Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes*, J. F. Malina, Jr. and F. G. Pohland, eds. Technomics Publishing, Lancaster, Pennsylvania, pp. 41–118, 1992.
24. Hall, E. R., ed., *Anaerobic Digestion VI, Water Science and Technology* **25**(7):1992.
25. Harada, H., K. Momonoi, S. Yamazaki, and S. Takizawa, Application of anaerobic-UF membrane reactor for treatment of a wastewater containing high strength particulate organics. *Water Science and Technology* **30**(12):307–319, 1994.
26. Harper, S. R. and Suidan, M. T., Anaerobic treatment kinetics—Discussers report. *Water Science and Technology* **24**(8):61–78, 1991.
27. Hawkes, F. R., A. J. Guwy, D. L. Hawkes, and A. G Rozzi, On-line monitoring of anaerobic digestion: application of a device for continuous measurement of bicarbonate alkalinity. *Water Science and Technology* **30**(12):1–10, 1994.
28. Hernandez, M. and D. Jenkins, The fate of *Nocardia* in anaerobic digestion. *Water Environment Research* **66**:828–835, 1994.
29. Hobson, P. N. and B. G. Shaw, Inhibition of methane production by *Methanobacterium formicicum*. *Water Research* **10**:849–852, 1976.
30. International Association on Water Pollution Research and Control, *International Specialized Workshop on Anaerobic Treatment Technology for Municipal and Industrial Wastewaters*, *Water Science and Technology* **24**(8):1991.
31. Iza, J., Fluidized-bed reactors for anaerobic waste-water treatment. *Water Science and Technology* **24**(8):109–132, 1991.
32. Iza, J., E. Colleran, J. M. Paris, W. M. Wu, International workshop on anaerobic treatment technology for municipal and industrial wastewaters—Summary paper. *Water Science and Technology* **24**(8):1–16, 1991.
33. Karlsson, I. and G. Smith, Pre-precipitation facilitates nitrogen removal without tank expansion. *Water Science and Technology* **23**(4/6):811–817, 1991.
34. Kennedy, K. J. and R. L. Droste, Anaerobic waste-water treatment in downflow stationary fixed film reactors. *Water Science and Technology* **24**(8):157–178, 1991.
35. Kim, I. S., J. C. Young, and H. H. Tabak, Kinetics of acetogenesis and methanogenesis in anaerobic reactions under toxic conditions. *Water Environment Research* **66**:119–132, 1994.
36. Kugelman, I. J. and K. K. Chin, Toxicity, synergism, and antagonism in anaerobic waste treatment processes. In *Anaerobic Biological Treatment Processes*, American Chemical Society Advances in Chemistry Series **105**:55–90, 1971.
37. Lawrence, A. W., Application of process kinetics to design of anaerobic processes. In *Anaerobic Biological Treatment Processes*, American Chemical Society Advances in Chemistry Series **105**:163–189, 1971.

38. Lawrence A. W. and P. L. McCarty, The role of sulfide in preventing heavy metal toxicity in anaerobic treatment. *Journal, Water Pollution Control Federation* **37**:392–409, 1965.
39. Lawrence, A. W., P. L. McCarty, and F. J. A. Guerin, The effect of sulfides on anaerobic treatment. *Proceedings of the 19th Industrial Waste Conference*, Purdue University Engineering Extension Series, No. 117, pp. 343–357, 1964.
40. Lema J. M., R. Mendez, J. Iza, P. Garcia, and F. Fernandezpolanco, Chemical reactor engineering concepts in design and operation of anaerobic treatment processes. *Water Science and Technology* **24**(8):61–78, 1991.
41. Lettinga, G, Treatment of raw sewage under tropical conditions. In *Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes*, J. F. Malina, Jr. and F. G. Pohland, eds. Technomics Publishing, Lancaster, Pennsylvania, pp. 147–166, 1992.
42. Lettinga, G. and L. W. Hulshoff, UASB process design for various types of wastewaters. *Water Science and Technology* **24**(8):87–108, 1991.
43. Lettinga, G. and L. W. Hulshoff, UASB process design for various types of wastewaters. In *Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes*, J. F. Malina, Jr. and F. G. Pohland, eds., Technomics Publishing, Lancaster, Pennsylvania, pp. 119–145, 1992.
44. Loewenthal, R. E., U. R. C. Kornmuller, and E. P. van Heerden, Modeling struvite precipitation in anaerobic treatment systems. *Water Science and Technology* **30**(12): 107–116, 1994.
45. Lue-Hing, C., D. R. Zenz, and R, Kuchenrither, eds., *Municipal Sewage Sludge Management: Processing, Utilization and Disposal* Technomics Publishing, Lancaster, Pennsylvania, 1992.
46. Malina, J. F., Jr., Anaerobic sludge digestion. In *Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes*, J. F. Malina, Jr. and F. G. Pohland, eds. Technomic Publishing, Lancaster, Pennsylvania, pp. 167–212, 1992.
47. Malina, J. F., Jr. and F. G. Pohland, eds., *Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes*, Technomic Publishing, Lancaster, Pennsylvania, 1992.
48. McCarty, P. L., Anaerobic waste treatment fundamentals. *Public Works* **95**(9):107–112; (10):123–126; (11):91–94; (12):95–99, 1964.
49. McCarty, P. L., Energetics and kinetics of anaerobic treatment. In *Anaerobic Biological Treatment Processes*, American Chemical Society Advances in Chemistry Series **105**: 91–107, 1971.
50. McCarty, P. L., One-hundred years of anaerobic treatment. In *Anaerobic Digestion, 1981*, D. E. Hughes and D. A. Stafford, eds., Elsevier Biomedical Press, New York, 1982.
51. McCarty, P. L. and F. E. Mosey, Modeling of anaerobic digestion processes (A discussion of concepts). *Water Science and Technology* **24**(8):17–34, 1991.
52. Metcalf & Eddy, Inc., *Wastewater Engineering: Treatment Disposal Reuse*, Third Edition, McGraw-Hill, New York, 1991.
53. Middleton, A. C. and A. W. Lawrence, Kinetics of microbial sulfate reduction. *Journal, Water Pollution Control Federation* **49**:1659–1670, 1977.
54. Moletta, R., Y. Excoffier, F. Ehlinger, J.-P. Coudert, and J.-P. Leyris, On-line automatic control system for monitoring an anaerobic fluidized-bed reactor: response to organic overload. *Water Science and Technology* **30**(12):11–20, 1994.
55. Monteith, H. D. and J. P. Stephenson, Mixing efficiencies in full-scale anaerobic digesters by tracer methods. *Journal, Water Pollution Control Federation* **53**:78–84, 1981.
56. Mosey, F. E. and D. A. Hughes, The toxicity of heavy metal ions to anaerobic digestion. *Water Pollution Control* **74**:18–39, 1975.
57. Nahle, C., The contact process for the anaerobic treatment of wastewater. *Water Science and Technology* **24**(8):170–192, 1991.

58. Parkin, G. F. and W. F. Owen, Fundamentals of anaerobic digestion of wastewater sludges. *Journal of Environmental Engineering* **112**:867–920, 1986.
59. Parkin, G. F., M. A. Sneve, and H. Loos, Anaerobic filter treatment of sulfate-containing wastewater. *Water Science and Technology* **23**(7/9):1283–1291, 1991.
60. Pavlostathis, S. G. and E. Giraldogomes, Kinetics of anaerobic treatment. *Water Science and Technology* **24**(8):35–60, 1991.
61. Pitman, A. R., S. L. Deacon, and W. V. Alexander, The thickening and treatment of sewage sludges to minimize phosphorus release. *Water Research* **25**:1285–1294, 1991.
62. Pohland, F. G., Jr., Anaerobic treatment: Fundamental concepts, applications, and new horizons. In *Design of Anaerobic Processes for the Treatment of Industrial and Municipal Wastes*, J. F. Malina, Jr. and F. G. Pohland, eds.Technomics Publishing, Lancaster, Pennsylvania, pp. 1–40, 1992.
63. Rantala, P. and A. Luonsi, eds., *Anaerobic Treatment of Forest Industry Wastewater*, *Water Science and Technology* **17**(1):1985.
64. Sasia, H., K. Kameyama, N. Sugimori, T. Itoh, H. Katsuura, M. Fujii, and T. Nakamura, An innovative option for nitrogen and phosphorus recovery in sludge treatment process. *Proceedings of the Water Environment Federation 68th Annual Conference & Exposition, Volume I, Wastewater Treatment Research and Municipal Wastewater Treatment*, pp. 745–756, 1995.
65. Siegrist, H., D. Renggli, and W. Gujer, Mathematical modeling of anaerobic mesophilic sewage sludge treatment. *Water Science and Technology* **27**(2):25–36, 1993.
66. Skalsky, D. S. and G. T. Daigger, Wastewater solids fermentation for volatile acids production and enhanced biological phosphorus removal. *Water Environment Research* **67**:230–237, 1995.
67. Song, K. and J. C. Young, Media design factors for fixed-bed anaerobic filters. *Journal, Water Pollution Control Federation* **58**:115–121, 1986.
68. Speece, R. E., *Anaerobic Biotechnology for Industrial Wastewater*, Archae Press, Nashville, Tennesee, 1996.
69. Tseng, S.-K. and C.-J. Yang, The reaction characteristics of wastewater containing nitrophenol, treated using an anaerobic biological fluidized bed. *Water Science and Technology* **30**(12):233–249, 1994.
70. U. S. Environmental Protection Agency, *Control of Pathogens and Vector Attraction in Sewage Sludge*, EPA/625/R-92/013, U. S. Environmental Protection Agency, Washington, D. C., 1992.
71. U. S. Environmental Protection Agency, *Operations Manual—Anaerobic Sludge Digestion*, EPA 430/9-76-001, U. S. Environmental Protection Agency, Washington, D.C., 1976.
72. U. S. Environmental Protection Agency, *Process Design Manual for Sludge Treatment and Disposal*, EPA 625/1-79-011, U. S. Environmental Protection Agency, Cincinnati, Ohio, 1979.
73. van Niekerk, A., J. Kawahigashi, D. Reichlin, A. Malea, and D. Jenkins, Foaming in anaerobic digesters, a survey and laboratory investigation. *Journal, Water Pollution Control Federation* **59**:249–353, 1987.
74. Vieira, S. M. M., J. L. Carvahlo, F. P. O. Barijan, and C. M. Rech, Application of the UASB technology for sewage treatment in a small community in Sumare, Sao Paulo State. *Water Science and Technology* **30**(12):203–210, 1994.
75. Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992.
76. Water Environment Federation, *Use of Fermentation to Enhance Biological Nutrient Removal*, Proceedings of the Conference Seminar, 67th Annual Water Environment Federation Conference & Exposition, Water Environment Federation, Alexandria, Virginia, 1994.

77. Water Pollution Control Federation, *Operation of Municipal Wastewater Treatment Plants*, Manual of Practice No. 11, Water Pollution Control Federation, Alexandria, Virginia, 1990.
78. Wentzel, M. C., R. E. Moosbrugger, P. A. L. N. S. Sam-Soon, G. A. Ekama, and G. v. R. Marais, Tentative guidelines for waste selection, process design, operation and control of upflow anaerobic sludge bed reactors. *Water Science and Technology* **30**(12):31–42, 1994.
79. Young, J. C., Factors affecting the design and performance of upflow anaerobic filters. *Water Science and Technology* **24**(8):133–156, 1991.
80. Young, J. C. and B. S. Sang, Design considerations for full-scale anaerobic filters. *Journal, Water Pollution Control Federation* **61**:1576–1587, 1989.
81. Young, J. C. and H. H. Tabak, Multi-level protocol for assessing the fate and effect of toxic organic chemicals in anaerobic reactions. *Water Environment Research* **65**:34–45, 1993.
82. Zoltec, J., Jr. and A. L. Gram, High-rate digester mixing study using radio-isotope tracer. *Journal, Water Pollution Control Federation* **47**:79–84, 1975.

This page intentionally left blank

14
Lagoons

The term lagoon refers to a diverse array of suspended growth biochemical operations with the common characteristic that they do not include downstream clarifiers and associated settled solids recycle. Their name comes from the technique historically used to construct them, as in-ground earthen basins that resemble shallow ponds. Lagoons are typically used to stabilize biodegradable organic matter, although nitrogen removal (by nitrification/denitrification and ammonia stripping) and phosphorus removal (by chemical precipitation) are observed in some instances. Several process options are available, depending on the type of metabolism occurring and the mechanism used to provide the terminal electron acceptor. This chapter provides an overview of lagoon options, with particular focus on aerobic lagoons.

14.1 PROCESS DESCRIPTION

Lagoons represent one of the oldest forms of biological wastewater treatment, having been used in some form for more than 3,000 years.[41] They have been used as the only means of treatment prior to discharge to surface waters and for pretreatment and/or storage prior to treatment in a conventional system or a wetland. A wide range of industrial and municipal wastewaters has been treated in lagoon systems.

Lagoons are mechanically simple, which often translates into low capital and operating costs. However, this mechanical simplicity masks a degree of physical, chemical, and biological complexity unparalleled by other biochemical operations, resulting in a poor understanding of the factors that affect process performance. As a consequence, the effluent quality from lagoons has often been relatively poor, relegating them to uses where high-quality effluent is not necessary. Algal growth is a particular problem. Algae generally settle slowly and, consequently, pass into the effluent where they increase the concentration of suspended solids and biodegradable organic matter. Recently, the rational application of fundamental principles, including understanding of the factors influencing algal growth, has led to the development of aerated lagoon systems with significantly improved effluent quality.[28, 32,35,37,45] Background will be provided in this chapter on the variety of lagoons used in practice, but emphasis will be placed on design of the newer generation of aerobic lagoons.

14.1.1 General Description

Figure 14.1 presents a schematic diagram of a lagoon. The structure is typically an earthen basin constructed with sloping sidewalls. To minimize construction costs, the

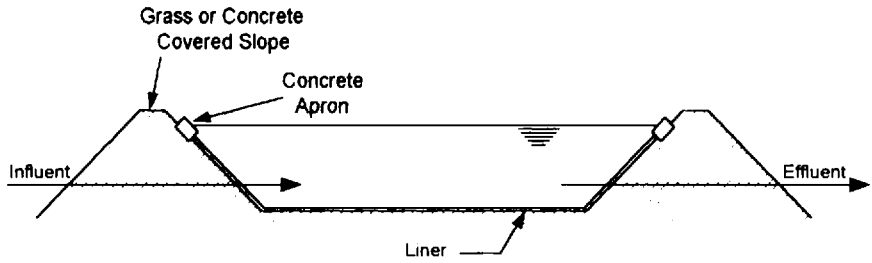


Figure 14.1 Schematic diagram of a lagoon (vertical dimension exaggerated).

lagoon is often configured so that the soil needed to construct the sidewalls is excavated from the interior, i.e., cut and fill are balanced. Natural sealing will occur to some extent as wastewater solids enter the pores of the soil and reduce the seepage rate. However, it is now common practice to provide a liner for positive seepage control. Materials used include: natural clays (such as bentonite), asphalt, synthetic membranes, and concrete. Regardless of the liner material used, a concrete apron is often provided at the water line to simplify maintenance. The remainder of the sidewalls above the water level are often covered by grass. Details on lagoon construction are provided elsewhere.[21,41] Influent and effluent structures complete the lagoon. Influent enters at one end, and the treated wastewater is collected in an effluent structure, generally located at the opposite end. No formal mechanisms are provided to retain biomass within the lagoon. Consequently, the solids retention time (SRT) approaches the hydraulic retention time (HRT), and therefore HRTs on the order of several days are typically used.

The term lagoon refers to a configuration, not to a set of controlled environmental conditions within a bioreactor. In fact, the environmental conditions depend on the process loading and operating conditions, causing significant variations in the physical, chemical, and biochemical conversions occurring, and in the treatment efficiency obtained. In short, lagoons can be used in a variety of ways, and a wide range of lagoon options exists.

Many approaches can be used to characterize lagoons.[6,21,22,41,44] We use a simplified method based on the type of metabolism occurring and the mechanism by which the terminal electron acceptor is supplied. In anaerobic lagoons, biodegradable organic matter is stabilized by its conversion to methane and carbon dioxide. In other lagoons, oxygen is provided as an electron acceptor, and biodegradable organic matter is stabilized by its conversion to carbon dioxide and water. Two principal mechanisms are used to provide oxygen: (1) the growth of algae, which produce oxygen through photosynthesis; and (2) mechanical means, such as with surface aerators, diffused aeration by blowers, and unique systems specifically designed for lagoons. These differences provide the basis for the three types of lagoons discussed in this chapter: (1) anaerobic, (2) facultative and facultative/aerated, and (3) aerobic.

14.1.2 Process Options

Anaerobic Lagoon. An anaerobic lagoon (ANL) is a low-rate anaerobic process (see Section 13.1.3) in which biodegradable organic matter is stabilized through

its conversion to carbon dioxide and methane. Compared to other lagoons, ANLs are constructed as relatively deep structures, typically ranging from 2 m to 6 m deep. This minimizes the lagoon surface area for a given volume, thereby minimizing oxygen transfer, odor release, and heat loss from the surface, which is important because ANLs are commonly not covered.[21,41,44] However, they can be covered to collect the methane gas produced and to eliminate odor release, as illustrated in Figure 13.5. Another alternative is to provide oxygen to the top layer of the lagoon so that the odoriferous compounds are oxidized to less odoriferous materials before they leave the lagoon. This aerated "cap" can be provided by directly aerating the top layer of the lagoon or by recirculating oxygenated process flow to it. Although some oxygen transfer can occur across the surface of uncovered ANLs, anaerobic conditions develop because the addition rate of biodegradable organic matter greatly exceeds the oxygen transfer rate. The loading rates also preclude the growth of algae, which would produce oxygen if they were present. Influent digestion chambers are sometimes incorporated into ANLs, particularly those treating more dilute wastewaters such as municipal wastewater.[21] Mechanical mixing is not provided, but gas evolution from the digesting organic matter provides some mixing. Nevertheless, settleable solids and floating materials accumulate in the lagoon, providing surfaces for microbial growth. The resulting retention of biomass makes the SRT greater than the HRT, but the difference is difficult to quantify. HRTs can be as high as 20 to 50 days, but many ANLs are designed and operated with HRTs less than 10 days.[21,41,44] As discussed in Chapter 13, an SRT in excess of 20 days (more likely approaching 40 or 50 days) is required to produce a stable, effective anaerobic treatment process, even when bioreactor temperature and other environmental conditions are carefully controlled. No such controls are provided in an anaerobic lagoon. Consequently, the fact that ANLs work suggests that significant retention of biomass occurs to achieve the required SRT.

Facultative and Facultative/Aerated Lagoon. As their name suggests, facultative and facultative/aerated lagoons (F/ALs) are systems in which biodegradable organic matter is stabilized by both anaerobic and aerobic processes. As illustrated in Figure 14.2, the lower portion of a facultative lagoon is anaerobic and biodegradable organic matter is stabilized there by anaerobic processes which convert it into methane and carbon dioxide. The upper portion of the lagoon is aerobic because oxygen is provided by algal growth (major contribution) and by surface reaeration across the air–liquid interface (minor contribution). Because of the aerobic environ-

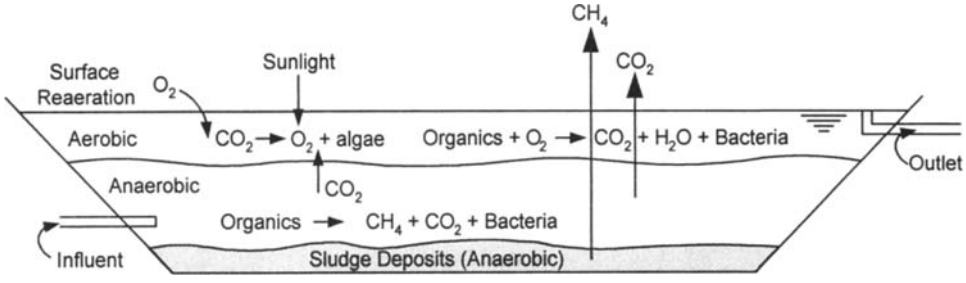


Figure 14.2 Facultative lagoon (vertical dimension exaggerated).

ment, biodegradable organic matter is stabilized by aerobic metabolism in the upper zone. The aerobic zone also provides an aerobic "cap" that oxidizes reduced compounds produced in the underlying anaerobic zone, minimizing odor release and oxidizing soluble oxygen-demanding compounds. Effluent is withdrawn from the aerobic zone.

A synergistic relationship exists between the bacteria and the algae in a facultative lagoon. Bacteria stabilize organic matter via anaerobic and aerobic metabolism, resulting in new biomass. During daylight hours, algae produce oxygen in the upper portion of the lagoon, and the bacteria there use it as their electron acceptor. Carbon dioxide produced by the bacteria serves as the carbon source for algal growth, while sunlight provides the necessary energy. However, when light is not available, algae use molecular oxygen to oxidize biodegradable organic matter and obtain energy by heterotrophic metabolism. Although the presence of algae produces most of the oxygen needed by the bacteria for aerobic metabolism, it also results in a decrease in waste stabilization because a portion of the carbon dioxide produced by the bacteria is converted back into particulate organic matter in the form of algal cells. Experience indicates that many algae do not settle well and pass into the lagoon effluent where they contribute biodegradable organic matter and total suspended solids.[28-30,32,35,45]

Algal growth is promoted by constructing facultative lagoons as shallow basins, generally 1 to 2 m deep, thereby allowing maximum exposure of the lagoon contents to sunlight; minimizing mixing, so that light can penetrate the upper layers of the lagoon; and balancing the organic loading with the production of oxygen by the algae. Because oxygen production is generally limited by the light available to the algae, and the light available is determined by the lagoon surface area, the organic loading is generally expressed as the mass of biodegradable organic matter applied per day per unit of surface area. The allowable organic loading rates generally result in HRTs of 25 days or more.

Diurnal variations in incident light cause significant changes in the environmental conditions within facultative lagoons.[21,41,44] During the day, when light is available and algae produce oxygen, the size of the aerobic zone is significant. During the night, however, when light is not available, the size of the aerobic layer is reduced, perhaps to zero. In addition, the diurnal variation in algal activity causes the carbon dioxide concentration to vary, which produces pH variations. During the day, the pH in the aerobic zone can reach values as high as 10 as carbon dioxide is consumed by the algae. During the night, on the other hand, the pH decreases to 7 or below as carbon dioxide is produced by both bacterial and algal respiration. The long HRTs in facultative lagoons, coupled with the high pH values, result in excellent pathogen destruction. In fact, in some instances, facultative lagoons have been used for the disinfection of municipal wastewater.[20,25] Sedimentation of nematode eggs is another important pathogen removal mechanism.[3]

Significant variations in facultative lagoon performance occur because of ambient conditions, which vary on both a seasonal and a geographical basis. For example, ambient temperatures vary, and this affects the temperature in the lagoon. The availability of sunlight also varies seasonally and geographically. Thus, wide ranges in environmental conditions can exist within a lagoon, resulting in a wide range in allowable loadings.[21] Consequently, care must be used in extrapolating al-

lowable loadings from one location to another. In some areas, lagoons freeze during the winter, which disrupts performance. This problem can be overcome by making the lagoon large enough to accumulate the wastewater during the portion of the year when performance is unsatisfactory, allowing its discharge only when the effluent quality is acceptable and the receiving water has sufficient assimilative capacity. Lagoons can also be designed to prevent surface water discharges; the water either seeps into the groundwater or evaporates.

Facultative lagoons can remove nitrogen and phosphorus from wastewaters. Nitrogen is removed by two mechanisms: nitrification and denitrification, and ammonia stripping. Because zones of both high and low oxygen concentration exist within a facultative lagoon, the environments required for both nitrification and denitrification are present. Ammonia stripping occurs because of the high pH in the aerobic zone of the lagoon. Elevated pH results in conversion of ammonium ion to free ammonia, as illustrated in Eq. 13.13. Free ammonia can be easily volatilized to the atmosphere. Although both mechanisms operate, their relative importance is not known.[14,24] Elevated pH can also result in the precipitation of phosphorus, thereby removing it from the liquid phase. Although these conversions occur in lagoons, they may not occur consistently. Consequently, effluent nutrient concentrations may fluctuate.

The organic loading on a facultative lagoon can be increased by providing additional oxygen by mechanical means. If only a low level of mixing energy is introduced by the oxygen transfer device, insufficient to completely mix the lagoon, the two zones will be maintained and light penetration will be sufficient for the algae to grow in the same fashion as in facultative lagoons. This provides the basis for facultative/aerated lagoons, which have operation and performance characteristics similar to facultative lagoons, but with somewhat higher allowable loadings.

Aerobic Lagoon. Aerobic lagoons (AELs) are designed and operated to exclude algae. This is accomplished by two means. First, sufficient mixing is used to keep all biomass in suspension, thereby providing turbidity that restricts penetration of light into the water column. The mixing also has the effect of making the SRT equal to the HRT. Second, the HRT is controlled to values less than the minimum SRT for algal growth (about 2 days).[28–30,32,35,45] Because algae are excluded, oxygen must be delivered by mechanical means.

Aerobic lagoon systems can be designed to meet a variety of objectives, including the removal of biodegradable organic matter through its conversion to biomass, the stabilization of organic matter (including synthesized biomass) by aerobic digestion, and the removal of synthesized biomass by gravity settling.[28–30,35,37,45] Figure 14.3 illustrates these process options. Regardless of the objective, the first step in an AEL system is a completely mixed aerated lagoon (CMAL) where sufficient mixing is provided to keep all biological solids in suspension. Just as in activated sludge systems, aerobic bacteria oxidize a portion of the biodegradable organic material into carbon dioxide and water, and convert a portion into new biomass. Consequently, the overall waste stabilization accomplished is the difference in the oxygen demands of the original wastewater and the synthesized biomass. As discussed in Chapter 5, this is equal to $1 - Y_{Hobs}$, which typically represents stabilization of about 40%. As discussed in Section 9.3.2, nearly complete conversion of biodegradable organic matter into biomass can be accomplished aerobically at SRTs on the order

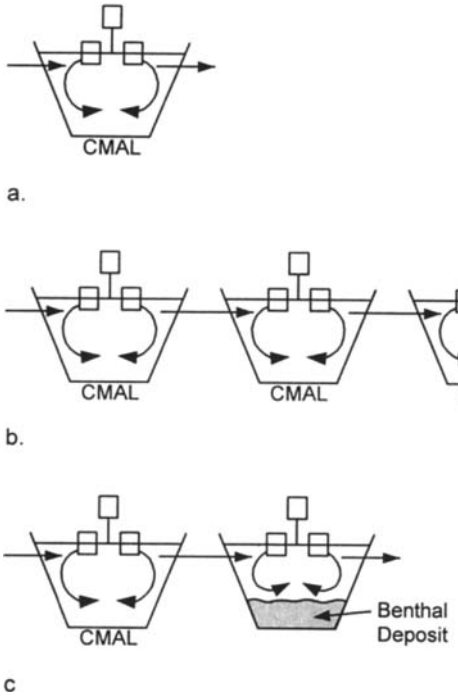


Figure 14.3 Types of aerobic lagoons (vertical dimension exaggerated): a. stabilization through conversion of biodegradable organic matter; b. organic matter conversion and aerobic digestion; and c. organic matter conversion plus benthal stabilization.

of 2 to 3 days. Experience confirms this for a CMAL, where the HRT equals the SRT.[28,29,31,35]

Further removal and stabilization of biodegradable organic matter can be accomplished in a variety of ways. One approach, illustrated in Figure 14.3b, is simply to provide a larger HRT to allow aerobic digestion of the synthesized biomass and any organic solids that entered via the influent. This can be accomplished by constructing a larger CMAL or by constructing several CMALs in series. Lagoons in series provide a slight benefit in terms of overall stabilization, as discussed in Section 12.2.5. Another approach, illustrated in Figure 14.3c, is to provide a lagoon with lower mixing energy in which the biosolids leaving the initial CMAL are removed by gravity sedimentation, stabilized by benthal processes, and stored for later disposal.[37] Benthal stabilization involves anaerobic digestion and the end products are methane, carbon dioxide, organic acids, and nutrients such as ammonia-N. If an oxygen concentration of at least 2 mg/L is maintained in the clear water zone overlying the benthal layer, the reduced products (such as organic acids) will be oxidized as they pass through the upper portion of the settled solids.[37] Steps must also be taken to minimize the growth of algae in the settling lagoon. This is generally accomplished by providing a minimal level of mechanical aeration and by limiting the HRT in the overlying clear water zone to a value less than the minimum SRT for algal growth.[30,40]

14.1.3 Comparison of Process Options

Table 14.1 summarizes the benefits and drawbacks of the various lagoon systems. Anaerobic lagoons are inexpensive, easy to operate, and effective. They can provide significant wastewater flow and load equalization for downstream treatment processes since their HRT is relatively large. Solids production is low, and methane is produced, which can be collected and used as an energy source. Effective destruction of pathogens is also obtained because of the relatively long SRTs, a benefit with many wastewater types. On the other hand, process control is poor because of the lack of mixing and biomass retention systems. Many ANL systems are not covered, except for scum and debris that accumulates at the lagoon surface, and thus they can be significant odor sources. Lagoons can be covered using membranes and other devices, but odor release can still occur from the reactor inlets and outlets. Furthermore, odors will be released when the cover is removed for periodic lagoon cleaning. Although significant removal of biodegradable organic matter can be accomplished with an anaerobic lagoon, effluent quality may be relatively poor, requiring further

Table 14.1 Lagoon Process Comparison

Process	Benefits	Drawbacks
Anaerobic lagoon (ANL)	• Simple construction • Low capital and O&M costs • Simple operation • Low solids production • Effective and efficient • Energy (methane) recovery possible • Flow and load equalization • Pathogen destruction	• Poor process control • Significant odor potential • Effluent may require further treatment • Significant land area required
Facultative and facultative/aerated lagoon (F/AL)	• Simple construction • Low capital and O&M costs • Simple operation • Flow equalization • Low solids production • Periodic solids management • Pathogen destruction	• Large land area required • Poor process control • Unreliable process performance • Odor potential
Aerobic lagoon (AEL)	• Simple construction • Low capital costs • High effluent quality • Simple operation • Periodic solids management • Low sludge production • Little odor production	• Significant land area required • Moderate O&M costs

treatment. Significant land area is also required due to the relatively low allowable process loadings.

Facultative and facultative/aerated lagoons are inexpensive to construct and operate, and are also simple to operate. Solids production is low because some of the biodegradable organic matter is stabilized anaerobically, and residual solids are stored in the lagoon. The large solids storage capacity allows solids to be managed on a periodic basis, which simplifies operations. Flow can also be equalized. Significant pathogen destruction can occur as a result of natural die-off and because the wastewater is exposed to sunlight. On the other hand, large land areas are required and process control is relatively poor. Furthermore, process performance is unreliable; while good quality effluent can be produced, these periods are interspersed between periods of excessive algal discharge with resulting poor effluent quality. Nutrient removal can be achieved in F/AL systems, but performance is not predictable and effluent quality is inconsistent. Odor potential is also significant because of the large land area and the diurnal variations in oxygen concentration caused by the diurnal availability of sunlight.

Aerobic lagoons are simple and inexpensive to construct, and are simple to operate. Operating costs are moderate, but higher than for anaerobic or facultative lagoons because of the aeration equipment. However, unlike the other lagoon processes, a high-quality effluent can be reliably produced. Periodic solids management is possible in systems that incorporate benthal stabilization basins. Solids production from such systems is also low because of anaerobic stabilization of the removed suspended solids. No solids management is required for systems that maintain solids in suspension and discharge them with the effluent, but effluent quality is much poorer. Odor production is minimal because aerobic conditions are maintained on a consistent basis. Land requirements are also significant, although much less than for a facultative lagoon.

14.1.4 Typical Applications

Lagoons are widely used for wastewater treatment, and their use is expected to continue. It has been estimated that nearly 5,000 lagoon municipal wastewater treatment plants exist in the U.S., out of a total of about 12,500 municipal wastewater treatment plants.[41] They are also widely used to treat industrial wastewater. Table 14.2 summarizes typical lagoon applications.

Because of their low cost and effectiveness, ANLs are often used to pretreat industrial and municipal wastewaters. The significant removal of biodegradable organic matter that occurs in an ANL results in a significant reduction in the size of a downstream aerobic biological treatment system. Significant reductions in energy requirements and solids production also result. Industrial wastewaters containing high concentrations of readily biodegradable organic matter are frequently pretreated in ANLs. The downstream treatment system can be located at the industry, or it may be a municipal system that the industry discharges to. Anaerobic lagoons are used less frequently to pretreat municipal wastewaters because of their lower strength, but such applications are known.[21,41,44] Anaerobic lagoons are also used frequently to pretreat and store animal wastes, such as from feedlots, prior to land application.

Facultative and facultative/aerated lagoons have been used historically to treat municipal and industrial wastewater.[41] Many municipalities served by lagoon systems

Table 14.2 Typical Lagoon Applications

Anaerobic lagoons	Facultative and facultative/aerated lagoons	Aerobic lagoons
Pretreatment of industrial wastewater prior to: downstream mechanical treatment system downstream natural treatment system discharge to municipal wastewater treatment system Pretreatment of municipal wastewater prior to: downstream F/AL system downstream natural treatment system	Pretreatment of municipal and industrial wastewater prior to: downstream mechanical treatment system downstream natural treatment system Treatment prior to surface water discharge (existing systems only)	Pretreatment of industrial wastewater prior to: discharge to municipal wastewater treatment system Pretreatment of municipal and industrial wastewater prior to: downstream natural treatment system Treatment prior to surface water discharge

are small (less than 7,500 m^3/day), although some relatively large systems exist (in excess of 75,000 m^3/day). A similar wide range in sizes exists for F/ALs serving industries. The storage capacity of F/ALs is particularly attractive for land application systems where significant wastewater storage may be required to accommodate cropping cycles and weather patterns. The treatment provided in the F/AL minimizes nuisances associated with wastewater storage, and the aerobic "cap" on the lagoon minimizes the discharge of odoriferous materials from the underlying anaerobic zone. Facultative and facultative/aerated lagoons also provide a measure of pathogen control for municipal wastewaters, a benefit for these applications. Facultative and facultative/aerated lagoons, sometimes coupled with ANLs, are frequently used for municipal and industrial wastewater treatment in developing countries because of their low cost and good pathogen control.

Given today's stringent surface water discharge standards, F/AL effluents must generally be further treated before they can be discharged to lakes and streams. U. S. water pollution control regulations recognize the generally poor effluent quality produced by F/AL systems and establish specific discharge limits for this technology.[41] However, these revised limits apply only to cases where the receiving stream is not water–quality-limited. Mechanical treatment systems such as dissolved air flotation and filters or microscreens have been applied to F/AL effluents, but with generally poor results. Intermittent sand filters and rock filters have been used successfully, along with wetland treatment systems.[21,36,41]

Aerobic lagoons were developed initially to treat wastewater from the pulp and paper industry, where they were referred to as aerated stabilization basins. Since then they have been used to treat a wide variety of industrial wastewaters.[1,8,10,16,19,22] Since the 1960s they have also been used extensively to treat municipal wastewaters.[4,7,15,21,22,41] Many of these systems are actually F/ALs because the aerator power is inadequate to fully suspend biosolids, resulting in significant algal growth and

poor effluent quality. However, improved understanding of the factors that control algal growth and affect effluent quality has made it possible to design AEL systems that produce an effluent acceptable for discharge to surface waters.[36] As a consequence, AEL systems are increasingly being used for complete treatment of municipal wastewater prior to surface water discharge, particularly those that couple a CMAL with a downstream benthal stabilization basin. They can also be used as pretreatment systems for industrial or municipal wastewaters.

14.2 FACTORS AFFECTING PERFORMANCE

The factors that affect the performance of lagoons are similar to those that affect the performance of other biochemical operations. This section summarizes the more important factors, especially those that distinguish lagoon process options.

14.2.1 Solids Retention Time/Hydraulic Retention Time

As with other biochemical operations, the primary factor determining the performance of a lagoon system is the SRT. Because a CMAL behaves like a CSTR without biomass recycle, its SRT is equal to its HRT, i.e., the reactor volume divided by the flow rate, as shown in Eq. 4.15. Consequently, CMAL performance relationships are often expressed as functions of the HRT. Experience confirms that the performance of a CMAL is as predicted by the simple model presented in Chapter 5. Typical performance is illustrated in Figure 14.4, which shows that the removal of soluble 5-day biochemical oxygen demand (BOD_5) is essentially complete for the lowest HRT tested, 5 days.[3] Total BOD_5, which includes both residual biodegradable organic matter and synthesized biomass, is removed to a much lower extent, although its removal increases with increasing HRT because of increased biomass decay and

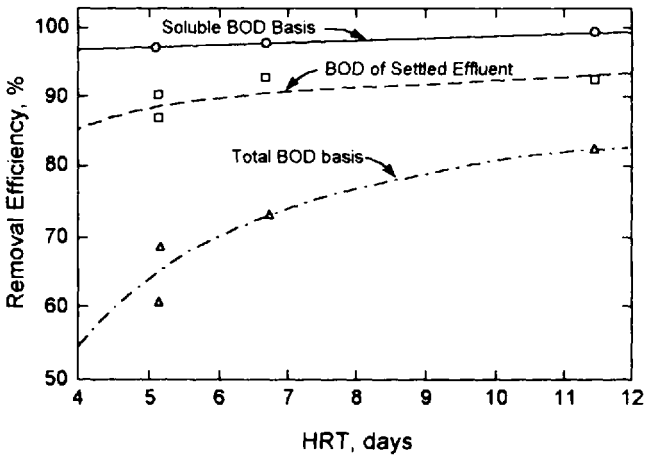


Figure 14.4 Performance of a CMAL treating municipal wastewater. (Based on data from Balasha and Sperber.[5])

destruction of biodegradable particulate organic matter. The BOD_5 of the settled effluent, which includes residual biodegradable soluble organic matter plus nonsettleable biomass and nonsettleable residual biodegradable particulate organic matter, is also reduced to a significant extent. Its removal increases only slightly with increasing HRT. Although a settling basin or pond would be required to produce the settled effluent, a significant improvement in effluent quality results from its addition.

Rich and coworkers[29, 32, 37, 45] have demonstrated that the concentration of biodegradable soluble organic matter can be accurately predicted using the simplified model presented in Chapter 5. This is demonstrated in Figure 14.5, which compares model predictions to full-scale CMAL performance data.[33] The comparison suggests that the model is conservative, i.e., the actual concentrations are less than or equal to predicted concentrations, thereby showing its utility in practice. Consistent with our understanding of other aerobic suspended growth systems, such comparisons also show that substantial removal of soluble organic matter can generally be obtained at SRTs on the order of 2 to 3 days.

Rich[35] further demonstrated that International Association on Water Quality (IAWQ) activated sludge model (ASM) No. 1 can be used to predict the fate of readily and slowly biodegradable organic matter in CMAL systems. The predictions of the fate of readily, i.e., soluble, biodegradable chemical oxygen demand (COD) are illustrated in Figure 14.6.[34] They demonstrate that its removal will be nearly complete at HRTs as short as 2 to 3 days, even at relatively low temperatures.

A further consideration for an AEL process in which settleable solids are to be removed in a settling basin is the production of a settleable biomass. Experience with activated sludge systems, as discussed in Chapters 9 and 10, suggests that this can occur at SRTs as low as about 1 day, and that effective flocculation will be accomplished at SRTs on the order to 2 to 3 days. Experience with full-scale AEL systems confirms these expectations.

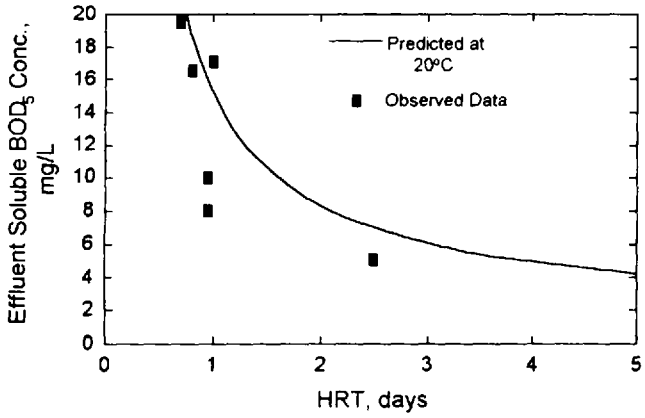


Figure 14.5 Comparison of simplified model of Chapter 5 to observed CMAL performance. The data shown were taken by L. G. Rich from H. R. Fleckseder and J. F. Malina, *Performance of the Aerated Lagoon Process*, Technical Report CRWR-71, Center for Research in Water Resources, University of Texas, Austin, TX, 1970. (From L. G. Rich, "Technical Note No. 1, Aerated Lagoons," Office of Continuing Engineering Education, Clemson University, Clemson, SC, 1993. Reprinted with permission.)

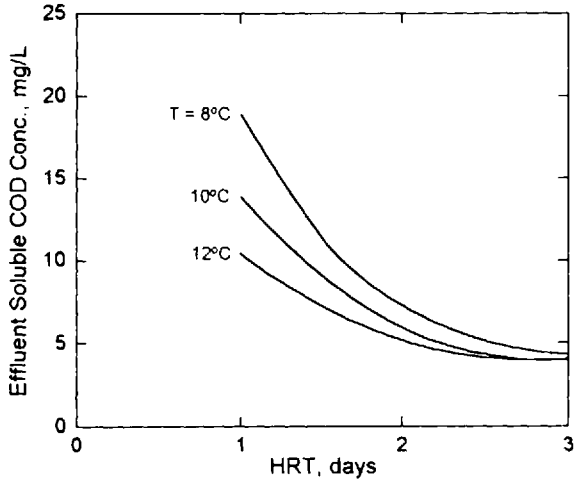


Figure 14.6 Effluent soluble COD from a CMAL as predicted by IAWQ ASM No. 1 (From L. G. Rich, "Technical Note No. 3, Aerated Lagoons," Office of Continuing Engineering Education, Clemson University, Clemson, SC, 1994. Reprinted with permission.)

The kinetics of algal growth in lagoon systems are not as well characterized as the kinetics of bacterial growth. However, data from Toms et al.[40] suggests an effective, average algal maximum specific growth rate of 0.48 day $^{-1}$, which corresponds to a minimum SRT of about 2 days. This is consistent with the observation that algal growth can typically be minimized in a single lagoon by maintaining an HRT of less than 2 days. Thus, an SRT (HRT) of 2 days is a reasonable "benchmark" of whether algae can grow.

Although the SRT is equal to the HRT for CMALs, the bioreactor configuration is not as well characterized or controlled for other types of lagoons, making calculation of the SRT less straightforward.[2,41] Anaerobic lagoons could be sized by assuming complete mixing with no biomass recycle or retention, but this would give a very conservative design because experience indicates that significant biomass retention occurs due to the settling of particulate matter, which provides surfaces for growth of anaerobic bacteria. This makes the SRT greater than the HRT. The mixing pattern in facultative and facultative/aerated lagoons is quite complex, and changes on a diurnal, daily, and seasonal basis. Consequently, it is not possible to calculate the SRT for such systems. However, since long HRTs are used, the SRTs are also quite long.

14.2.2 Volumetric Organic Loading Rate

Just as for anaerobic processes, the volumetric organic loading (VOL) is the mass of biodegradable organic matter applied to the lagoon system per unit time per unit volume. It is denoted symbolically by Γ_{VS} and is expressed in units of kg COD or $BOD_5/(m^3 \cdot day)$. It is calculated with Eq. 13.1 in terms of the flow rate and reactor volume or with Eq. 13.2 in terms of the HRT. As illustrated by Eq. 13.4, the VOL for a suspended growth process is related to the SRT by the concentration of active

biomass in the reactor and the biomass net process yield. For processes such as anaerobic and facultative lagoons, it is not possible to determine the active biomass concentration. However, the VOL can be used to characterize process performance just as it was used for the various anaerobic processes described in Chapter 13 for which biomass concentrations could not be determined. Thus, it is useful as a design parameter for these more poorly defined systems. Volumetric organic loadings for anaerobic lagoons generally range from 0.2 to 1 kg $BOD_5/(m^3 \cdot day)$,[21,41,44] although somewhat higher loadings can be used in some instances. For facultative and facultative/aerated lagoons they generally range from 0.1 to 0.3 kg $BOD_5/(m^3 \cdot day)$.[21,41,44] Since the SRT is well defined for CMAL processes, VOL need not be used to design them. However, the VOL is related to the volumetric oxygen requirement, and can be used to estimate whether adequate oxygen transfer can be achieved.

Lagoon performance will deteriorate slightly as the VOL is increased within the applicable loading range, but will deteriorate rapidly as the VOL is increased beyond the applicable range. This is consistent with the typical relationship observed between process performance and SRT. Process performance is relatively independent of the SRT for values significantly in excess of the minimum SRT for the process, but becomes quite sensitive to SRT as it approaches the minimum value. Limiting VOLs corresponds to ensuring adequate SRTs.

14.2.3 Areal Organic Loading Rate

The areal organic loading (AOL) is the mass of biodegradable organic matter applied per unit time per unit lagoon surface area, A_f. It is denoted symbolically as $\Gamma_{A,S}$, and is typically expressed in units of kg COD or $BOD_5/(ha \cdot day)$. It is calculated as:

$$\Gamma_{A,S} = \frac{F(S_{SO} + X_{SO})}{A_f} \tag{14.1}$$

The logic behind this loading parameter is made obvious when it is recognized that it is used for lagoons in which the growth of algae is necessary. It represents the balance between the loading of biodegradable organic matter and the production of oxygen by algal growth, which is dependent on the penetration of light across the lagoon surface area. Consequently, it is typically used to characterize the performance of facultative and facultative/aerated lagoons.[21,41,44]

As with the other performance parameters, facultative lagoon performance is relatively insensitive to AOL over a wide range of loadings, but deteriorates rapidly as the normal range is exceeded. Acceptable values of AOL vary widely from one geographic location to another. For example, values on the order of 50 to 70 kg $BOD_5/(ha \cdot day)$ can be used quite successfully in the Southeastern portion of the United States, while loadings of no more than 20 to 40 kg $BOD_5/(ha \cdot day)$ can be used in the more Northern regions. This range reflects variations in ambient temperature and solar radiation over the annual cycle. Local experience and practice should be consulted to select the appropriate AOL for a particular application.

The AOL can also be used to characterize loadings when particulate biodegradable organic matter is being stabilized anaerobically in benthal deposits. Again, the process oxygen demands are being balanced with the supply of oxygen to the process. Typically, the AOL should not exceed 80 g biodegradable $VSS/(m^2 \cdot day)$ [about 115 g biodegradable $COD/(m^2 \cdot day)$] for AEL processes with benthal stabi-

lization.[15,37] Mechanical aeration is used to maintain an oxygen concentration in the overlying liquid of 2 mg/L or more. At this loading and residual oxygen concentration, the reduced products (such as organic acids, hydrogen sulfide, etc.) diffusing from the anaerobic benthal deposits are oxidized by a thin aerobic layer at the top of the deposits. Higher loadings or lower oxygen concentrations result in anaerobic conditions throughout the benthal deposit and the release of odoriferous compounds to the overlying water column and, subsequently, to the atmosphere. Likewise, facultative solids storage lagoons, which provide long-term storage of anaerobically digested solids and maintain an aerobic "cap" by algal growth in the overlying clear fluid, are typically sized based on an AOL of 22.5 kg VS/(ha·year).[41]

14.2.4 Mixing

The degree of mixing provided in lagoons differs dramatically from type to type. These differences, coupled with variations in lagoon configuration, result in differences in flow patterns.[21,23]

Mechanical mixing is generally not provided in anaerobic lagoons, although some mixing is contributed by gas evolution from the digesting material. The absence of intentional mixing results in settling of solids within the lagoon and the retention of biomass. However, the absence of controlled mixing also causes complex flow patterns and the potential for short-circuiting. Inlet and outlet locations and configurations can be selected to minimize short-circuiting.[21,23]

Mixing is provided in facultative lagoons by several mechanisms, including wind action, gas evolution, and thermal gradients caused by diurnal heating and cooling of the lagoon surface. As with anaerobic lagoons, the uncontrolled nature of the mixing results in poorly defined flow patterns and the potential for short-circuiting. Facultative lagoons tend to be well mixed, and must be configured as tanks in series to achieve some degree of plug flow. As with anaerobic lagoons, short-circuiting is controlled by proper selection of the inlet and outlet location and configuration. The reader is referred elsewhere for a more complete discussion of facultative lagoon configuration.[21,28,44]

Mixing is provided in F/ALs by an oxygen transfer device. The power input to that device generally determines the degree of mixing, but different volumetric power inputs are required to achieve complete mixing of soluble versus particulate constituents. A volumetric power input of about 1 kW/1000 m^3 is adequate to achieve a uniform concentration of dissolved species such as oxygen within the zone of influence of the aerator,[30] although this power input may not produce a residence time distribution reflective of complete mixing. In contrast, a volumetric power input on the order of 6 to 10 kW/1000 m^3 may be required to achieve uniform concentrations of settleable solids, with the input depending on the concentration.[31] Another consideration is the volumetric power input required to suspend settleable solids. At power densities below 2 kW/1000 m^3 all settleable solids will be removed from suspension, leaving only nonsettleable solids.[28,29,31,32] The impacts of this are illustrated in Figure 14.7.[29] There it can be seen that algal concentrations in aerated lagoons (as indicated by the chlorophyll a concentration) decrease significantly at power densities above 2 kW/1000 m^3. This is because light penetration is reduced by the solids in suspension. A volumetric power input of at least 6 kW/1000 m^3 is

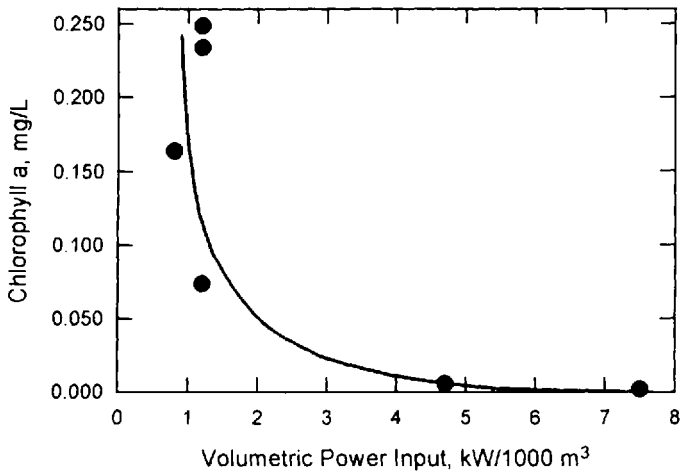


Figure 14.7 Relationship between aerator power density and chlorophyll a concentration for aerated lagoons. (From L. G. Rich, "Solids control in effluents from aerated lagoon systems," Report No. 73, Water Resources Research Institute, Clemson University, Clemson, SC, 1978. Reprinted with permission.)

required to completely suspend settleable solids when domestic wastewater is being treated.[5,9]

A more precise estimate of the volumetric power input required to suspend settleable solids is obtained from:

$$\Pi = 0.004X_{M,T} + 5 \tag{14.2}$$

where Π is the volumetric power input in kW/1000 m^3 and $X_{M,T}$ is the suspended solids concentration in mg/L as TSS.[29,31] This equation is valid for suspended solids concentrations less than about 2,000 mg/L. At a suspended solids concentration of 250 mg/L, Eq. 14.2 stipulates that a volumetric power input of 6 kW/1000 m^3 is needed.

Volumetric power inputs to facultative/aerated lagoons are generally below the level required to maintain solids in suspension. As a consequence, sunlight can penetrate the water column and algae will grow. In an aerobic lagoon, on the other hand, the volumetric power input will be sufficiently high to maintain solids in suspension, light penetration will be prevented, and algae will not grow. Thus, the volumetric power input dramatically affects the environmental conditions in the lagoon, and the resulting treatment efficiency. As discussed in Section 14.1, algae do not generally settle in lagoons and pass into the effluent, deteriorating its quality.

The presence of mechanical mixing in aerobic lagoons results in more controlled mixing conditions, and these bioreactors can be characterized as completely mixed. Aerobic lagoons are generally constructed as rectangular structures with sloping sidewalls. Length-to-width ratios between 1:1 and 3:1 are typical. The reader is referred to Murphy and Wilson[23] for further discussion of the factors affecting mixing patterns in aerobic lagoons.

The number and placement of aerators also affects the performance of facultative/aerated and aerobic lagoons. The turnover time for an aerator, which is an

indication of its mixing ability, is a function of its size. The lower the turnover time, the more closely the lagoon approaches complete mixing. The turnover time of the aerator is determined, in turn, by its pumping capacity. The pumping capacity for a surface aerator varies inversely with its size and approaches a maximum value of about 4.2×10^{-3} $m^3/(sec \cdot kW)$.[22] Thus, it is normally better to use several small units, rather than one large unit, as this will result in a higher overall pumping rate and a smaller turnover time. Surface aerators also have an effective volume and distance, which are the volume and horizontal distance they can mix effectively. The effective distance also varies with the aerator size[12] and has a reported maximum value of 0.8 m/kW. Regardless of the aerator size, the distance between adjacent aerators should not exceed about 75 m. Minimum spacings should also be maintained to avoid interactions between adjacent aerators.[20,46] Surface aerators also have effective depths. Consequently, as the depth increases, a draft tube may be required to allow the aerator to mix the entire lagoon depth. Conversely, in a shallow lagoon a concrete pad may be needed beneath the aerator to prevent erosion of soil or destruction of the lagoon liner. Surface aerator manufacturers can provide further information on the specific characteristics of their aerators.

14.2.5 Temperature

Temperature affects the biological activity in a lagoon in the same fashion as in other biochemical operations. A primary difference is that the large lagoon surface area results in higher heat losses and, typically, in lower temperatures. This is exacerbated in facultative/aerated and aerobic lagoons by the action of the oxygen transfer device if a mechanical surface aerator is used. Freezing will typically occur in cold climates, resulting in a significant decline in treatment efficiency. Significant cooling will occur even in warm climates, resulting in reduced reaction rates, which must be compensated for in the process design.

A heat balance for a lagoon requires consideration of a number of components. Potential heat inputs include the influent wastewater, solar radiation, the aeration device (if one is provided), and biological reactions. Potential heat losses include the effluent wastewater, radiation to the atmosphere, evaporation, surface convection, convection/conduction through the wall of the lagoon, and the aeration device. Nevertheless it is possible to obtain a preliminary estimate of lagoon temperature by considering just the lagoon surface area and the appropriate temperatures[11]:

$$\frac{T_L}{T_i} = \frac{62.4(F/A_L) + H'(T_A - 2)/T_i}{62.4(F/A_L) + H'} \tag{14.3}$$

This is a dimensional equation in which T_L is the weekly average lagoon temperature in °F, T_i is the weekly average influent wastewater temperature in °F, T_A is the weekly average air temperature in °F, F is the influent wastewater flow rate in ft^3/day, A_L is the surface area of the lagoon in ft^2, and H′ is the average heat transfer coefficient, which is generally taken to be 145 BTU/($ft^2 \cdot$day $\cdot$ °F). Experience indicates that predictions using this model are relatively insensitive to the specific value of the heat transfer coefficient.

Other workers have developed more sophisticated lagoon temperature models that require more input data; examples include the models of Talati and Stenstrom[39] and Argaman and Adams.[2] Table 14.3 summarizes the required inputs for the model of Talati and Strenstrom.[39] These models can provide quite precise estimates of bioreactor temperature when adequate information is available.[38]

14.2.6 Other Factors

Like other biochemical operations, the performance of lagoons is affected by other environmental conditions such as pH and nutrients. Lagoons are often used to treat industrial wastewaters, which can be nutrient deficient, and nutrients may have to be added. One advantage of many lagoon processes is that the produced solids are accumulated and anaerobically digested, which will allow the removed nutrients to be recycled. However, the rate of digestion may vary seasonally, with increased rates during warmer temperatures and reduced rates during colder temperatures, so seasonal variations in nutrient recycle may occur. This can result in seasonal variations in net nutrient requirements. Significant pH variations can be observed on a diurnal basis in lagoons where algae are active. In such systems, pH values can easily range from 7 to 10 on a diurnal basis. Such pH variations do not appear to inhibit the biological activity within a lagoon, although operation with sustained pH values greater than 8 or less than 6 should be avoided. The presence of toxic materials will adversely impact the performance of a lagoon. The large volume and HRT of a

Table 14.3 Inputs to the Heat Loss Model of Talati and Stenstrom[39]

Site-specific data	Process data	Physical properties
• Latitude of plant site, degrees • Ambient air temperature, °C • Wind speed, m/s • Relative humidity, percent • Cloud cover, tenths • Atmospheric radiation factor	• Tank dimensions (L × W × H), m • Wastewater flow rate, m^3/day • Influent temperature, °C • Airflow rate (for diffused air), m^3/s • Number of aerators • Aerator spray area, m^2 • Power input to aerator/compressor, hp • Efficiency of air compressor (for diffused aeration), percent • Substrate removal rate, kg COD removed/day • Overall heat transfer coefficient for tank walls, cal/(m^2 · day) • Humidity factor for exit air	• Air density, kg/m^3 • Water density, kg/m^3 • Specific heat of air, cal/(kg · °C) • Specific heat of water, cal/(kg · °C) • Emissivity of water • Reflectivity of water

typical lagoon provides significant equalization, which dampens adverse short-term variations in the influent wastewater characteristics.

14.3 PROCESS DESIGN

The design of anaerobic, facultative, and facultative/aerated lagoons is based on experience and various rules of thumb, and will not be discussed here. Such a design approach must be used because of the poorly defined bioreactor flow patterns and the complexity of the biochemical processes occurring. The design of anaerobic lagoons is discussed in Chapter 13 and the design of facultative and facultative/aerated lagoons is discussed in various design manuals.[21,41,44]

Aerobic lagoons use well mixed bioreactors, and the biochemical reactions occurring within them are primarily bacterial. As a consequence, the principles presented previously in this book can be used for their design. This is the focus of this section. The examples will consider a wastewater with the characteristics listed in Table E14.1. Sections 8.6 and 8.7 discuss how to translate between the various ways of expressing wastewater characteristics and parameters.

14.3.1 Completely Mixed Aerated Lagoons

A CMAL (illustrated in Figure 14.3a) can be characterized as a continuous stirred tank reactor (CSTR) with no biomass recycle. It is normally sized to produce an effluent with a specified concentration of soluble organic substrate. The design is typically based on the simplified model presented in Chapter 5, although IAWQ ASM No. 1 can also be used.[35] The process design of a CMAL requires three decisions: (1) the bioreactor volume required, (2) the amount of oxygen that must be supplied, and (3) the minimum power input required to achieve adequate mixing and oxygen transfer. The basis for these decisions is discussed in Chapter 5 and Section 10.3, and it is reviewed in this section.

For a CMAL the SRT is equal to the HRT. Equation 5.13 provides the effluent soluble substrate concentration from a CSTR as a function of the SRT. Setting the SRT equal to the HRT, inserting Eq. 4.15 for the HRT, and rearranging Eq. 5.13 gives:

$$V = \frac{F(K_S + S_S)}{S_S(\hat{\mu}_H - b_H) - K_S b_H} \tag{14.4}$$

The design volume is generally determined for wintertime operating temperatures because biological activity will be the lowest then and the largest volume will be required. Because the temperature in the lagoon depends on its size and the kinetics of biomass growth depend on the temperature, an iterative procedure is required using Eq. 14.4 to calculate the bioreactor volume with temperature corrected coefficients and Eq. 14.3 to calculate the bioreactor temperature. Example 14.3.1.1 illustrates the use of the simplified model to determine the required bioreactor volume.

Example 14.3.1.1

Consider a wastewater with the characteristics given in Table E14.1. A CMAL is to be designed to reduce the soluble organic substrate concentration to 10

Table E14.1 Wastewater Strength and Treatment Characteristics for Design Examples

Wastewater strength		Kinetic and stoichiometric coefficients	
X_{SO}	84 mg/L as COD	$\hat{\mu}_{H,20}$	10 day $^{-1}$, θ = 1.08
S_{SO}	440 mg/L as COD	K_S	120 mg/L as COD
X_{IOI}	45 mg/L as TSS	$Y_{H,I}$	0.521 mg biomass TSS/mg biode-
S_{NHO}	35 mg/L as N	$b_{H,20}$	gradable COD
X_{NSO}	7 mg/L as N	f_D	0.15 day $^{-1}$, θ = 1.04
Temperature	25°C	$i_{O\,XB,I}$	0.2 mg debris TSS/mg biomass TSS
Flow	Equalized to 2,400	$\hat{\mu}_{A,20}$	1.2 mg COD/mg TSS
	m^3/day		0.8 day $^{-1}$, θ = 1.11
Air temperature	−5°C minimum	$K_{NH,20}$	1.0 mg/L as N, θ = 1.14
	weekly average	$Y_{A,I}$	0.20 mg biomass TSS/mg N
		$b_{A,20}$	0.5 day $^{-1}$, θ = 1.04

mg/L as COD under all conditions. The lagoon is to have an average depth of 3.0 m. Determine the lagoon volume needed to meet the effluent requirement year round.

a. What bioreactor volume would be required at 20°C?
Since the kinetic parameters are given for 20°C, it is convenient to start the process by determining the volume required at that temperature. Using Eq. 14.4 and the values from Table E14.1,

$$V = \frac{(2{,}400)(120 + 10)}{10(10 - 0.15) - (120)(0.15)} = 3{,}880 \text{ m}^3$$

Since the flow rate is 2,400 m^3/day, this provides an HRT of 1.62 days. In addition, the surface area would be 1,293 m^2.

b. What would the lagoon temperature be when the air temperature is at its minimum weekly average value of −5°C?
A rough estimate of the lagoon temperature can be obtained with Eq. 14.3, which is dimensional using English units. Therefore, the inputs must be converted using the factors in Appendix C.

$$F = 2{,}400 \text{ m}^3\text{/day} = 84{,}755 \text{ ft}^3\text{/day}$$

$$A_I = 1{,}293 \text{ m}^2 = 13{,}917 \text{ ft}^2$$

$$T_I = 25°C = 77°F$$

$$T_A = -5°C = 23°F$$

Taking the heat transfer coefficient H′ to be 145 BTU/(ft^2 · day · °F), Eq. 14.3 gives:

$$\frac{T_I}{77} = \frac{62.4(84{,}755/13{,}917) + 145(23 - 2)/77}{62.4(84{,}755/13{,}917) + 145} = 0.799$$

Therefore, the minimum weekly average lagoon temperature will be:

$$T_I = 61.5°F = 16.4°C$$

This is less than 20°C, the value that was used to calculate the bioreactor volume. Consequently, an iterative procedure must be used until the assumed and calculated temperatures are equal.

c. What bioreactor volume is required for the minimum weekly average temperature?
The last iteration, with an assumed temperature of 12°C (54°F), is summarized below. The kinetic parameter values are corrected for temperature using the temperature coefficients in Table E14.1 and Eq. 3.95.

$$\hat{\mu}_{H,12} = (10)(1.08)^{12-20} = 5.40 \text{ day}^{-1}$$

$$b_{H,12} = (0.15)(1.04)^{12-20} = 0.11 \text{ day}^{-1}$$

$$V = \frac{(2{,}400)(120 + 10)}{[10(5.40 - 0.11) - (120)(0.11)]} = 7{,}860 \text{ m}^3$$

This gives an HRT of 3.28 days.

$$A_L = \frac{7{,}860}{3} = 2{,}560 \text{ m}^2 = 27{,}556 \text{ ft}^2$$

$$\frac{T_L}{77} = \frac{62.4(84{,}755/27{,}556) + 145(23 - 2)/77}{62.4(84{,}755/27{,}556) + 145} = 0.687$$

$$T_L = 53°\text{F} \ (12°\text{C})$$

Given the approximate nature of Eq. 14.3, this is close enough to the assumed value. Thus, the calculated lagoon volume can be used. For design purposes use a volume of 8,000 m^3, which corresponds to an HRT of 3.3 days. The difference between this volume and the one calculated in step a demonstrates that temperature has an important impact on the design.

The oxygen requirement for heterotrophic activity must include both the average (steady state) and transient state requirements, just as required for activated sludge design, as discussed in Sections 10.3.2 and 10.3.3. The average oxygen requirement for the heterotrophs, RO_H, can be calculated using Eq. 9.13 with the HRT substituted for the SRT. The resulting relationship is:

$$RO_H = F(S_{SO} + X_{XO} - S_S)\left[1 - \frac{(1 + f_D \cdot b_H \cdot \tau)Y_{H,T} i_{O\,XB,T}}{1 + b_H \cdot \tau}\right] \tag{14.5}$$

The transient state oxygen requirement for the heterotrophs, $RO_{H,TS}$, is calculated using Eq. 10.12.

If the HRT is sufficiently long for the operating temperature, nitrifying bacteria will be able to grow and oxygen must be provided to meet their metabolic requirements. If this is not done, adequate oxygen concentrations will not be maintained, resulting in impaired performance and potential odor problems. Just as for activated sludge design, the minimum SRT required for nitrification can be estimated from Figure 9.4 or calculated with Eq. 5.16a, which appears in Section 10.3.2. Recognizing that for a CMAL the minimum SRT is equal to the minimum HRT allows Eq. 5.16a to be modified to:

$$\tau_{min} = \frac{K_{NH} + S_{NHO}}{S_{NHO}(\hat{\mu}_A - b_A) - K_{NH} \cdot b_A} \tag{14.6}$$

If the HRT exceeds τ_{min}, then the effluent ammonia-N concentration can be calculated using Eq. 10.18, with the HRT substituted for the SRT. The effects of dissolved oxygen (DO) concentration on the effluent ammonia-N concentration are considered in Eq. 10.18 through inclusion of the DO safety factor, ζ_{DO}. Unlike activated sludge systems, however, engineering control is not often exerted over the oxygen concentration in a CMAL. Consequently, the DO safety factor is normally set equal to 1.0 when calculating average effluent ammonia-N concentrations from CMALs. Incorporating these changes into Eq. 10.18 gives:

$$S_{NH} = \frac{K_{NH}(1/\tau + b_A)}{\hat{\mu}_A - (1/\tau + b_A)} \tag{14.7}$$

The average oxygen requirement for nitrification, RO_A, can be calculated using Eq. 10.16 with the HRT inserted in place of the SRT. The result is:

$$RO_A = F(S_{N,a} - S_{NH})\left[4.57 - \frac{(1 + f_D \cdot b_A \cdot \tau)Y_{A,T} i_{O\,XB,T}}{1 + b_A \cdot \tau}\right] \tag{14.8}$$

The nitrogen available to the nitrifiers, $S_{N,a}$, is calculated as discussed in Section 10.3.3 and given by Eq. 10.17. The transient state oxygen requirement for the nitrifiers is calculated using Eq. 10.13 or 10.15, as appropriate. The total oxygen requirement for nitrification is the sum of the average and transient state oxygen requirements.

As with other aerobic processes, the oxygen requirements are calculated for the high summertime temperature, because that is when biological activity is greatest, resulting in the highest requirements. Calculation of the oxygen requirement is illustrated in the following example.

Example 14.3.1.2

The oxygen requirement for the CMAL considered in Example 14.3.1.1 is to be calculated. Even though the flow is equalized, transient loadings result in elevated pollutant concentrations in the CMAL influent. The peak influent concentrations are:

$X_{SO} = 84$ mg/L as COD

$S_{SO} = 677$ mg/L as COD

$S_{NHO} = 54$ mg/L as N

$X_{NSO} = 7$ mg/L as N

The oxygen requirement must be calculated at the high summertime temperature. To be conservative, assume that no cooling of the influent wastewater occurs, resulting in a lagoon temperature of 25°C.

a. What is the average oxygen requirement for the heterotrophs?
Equation 14.5 is used. The influent slowly and readily biodegradable organic matter concentrations are 84 and 440 mg/L as COD, respectively, for a total of 524 mg/L (g/m^3). During the wintertime, the effluent soluble substrate concentration is 10 mg/L as COD, and it will be even smaller at the higher summertime temperatures. Thus, S_S can be neglected relative to the influent substrate concentration when calculating the oxygen requirement. The value

of the heterotrophic decay coefficient, b_H, must be adjusted to 25°C by using the temperature coefficient given in Table E14.1 and Eq. 3.95:

$$b_{H,25} = (0.15)(1.04)^{25-20} = 0.18 \text{ day}^{-1}$$

The steady state heterotrophic oxygen requirement is then calculated with Eq. 14.5 using an HRT of 3.3 days.

$$RO_H = (2{,}400)(524)\left[1 - \frac{[1 + (0.20)(0.18)(3.3)](0.521)(1.20)}{1 + (0.18)(3.3)}\right]$$

$$= 706{,}000 \text{ g } O_2/\text{day} = 706 \text{ kg } O_2/\text{day}$$

b. What is the transient oxygen requirement for the heterotrophs?
The transient state oxygen requirement for the heterotrophs is calculated using Eq. 10.12. This requires knowledge of the transient state increases in the mass flow rates of readily and slowly biodegradable substrate, $\Delta(F \cdot S_{SO})$ and $\Delta(F \cdot X_{SO})$, respectively:

$$\Delta(F \cdot S_{SO}) = (2{,}400)(677) - (2{,}400)(440) = 569{,}000 \text{ g COD/day}$$

$$= 569 \text{ kg COD/day}$$

$$\Delta(F \cdot X_{SO}) = (2{,}400)(84) - (2{,}400)(84) = 0.0 \text{ kg COD/day}$$

Substitution into Eq. 10.12 gives:

$$RO_{H,TS} = [1 - (0.521)(1.2)](569) = 213 \text{ kg } O_2/\text{day}$$

c. What is the total heterotrophic oxygen requirement?
The total heterotrophic oxygen requirement is $RO_H + RO_{H,TS}$, or

$$706 + 213 = 919 \text{ kg } O_2/\text{day}$$

d. Will the process nitrify in the summer?
Determine whether autotrophs will grow in the summer using Eq. 14.6 and the average influent ammonia-N concentration. Since the summer lagoon temperature is 25°C, the kinetic parameters for the autotrophs must be corrected to that value with the temperature coefficients in Table E14.1 and Eq. 3.95:

$$\hat{\mu}_{A,25} = (0.8)(1.11)^{25-20} = 1.35 \text{ day}^{-1}$$

$$K_{NH,25} = (1.0)(1.14)^{25-20} = 1.92 \text{ mg/L as N}$$

$$b_{A,25} = (0.05)(1.04)^{25-20} = 0.06 \text{ day}^{-1}$$

Substitution of these values into Eq. 14.6 gives:

$$\tau_{min} = \frac{1.92 + 35}{35(1.35 - 0.06) - (1.92)(0.06)} = 0.82 \text{ day}$$

Since this is well below the HRT of 3.3 days, nitrification will occur.

e. What will the effluent ammonia-N concentration be?
Calculate the effluent ammonia-N concentration with Eq. 14.7:

$$S_{NH} = \frac{1.92(1/3.3 + 0.06)}{1.35 - (1/3.3 + 0.06)} = 0.7 \text{ mg/L as N}$$

f. What is the average oxygen requirement for nitrification?
The average oxygen requirement for nitrification can be calculated with Eq. 14.8. This requires knowledge of the concentration of nitrogen available to the nitrifiers, which can be calculated with Eq. 10.17. The use of Eq. 10.17 requires knowledge of the nitrogen requirement for the heterotrophs, which can be calculated with Eq. 5.36a as given in Section 10.3.3:

$$NR = 0.087\left[\frac{[1 + (0.2)(0.18)(3.3)](0.521)(1.20)}{1 + (0.18)(3.3)}\right]$$
$$= 0.038 \text{ mg N/mg COD removed.}$$

This can now be used in Eq. 10.17. Recognizing that the effluent soluble organic substrate concentration is negligible compared to the influent gives:

$$S_{N,e} = 35 + 7 - (0.038)(440 + 84) = 22 \text{ mg/L as N}$$

This, then, can be substituted into Eq. 14.8 to calculate the steady state autotrophic oxygen requirement, RO_A:

$$RO_A = (2{,}400)(22 - 0.7)\left[4.57 - \frac{[1 + (0.20)(0.06)(3.3)](0.24)}{1 + (0.06)(3.3)}\right]$$
$$= 223{,}000 \text{ g } O_2\text{/day} = 223 \text{ kg } O_2\text{/day}$$

g. What is the transient oxygen requirement for nitrification?
The transient oxygen requirement for nitrification is calculated using Eq. 10.13, which requires knowledge of the change in the nitrogen available to the nitrifiers, $\Delta(F \cdot S_N)_{A,TS}$. It must be calculated with Eq. 10.14, which requires that several changes in mass flow rates be known:

$$\Delta(F \cdot S_{NIO}) = (2{,}400)(54) - (2{,}400)(35) = 45{,}600 \text{ g N/day} = 45.6 \text{ kg N/day}$$

$$\Delta(F \cdot S_{NSO}) = (2{,}400)(0) - (2{,}400)(0) = 0.0 \text{ kg N/day}$$

$$\Delta(F \cdot X_{NSO}) = (2{,}400)(7) - (2{,}400)(7) = 0.0 \text{ kg N/day}$$

$$\Delta(F \cdot S_{SO}) = (2{,}400)(677) - (2{,}400)(440) = 569{,}000 \text{ g COD/day}$$
$$= 569 \text{ kg COD/day}$$

$$\Delta(F \cdot X_{SO}) = (2{,}400)(84) - (2{,}400)(84) = 0.0 \text{ kg COD/day}$$

Substitution of these values into Eq. 10.14 gives:

$$\Delta(F \cdot S_N)_{A,TS} = 45.6 - (0.087)(0.521)(1.20)(569) = 14.7 \text{ kg N/day}$$

Substitution of this value into Eq. 10.13 gives the transient state oxygen requirement associated with nitrification:

$$RO_{A,TS} = [(4.57 - (0.20)(1.20)](14.7) = 63.4 \text{ kg } O_2\text{/day}$$

h. What is the total oxygen requirement for nitrification?
The total oxygen requirement for nitrification is $RO_A + RO_{A,TS}$, or

$$223 + 63.4 = 286.4 \text{ kg } O_2\text{/day}$$

i. What is the total oxygen requirement for the lagoon?
The total oxygen requirement is the sum of the heterotrophic and autotrophic requirements, or:

$$919 + 286 = 1{,}205 \text{ kg } O_2/\text{day}$$

After the lagoon volume and oxygen requirement are known, the power input required to achieve complete mixing and suspension of settleable solids is then calculated and compared to the power input required to transfer the needed oxygen. The required volumetric power input for mixing is calculated with Eq. 14.2 and multiplied by the lagoon volume to obtain the total power input for mixing. Equation 14.2 requires knowledge of the bioreactor suspended solids concentration. That concentration can be calculated using Eq. 9.11 after simplification by recognizing that, for a CMAL, the SRT and HRT are identical. The modified equation is:

$$X_{M,T} = X_{IO,T} + \frac{(1 + f_D \cdot b_H \cdot \tau)Y_{H,T}(S_{SO} + X_{SO} - S_S)}{1 + b_H \cdot \tau} \qquad (14.9)$$

Suspended solids concentrations should be calculated for the lowest sustained bioreactor temperature since that will result in the highest value because of the low biomass decay. If sufficient information is not available to use Eq. 14.9, a volumetric power input of 6 KW/1000 m^3 can be used as an approximation.

The power input for oxygen transfer can be calculated based on the efficiency of the oxygen transfer device used. High-speed surface aerators are often used in CMALs, and these devices have in-process efficiencies on the order of 1.0 to 1.1 kg O_2/(kW · hr). Values for other oxygen transfer devices are available from the manufacturers of the equipment and in various manuals of practice.[42,43] These calculations are illustrated in the following example.

Example 14.3.1.3
Determine the power that must be supplied to the CMAL considered in Examples 14.3.1.1 and 14.3.1.2. This must be done by calculating the power required to mix the lagoon and keep the solids in suspension and the power required to deliver the needed amount of oxygen, and then choosing the larger of the two.

a. What is the total suspended solid (TSS) concentration?
The TSS concentration is calculated using Eq. 14.9 for the average process loading condition and the low weekly CMAL temperature of 12°C. We saw in Example 14.3.1.1 that b_H has a value of 0.11 day^{-1} at that temperature. Using the HRT of 3.3 days, Eq. 14.9 gives:

$$X_{M,T} = 45 + \frac{[1 + (0.20)(0.11)(3.3)](0.521)(440 + 84 - 10)}{1 + (0.11)(3.3)}$$

$$= 256 \text{ mg/L as TSS}$$

b. What is the power input required for mixing and biomass suspension?
The volumetric power input for mixing and biomass suspension can be estimated with Eq. 14.2 using the solids concentration of 256 mg/L:

$$\Pi = (0.004)(256) + 5 = 6 \text{ kW/1000 } m^3$$

At a bioreactor volume of 8,000 m^3, the total power required for mixing is:

$$(8{,}000)(6) = 48 \text{ kW}$$

c. What power input is required for oxygen transfer?
The peak oxygen requirement, as calculated in Example 14.3.1.2, is 1,205 kg O_2/day, or 50.2 kg O_2/hr. Assuming the use of an aerator with an efficiency of 1.0 kg O_2/(kW · hr), the power required for oxygen transfer is 50.2/1.0 = 50.2 kW. Consequently, the power required for oxygen transfer will control. When that amount of power is supplied, the lagoon will be adequately mixed. The selection of specific aerators and the determination of their locations in the actual lagoon geometry may result in the need for some adjustment to the design and power input.

14.3.2 Completely Mixed Aerated Lagoon with Aerobic Solids Stabilization

Conversion of organic matter into biomass results in stabilization of a portion of the oxygen demand contained in the wastewater. This occurs because a fraction of the organic matter, roughly $(1 - Y_H)$, must be oxidized to provide energy for biomass synthesis. Since the numerical value of Y_H is on the order of 0.6 mg biomass COD synthesized/mg COD removed, about 40% of the biodegradable organic matter in the influent wastewater will be stabilized by a CMAL sized for conversion of organic matter to biomass. As demonstrated above, organic matter can be converted to biomass at relatively short HRTs in CMALs. Additional stabilization of organic matter can be obtained by increasing the HRT (SRT) to allow the synthesized biomass to decay.

Aerobic stabilization is accomplished by oxidation of biodegradable organic matter, whether in the form of the original substrate or in the form of biomass derived from that substrate. Thus, the efficiency of stabilization in a lagoon can be calculated as the steady-state oxygen requirement associated with the oxidation of organic matter divided by the loading of biodegradable organic matter to the lagoon. The fraction of the influent biodegradable organic substrate aerobically stabilized, Ξ, can be calculated as follows:

$$\Xi = \frac{RO_H}{F(S_{SO} + X_{SO})} \tag{14.10}$$

The bioreactor volume required to achieve a specified degree of stabilization can be calculated with the following equation, which follows from substituting Eq. 14.5 into Eq. 14.10 and assuming that the residual soluble substrate concentration is negligible:

$$V = \frac{F[(Y_{H,T} \cdot i_{O\,XB,T}) - (1 - \Xi)]}{b_H[1 - \Xi - (f_D \cdot Y_{H,T} \cdot i_{O\,XB,T})]} \tag{14.11}$$

Again, the bioreactor volume is calculated at the lowest sustained operating temperature since reactions rates are slowest then. Bioreactor volumes can escalate significantly if a high degree of stabilization is desired under all weather conditions because larger lagoons lead to more cooling, which decrease reaction rates. Sizing a CMAL for solids stabilization is illustrated in the following example.

Example 14.3.2.1

Reconsider the wastewater used in the previous examples and size a CMAL to stabilize 55% of the biodegradable organic matter contained in the wastewater.

a. What degree of stabilization is achieved in the summer in the lagoon designed in Examples 14.3.1.1–14.3.1.3, and what does it suggest about the operating temperature in this situation?

The degree of stabilization can be estimated with Eq. 14.10 by using the RO_{II} value of 706,000 g O_2/day calculated in Example 14.3.1.2:

$$\Xi = \frac{706{,}000}{(2{,}400)(440 + 84)} = 0.56$$

Thus, 56% stabilization is achieved under summer operating conditions. Since the new design must achieve almost this degree of stabilization in the winter, a larger lagoon will be required, which will make the winter operating temperature lower. Since the winter temperature was 12°C in the examples of Section 14.3.1, assume a winter temperature of 5°C for this situation. This assumption will have to be checked.

b. What size CMAL will be required to achieve 55% stabilization in winter?

The decay coefficient is the only kinetic parameter appearing in Eq. 14.11, so only its value must be adjusted to the assumed temperature of 5°C:

$$b_{H,5} = (0.15)(1.04)^{(5-20)} = 0.08 \text{ day}^{-1}$$

That coefficient is then used in Eq. 14.11 to calculate the required volume.

$$V = \frac{(2{,}400)[(0.521)(1.20) - (1 - 0.55)]}{(0.08)[1 - 0.55 - (0.2)(0.521)(1.20)]} = 16{,}150 \text{ m}^3$$

This is approximately twice as large as the CMAL required to give an effluent soluble substrate concentration of 10 mg/L, as determined in Example 14.3.1.1. The larger volume gives an HRT of 6.7 days.

Before the calculated volume can be considered to be acceptable, the assumed temperature must be verified through use of Eq. 14.3, just as was done in Example 14.3.1.1. Such an exercise predicts that the lagoon temperature will be 6°C in the winter. This is sufficiently close to the assumed temperature to be acceptable. As with Example 14.3.1.1, if the calculated temperature had not been consistent with the assumed temperature an iterative procedure would have been required to calculate the required volume and resulting temperature.

Oxygen requirements and mixing requirements are calculated in the same manner as in Section 14.3.1. Oxygen requirements are calculated for summertime operating conditions, and the long HRTs required to achieve solids stabilization generally mean that nitrification will occur. The increased volume required for solids stabilization also means that bioreactor power inputs are more likely to be controlled by mixing and settleable solids suspension than by oxygen transfer.

Completely mixed aerated lagoon processes for solids stabilization can be configured as either a single CSTR or as CSTRs in series, as illustrated in Figure 14.3b. Significant improvements in performance result by using two lagoons in series rather than a single lagoon, but additional improvements are modest as the number of

lagoons is increased further, just as was the case for aerobic digesters, as illustrated in Figure 12.14. The treatability of the wastewater will generally decrease as it moves from lagoon to lagoon because the more easily biodegradable organic matter is removed in the upstream lagoons. Consequently, the potential improvement in performance may be overestimated by equations that assume constant kinetics throughout a staged system,[22] such as those used to generate Figure 12.14. Thus, a maximum of three lagoons in series is generally used. Caution should be exercised to avoid an initial stage that is so small that the oxygen requirement controls the power input, rather than the mixing requirement. It is not unusual to use unequal size lagoons in series, with the initial CMAL being larger so that power requirements for mixing and oxygen transfer are balanced. The initial CMAL may be 50 to 60% of the total lagoon volume, with the subsequent two lagoons of equal size.

14.3.3 Completely Mixed Aerated Lagoon with Benthal Stabilization and Storage

A CMAL with benthal stabilization, illustrated in Figure 14.3c, represents another approach to incorporating solids stabilization into aerobic lagoons. It offers the further benefit of providing a clarified effluent since settleable solids are removed in the benthal stabilization basin. The CMAL, i.e., the first basin, is designed as described in Section 14.3.1, while the benthal stabilization basin is designed for solids sedimentation, stabilization, and storage.

As illustrated in Figure 14.3c, settleable solids accumulate in the lower portion of the benthal stabilization basin where they are anaerobically digested. Aeration and mixing are provided to the upper portion to supply oxygen to serve as the electron acceptor for bacteria degrading the organic matter released from the anaerobically digesting solids. Aeration of the overlying water column also allows aerobic biodegradation of residual organic matter that escapes the upstream CMAL. Mixing promotes fluid movement, which minimizes stagnant regions where algae can grow. However, mixing levels must be kept low enough so that settleable solids do not remain in suspension. Furthermore, the HRT in the clear water zone above the benthal layer must be sufficiently short to minimize the growth of algae. From considering the above, it can be seen that requirements in the design of a benthal stabilization basin include: providing sufficient volumetric power inputs to mix the upper portion of the basin and to meet oxygen requirements, while still allowing the sedimentation of settleable solids; supplying a sufficiently large clear water zone to allow sedimentation of suspended solids, while also making it small enough to avoid algal growth; and furnishing a sufficient solids digestion and storage zone for stabilization of the particulate organic matter applied. Experience indicates that these multiple, conflicting objectives can generally be achieved.

Although sedimentation of settleable solids is an important function of the clear water zone, it does not generally control the sizing of the basin since the basin's HRT and depth result in overflow rates that are at least an order of magnitude lower than those required for good solids sedimentation. Thus, other factors control its size.

Consider first the solids stabilization and storage requirements. The input of suspended solids into the benthal layer can be thought to consist of two components, one biodegradable and the other nonbiodegradable. The surface area of the benthal stabilization basin is determined by the input of biodegradable solids, while the depth

of the benthal zone is determined primarily by the input and storage of nonbiodegradable solids.

First consider the surface area. If the biodegradable solids loading is sufficiently small, then those solids will not accumulate during the warm months when biodegradation is proceeding at its maximal rate. Rather, they will be anaerobically degraded into carbon dioxide and methane in the benthal zone. However, during cold winter months, biodegradable solids may accumulate because the lagoon temperature may be too cold for effective anaerobic decomposition. As a consequence, areal solids loadings are generally made small enough to allow time for biodegradable solids deposited during the winter to decompose during the subsequent summer. The steady-state areal loading rate for biodegradable suspended solids, $\Gamma_{A,XB}$, depends on the amount of biomass synthesized in the preceding CMAL. Consequently:

$$\Gamma_{A,XB} = \frac{F\left[\dfrac{(S_{SO} + X_{SO} - S_{SI})Y_{H,T} \cdot i_{O,XB,T}}{1 + b_H \cdot \tau_1}\right]}{A_1} \tag{14.12}$$

The units on $\Gamma_{A,XB}$ are g biodegradable COD/($m^2 \cdot$ day). The bracketed term in the numerator of Eq. 14.12 is the active heterotrophic biomass concentration (as COD) in the CMAL preceding the benthal stabilization basin, where τ_1 is the HRT of that CMAL. Settleable solids are generally deposited on both the bottom and the sloping sides of the benthal stabilization basin. Consequently, the entire water surface of that basin, A_1, is used in the calculation of the biodegradable areal solids loading. The required surface area, A_1, is calculated by adopting a reasonable value for the areal loading and using a rearranged form of Eq. 14.12. A conservative value of about 115 g biodegradable COD/($m^2 \cdot$ day) [80 g biodegradable VSS/($m^2 \cdot$ day)] has been recommended by Rich[28,29,31,32,37] based on pilot tests and experience with full-scale systems.

Inert solids entering the benthal stabilization basin will not be biodegraded and will accumulate. The depth of the benthal zone depends on their rate of accumulation, the length of time over which they are allowed to accumulate, their concentration in the benthal layer, and the surface area of the basin. Inert solids include those present in the influent wastewater as well as those produced by decay of biomass produced in the upstream CMAL. The steady-state input (accumulation) rate for inert organic solids as TSS, $RX_{I,T}$, can be calculated as:

$$RX_{I,T} = F\left[X_{IO,T} + \frac{f_D \cdot b_H \cdot \tau_1 \cdot Y_{H,T}(S_{SO} + X_{SO} - S_{SI})}{1 + b_H \tau_1}\right] \tag{14.13}$$

in which the right term within the bracket represents biomass debris arising from biomass grown in the preceding CMAL. Multiplication of this accumulation rate by the time period over which accumulation will occur gives the mass of inert organic solids to be stored. That time period depends on the means by which the solids are disposed of, as well as on the cost of land for the CMAL and benthal stabilization basin. However, three years is a common value. Space must also be provided for accumulation of the biodegradable suspended solids during the winter months when colder temperatures reduce anaerobic activity. Typical practice is to remove solids from a benthal stabilization basin in the fall before biodegradable solids start to accumulate because that provides the minimum quantity on unstabilized solids. Con-

sequently, only enough storage volume is required for six months accumulation of biodegradable solids. The accumulation rate of biodegradable solids during the winter in COD units is just $\Gamma_{A,XB} \cdot A_l$ from Eq. 14.12. It can be converted to TSS units by division by the conversion factor $i_{O\ XB,T}$. After the mass of accumulated solids is known, the volume required to house them can be calculated by dividing by their concentration in the benthal layer. Concentrations in the 30 to 50 g/L (kg/m^3) range are commonly obtained. After the required volume is known, the depth can be calculated by dividing the volume by the basin area, A_l, calculated with Eq. 14.12. Finally, some additional depth must be provided to allow space for the digesting biodegradable suspended solids. The following example illustrates the sizing of a benthal stabilization basin.

Example 14.3.3.1

Consider the CMAL from Example 14.3.1.1, which has an HRT of 3.3 days. A benthal solids stabilization basin is to be provided downstream to remove and stabilize settleable suspended solids. Determine the required surface area for the basin as well as the depth of the benthal zone if the basin is to be cleaned every three years in the early fall. Base the design on an areal loading rate of 115 g biodegradable COD/(m^2·day), a benthal solids concentration of 40 g/L as TSS of organic solids, and a temperature of 20°C.

a. What is the surface area of the benthal stabilization basin?
The basin surface area can be calculated with Eq. 14.12 by using a value of 115 g biodegradable COD/(m^2·day) for $\Gamma_{A,XB}$. The value of S_s may be neglected.

$$A_l = \frac{(2{,}400)\left[\dfrac{(440 + 84)(0.521)(1.20)}{1 + (0.15)(3.3)}\right]}{115} = 4{,}570 \text{ m}^2$$

b. What mass of inert organic suspended solids must be stored if the basin is cleaned every three years?
The mass of inert organic solids to be stored is the steady state input rate times the storage period. The steady state input rate of inert organic solids can be calculated with Eq. 14.13:

$$RX_{I,I} = (2{,}400)\left[45 + \frac{(0.2)(0.15)(3.3)(0.521)(440 + 84)}{1 + (0.15)(3.3)}\right]$$
$$= 151{,}000 \text{ g/day} = 151 \text{ kg/day}$$

Since the lagoon will be cleaned every three years, the total mass accumulated is:

$$(151)(3)(365) = 165{,}000 \text{ kg of inert organic solids as TSS.}$$

c. What mass of biodegradable solids must be provided for?
Space must be provided for six months accumulation of biodegradable organic solids. Because this space must be available during the winter and the basin will be cleaned in the fall, not all of the inert solids will have accumulated when this space is required. However, by adding the mass of bio-

degradable solids to the total mass of inert solids calculated above, additional space will be provided to accommodate digesting solids.

The accumulation rate of biodegradable solids in COD units is just $\Gamma_{A,XB} \cdot A_L$. Dividing this by the conversion factor between COD and TSS units, $i_{O,XB,T}$, gives the accumulation rate in TSS units. Using this concept, the accumulation rate of biodegradable solids during the winter is:

$$\frac{(115)(4{,}570)}{1.20} = 438{,}000 \text{ g/day} = 438 \text{ kg/day as TSS}$$

Since these solids will accumulate for up to six months, the total mass accumulated will be:

$$(438)(0.5)(365) = 80{,}000 \text{ kg of biodegradable solids as TSS.}$$

d. What will the volume of the solids storage zone be?
Solids in the digesting benthal solids blanket will thicken to approximately 40 g/L (kg/m^3). Thus, the maximum solids volume will be approximately:

$$\frac{165{,}000 + 80{,}000}{40} = 6{,}100 \text{ m}^3$$

e. What will the depth of the solids storage zone be?
Since the basin surface area is 4,570 m^2 the solids will accumulate to an average depth of:

$$\frac{6{,}100}{4{,}570} = 1.3 \text{ m}$$

f. What mass of solids must be disposed of?
The mass of solids removed at the end of three years is three years accumulation of inert solids. Because the basin will be desludged at the end of the summer, all of the biodegradable solids will have been digested. The mass of inert solids accumulated over three years was calculated in step b. Thus, 165,000 kg as TSS of inert organic solids must be disposed of.

The clear water zone above the benthal solids is sized next. Several factors determine the volume of this zone. One is the trade-off between oxygen transfer and mixing. Sufficient oxygen must be transferred to meet peak demands exerted by the feedback of reduced compounds from the digesting benthal solids layer. Mixing levels must also be sufficiently low so that settleable solids do not remain in suspension.

Peak oxygen demands will generally be experienced in the spring and the fall of the year as the water temperature reaches about 16°C. Experience indicates a peak oxygen requirement during these periods of about 80 g O_2/(m^2 · day), again based on the lagoon surface area.[28,29,31,32,37] The power input required to meet this demand will depend on the efficiency of the aerator used. Just as with a CMAL, the mechanical surface aerators typically have an oxygen transfer efficiency on the order of 1.0 to 1.1 kg O_2/(kW · hr).

Once the power input for oxygen transfer is known, the volume of the clear water zone is chosen to keep the volumetric power input between 1.0 and 2.0 kW/1000 m^3. As discussed previously, at volumetric power inputs below 1.0 kW/1000 m^3 dissolved oxygen and other soluble constituents will not be uniformly

distributed, while at volumetric power inputs above 2.0 kW/1000 m^3 settleable solids will remain in suspension.

The sizing of the clear water zone based on these considerations is illustrated in the following example.

Example 14.3.3.2

Size the clear water zone for the benthal stabilization basin considered in Example 14.3.3.1. Assume that the peak oxygen demand is 80 g $O_2/(m^2 \cdot day)$ and that the oxygen transfer device has a transfer efficiency of 1.0 kg $O_2/(kW \cdot hr)$. Maintain the volumetric power input between 1.0 and 2.0 kW/1000 m^3 during all operating conditions.

a. How much power is needed for oxygen transfer?
The lagoon surface area is 4,570 m^2. At a peak oxygen demand of 80 g $O_2/(m^2 \cdot day)$, the peak oxygen requirement is:

$$(4,570)(80) = 365,600 \text{ g } O_2/\text{day} = 15.2 \text{ kg } O_2/\text{hr}.$$

Assuming the use of high speed surface aerators with an efficiency of 1.0 kg $O_2/(kW \cdot hr)$, the power required is 15.2 kW.

b. What is the volume of the clear water zone?
The minimum clear water zone volume will result when the volumetric power input to the clear water zone is maximized. At a maximum volumetric power input of 2 kW/1000 m^3, the volume of the clear water zone is 15.2 ÷ 2 = 7.6, in units of 1000 m^3. Thus, the volume is 7,600 m^3.

c. What is the volumetric power input immediately after cleaning before solids have accumulated?
Before solids have accumulated, the entire basin acts like a clear water zone. The total basin volume is the volume of the clear water zone calculated in step b (7,600 m^3) plus the maximum volume of the benthal solids zone (6,100 m^3) for a total of 13,700 m^3. Thus, the initial volumetric power input will be 15.2 ÷ 13.7 x 10^3 = 1.1 kW/1000 m^3. This is sufficiently high to uniformly disperse oxygen throughout the basin.

A final consideration is the HRT in the clear water zone. An excessive HRT can lead to the growth of algae, which will pass into the effluent and deteriorate effluent quality. As discussed in Section 14.2.1, an HRT of 2 days or greater will generally result in excessive algal growth. In cases where the HRT exceeds 2 days, a series of completely mixed cells can be used to prevent the HRT in any one cell from exceeding 2 days. This works well from a practical perspective since multiple benthal stabilization basins are typically provided so that one can be taken out of service to allow the removal of stabilized solids while the remaining cells continue to remove solids coming from the upstream CMAL. Based on the work of Toms et al.,[40] Rich[30,35] developed an equation for the growth of algae in equal sized completely mixed cells in series:

$$\frac{X_{PN}}{X_{PO}} = \frac{1}{\left[1 - \mu_P \frac{\tau}{N}\right]^N} \tag{14.14}$$

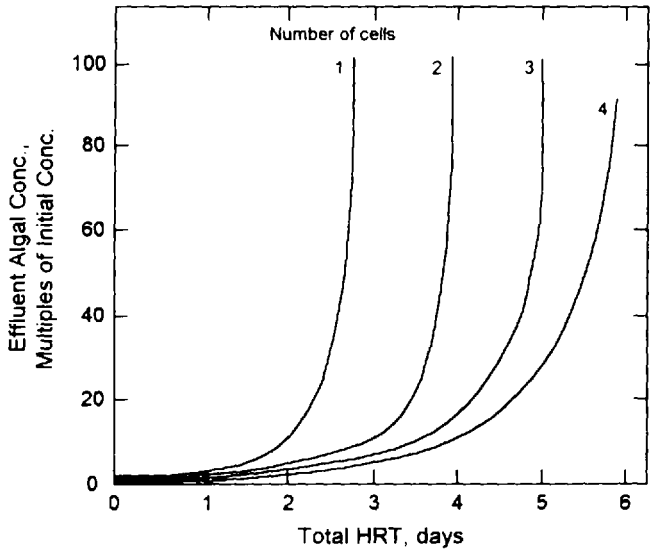


Figure 14.8 Effects of total clear water zone HRT and the number of equal size completely mixed cells in series on algal growth in lagoons. Theoretical curves from Eq. 14.14. (From L. G. Rich, "Modification of design approach to aerated lagoons," *Journal of Environmental Engineering*, **122**, 149–153, 1996. Copyright © American Society of Civil Engineers; reprinted with permission.)

where X_{PN} and X_{PO} are the concentrations of photosynthetic microorganisms (algae) in the effluent from the last completely mixed cell and the influent to the first completely mixed cell, respectively, N is the number of equal-sized completely mixed cells, μ_P is the growth rate of the photosynthetic microorganisms (taken to be 0.48 day^{-1}), and τ is the total HRT of the clear water zones of the series of cells. This equation is applicable for values of individual cell HRTs (τ/N) less than $1/\mu_P$, or 2.08 days. For values of individual cell HRTs greater than this, algal growth will not be controlled and X_{PN}/X_{PO} cannot be predicted by this equation. Based on experience with CMAL systems followed by benthal stabilization basins, Rich[35] recommended that X_{PN}/X_{PO} ratios be kept less than 25. Equation 14.14 is plotted in Figure 14.8 where it illustrates that the total HRT in the clear water zone can be increased significantly when multiple cells in series are used while still minimizing the growth of algae. The effect is not linear, since an increase from one to two cells produces a much larger increase in allowable total basin clear water zone HRT than does an increase from two to three cells, and so on. The following example illustrates the use of this equation to configure a benthal stabilization basin.

Example 14.3.3.3

Determine what configuration should be used for the benthal stabilization basin sized in Examples 14.3.3.1 and 14.3.3.2 to minimize algal growth.

a. What are the HRTs in the basin and what do they suggest about the required configuration of the system?

Since the flow to be treated is 2,400 m^3/day, the HRT in the clear water zone is 7,600 ÷ 2,400 = 3.1 days and the HRT based on the total basin volume is 13,700 ÷ 2,400 = 5.7 days. Since both of these exceed a value of 2 days, excessive algal growth is likely if the basin is configured as a single cell.

b. What is the minimum number of cells in series required to control algal growth?
We saw in Section 14.2.1 that the minimum SRT for algal growth was around 2 days. This means that the maximum SRT per cell is around 2 days. Consequently, considering only the clear water zone, the stabilization basin would have to be divided into at least two equal size cells. Immediately after cleaning, light could penetrate throughout the basin, so the entire volume would be available for algal growth. Since the total HRT of the basin is 5.7 days, it would have be divided into at least three equal sized cells to keep the HRT in any given cell below 2 days.

c. Based on Eq. 14.14, how many equal sized cells in series should be used to prevent excessive algae growth?
An HRT of 2 days in each cell does not guarantee that the effluent algal concentration will be acceptable. Equation 14.14 must be used to evaluate the effect of the number of cells on algal growth potential for the full range of operating conditions. Consequently, use it to calculate the X_{PN}/X_{PO} ratio based on the clear water zone and total basin volume for 3, 4, 5 and 6 cells in series. For example, using the clear water zone volume, which gives a total HRT of 3.1 days, and 3 cells in series:

$$\frac{X_{PN}}{X_{PO}} = \frac{1}{\left[1 - 0.48\left(\frac{3.1}{3}\right)\right]^3} = 7$$

which is acceptable. The values of X_{PN}/X_{PO} are calculated for HRTs of 3.1 and 5.7 days and summarized in Table E14.2. The results shown there indicate that effective control of algae (X_{PN}/X_{PO} less than 25) can be achieved using three cells if the clear water zone HRT is restricted to 3.1 days. Algae control, on the other hand, is quite difficult if the full basin volume is used, i.e., such as immediately after cleaning. Consequently, use three cells along with the capability to adjust the lagoon depth to control the clear water zone HRT. The combination of these features should provide ample flexibility to control algal growth.

Table E14.2 Evaluation of the Potential for Algal Growth in the Benthal Stabilization Basin Considered in Example 14.3.3.3

Total HRT	X_{PN}/X_{PO}			
	N			
days	3	4	5	6
3.1	7	6	6	6
5.7	2,680	117	58	42

This example illustrates the need for flexibility in the design of a benthal stabilization basin. It may be desirable to reduce the basin depth (and consequently the volume) immediately after solids have been removed from the basin so that the clean water zone HRT, and the corresponding potential for the growth of algae, is reduced. The depth would be increased later as solids accumulated to maintain the necessary clear water zone.

While degradation of most of the readily biodegradable organic matter occurs in the CMAL, further biodegradation will occur in the clear water zone of the benthal stabilization basin. This can be significant when a high-quality effluent is desired. Although the bulk of the active heterotrophs are removed from the clear water zone by sedimentation, a residual concentration of about 4 to 6.5 mg/L as TSS remains. With the long HRT in the clear water zone, and particularly when the benthal stabilization basin consists of multiple cells in series, effective removal of readily biodegradable substrate can be obtained. Because of the already low concentration of biodegradable organic matter, the first order approximation of the Monod equation (Eq. 3.38) can be used and applied to the clear water zone of the multi-cell benthal stabilization basin for prediction of the soluble substrate concentration in the Nth cell, S_{SN}. The relationship is:

$$S_{SN} = \frac{S_{SCMAL}}{\left[1 + \left(\frac{\hat{\mu}_H}{Y_{H,T} \cdot K_S}\right) X_{B,H} \cdot \frac{\tau}{N}\right]^N} \tag{14.15}$$

where S_{SCMAL} is the concentration of readily biodegradable organic matter in the effluent from the upstream CMAL, τ is the total HRT of the clear water zone of the multi-cell benthal stabilization basin, and N is the number of cells in the basin. Again, $X_{B,H}$ will be about 4 to 6.5 mg/L as TSS.

14.4 PROCESS OPERATION

Lagoon operation is quite straightforward. It generally involves placing the required number of units in service to achieve the desired VOL or AOL and performing necessary maintenance so that facilities are available for service.

One of the principal difficulties experienced with lagoon processes other than CMALs is odors. The potential for odor production is an inherent part of anaerobic and facultative lagoons because the anaerobic metabolism occurring in them forms odoriferous organic and inorganic end products. Emission of these products can be controlled by minimizing transport to a receptor. For some wastewaters, scum and other materials accumulate on the surface of the lagoon, sealing it and restricting odor emissions. In other cases an overlying layer of relatively clean, oxygenated water is provided by encouraging the growth of algae in the overlying clear water (facultative lagoons), by providing devices that transfer oxygen into the overlying clear water (benthal stabilization basins), or by recirculating oxygenated water from a downstream aerobic unit. This latter approach is typically used when an anaerobic lagoon is followed by a facultative lagoon. Covers can also be added to anaerobic lagoons to contain odoriferous materials and collect the methane gas.

Organic overloads can cause odor problems when oxygen demands exceed oxygen supplies in any lagoon. Such upsets require that additional electron acceptor

be added to match the increased organic loading. Effluent recirculation and the addition of mechanical aeration are two options. Other options include addition of hydrogen peroxide, which degrades to water and oxygen, or nitrate-N, which serves directly as an alternative electron acceptor. The addition of nitrate-based fertilizers offers an easily implemented and effective option. The expense of these latter approaches relative to mechanical aeration generally relegates them to temporary use.

Solids accumulations occur in most lagoons, and they must be removed when they begin to interfere with performance. One method of removing solids involves taking the lagoon out of service, removing the relatively clean water, and allowing the remaining solids to dry. Although a dry product is produced for ultimate disposal, this method will keep the lagoon out of service for several weeks, which may adversely impact overall wastewater treatment. Alternatively, dredging can be used to remove accumulated solids while the lagoon remains in service. Odor emissions are generally a problem during lagoon cleaning since the solids are anaerobic and are exposed to the atmosphere. Solids accumulations can also occur in CMALs if the volumetric power inputs are inadequate or if the location and arrangement of aerators are inappropriate.[3,16,18]

The control of mixing energy levels and HRTs to control the growth of algae is particularly critical in aerobic lagoons. Sufficient mixing must be provided in CMALs to maintain settleable solids in suspension and block light penetration to minimize algal growth. In basins where suspended solids are allowed to settle and light can penetrate the water column, algal growth is controlled by maintaining the HRT of each cell in the basin sufficiently low so that algae are washed out of the system. Techniques available to control the HRT include adjusting the number of cells in service and controlling the basin depth. If the guidelines presented in Section 14.3.3 are followed, algal growth can be controlled and the associated adverse impacts on effluent quality can be avoided.

Icing of aerators can be a particular problem in cold climates.[39] Mechanical surface aerators spray bioreactor contents into the air, and mist can condense and freeze on the aerator floats. Accumulations of ice can result in unbalanced loads on the floats, which cause them to tip over and sink. Heat tracing can be added to the aerator to prevent such accumulations.

Further information on the operation of lagoons is provided in various standard references.[21,41,44]

14.5 KEY POINTS

1. The term lagoon refers to a diverse array of suspended growth biochemical operations with no downstream clarifier. They are typically constructed as in-ground earthen basins, resulting in their name.
2. Lagoons are mechanically simple, but this masks a degree of physical, chemical, and biological complexity that exceeds that of other biochemical operations. They stabilize biodegradable organic matter by aerobic and anaerobic processes; remove nitrogen by nitrification, denitrification, and stripping; and remove phosphorus by precipitation.
3. In anaerobic lagoons (ANLs) organic matter is stabilized by conversion to carbon dioxide and methane. Mixing is provided by gas evolution, and

this results in settling and retention of suspended solids. Consequently, the solids retention time (SRT) is longer than the hydraulic retention time (HRT).

4. In facultative and facultative/aerated lagoons organic matter is stabilized by both aerobic and anaerobic metabolism. Oxygen is provided by algae and, in some instances, also by mechanical means. Anaerobic metabolism occurs in the lower portion of the lagoon, and odoriferous compounds formed there are oxidized in the aerobic upper portion.
5. In aerobic lagoons organic matter is stabilized using oxygen as the electron acceptor. The growth of algae is limited by maintaining solids in suspension to limit light penetration and by maintaining the HRT in the clear water zone below the minimum value required for algal growth (generally about 2 days).
6. Lagoons are widely used to pretreat municipal and industrial wastewaters prior to treatment in a downstream treatment system.
7. In a completely mixed aerated lagoon (CMAL) the SRT and HRT are equal. Its performance can be characterized by the relationships presented in Chapter 5 or by International Association on Water Quality (IAWQ) activated sludge model (ASM) No 1.
8. The volumetric organic loading rate (VOL) is used to characterize the performance of processes such as ANLs where mixing conditions allow the settlement and accumulation of biomass.
9. The areal organic loading rate (AOL) is used to characterize the performance of lagoons in which algae provide the oxygen. It is also used to characterize benthal stabilization processes where odoriferous products released to the overlying clear water zone must be oxidized.
10. The volumetric power input provided by mechanical aeration systems determines the degree of mixing and the suspension of solids in an aerobic lagoon. A value of 1 kW/1000 m^3 provides uniform dispersion of dissolved species such as oxygen. A value less than 2 kW/1000 m^3 allows settleable suspended solids to settle, while 6 to 10 kW/1000 m^3 is generally needed to maintain all suspended solids in suspension.
11. The large surface area and long HRT generally used with lagoons result in significant heat loss during cold weather. Reduced temperatures cause lower biological activity and decreased treatment efficiency.
12. Three decisions must be made when a CMAL is designed: (1) the bioreactor volume, (2) the amount of oxygen to be supplied, and (3) the minimum power input required to achieve adequate mixing.
13. The effluent from a CMAL can be treated in a benthal stabilization basin where settleable solids are removed by sedimentation and stabilized by anaerobic processes.
14. Proper operation of a lagoon is achieved by maintaining appropriate process loadings. Temporary overloads can be mitigated by the addition of alternative electron acceptors such as hydrogen peroxide and nitrate-N. Solids are disposed of on a periodic basis, but accumulated solids must be removed when they begin to interfere with lagoon performance. Icing of aerators can be a particular problem in cold climates.

14.6 STUDY QUESTIONS

1. Prepare a table summarizing the benefits and drawbacks of lagoon processes. List the characteristics of the various lagoon processes and discuss where they are typically used.
2. Summarize the design approaches and types of design criteria used to size anaerobic lagoons, facultative and facultative/aerated lagoons, and aerobic lagoons. List typical values of the design criteria for each process.
3. Prepare a figure illustrating the impact of volumetric power input on mixing in an aerated lagoon. Distinguish between various mixing regimes and discuss the potential for algal growth in each of these regimes. What other approaches are available to control algal growth?
4. Using the wastewater characteristics listed in Table E8.4, the stoichiometric and kinetic parameters in Table E8.5, and the temperature correction factors in Table E10.1, size a CMAL to treat an average flow rate of 10,000 m^3/day to reduce the concentration of readily biodegradable substrate to 10 mg/L as COD. The range of weekly average air temperatures is 0°C to 35°C, with an average of 20°C. The diurnal flow and pollutant concentration variations are as illustrated in Figure 6.2. Determine the required volume of the CMAL.
5. Determine the average and diurnal peak oxygen requirements and required power input for aeration for the CMAL sized in Study Question 4. Assume an oxygen transfer efficiency for the oxygen transfer device of 1.1 kg O_2/(kW·hr). Also determine the power required to adequately mix the CMAL. Which power input controls the design?
6. Use a simulation package implementing IAWQ ASM No. 1 to evaluate the design developed in Study Questions 4 and 5.
7. The CMAL designed in Study Questions 4 and 5 must be upgraded by adding another CMAL in series with the existing unit to achieve 50% stabilization of biodegradable organic matter at the minimum weekly average temperature. Determine the size of the CMAL that must be added. What will the efficiency of stabilization of biodegradable organic matter be under average temperature conditions; under maximum weekly temperature conditions?
8. Determine the oxygen requirement and required aeration power input for the new CMAL considered in Study Question 7. Assume an oxygen transfer efficiency for the oxygen transfer device of 1.1 kg O_2/(kW·hr). Also determine the power required to adequately mix the CMAL. Which power input controls the design?
9. Use a simulation package implementing IAWQ ASM No. 1 to evaluate the design developed in Study Questions 7 and 8.
10. The CMAL designed in Study Questions 4 and 5 is to be upgraded by the addition of a benthal stabilization basin. Size and configure the basin and determine the power input required to meet oxygen and mixing requirements.
11. Develop a figure that demonstrates the factors affecting the heat balance and resulting temperature of a lagoon. The work of Argaman and Adams[3]

and of Stenstrom and coworkers[38,39] may be helpful. Discuss the relative importance of the various components of the heat balance.

12. Prepare a table comparing the benefits and drawbacks of the three lagoon processes considered in this chapter relative to their ability to meet effluent quality requirements for discharge to surface waters.

REFERENCES

1. Amberg, H. R., Aerated stabilization of board mill white water. *Proceedings of the 20th Industrial Waste Conference*, Purdue University Engineering Extension Series, No. 118, pp. 525–537, 1965.
2. Argaman, Y. and C. F. Adams, Jr., Comprehensive temperature model for aerated biological systems. *Progress in Water Technology* **9**(1/2):397–409, 1977.
3. Ayres, R. M., G. P. Alabaster, D. D. Mara and D. L. Lee, A design equation for human intestinal nematode egg removal in waste stabilization ponds. *Water Research* **26**:863–865, 1992.
4. Balasha, E. and H. Sperber, Treatment of domestic wastes in an aerated lagoon and polishing pond. *Water Research* **9**:43–49, 1975.
5. Barnhart, E. L., The treatment of chemical wastes in aerated lagoons. *Chemical Engineering Progress Symposium Series* **64**(90):111–114, 1968.
6. Barson, G., *Lagoon Performance and the State of Lagoon Technology*, U. S. Environmental Protection Agency Environmental Protection Technology Series, Report No. EPA-R2-73-144, June, 1973.
7. Bartsch, E. H. and C. W. Randall, Aerated lagoons—A report on the state of the art. *Journal, Water Pollution Control Federation* **43**:699–708, 1971.
8. Bess, F. D. and R. A. Conway, Aerated stabilization of synthetic organic chemical wastes. *Journal, Water Pollution Control Federation* **38**:939–956, 1966.
9. Beychock, M. R., Performance of surface-aerated basins. *Chemical Engineering Progress Symposium Series* **67**(107):322–339, 1971.
10. Boyle, W. C. and L. B. Polkowski, Treatment of cheese processing wastewater in aerated lagoons. *Proceedings of the Third National Symposium on Food Processing Wastes*, U.S. Environmental Protection Agency Environmental Protection Technology Series, Report No. EPA-R2-72-018, pp. 323–370, Nov. 1972.
11. City of Austin Texas and the Center for Research in Water Resources, *Design Guides for Biological Wastewater Treatment Processes*, U. S. Environmental Protection Agency Water Pollution Control Research Series, Report No. 11010 ESQ 08/71, August 1971.
12. Crown Zellerbach Corp., *Aerated Lagoon Treatment of Sulfite Pulping Effluents*, U. S. Environmental Protection Agency Water Pollution Control Research Series, Report No. 12040 ELW 12/70, Dec. 1970.
13. Fall, E. G., Jr., Retention time improves activated sludge effluent quality. *Journal, Water Pollution Control Federation* **37**:1194–1202, 1965.
14. Ferrara, R. A. and C. B. Avci, Nitrogen dynamics in waste stabilization ponds. *Journal, Water Pollution Control Federation* **54**:361–369, 1982.
15. Fleckseder, H. R. and J. F. Malina, *Performance of the Aerated Lagoon Process*, Technical Report CRWR-71, Center for the Research in Water Resources, University of Texas, Austin, Texas, 1970.
16. Haynes, F. D., Three years operation of aerated stabilization basins for paperboard mill effluent. *Proceedings of the 23rd Industrial Waste Conference*, Purdue University Engineering Extension Series, No. 132, pp. 361–373, 1968.

17. Kantardjieff, A. and J. P. Jones, Removal of toxicity and some nonconventional pollutants by a dual power multicellular lagoon system. *Water Environment Research* **65**: 819–826, 1993.
18. Laing, W. M., New secondary aerated stabilization basins at the Moraine Division. *Proceedings of the 23rd Industrial Waste Conference*, Purdue University Engineering Extension Series, No. 132, pp. 484–492, 1968.
19. Mancini, J. L. and E. L. Barnhart, Industrial waste treatment in aerated lagoons. In *Advances in Water Quality Improvement*, E. F. Gloyna and W. W. Eckenfelder, Jr., eds. University of Texas Press, Austin, pp. 313–324, 1968.
20. Marais, G. v. R., Fecal bacterial kinetics in stabilization ponds. *Journal of the Environmental Engineering Division, ASCE* **100**:119–139, 1974.
21. Middlebrooks, E. J., C. H. Middlebrooks, J. H. Reynolds, G. Z. Watters, S. C. Reed and D. B. George, *Wastewater Stabilization Lagoon Design, Performance and Upgrading*. Macmillan Publishing, New York, 1982.
22. Missouri Basin Engineering Health Council, *Waste Treatment Lagoons—State of the Art*, U. S. Environmental Protection Agency Water Pollution Control Research Series, Report No. 17090 EHX 07/71, July 1971.
23. Murphy, K. L. and A. W. Wilson, Characterization of mixing in aerated lagoons. *Journal of the Environmental Engineering Division, ASCE* **100**:1105–1117, 1974.
24. Pano, A. and E. J. Middlebrooks, Ammonia nitrogen removal in facultative wastewater stabilization ponds. *Journal, Water Pollution Control Federation* **54**:344–351, 1982.
25. Polprasert, C., M. G. Dissanayake, and N. C. Thanh, Bacterial die-off kinetics in waste stabilization ponds. *Journal, Water Pollution Control Federation* **55**:285–296, 1983.
26. Price, K. S., R. A. Conway, and A. H. Cheely, Surface aerator interactions. *Journal of Environmental Engineering Division, ASCE* **99**:283–300, 1973.
27. Rich, L. G., *Solids Control in Effluents from Aerated Lagoon Systems*, Report No. 73, Water Resources Research Institute, Clemson University, Clemson, South Carolina, 1978.
28. Rich, L. G., *Low-Maintenance, Mechanically Simple Wastewater Treatment Systems*. McGraw-Hill, New York, 1980.
29. Rich, L. G., Design approach to dual-power aerated lagoons. *Journal of the Environmental Engineering Division, ASCE* **108**:532–548, 1982.
30. Rich, L. G., Influence of multicellular configurations on algal growth in aerated lagoons. *Water Research* **16**:929–931, 1982.
31. Rich, L. G., Mathematical model for dual-power level, multicellular (DPMC) aerated lagoon systems. In *Mathematical Models in Biological Waste Water Treatment*, S. E. Jorgensen and M. J. Gromiec, eds. Elsevier, Amsterdam, pp. 147–168, 1985.
32. Rich, L. G., Troubleshooting aerated lagoon systems. *Public Works* **120**(10):50–52, 1989.
33. Rich, L. G., Technical Note No. 1, Aerated Lagoons. Office of Continuing Engineering Education, Clemson University, Clemson, South Carolina, 1993.
34. Rich, L. G., Technical Note No. 3, Aerated Lagoons. Office of Continuing Engineering Education, Clemson University, Clemson, South Carolina, 1994.
35. Rich, L. G., Modification of design approach to aerated lagoons. *Journal of Environmental Engineering* **122**:149–153, 1996.
36. Rich, L. G., Low-tech systems for high levels of BOD_5 and ammonia removal. *Public Works* **127**(4):41–42, 1996.
37. Rich, L. G. and B. W. Connor, Benthal stabilization of waste activated sludge. *Water Research* **16**:1419–1423, 1982.
38. Sedory, P. E. and M. K. Stenstrom, Dynamic prediction of wastewater aeration basin temperature. *Journal of Environmental Engineering* **121**:609–618, 1995.

39. Talati, S. N. and M. K. Stenstrom, Aeration-basin heat loss. *Journal of Environmental Engineering* **116**:70–86, 1990.
40. Toms, I. P., M. Owens, J. A. Hall, and M. J. Mindenhall, Observations on the performance of polishing lagoons at a large regional works. *Water Pollution Control* **74**: 383–401, 1975.
41. U. S. Environmental Protection Agency, *Design Manual, Municipal Wastewater Stabilization Ponds*, EPA-625/1-83-015, U. S. Environmental Protection Agency, Cincinnati, Ohio, 1983.
42. Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992.
43. Water Pollution Control Federation, *Aeration in Wastewater Treatment*, Manual of Practice No. FD-13, Water Pollution Control Federation, Alexandria, Virginia, 1988.
44. Water Pollution Control Federation, *Operation of Municipal Wastewater Treatment Plants*, Manual of Practice 11, Water Pollution Control Federation, Alexandria, Virginia, 1990.
45. White, S. C. and L. G. Rich, How to design aerated lagoon systems to meet 1977 effluent standards—Experimental studies. *Water and Sewage Works* **123**(3):85–87, 1976.
46. Williams, G. F., Discussion to 'Surface aerator interactions.' *Journal of the Environmental Engineering Division. ASCE* **100**:768–770, 1974.

Part IV
Theory: Modeling of Ideal Attached Growth Reactors

All of the biochemical operations considered in the preceding chapters have been suspended growth systems in which the biomass is suspended uniformly throughout the liquid phase. One assumption made in their modeling is that they behave as homogeneous systems; that is, that all microorganisms experience the dissolved constituents at the concentration of the liquid phase surrounding them. Even though the biomass exists in floc particles in most of those operations, no consideration was given to changes in concentration within the floc. This approach was acceptable because the effects of concentration gradients within floc particles are accounted for through quantification of the half-saturation coefficient, K_s, provided that the physical characteristics of the biofloc used in kinetic testing mimic closely the characteristics in the system to be modeled. The modeling of attached growth systems is more complicated because this simplifying assumption cannot be made. The biomass in these systems grows as a biofilm attached to a solid support (usually impermeable) with the result that substrate and other nutrients can only get to the bacteria within the biofilm by mass transport mechanisms. Consequently, biofilms must be treated as heterogenous systems in which the combined effects of reaction and transport are explicitly accounted for. In Chapter 15, we explore two approaches for doing this, the effectiveness factor and the pseudoanalytical approaches. In Chapters 16 and 17 we use the effectiveness factor approach to investigate the performance of packed towers, which behave like plug–flow systems, and rotating disc reactors, which behave like biofilms in CSTRs, respectively. Finally, in Chapter 18, we examine the theoretical performance of fluidized bed bioreactors, in which the biomass grows on particles suspended in the liquid phase, requiring us to also consider the hydrodynamics of particle fluidization.

This page intentionally left blank

15
Biofilm Modeling

The biochemical operations considered so far have all employed suspended growth cultures of microorganisms. However, as discussed in Section 1.2.3, attached growth bioreactors have been used extensively in environmental engineering practice. As the name implies, microorganisms in such bioreactors grow attached to a solid surface rather than being freely suspended in the wastewater undergoing treatment. That surface may be fixed in space with the wastewater flowing over it in thin sheets, such as in a packed tower, it may rotate about an axis, thereby moving through the fluid in the bioreactor, as in a rotating disc reactor, or it may be in the form of small particles that are held in suspension by the upward flow of water, as in a fluidized bed bioreactor. In all cases, however, the key distinguishing characteristic is that the microorganisms live in a biofilm attached to a surface. This means that the electron donor, the electron acceptor, and all other nutrients must be transported to the microorganisms within the biofilm by diffusional and other mass transport processes. It is the necessity to consider the combined effects of mass transport and reaction that makes the modeling of biofilm systems different from and more complicated than the modeling of suspended growth systems. In this chapter, we briefly examine the structure of biofilms and then review the ways in which the combined effects of transport and reaction are considered during modeling.

15.1 NATURE OF BIOFILMS

Biofilms are very complex, both physically and microbiologically. In fact, they are so complex that it is impossible to fully explore all aspects of them in the space available here. Therefore, those interested in a more detailed explanation of their properties should consult the treatise edited by Characklis and Marshall[6] and the reviews by Costerton et al.[9] and Lazarova and Manem.[23]

The basic conceptualization of a biofilm system is shown in Figure 15.1.[7] The biofilm grows attached to a solid support, which is usually impermeable, although it need not be. In this book, only impermeable supports will be considered. The solid support may be natural material, such as rock in old trickling filters, or it may be synthetic, such as the plastic packing in modern ones. Furthermore, it may range in configuration from corrugated sheets, such as in packed towers, to small particles, such as in fluidized beds. In general, the biofilm can be divided into two zones, the base film and the surface film. Both contain an assemblage of microorganisms and other particulate material bound together by a matrix of extracellular polymers. Those polymers, which are excreted by the microorganisms, are thought to be the same as

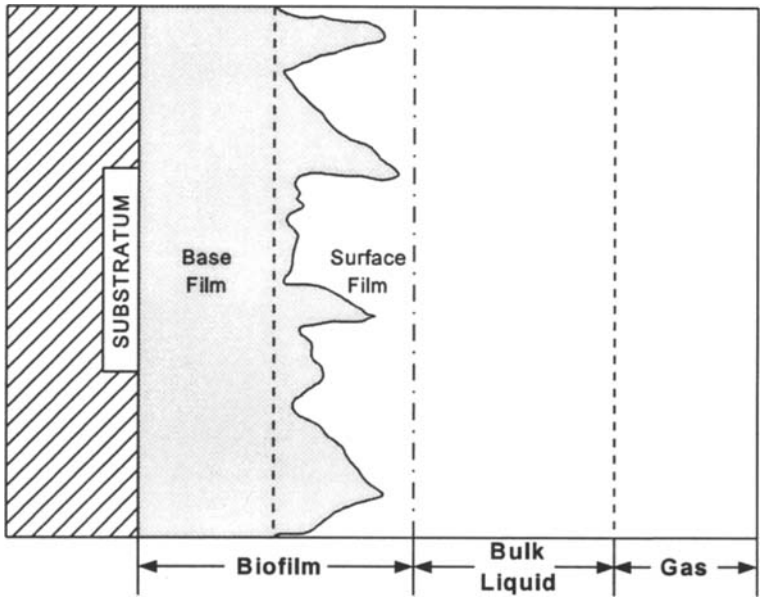


Figure 15.1 Conceptualization of a biofilm system. The base film and the surface film constitute the biofilm. (After Characklis and Marshall.[7])

the polymers involved in bioflocculation (see Section 10.2.1). The base film consists of a structured accumulation, with well-defined boundaries. Transport in the base film has historically been viewed as being by molecular processes (diffusion), although, as we will see later, that view is changing. The surface film provides a transition between the base film and the bulk liquid, and transport within it is dominated by advection. The relative thicknesses of the base and surface films depend largely on the hydrodynamic characteristics of the system, but also on the nature of the microorganisms in the biofilm. Consequently, one biofilm may have almost no surface film whereas another may be entirely surface film. There is normally relative motion between the biofilm and the bulk liquid, with the one moving depending upon the configuration of the attached growth process. For example, in packed towers the bulk fluid moves down over the biofilm in a thin sheet, whereas in a rotating disc reactor the biofilm support moves through the bulk liquid. In either case, however, mass transfer from the bulk fluid to the biofilm depends on the hydrodynamic regime. Finally, some biofilm systems contain a gas phase that provides oxygen or serves as a sink for gaseous products.

Most mathematical models of biofilm systems consider the surface film to be negligible, and therefore consider only the base film. Furthermore, unless they are specifically trying to model a variety of events such as carbon oxidation, nitrification, and denitrification, they usually reflect a single-species biofilm. Figure 15.2[43] shows such a biofilm. The bacterial cells can be seen to be suspended within a polymeric matrix, much the way fruit is held in a Jell-O™ salad. From such pictures the concept developed that the transport of substrates, nutrients, electron acceptor, etc. to and from the bacteria within the biofilm is by molecular diffusion alone.[18] Transport

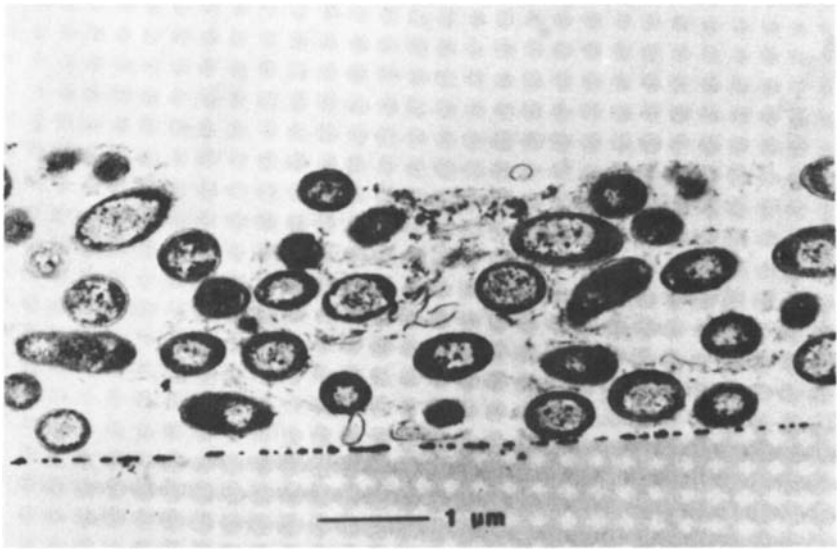


Figure 15.2 A transmission electron micrograph of a *Pseudomonas aeruginosa* biofilm consisting almost entirely of a base film. (From P. A. Wilderer and W. G. Characklis, Structure and function of biofilms. In *Structure and Function of Biofilms*, W. G. Characklis and P. A. Wilderer, eds. Wiley, New York, pp. 5–17, 1989. Reprinted by permission of John Wiley & Sons, Inc.)

between the bulk fluid and the biofilm, on the other hand, is dominated by advection and turbulent diffusion.[18] These concepts dominate all mathematical models today.

Due to the development of new tools for the study of biofilms, a different picture of the internal structure of the base film is emerging.[9,10,12] Figure 15.3 is an artist's conceptualization of the architecture of a biofilm based on the observations of several researchers.[9] Biofilms now appear to be nonuniform structures consisting of discrete cell clusters attached to each other and to the solid support with extracellular polymeric material.[12,34] The spaces between clusters form vertical and horizontal voids, with the vertical voids acting as pores and the horizontal voids acting as channels. As a result, biomass distribution within a biofilm is not uniform,[9,23] nor are physical factors such as porosity and density.[46] The cell clusters are microbial aggregates cemented with extracellular polymeric material, whereas the voids are open structures relatively free of it. The significance of the voids is that liquid can flow through them.[10] This has a profound effect on mass transfer in the biofilm because it suggests that it can occur by both diffusion and advection, with diffusion dominating in the cell clusters. However, because advection brings materials to the clusters, diffusion can occur from almost any direction into a cluster, rather than just from the liquid–biofilm interface. In addition, it also appears that the cell clusters have small conduits through them, adding another level of complexity to the biofilm.[44] Finally, many factors, such as the texture of the substratum, the nature of the flow past the biofilm, and the geometry of the bioreactor, influence the heterogeneity of the biofilm that develops.[15] These observations suggest that the commonly accepted use of a single transport parameter, such as an effective diffusion coefficient,

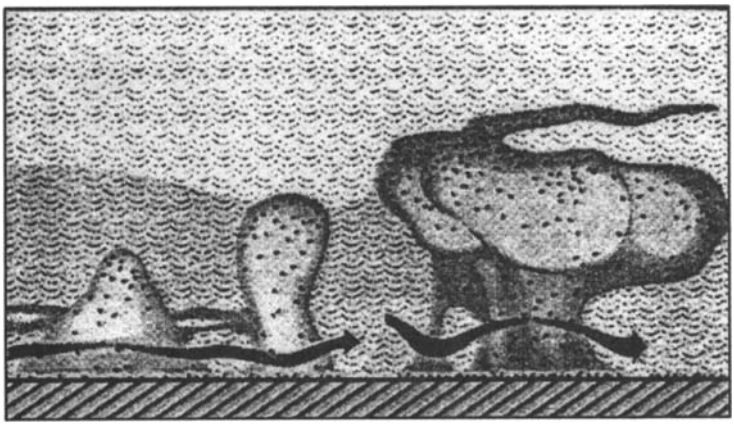


Figure 15.3 Artist's conceptualization of the architecture of a biofilm. (From J. W. Costerton, Z. Lewandowski, D. E. Caldwell, D. R. Korber, and H. M. Lappin-Scott, Microbial biofilms. *Annual Review of Microbiology* **49**:711–745, 1995. Copyright © Annual Reviews, Inc.; reprinted with permission.)

for describing the transport of substrate, electron acceptor, etc. within a biofilm is inadequate.[11,12] In fact, various researchers have shown that effective diffusion coefficients vary with biofilm depth,[33,46,49] which is consistent with changes in the structure of the biofilm. Nevertheless, because most current mathematical models for biofilm reactors assume that transport within a biofilm is by diffusion alone with a constant diffusion coefficient, that is the approach we will take herein. However, the reader should be aware of the limitations of such an approach. Recent mathematical models attempt to consider the variability in the diffusion coefficient,[41] suggesting that different approaches to modeling transport within the biofilm will be used in the future.

The conceptual models presented above are for a simple heterotrophic biofilm in which the bacteria are using a single electron donor with a single electron acceptor. However, just as heterotrophic and autotrophic bacteria can grow together in suspended growth bioreactors, they can also grow together in attached growth reactors. In this instance they have different electron donors (organic matter and ammonia-N), but compete for the same electron acceptor (oxygen). They also must compete for space in the biofilm. The assumed spatial arrangements of the competing species within the biofilm can take several forms in mathematical models.[18] However, the most realistic approach assumes that all types of bacteria are available for growth at any point within a biofilm, but that their ultimate distribution is determined by their competition for shared nutrients and space,[18,22,24,27,39,40] which is consistent with observation.[48] Although the mathematical models for this competition were developed before the advent of the conceptual model in the preceding paragraph, we will consider them because of the importance of the interactions between heterotrophs and autotrophs in attached growth reactors. Multispecies models have also been developed for methanogenic cultures containing three trophic levels.[37] However, because of space constraints, they will not be covered here.

The importance of competition for space in determining the ultimate distribution of competing species within a biofilm can be visualized by considering the

traditional conceptualization of a base biofilm. Consider first a single species biofilm. Because substrate can only move into the biofilm by diffusion, a substrate concentration gradient will exist through the biofilm as illustrated by Figure 15.4. This means that bacteria near the liquid–biofilm interface are growing faster than those in the interior. However, as bacteria in the interior grow, they occupy more space, pushing those that are closer to the liquid–biofilm interface further away from the solid support. In addition, all of the bacteria are subject to decay, regardless of their position in the biofilm, resulting in the accumulation of biomass debris. The net effect of both processes is to cause a migration of particles from the interior of the film to the exterior where surface shear forces remove them, allowing a biofilm of constant thickness to develop. Even for a single species biofilm, however, the distribution of active organisms will not be the same throughout the depth of the biofilm.[46] Rather, active biomass will predominate in the outer regions of the film and biomass debris in the inner regions, as shown by the simulation results in Figure 15.5.[39]

If we have two species that do not compete for any nutrient, but only for space, their ultimate distribution will depend upon their relative specific growth rates at any point within the biofilm. Consider two species, A and B, growing on different substrates, but sharing oxygen as the electron acceptor. Oxygen is assumed to be present in excess, so as not to limit either species. Species A has a higher maximum specific growth rate coefficient on its substrate than species B does on its substrate. Species A will dominate the outer regions and species B will dominate the inner regions, as shown by the simulation results in Figure 15.6.[27] Species B is confined to the inner regions because there the substrate concentration for species A will have been diminished sufficiently to allow species B to grow as fast as, or faster than, species A. When the two species compete for a resource, such as oxygen, the distribution of organisms can become even more complex, depending upon the relative K_S values for the shared resource, as well as the growth kinetics of each species on its individual substrate.

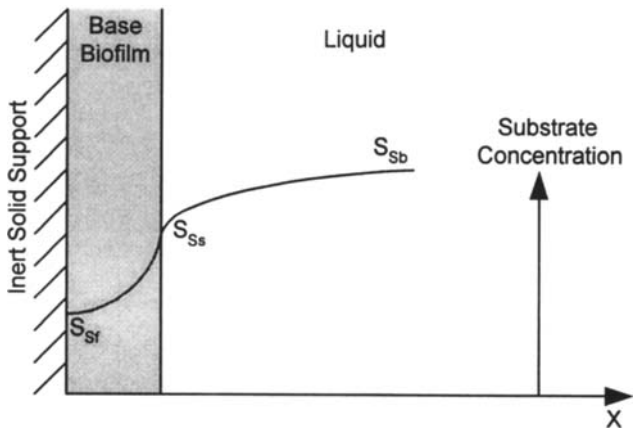


Figure 15.4 Traditional conceptualization of a base biofilm showing a typical concentration profile for a single limiting nutrient.

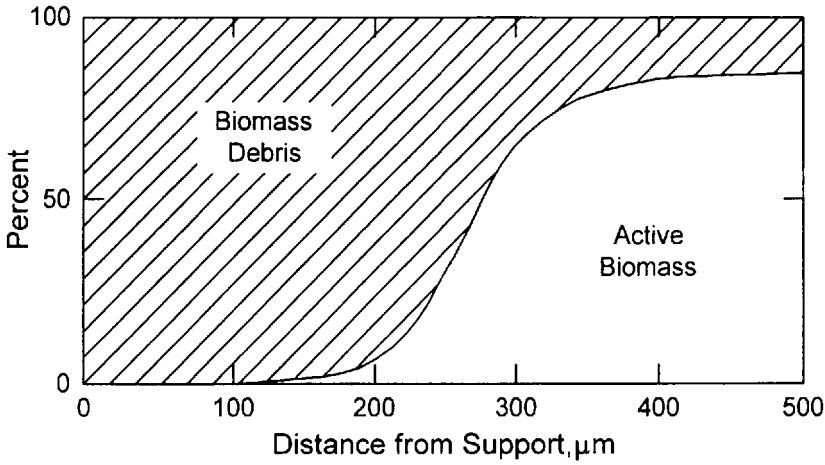


Figure 15.5 Simulation results showing the relative distribution of active biomass and biomass debris in a single species biofilm. (After Wanner and Gujer.[39])

Although the picture presented in the preceding paragraph is very logical given our current understanding of biofilms, the reader should recognize that it is only a conceptual model. It has not been proven by direct observation, although indirect observations suggest it is correct.[48] However, the new microscopy techniques that made it possible to develop the conceptual model of a biofilm illustrated in Figure 15.3, when combined with new molecular microbiology tools, make it possible to gather experimental evidence about the distribution of individual species in a multispecies biofilm. Such information will allow us to evaluate for the first time the qualitative accuracy of the multispecies biofilm conceptual model. Although we will

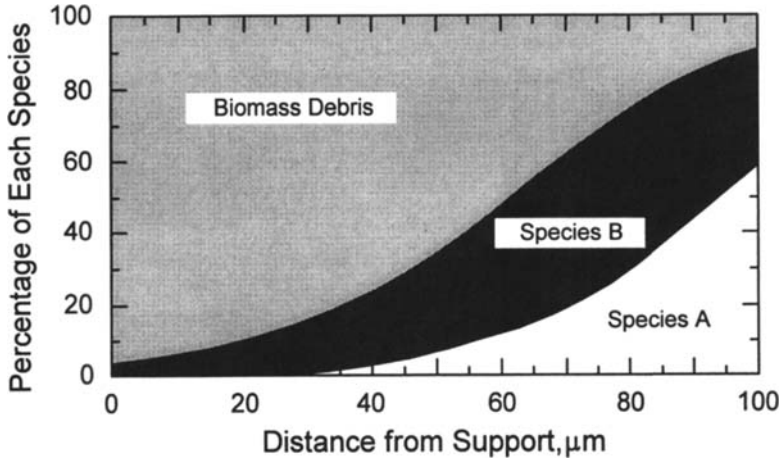


Figure 15.6 Simulation results showing the relative distributions of rapidly (Species A) and slowly (Species B) growing bacteria in a dual-species biofilm when their only shared resource is space. (After Rittmann and Manem.[27])

assume the validity of those concepts herein, the reader should be aware that future research may change our picture.

15.2 EFFECTS OF TRANSPORT LIMITATIONS

All current mathematical models for biofilms assume that the electron donor, the electron acceptor, and all nutrients are transported to the biomass within the biofilm by diffusional processes alone. In addition, consideration must also be given to the transport of those constituents from the bulk fluid to the biofilm. In this section, we will examine these processes and the techniques used to model them. This examination will be limited to transport of a single electron donor, i.e., the substrate, to one type of biomass. It will be assumed that the electron acceptor and all other nutrients are provided in sufficiently high concentration in the bulk liquid so as not to be limiting within the biofilm.

15.2.1 Mass Transfer to and Within a Biofilm

Consider a flat plate covered with a base biofilm. If this plate is placed into a substrate solution, the concentration of the substrate at the surface of the biofilm will be less than the concentration in the bulk of the fluid because of the substrate consumption by the microorganisms within the biofilm. Furthermore, because of that consumption, the substrate concentration will continue to drop with depth in the biofilm. In order for the consumption to continue, substrate must be transported from the bulk fluid to the liquid–biofilm interface by molecular and turbulent diffusion. It must also be transported within the biofilm. As discussed above, although both diffusion and advection are involved in internal transport, the phenomenon is modeled as if it were due to diffusion alone. Nevertheless, the net effect of these events is to cause a substrate concentration profile that looks something like the one in Figure 15.4. In this situation, the observed substrate consumption rate depends on the rate of mass transport external to and within the biofilm as well as on the true, intrinsic substrate consumption rate of the biomass, i.e., the true reaction rate without any mass transfer limitations. Consequently, if one were to observe the substrate consumption rate of a biofilm as a function of the substrate concentration in the bulk liquid, it would differ from the intrinsic relationship between substrate consumption rate and substrate concentration that could be measured when the microorganisms were dispersed throughout the liquid phase (thereby eliminating mass transfer effects). Thus, the effects of mass transfer obscure the true reaction rate relationship in a biofilm and any attempt to model the situation without incorporating the effects of mass transfer would be futile.

External mass transfer is typically modeled by idealizing the substrate concentration profile in the bulk liquid as shown in Figure 15.7. The variation in substrate concentration is restricted to a hypothetical stagnant liquid film of thickness L_s, through which substrate must be transported to reach the biofilm. As a consequence, the substrate concentration throughout the remaining fluid, i.e., the bulk liquid phase, is constant. All resistance to mass transfer from the bulk fluid to the biofilm is assumed to occur in the stagnant liquid film.

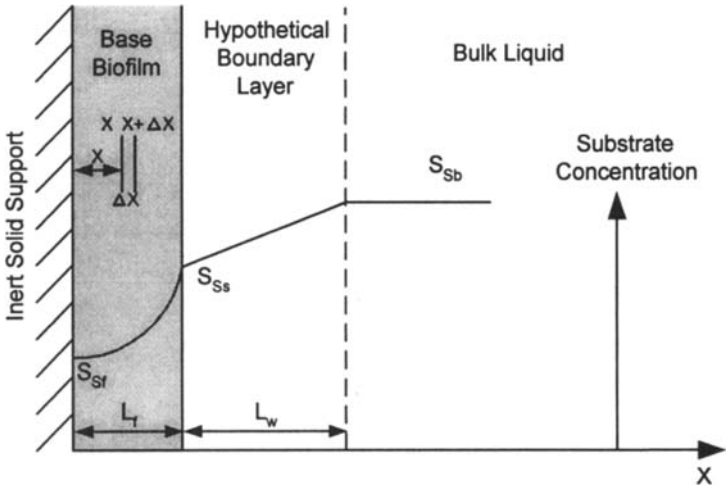


Figure 15.7 Traditional conceptualization of a base biofilm showing an idealized concentration profile for a single limiting nutrient.

Two approaches are commonly used to model external mass transfer. One assumes that transport across the liquid layer is by molecular diffusion, with diffusivity D_w. In that case, the thickness L_w is defined as the equivalent depth of liquid through which the actual mass transfer can be described by molecular diffusion alone. Consequently, the flux, J_S, or mass of substrate transported per unit area per unit time, is given by:

$$J_S = \frac{D_w}{L_w}(S_{Sb} - S_{Ss}) \tag{15.1}$$

Because the diffusivity is an intrinsic characteristic of the material being transported (the fluid is assumed to be water), L_w becomes the parameter that must be evaluated before Eq. 15.1 can be used to depict the rate of transport of the substrate to the biofilm. Its value must be deduced from Eq. 15.1 using measured fluxes coupled with known diffusivities and concentration gradients. The second approach employs a liquid phase mass transfer coefficient, k_L, that incorporates all of the effects of diffusive and advective mass transfer into one parameter. In that approach:

$$J_S = k_L(S_{Sb} - S_{Ss}) \tag{15.2}$$

The value of k_L must also be deduced from measured fluxes and concentration gradients. It is apparent from comparison of Eqs. 15.1 and 15.2 that:

$$k_L = \frac{D_w}{L_w} \tag{15.3}$$

Thus, measured values of k_L may be used to estimate L_w, and vice versa. The value of k_L (and L_w) will depend on the properties of the fluid (such as its viscosity, μ_w, and its density, ρ_w), the diffusivity of the substrate in the fluid, and the nature of the turbulence, which can be represented in part by the bulk fluid velocity past the biofilm, v. Figure 15.8[17] illustrates how that velocity influences the gradient in the

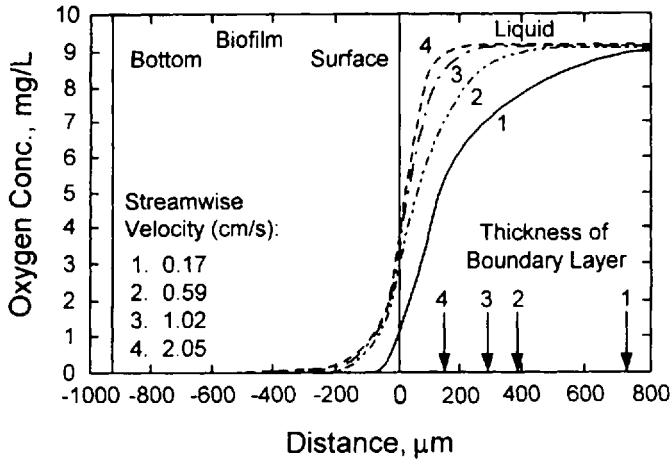


Figure 15.8 Effect of the fluid velocity past a biofilm on the thickness of the boundary layer for external mass transfer. (From T. C. Zhang and P. L. Bishop, Experimental determination of the dissolved oxygen boundary layer and mass transfer resistance near the fluid-biofilm interface. *Water Science and Technology* **30**(11):47–58, 1994. Copyright © Elsevier Science Ltd., reprinted with permission.)

bulk fluid. Here, the material being transported is oxygen, which is being used by the biofilm as it consumes the substrate. The numbered arrows in the figure show how the fluid velocity affects the thickness of the actual boundary layer, and thus illustrates the impact on L_w or k_L. Many relationships are available for relating k_L to the system characteristics. They are usually defined in terms of the Reynolds number ($v\rho_w d/\mu_w$) and the Schmidt number ($\mu_w/\rho_w D_w$) and are discussed in texts covering mass transfer (for example, see Weber and DiGiano[42]), as well as elsewhere.[5] Examination of common relationships reveals that many predict that k_L will increase with the square root of the fluid velocity. However, because flow situations in attached growth reactors are complex, it is usually necessary to determine experimentally how k_L depends on fluid velocity, or some factor affecting it, like mixing intensity in an agitated vessel or speed of rotation of a disc in a quiescent fluid.

Mass transfer within the biofilm is normally characterized by Fick's law, which for free diffusion in an aqueous solution is:

$$J_S = D_w \frac{dS_S}{dx} \tag{15.4}$$

in which D_w is a diffusivity and dS_S/dx is a concentration gradient. It is obvious from the previous discussion, however, that a biofilm is more complex than the situation of free diffusion for which Fick's law was developed. Therefore, the approach generally taken by modelers is to retain Fick's law as the governing equation, but to replace the diffusivity with an effective diffusivity, D_e:

$$J_S = D_e \frac{dS_S}{dx} \tag{15.5}$$

The effective diffusivity is usually smaller than the free diffusivity due to the pres-

ence of the extracellular polymer surrounding the cells in the biofilm. However, some researchers have measured effective diffusivities that are greater than the corresponding free diffusivities,[21] which is consistent with the presence of advection within the biofilm as discussed in Section 15.1. Thus, while Eq. 15.5 continues to be used to describe transport within the biofilm, D_e should be thought of as being due to more than diffusion alone.

Having determined how transport to and within the biofilm can be modeled, the next step is to combine transport with the reactions occurring within the biofilm to establish the relationship between the bulk substrate concentration and the rate of substrate removal by the biofilm. The resulting relationship can then be combined with the appropriate process model to simulate the performance of an attached growth bioreactor. Three techniques are in common use for combining the biofilm model with a process model: (1) the direct technique, (2) the effectiveness factor technique, and (3) the pseudoanalytical technique.[16] In the direct technique, the differential equations describing reaction within the biofilm are combined directly with the differential equations describing the bioreactor, giving a set of differential equations that must be solved by numerical methods. This technique is commonly employed for modeling systems involving multiple species carrying out multiple reactions, such as carbon oxidation, nitrification, and denitrification. The effectiveness factor technique pretends that the reaction rate at any point in a bioreactor can be defined by the intrinsic reaction rate expressed in terms of the bulk substrate concentration, multiplied by a factor (the effectiveness factor) that corrects for the effects of mass transport. Relationships between the effectiveness factor and the system characteristics are then coupled with the differential equations of the process model to allow simulation of bioreactor performance. The pseudoanalytical technique is similar to the effectiveness factor technique in concept, in that it develops a relationship between reaction rate and bulk substrate concentration that can then be used in the bioreactor model. In this case, however, the differential equations representing transport and reaction within the biofilm are solved numerically and the output is used to develop simplified general relationships for transport and reaction that can be solved analytically, thereby allowing them to be coupled with the process equations. In the following two sections we will investigate the effectiveness factor and pseudoanalytical techniques.

15.2.2 Modeling Transport and Reaction: Effectiveness Factor Approach[17]

Effectiveness Factor. The basic concept in the modeling of biofilms is that the flux of substrate to and through the liquid–biofilm interface must equal the overall utilization rate per unit of biofilm planar area. Because the local substrate utilization rate depends on the substrate concentration at that location, it is clear from Figure 15.7 that the utilization rates at various points in the biofilm will be different. The overall utilization rate by the biofilm must consider this by integrating the reaction rate over the biofilm depth. Because of this averaging and because of the requirement for substrate transport from the bulk fluid to the biofilm surface, any observed relationship between the overall substrate removal rate and the bulk substrate concentration will be different from the intrinsic reaction rate expression for substrate removal. It is often convenient, however, to express the substrate removal

rate as a function of the bulk substrate concentration, S_{Sb}, using a correction factor that takes into account the effects of transport. The correction factor is called the effectiveness factor[1] and is given the symbol η_e. If the Monod equation (Eq. 3.36) expresses the intrinsic relationship between the specific substrate removal rate, q, and the substrate concentration, this can be expressed as:

$$J_S = \eta_e \cdot X_{B,Hf} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \right) \tag{15.6}$$

in which $X_{B,Hf}$ is the mass of biomass per unit volume of biofilm and L_f is the biofilm thickness. Note that the substrate concentration is expressed in terms of the bulk substrate concentration, S_{Sb}. To use Eq. 15.6 in mass balance equations for various types of biofilm reactors, information must be available about the effectiveness factor, η_e.

The nature of η_e can be determined by writing the mass balance equation for substrate for a differential element within the biofilm (Figure 15.7) and solving it to obtain the actual flux of substrate into the biofilm. The actual flux can then be used with Eq. 15.6 to deduce the effects of the system kinetic and transport parameters on η_e. If the transport parameters include transport both to and within the biofilm, the effectiveness factor is called an overall effectiveness factor, denoted as η_{eo}. Assume that the biofilm in Figure 15.7 has reached a steady state in which it has a constant thickness L_f, a constant biomass concentration $X_{B,Hf}$, and uses substrate at a constant rate when exposed to a bulk substrate concentration S_{Sb}. A mass balance on substrate around a differential element in the biofilm yields:

$$-D_e A_\perp \left.\frac{dS_S}{dx}\right|_x + D_e A_\perp \left.\frac{dS_S}{dx}\right|_{x+\Delta x} - X_{B,Hf} \cdot A_\perp \cdot \Delta x \left(\frac{\hat{q}_H \cdot S_S}{K_S + S_S} \right) = 0 \tag{15.7}$$

where $A_\perp$ is the planar surface area normal to the direction of diffusion and x is the distance into the biofilm from the inert solid support. If D_e is constant, dividing both sides by $A_\perp$ and Δx, and taking the limit as Δx approaches zero yields:

$$D_e \frac{d^2 S_S}{dx^2} - X_{B,Hf} \left(\frac{\hat{q}_H \cdot S_S}{K_S + S_S} \right) = 0 \tag{15.8}$$

which must be solved with two boundary conditions, one at the biofilm-support interface (x = 0) and the other at the liquid–biofilm interface ($x = L_f$). At the biofilm-support interface there is no transfer of substrate because the solid support is inert and impermeable. Thus, the appropriate boundary condition is:

$$\frac{dS_S}{dx} = 0 \text{ at } x = 0 \tag{15.9}$$

As mentioned previously, having a permeable support is possible, but that case is not considered herein. It is apparent, however, that a different boundary condition would be required. The boundary condition at the liquid–biofilm interface is more complicated. It is written by recognizing that the substrate flux at that interface must equal the substrate flux through the stagnant liquid layer. Consequently, the appropriate boundary condition is:

$$J_S = D_e \frac{dS_S}{dx} = k_L(S_{Sb} - S_{Ss}) \text{ at } x = L_f \tag{15.10}$$

The development of an equation for the overall effectiveness factor, η_{eO}, requires the solution of Eq. 15.8 with Eqs. 15.9 and 15.10 as boundary conditions, giving substrate flux, J_S, and hence the substrate removal rate per unit of biofilm planar area, as a function of the bulk substrate concentration, S_{Sb}. The resulting relationship can then be used in Eq. 15.6 to obtain η_{eO}.

In order to develop a generalized relationship between the overall effectiveness factor and the physical and biochemical characteristics of a biofilm system, Fink et al.[13] solved Eq. 15.8 with its associated boundary conditions. They did this by a transformation of the two-point boundary value problem into an initial value problem. In doing so they used the following dimensionless quantities:

$$Bi = \frac{k_l \cdot L_f}{D_e} \tag{15.11}$$

$$\phi = \left(\frac{X_{B,H} \cdot \hat{q}_H \cdot L_f^2}{K_S \cdot D_e}\right)^{0.5} \tag{15.12}$$

$$\kappa = \frac{S_{Sb}}{K_S} \tag{15.13}$$

$$\phi_l = \phi \left(\frac{1}{1 + \kappa}\right)^{0.5} \tag{15.14}$$

Bi is a Sherwood number, called the Biot number.[4] Recalling that a diffusivity divided by a length is equivalent to a mass transfer coefficient, it can be seen that the term D_e/L_f represents an internal mass transfer coefficient. Thus, the Biot number is the ratio of the external mass transfer rate to the internal mass transfer rate. This means that when Bi is large, the external mass transfer coefficient is large relative to the internal coefficient so that all resistance to mass transfer can be considered to reside within the biofilm. In other words, the external resistance to mass transfer is negligible. This situation can arise when the flow rate past the biofilm is high. Conversely, when Bi is small, all of the resistance to mass transfer can be considered to be external to the biofilm. This situation can arise when the biofilm is very thin.

ϕ is a Thiele modulus. The physical significance of the Thiele modulus may be seen by squaring it, multiplying both the numerator and the denominator by $(S_{Sb} \cdot A_c)$, and rearranging.

$$\phi^2 = \left(\frac{X_{B,H} \cdot \hat{q}_H \cdot L_f^2}{K_S \cdot D_e}\right)\left(\frac{S_{Sb} \cdot A_c}{S_{Sb} \cdot A_c}\right)$$

$$= \frac{(A_c \cdot L_f \cdot X_{B,H})(\hat{q}_H/K_S)S_{Sb}}{A_c(D_e/L_f)S_{Sb}} \tag{15.15}$$

The term $(\hat{q}_H/K_S)S_{Sb}$ in the numerator is the first order approximation of the Monod equation (Eq. 3.38) for the specific substrate removal rate. For these first order kinetics, the maximum possible removal rate will occur when the substrate concentration surrounding the bacteria is the bulk substrate concentration. Consequently, the numerator represents the maximum possible first order reaction rate. Likewise, the maximum possible diffusion rate will occur when the gradient is maximized, so that the denominator represents a maximum diffusion rate within the biofilm. Therefore the Thiele modulus is the ratio of the maximum first order reaction rate to the

maximum diffusion rate. A large value of the Thiele modulus represents a situation in which the reaction rate is large relative to the diffusion rate. Such a situation is said to be diffusion limited. Conversely, a small value of ϕ represents a situation in which the diffusion rate is larger than the reaction rate. Such a situation is said to be reaction limited.

ϕ_f is a modified Thiele modulus. The purpose of the parameter κ is to take into consideration the deviation of the Monod equation from first order kinetics, which were the basis for the Thiele modulus. It will be recalled from Section 3.2.2 that when the substrate concentration is small relative to the half-saturation coefficient, the Monod equation simplifies to an expression that is first order with respect to substrate concentration, which is consistent with the basis of the Thiele modulus. Thus, when κ is small, the substrate removal rate behaves in a first order manner so that ϕ_f equals ϕ, and the Thiele modulus adequately describes the relative importance of reaction versus diffusion. On the other hand, when κ is large, the substrate concentration will be large relative to the half-saturation coefficient and the Monod equation will not behave in a first order manner. In that situation, the deviation from first-order kinetics will be large and ϕ_f will be smaller than ϕ.

The results of Fink et al.[13] giving the overall effectiveness factor as a function of these dimensionless groups are shown in Figure 15.9. These values of η_{eO} may be used to calculate the overall flux of substrate into a biofilm of thickness L_f containing microorganisms at concentration $X_{B,Hf}$ under conditions where both internal and external mass transfer resistances exist. Thus:

$$J_S = \eta_{eO} \cdot X_{B,Hf} \cdot L_f \left(\frac{\hat{q}_{Hf} \cdot S_{Sb}}{K_S + S_{Sb}} \right) \tag{15.16}$$

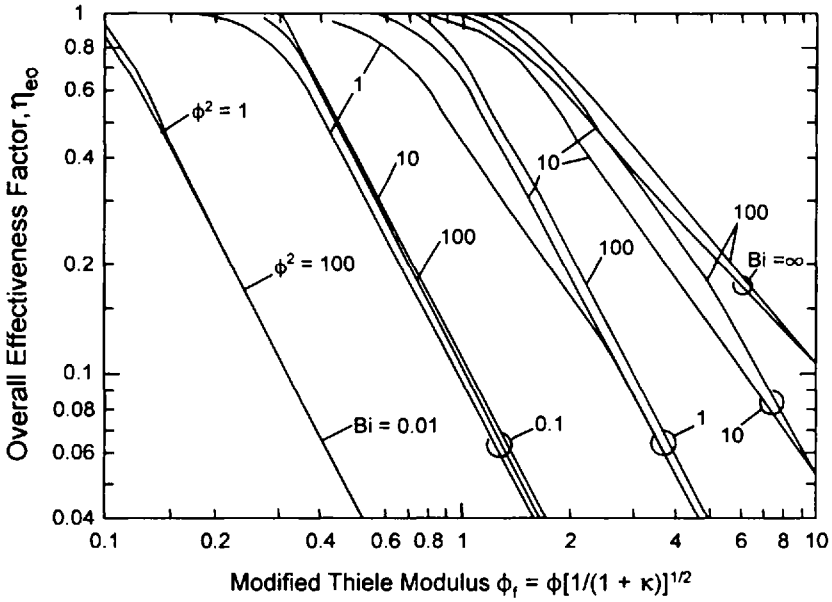


Figure 15.9 Overall effectiveness factor for Monod kinetics within a flat biofilm with external mass transfer resistance. (Adapted from Fink et al.[13])

The two curves for Bi = ∞ represent the case when the rate of external mass transfer is much higher than the rate of internal mass transfer, whereas the two curves for Bi = 0.01 represent the case where internal mass transfer is much more rapid than external mass transfer, due to a large external mass transfer resistance. Comparison of two groups of curves with different Bi values but the same ϕ_1 value demonstrates that the existence of external mass transfer resistance has a strong effect on the overall effectiveness factor. For example, when ϕ_1 = 1.0, a ten-fold decrease in Bi (from 1.0 to 0.1) results in almost a ten-fold reduction in the overall effectiveness factor. Moreover, comparison of curves with the same Biot number but different values of the Thiele modulus shows the relative importance of reaction versus diffusion. When Bi = 0.01, the external mass transfer coefficient is much smaller than the internal mass transfer coefficient. Consequently, external mass transfer controls and the effectiveness factor is influenced little by the relative importance of reaction versus diffusion. Thus, the value of the Thiele modulus, ϕ, has little effect. Under these circumstances, the effectiveness factor is dominated by external mass transfer resistance and is often called an external effectiveness factor and given the symbol η_{el}. On the other hand, when Bi = ∞, there is no external resistance to mass transfer and thus the relative importance of reaction versus diffusion has a strong impact on η_{el}, as evidenced by the strong impact of the Thiele modulus. Under circumstances when Bi = ∞, the effectiveness factor is often called an internal effectiveness factor and given the symbol η_{el}.

Although graphical representations like Figure 15.9 are convenient for some applications, for most occasions being able to determine the effectiveness factor analytically would be better. Consequently, it is common for investigators to develop functional relationships for the limited range of conditions they are interested in and this has been done for external, internal, and overall effectiveness factors. For example, for the case in which external mass transfer resistance is negligible, i.e., Bi is very large, Atkinson and Davies[2] developed both complex and simplified functional relationships for the internal effectiveness factor that agree quite well with the numerical results. In the interest of brevity, their equations will not be presented here. Rather, the reader is referred to other sources.[1,2]

Application of Effectiveness Factor. Equation 15.16 can be used with Figure 15.9 to determine the performance of a bioreactor containing a biofilm of known depth and biomass density. To illustrate how this is done, we will consider a continuous stirred tank reactor (CSTR) containing a biofilm.

Assume that steady-state conditions prevail over a reasonable time in a CSTR containing a solid surface covered by a biofilm of thickness L_f containing biomass at concentration $X_{B,Hf}$. To maintain a constant biofilm thickness, the cells generated by substrate consumption must be detached from the surface, dispersed throughout the liquid phase, and washed out in the bioreactor effluent. Because cells are in the bulk of the liquid as well as the biofilm, they are consuming substrate from both locations. Thus, the steady-state mass balance equation for substrate is:

$$F \cdot S_{S0} - F \cdot S_{Sb} - J_S A_s - q_H \cdot X_{B,Hb} \cdot V = 0 \qquad (15.17)$$

where J_S is the substrate consumption rate per unit surface area of biofilm, which is equivalent to the substrate flux, A_s is the biofilm surface area in the reactor, $q_H \cdot X_{B,Hb}$ is the substrate consumption rate per unit volume of reactor by dispersed bacteria, V is the bioreactor volume, F is the flow rate of influent and effluent, S_{S0} is the

influent substrate concentration, and S_{Sb} is the effluent or bulk liquid substrate concentration. Substitution of Eq. 15.16 for J_S and the Monod equation (Eq. 3.36) for q_H yields:

$$\frac{1}{\tau}(S_{SO} - S_{Sb}) - \eta_{eO} \cdot X_{B,H} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}}\right) \frac{A_s}{V} - X_{B,Hb} \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}}\right) = 0 \tag{15.18}$$

where τ is the HRT and $X_{B,Hb}$ is given by:

$$X_{B,Hb} = Y_{Hobs}(S_{SO} - S_{Sb}) \tag{15.19}$$

Equation 15.19 assumes that the influent contains no biomass. It is an approximation of the biomass concentration in a CSTR without biomass recycle in which Y_{Hobs} is an observed yield that accounts for decay in both the biofilm and the dispersed bacteria. The rationale for its use is that biomass can only arise from the utilization of substrate and that the biofilm is at steady state. Therefore, Eq. 15.18 may be rewritten as:

$$\frac{1}{\tau} = \eta_{eO} \cdot X_{B,H} \cdot L_f \left(\frac{A_s}{V}\right) \left[\frac{\hat{q}_H \cdot S_{Sb}}{(K_S + S_{Sb})(S_{SO} - S_{Sb})}\right] + \frac{Y_{Hobs} \cdot \hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \tag{15.20}$$

The solution of Eq. 15.20 to determine the value of S_{Sb} associated with a given hydraulic residence time (HRT) requires an iterative approach since the value of η_{eO} depends on S_{Sb}. Thus, a value must be assumed for S_{Sb} and the value of η_{eO} determined from Figure 15.9 or an associated approximate equation. Equation 15.20 can then be solved for S_{Sb} and the solution compared with the assumed value. The procedure is repeated until the assumed and calculated values of S_{Sb} agree. An iterative solution is not required when the bulk substrate concentration is fixed at a desired value and the bioreactor HRT or biofilm surface area per unit volume (A_s/V) required to achieve that value is being calculated. Under that circumstance, the effectiveness factor may be determined directly from Figure 15.9 for use in Eq. 15.20.

To show the effect of external mass transfer resistance on the performance of a CSTR containing a biofilm, Figure 15.10 was prepared by using Eq. 15.20 with Figure 15.9 and the parameter values in Table 15.1.[17] As shown in Table 15.1, the value of D_e used to generate Figure 15.10 was extremely large ($\approx D_w \times 10^4$) to remove all internal mass transfer resistance. Three curves are presented, one with a very large k_L value to represent the absence of external mass transfer resistance, and two with k_L values that might be encountered in practice. Each of these curves represents effects that might be caused by changes in the velocity of the fluid past the biofilm. Examination of Figure 15.10a shows that the effect of a decrease in k_L is to reduce the activity of the microbial film, thereby making the effluent substrate concentration greater than it would be in a bioreactor with less mass transfer resistance. Figure 15.10b shows the effect of the surface area of microbial film available in a bioreactor with an external mass transfer coefficient of 20 cm/hr. There it can be seen that reactors with more biofilm area remove more substrate, but that the effects diminish as the HRT is made larger because of the effects of the bulk substrate concentration on the effectiveness factor and substrate removal by suspended bio-

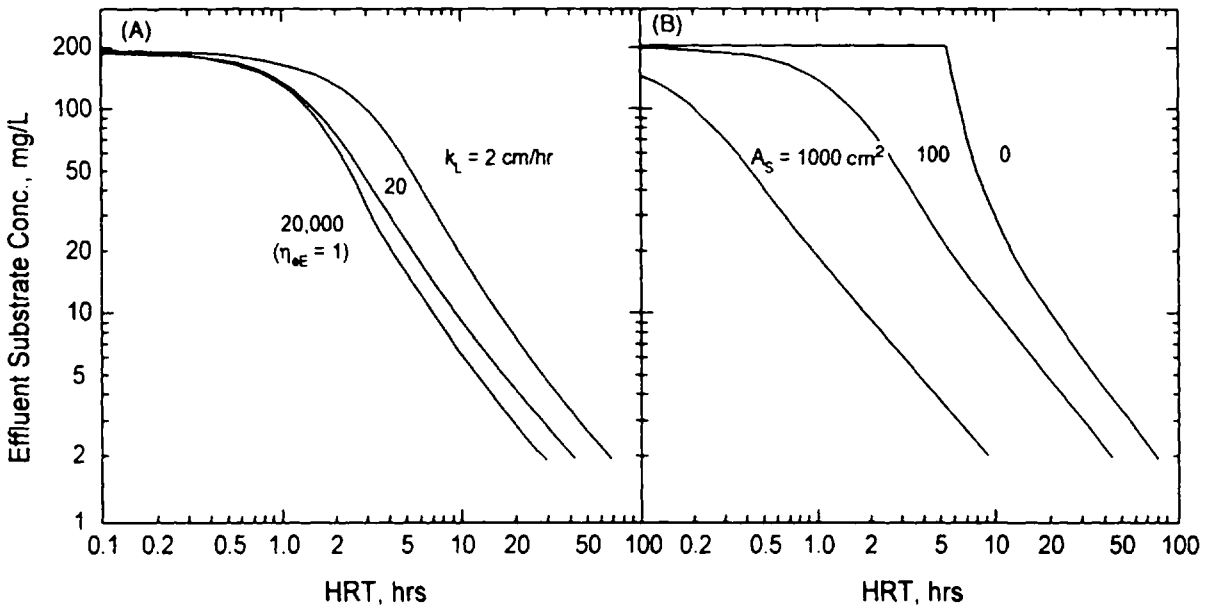


Figure 15.10 Effects of external mass transfer (a) and biofilm area (b) on the removal of soluble substrate by a CSTR containing a biofilm.

Table 15.1 Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used to Generate Figures 15.10 and 15.11[17]

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD · hr)	0.44
K_S	mg/L as COD	30
$Y_{H,obs}$	mg biomass COD/mg substrate COD	0.50
k_l	cm/hr	as noted in Fig. 15.10a 20 in Fig. 15.10 20,000 in Fig. 15.11
D_e	cm^2/hr	2,484 in Fig. 15.10 as noted in Fig. 15.11
D_w	cm^2/hr	0.2484
$X_{H,fil}$	mg biomass COD/cm^3	32
L_f	cm	0.05
V	cm^3	1,000
A_s	cm^2	100 in Figs. 15.10 and 15.11 as noted in Fig. 15.10b
S_{SO}	mg/L as COD	200

mass. Another important point to be seen in Figure 15.10b is that the presence of a biofilm prevents washout of the bioreactor. The curve labeled $A_s = 0$ represents a CSTR with no biofilm. Washout occurs at an HRT of a little more than five hours, making the substrate concentration equal to the influent. However, the presence of a biofilm allows substantial substrate removal at that HRT. Consequently, CSTRs with biofilms can remove substrate at HRTs well below those that would cause washout in a simple CSTR. Furthermore, the larger the surface area of biofilm, the shorter the HRT may be.

Figure 15.11 was prepared to show the effects of internal mass transfer resistance, again by using Eq. 15.20 with Figure 15.9 and the parameter values in Table 15.1. In this case, however, the value of k_l was made very large to remove all external mass transfer resistance. As in Figure 15.10a, three curves are presented. One curve has a very large D_e value to represent the absence of internal mass transfer resistance. It is essentially the same as the curve with a k_l value of 20,000 cm/hr in Figure 15.10a and can be used for comparing the relative effects of the two types of resistance. In other words, that curve represents a situation in which the overall effectiveness factor is 1.0 for all conditions. Examination of Figure 15.11 shows that the general effects of internal mass transfer resistance are similar to those of external mass transfer resistance, i.e., a reduction in the amount of substrate that can be removed by the biofilm. One difference that should be noted, however, is that whereas the external mass transfer resistance is subject to change by engineering factors such as the velocity of flow past the biofilm, the internal mass transfer resistance is not. Instead, it depends on the physical and chemical properties of the wastewater and the microorganisms in the system.

Many factors can influence the values of the mass transfer coefficients in biofilm systems. Unfortunately, space does not permit a discussion of them here. Never-

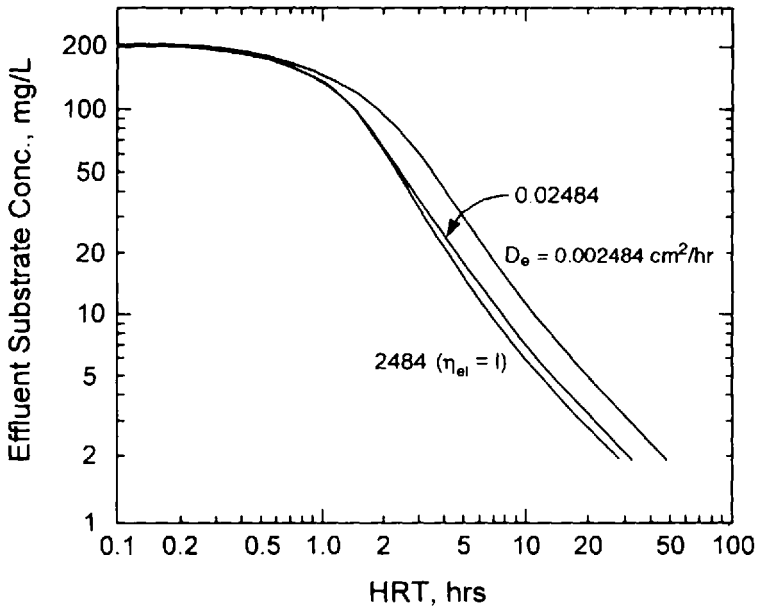


Figure 15.11 Effects of internal mass transfer on the removal of soluble substrate by a CSTR containing a biofilm.

theless, it is apparent from the above that accurate estimation of the coefficients is a requirement for proper application of mathematical models. Consequently, readers should consult the work of others for more information on this important topic.[8,18,33]

Equation 15.16 can also be used in the mass balance on substrate in a plug–flow bioreactor. The mass balance must be written around an infinitesimal section and the limit taken to get the differential equation describing the change of substrate concentration along the bioreactor length. In this case, because the substrate concentration varies along the bioreactor length, so will the overall effectiveness factor. Thus, solution of the equation requires a functional relationship for η_{eo} that can be used with numerical methods to solve the problem. Applications of this method are presented elsewhere[17] and the reader is referred there for more information. It is important to note, however, that the change in the effectiveness factor can be appreciable from one end of a plug–flow bioreactor to the other. Thus, assuming a constant effectiveness factor throughout the bioreactor is inappropriate.

Many advantages are associated with the use of effectiveness factors for modeling attached growth systems and they have found reasonably wide use, particularly to model fluidized bed systems.[16] Unfortunately, the effectiveness factor approach requires knowledge of the steady-state biofilm thickness. Thus, the biofilm and bioreactor models must be coupled with some means of obtaining the steady-state biofilm thickness. In addition, the effectiveness factor approach gets much more complex when one needs to consider dual nutrient limitation and competition for space by multiple bacterial types. These situations are more easily handled by an alternative approach. We will examine one, the pseudoanalytical approach.

15.2.3 Modeling Transport and Reaction: Pseudoanalytical Approach

Pseudoanalytical Approach. The pseudoanalytical approach uses simple algebraic expressions for the flux of substrate into a biofilm. Those expressions are based on an analysis of the results from the numerical solution of the differential equations describing transport and reaction in a biofilm. The availability of simple algebraic equations eliminates the need to repetitiously solve numerically a set of nonlinear differential equations while modeling the performance of a biofilm reactor. Pseudoanalytical solutions have been developed by several authors,[28,30,32,35] but the approach of Sácz and Rittmann[30,32] is particularly accurate.

A key characteristic of the pseudoanalytical approach is that it allows calculation of the bulk substrate concentration and the biofilm thickness for a steady-state biofilm. A steady-state biofilm is one in which the gains in biofilm mass due to biomass growth are just balanced by the losses in biofilm mass due to the combined effects of microbial decay within the biofilm and detachment by shear at the liquid–biofilm interface.[28] Decay is treated in the traditional manner as presented in Chapter 5. Because of the balance between growth and loss, the biofilm attains a uniform thickness, L_f. For a purely heterotrophic biofilm, that thickness is given by:[28]

$$L_f = \frac{J_S Y_H}{(b_H + b_D)X_{B,H}} \tag{15.21}$$

where b_D is the loss coefficient due to detachment caused by surface shear and the other parameters have their usual meaning. The value of the detachment coefficient varies with the shear stress on the biofilm, which depends on the hydrodynamic regime surrounding the biofilm.[26,38]

An important characteristic of a steady-state biofilm is the existence of a minimum bulk substrate concentration below which the biofilm cannot be maintained.[28] If the bulk substrate concentration is below that value, growth cannot occur rapidly enough to replace the losses to decay and detachment and the biofilm will decrease in thickness until it ceases to exist. The minimum bulk substrate concentration, S_{Sbmin}, is given by:[28]

$$S_{Sbmin} = \frac{K_S(b_H + b_D)}{Y_H \cdot \hat{q}_H - (b_H + b_D)} \tag{15.22}$$

Examination of Eq. 15.22 reveals that it is analogous to Eq. 5.14, the minimum attainable substrate concentration in a CSTR. This is because both represent the substrate concentration required to drive the growth reactions at a rate that will just balance loss by decay (and detachment, in the case of the biofilm). Since S_{Sbmin} is determined solely by parameters that depend on the biomass and substrate ($\hat{q}_H$, K_S, Y_H, and b_H) and the fluid regime (b_D), it takes on special significance as a parameter in the pseudoanalytical approach.

The equations upon which the pseudoanalytical solution of steady-state biofilm kinetics is based differ somewhat from those used to develop the effectiveness factor approach and are the result of the necessity to compute the biofilm thickness, L_f. They are:[30]

$$D_e \frac{d^2S_S}{dx^2} - X_{B,H}\left(\frac{\hat{q}_H S_S}{K_S + S_S}\right) = 0 \tag{15.23}$$

$$\frac{dS}{dx} = 0 \text{ at } x = 0 \tag{15.24}$$

$$S_S = S_{Ss} \text{ at } x = L_f \tag{15.25}$$

$$\frac{dL_f}{dt} = \int_0^{L_f} \left[\frac{Y_H \cdot \hat{q}_H \cdot S_S}{K_S + S_S} - (b_H + b_D) \right] dx \tag{15.26}$$

$$J_S = D_e \left. \frac{dS_S}{dx} \right|_{x=L_f} \tag{15.27}$$

$$S_{Sb} = S_{Ss} + \frac{J_S}{k_l} \tag{15.28}$$

where t is time. All other symbols are as defined previously. Equation 15.28 is just a rearranged form of Eq. 15.2, the flux across the hypothetical boundary layer. After the introduction of several dimensionless variables, Sáez and Rittmann[30,32] solved Eqs. 15.23 through 15.28 using the numerical method of orthogonal collocation.[34] This was done for 500 initial conditions, covering the entire region of feasible solutions.[32] The output was then used to develop the pseudoanalytical solution.

The pseudoanalytical solution is based on the flux into a deep biofilm, which is defined as one in which the substrate concentration at the biofilm-support interface is zero.[19] The reason for using a deep biofilm as the reference case is because the dimensionless flux into a deep biofilm, $J^*_{S,deep}$ can be calculated analytically with:[19]

$$J^*_{S,deep} = \{2[S^*_{Ss} - \ln(1 + S^*_{Ss})]\}^{0.5} \tag{15.29}$$

where S^*_{Ss} is the dimensionless substrate concentration at the liquid–biofilm interface:

$$S^*_{Ss} = \frac{S_{Ss}}{K_S} \tag{15.30}$$

Thus, for any value of S_{Ss}, $J^*_{S,deep}$ can be calculated. Once that has been done, the dimensionless flux into an actual biofilm with that value of S_{Ss}, J^*_S, can be computed as some function of $J^*_{S,deep}$:

$$J^*_S = \xi J^*_{S,deep} \tag{15.31}$$

provided an expression is available for ξ. The dimensionless flux into the actual biofilm, J^*_S, is defined as:

$$J^*_S = \frac{J_S}{(K_S \cdot \hat{q}_H \cdot X_{B,H} \cdot D_e)^{0.5}} \tag{15.32}$$

Consequently, once the dimensionless flux has been determined from Eq. 15.31, the actual flux, J_S, associated with the liquid–biofilm interface substrate concentration, S_{Ss}, can be calculated with Eq. 15.32 since all of the parameters in the relationship are known.

In developing the pseudoanalytical approach, Sáez and Rittmann[30] defined a new dimensionless group, S^*_{Sbmin}, the dimensionless minimum bulk substrate concentration. Its value is given by:

$$S^*_{Sbmin} = \frac{S_{Sbmin}}{K_S} = \frac{b_H + b_D}{Y_H \cdot \hat{q}_H - (b_H + b_D)} \tag{15.33}$$

S^*_{Sbmin} is important because of its physical significance.[19] Recognition of the fact that $Y_H \cdot \hat{q}$ is equal to $\hat{\mu}$ (Eq. 3.44), suggests that the value of S^*_{Sbmin} is an indication of the relative importance of biomass loss (by decay and detachment) and biomass growth. A large value of S^*_{Sbmin} (>1) implies that the maximum specific growth rate is not much larger than the specific loss rate of biomass from the biofilm, suggesting that the biofilm may be difficult to maintain. A small value (<1), on the other hand, suggests a potentially high net growth rate relative to losses, thereby making the biofilm easy to maintain. It is important to recognize that the value of b_D, the specific detachment coefficient, depends on the flow velocity past the face of the biofilm, and thus is under engineering control. Consequently, the term S^*_{Sbmin} represents both biological and physical factors. Because the significance of this term to the pseudoanalytical approach is similar to the significance of the dimensionless groups in the effectiveness factor approach, we think that it should be a named dimensionless group. Therefore, we propose and use the name Rittmann number, with the symbol Ri:

$$Ri = \frac{b_H + b_D}{Y_H \cdot \hat{q}_H - (b_H + b_D)} \tag{15.34}$$

Thus, the Rittmann number is the ratio of the specific loss rate of biomass from a biofilm to the net potential growth rate. As stated above, a large value means that a biofilm will be difficult to maintain, whereas a small value means that it will be easy to maintain.

Examination of the results from the 500 conditions studied by Sáez and Rittmann,[32] revealed that ξ could be adequately represented by:[32]

$$\xi = \tanh\left[\alpha'\left(\frac{S^*_{Ss}}{Ri} - 1\right)^{\beta'}\right] \text{ where} \tag{15.35}$$

$$\alpha' = 1.5557 - 0.4117 \tanh[\log_{10}Ri] \tag{15.36}$$

and

$$\beta' = 0.5035 - 0.0257 \tanh[\log_{10}Ri] \tag{15.37}$$

Examination of Eqs. 15.35 through 15.37 shows that they all depend upon the Rittmann number, showing why it is an important parameter in the pseudoanalytical solution. Sáez and Rittmann[32] examined the accuracy of the pseudoanalytical solution technique by using it to compute the flux into the biofilm for each of the 500 initial conditions and comparing the values with those obtained with the full numerical solution for each of the same conditions. The error depended somewhat on the value of the Rittmann number, but gave a standard error on the order of 2%, with 2.6% being the greatest observed. Thus, the pseudoanalytical solution is quite accurate.

To summarize, the calculation of the substrate flux associated with a given substrate concentration at the liquid–biofilm interface proceeds in the following manner. First, that concentration is put into dimensionless form with Eq. 15.30, allowing calculation of the dimensionless flux into a deep biofilm, $J^*_{S,deep}$, with Eq. 15.29. Then the Rittmann number is calculated with Eq. 15.34, allowing the parameter ξ to be determined with Eqs. 15.35–15.37. Once ξ is known, the dimensionless flux into the biofilm can be calculated with Eq. 15.31, allowing the actual flux to be determined with Eq. 15.32.

Application of Pseudoanalytical Approach. The pseudoanalytical approach allows direct calculation of the flux into a steady-state biofilm associated with a given liquid–biofilm interface substrate concentration. However, what we really want to know is the flux associated with a given bulk substrate concentration, since it is the concentration that can be measured. The pseudoanalytical approach can be used to calculate it, as well as the biofilm's thickness, by combining Eq. 15.28 (in dimensionless form) with Eqs. 15.29, 15.31, and 15.35, giving:

$$S_{Sb}^* = S_{S\ell}^* + \frac{\tanh\left[\alpha'\left(\frac{S_{S\ell}^*}{Ri} - 1\right)^{\beta'}\right]\{2[S_{S\ell}^* - \ln(1 + S_{S\ell}^*)]\}^{0.5}}{k_f^*} \tag{15.38}$$

where the Rittmann number, Ri, the dimensionless bulk substrate concentration, S_{Sb}^*, and the dimensionless external mass transfer coefficient, k_f^*, are calculated with Eqs. 15.34, 15.39, and 15.40, respectively.

$$S_{Sb}^* = \frac{S_{Sb}}{K_S} \tag{15.39}$$

$$k_f^* = k_L\left[\frac{K_S}{\hat{q}_H \cdot X_{B,H} \cdot D_c}\right]^{0.5} \tag{15.40}$$

After calculating α' and β' with Eqs. 15.36 and 15.37, respectively, Eq. 15.38 can be solved iteratively for $S_{S\ell}^*$ using Newton's root-finding technique, which converges rapidly.[30] Sáez and Rittmann[30] recommend using Ri + 10^{-6} as the initial guess for $S_{S\ell}^*$. Once $S_{S\ell}^*$ is known, $S_{S\ell}$ can be calculated from its definition (Eq. 15.30), allowing the flux into the biofilm, J_S, to be calculated from Eq. 15.2. Finally, the biofilm thickness, L_f, can be calculated with Eq. 15.21.

Example 15.2.3.1 (Adapted from Sáez and Rittmann.[30])

A steady-state biofilm, described by the kinetic and stoichiometric coefficients given in Table E15.1, exists in an environment where the bulk substrate concentration is 0.5 mg/L. Determine the substrate flux into the biofilm, i.e., the substrate utilization rate per unit area of biofilm, and the biofilm thickness.

Because k_L, D_c, and $X_{B,H}$ all use cm in their units, S_{Sb} and K_S should be expressed as mg/cm^3 for consistency. Thus, S_{Sb} has a value of 0.0005 mg/cm^3 and K_S has a value of 0.01 mg/cm^3. The values of S_{Sb}^*, k_f^*, and Ri are calculated with Eqs. 15.39, 15.40, and 15.34, respectively, yielding:

$$S_{Sb}^* = \frac{0.0005}{0.01} = 0.05$$

$$k_f^* = (3.33)\left[\frac{0.01}{(0.2667)(50)(0.0266)}\right]^{0.5} = 0.559$$

$$Ri = \frac{0.0025 + 0.0017}{(0.625)(0.2667) - (0.0025 + 0.0017)} = 0.025641$$

Equations 15.36 and 15.37 are used to calculate the parameters α' and β'.

$$\alpha' = 1.5557 - 0.4117 \tanh[\log_{10}0.025641] = 1.9346$$

$$\beta' = 0.5035 - 0.0257 \tanh[\log_{10}0.025641] = 0.5272$$

Table E15.1 Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used in Example 15.2.3.1[30]

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD · hr)	0.2667
K_S	mg/L as COD	10
Y_H	mg biomass COD/mg substrate COD	0.625
b_H	hr^{-1}	0.0025
$X_{B,H}$	mg biomass COD/cm^3	50
k_l	cm/hr	3.333
D_e	cm^2/hr	0.02667
b_D	hr^{-1}	0.0017

Equation 15.38 is then solved iteratively, giving a value of S_S^* of 0.027577. The dimensionless substrate concentration at the liquid-biofilm interface may be transformed into the physical domain by using the definition of the dimensionless variable, Eq. 15.30.

$$S_{S_b} = (0.01)(0.027577) = 0.00027577 \text{ mg COD/cm}^3 = 0.2758 \text{ mg COD/L}$$

The flux of substrate into the biofilm can then be calculated with Eq. 15.2.

$$J_S = 3.333(0.0005 - 0.0002758) = 0.000747 \text{ mg COD/(cm}^2 \cdot \text{hr)}$$

Finally, the steady-state biofilm thickness can be calculated with Eq. 15.21.

$$L_f = \frac{(0.000747)(0.625)}{(0.0025 + 0.0017)(50)} = 0.00222 \text{ cm} = 22.2 \ \mu\text{m}$$

Once the flux of substrate into a biofilm is known, it is a simple matter to calculate the biofilm area required to achieve the desired bulk substrate concentration in a CSTR, provided that all substrate removal is due to the biofilm alone. In other words, the contribution of suspended biomass to substrate removal is assumed to be negligible, which is a reasonable assumption given the HRTs normally associated with attached growth reactors. Under that condition, a steady-state mass balance on substrate gives:

$$A_s = \frac{F(S_{S0} - S_{Sb})}{J_S} \tag{15.41}$$

Biofilm media is generally characterized by its specific surface area, i.e., the surface area per unit of bioreactor volume, a_s:

$$a_s = \frac{A_s}{V} \tag{15.42}$$

where V is the volume of the bioreactor actually containing the media. Thus, once the total required surface area is known, calculating the bioreactor volume required for a given media is easy.

Example 15.2.3.2

A synthetic wastewater with a biodegradable chemical oxygen demand (COD) of 10 mg/L (0.010 mg/cm^3) is flowing at a rate of 1.0 L/hr (1000 cm^3/hr) into a CSTR containing a biofilm media with a specific surface area of 90 m^2/m^3 (0.90 cm^2/cm^3). The wastewater, the bioreactor, and the associated biofilm have the characteristics listed in Table E15.1. What total surface area of biofilm would be required to reduce the biodegradable COD to 0.5 mg/L (0.0005 mg/cm^3)? What bioreactor volume is required to house the media?

Because the bioreactor is a CSTR, the bulk substrate concentration is 0.5 mg COD/L throughout, and thus the substrate flux into the biofilm is the same as in Example 15.2.3.1. As a result, the total biofilm area can be calculated with Eq. 15.41:

$$A_s = \frac{(1{,}000)(0.010 - 0.0005)}{0.000747} = 12{,}700 \text{ cm}^2$$

The bioreactor volume can then be calculated from a rearranged form of Eq. 15.42:

$$V = \frac{12{,}700}{0.90} = 14{,}100 \text{ cm}^3 = 14.1 \text{ L}$$

A simple CSTR without biomass recycle or biofilm would require a volume of 184 L to achieve the same effluent substrate concentration. Thus, the benefit of the biofilm is apparent.

As with other systems we have studied, biofilm systems also benefit from being housed in a plug–flow or tanks-in-series configuration. The pseudoanalytical approach can also be used to determine the performance of such systems by considering the plug–flow systems to be a series of completely mixed compartments.[19] Although the number of compartments required to adequately simulate a plug–flow system depends on its hydraulic characteristics, considering the bioreactor to contain six is usually adequate.[19] Regardless of the situation considered, a number of compartments in series should be assumed, as well as the volume of each compartment. Starting with the last compartment and the desired effluent substrate concentration, it is possible to calculate the flux into the biofilm using the procedure in Example 15.2.3.1. Once that flux is known, it can be used to determine the influent substrate concentration that could be treated with a given media by using the procedure in Example 15.2.3.2. Since the influent to the last compartment is the effluent from the next-to-last compartment, the procedure can be repeated to determine the concentration of organic matter in the influent to the next-to-last compartment, etc. The procedure is repeated until the influent to the first compartment is calculated. If it is equal to the known influent concentration, the system is the correct size. If it is larger than the known concentration, the system is larger than required and the calculations should be repeated with a smaller volume for each compartment. (Alternatively, the excess volume may be acceptable as a factor of safety.) If the calculated influent substrate

concentration is less than the known value, the system is too small and the calculations must be repeated with larger compartment volumes.

While the procedure in the preceding paragraph is straightforward, it is tedious because of the need to solve Eq. 15.38 repeatedly. Consequently, Heath et al.[19,20] have proposed an even simpler method based on normalized loading curves.

Normalized Loading Curves. A graphical representation forms the basis for the normalized loading curve approach to biofilm reactor analysis and design. Heath et al.[19,20] used the pseudoanalytical approach to solve for the substrate flux associated with various bulk substrate concentrations. This was done for many conditions and the results were presented in generalized form by normalizing the bulk substrate concentration relative to S_{Sbmin} and the flux relative to a reference flux, J_{SR}. The curves are called loading curves because the flux to a biofilm is approximately equal to the rate of substrate input per unit of biofilm, i.e., the loading. The reference flux is the minimum flux just required to maintain a steady-state biofilm that is deep. For computational purposes, it is defined as the flux resulting when $\xi = 0.99$ in Eq. 15.31.[19] The reference flux depends on the value of the Rittmann number and can be presented in generalized form by a plot of J^*_{SR}/Ri versus Ri, where J_{SR} is made dimensionless in the same way as J_S, as indicated in Eq. 15.32.[5] Figure 15.12 shows such a plot.[32] It can be entered with a known value of the Rittmann number, giving the corresponding value of J^*_{SR}, which can be put back into the physical domain, i.e., J_{SR}, by using Eq. 15.32.

Normalized loading curves were plotted for fixed values of Ri and k_f^*.[20] The Rittmann number is used as a parameter because of its fundamental importance, as discussed previously. The dimensionless external mass transfer coefficient, k_f^*, was chosen as the second parameter because it represents the importance of external mass transfer resistance in the performance of the biofilm. As can be seen in Eq. 15.40, if k_f^* is large (>10), external mass transfer resistance is of little importance relative to reaction and internal mass transfer. Conversely, when k_f^* is small (<1) external mass transport is likely to play an important role in biofilm performance. Figures 15.13 through 15.17 present normalized loading curves for Ri values of 0.01, 0.1, 1, 10, and 100, respectively.[20] It should be noted that S_{Sb}/S_{Sbmin} is equivalent to S^*_{Sb}/Ri and J_S/J_{SR} is equivalent to J^*_S/J^*_{SR}.

The use of the normalized loading plots is very straightforward for a biofilm in a completely mixed bioreactor or bioreactor compartment. For a given situation, the value of Ri is calculated with Eq. 15.34 and Figure 15.12 is used to determine the dimensionless reference flux, J^*_{SR}. The family of normalized loading curves corresponding most closely to the value of Ri is then used, with the particular curve depending on the value of k_f^*. If no curve corresponds exactly to the Ri and k_f^* values for the system, then interpolation can be used.[19] Using the desired bulk substrate concentration, S_{Sb}, the value of S_{Sb}/S_{Sbmin} can be calculated and used to determine J_S/J_{SR} (or J^*_S/J^*_{SR}) from the appropriate curve. Since the value of J^*_{SR} is known, the value of J_S is fixed. Equation 15.41 can then be used to calculate the required biofilm area, A_S, and Eq. 15.42 to calculate the associated bioreactor volume. This gives a direct solution for a single CSTR, or allows the iterative procedure discussed previously to be used for a plug-flow system approximated as compartments in series.

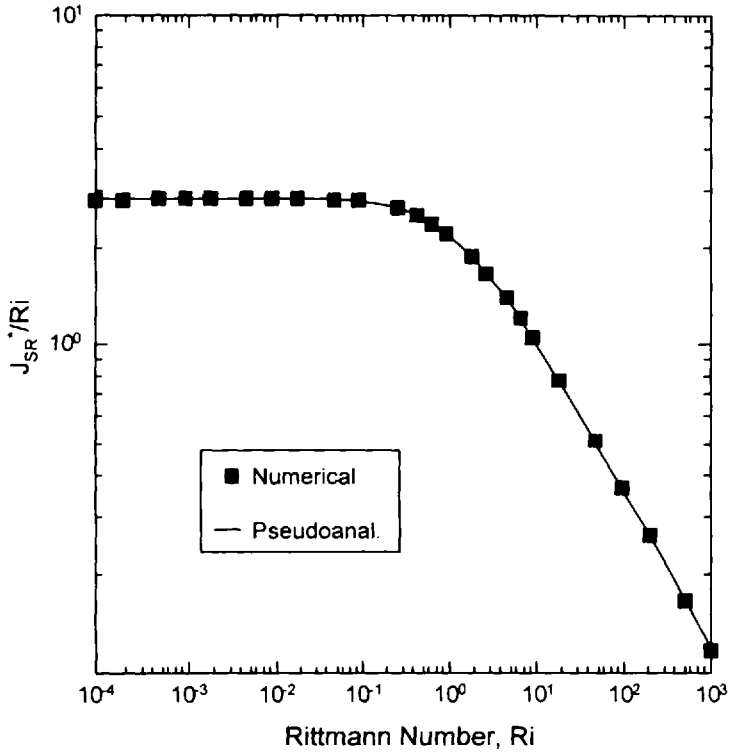


Figure 15.12 Curve for the determination of the dimensionless reference flux, J_{SR}^*, from the Rittmann number. (From P. B. Sáez and B. E. Rittmann, Accurate pseudoanalytical solution for steady-state biofilms. *Biotechnology and Bioengineering* **39**:790–793, 1992. Copyright © John Wiley & Sons; reprinted with permission.)

Example 15.2.3.3

A synthetic wastewater with a biodegradable COD of 10 mg/L (0.010 mg/cm^3) is flowing at a rate of 1.0 L/hr (1000 cm^3/hr) into a CSTR containing a biofilm media with a specific surface area of 90 m^2/m^3 (0.90 cm^2/cm^3). The wastewater, the bioreactor, and the associated biofilm have the characteristics listed in Table E15.2. What total surface area of biofilm would be required to reduce the biodegradable COD to 1.5 mg/L (0.0015 mg/cm^3)? What bioreactor volume is required to house the media?

The first task is to determine the dimensionless reference flux, J_{SR}^* from Figure 15.12. This requires calculation of Ri with Eq. 15.34.

$$Ri = \frac{0.0075 + 0.0076}{(0.625)(0.2667) - (0.0075 + 0.0076)} = 0.10$$

Entering Figure 15.12 with Ri = 0.10 gives a value of J_{SR}^*/Ri of 2.8, thereby fixing J_{SR}^* at 0.28.

The next task is to determine J_S from the appropriate normalized loading curve. This requires calculation of S_{Sb}/S_{Sbmin} and k_f^*. S_{Sb} is given as 1.5 mg COD/L. S_{Sbmin} can be calculated with Eq. 15.22, or by multiplying Ri by K_S (See Eqs.

Table E15.2 Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used in Example 15.2.3.3

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD · hr)	0.2667
K_S	mg/L as COD	10
Y_H	mg biomass COD/mg substrate COD	0.625
b_H	hr^{-1}	0.0075
$X_{H,th}$	mg biomass COD/cm^3	50
k_l	cm/hr	6.00
D_e	cm^2/hr	0.02667
b_D	hr^{-1}	0.0076

15.33 and 15.34), which is 10 mg COD/L, giving S_{Sbmin} = 1.0 mg COD/L. Thus, S_{Sb}/S_{Sbmin} is 1.5. The value of k_l^* is calculated with Eq. 15.40.

$$k_l^* = 6.00\left[\frac{0.01}{(10.2667)(50)(0.02667)}\right]^{0.5} = 1.0$$

Thus, the appropriate curve to use to obtain J_S is the curve for k_l^* = 1.0 in Figure 15.14. Entering that curve with S_{Sb}/S_{Sbmin} = 1.5 gives J_S/J_{SR} (or J_S^*/J_{SR}^*) = 0.2. Since J_{SR}^* is 0.28, J_S^* is 0.056. The value of J_S can then be determined from Eq. 15.32.

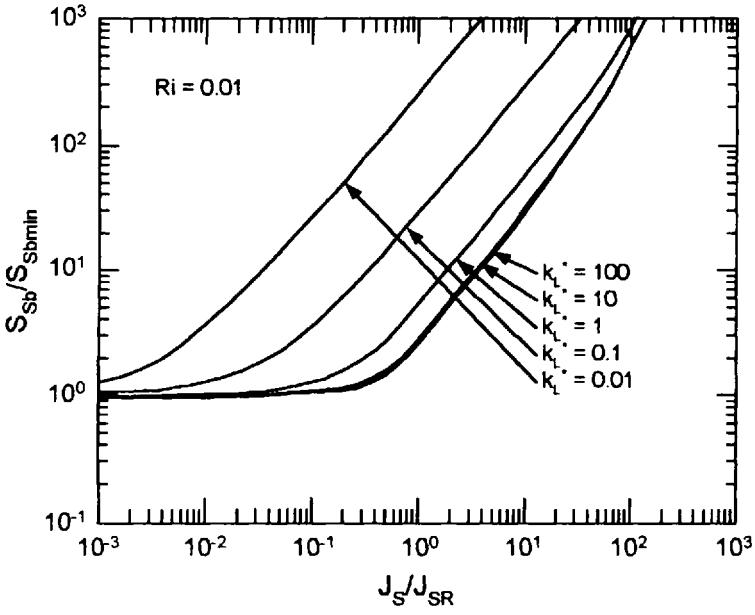


Figure 15.13 Normalized loading curves for Ri = 0.01. (From M. S. Heath, S. A. Wirtel, B. E. Rittmann, and D. R. Noguera, Closure to discussion of "Simplified design of biofilm processes using normalized loading curves." *Research Journal, Water Pollution Control Federation* **63**:91–92, 1991. Copyright © Water Environment Federation; reprinted with permission.)

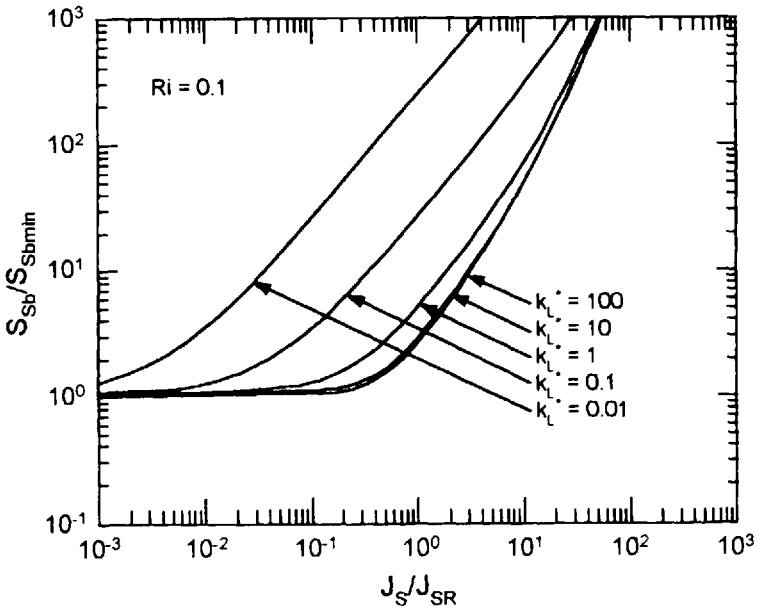


Figure 15.14 Normalized loading curves for Ri = 0.1. (From M. S. Heath, S. A. Wirtel, B. E. Rittmann, and D. R. Noguera, Closure to discussion of 'Simplified design of biofilm processes using normalized loading curves'. *Research Journal, Water Pollution Control Federation* **63**:91–92, 1991. Copyright © Water Environment Federation; reprinted with permission.)

$$J_s = 0.056[(0.010)(0.2667)(50)(0.02667)]^{1/2} = 0.00334 \text{ mg COD/(cm}^2 \cdot \text{hr)}$$

Once the flux is known, the required biofilm surface area can be calculated with Eq. 15.4.

$$A_s = \frac{(1{,}000)(0.010 - 0.0015)}{0.00334} = 2{,}545 \text{ cm}^2$$

The bioreactor volume can then be calculated from a rearranged form of Eq. 15.42:

$$V = \frac{2{,}545}{0.90} = 2{,}840 \text{ cm}^3 = 2.83 \text{ L}$$

A simple CSTR without biomass recycle or biofilm would require a volume of 70 L to achieve the same effluent substrate concentration. As in Example 15.2.3.2, the benefit of the biofilm is apparent.

Parameter Estimation. Before the pseudoanalytical approach can be used, values must be available for the parameters in the model. Since the Monod kinetic parameters represent the kinetics of the biomass in the absence of mass transfer limitations, theoretically, one could use kinetic parameters determined with suspended growth cultures. However, the growth of biomass in a biofilm alters its physiological state.[9] Consequently, the parameters should be determined for biomass

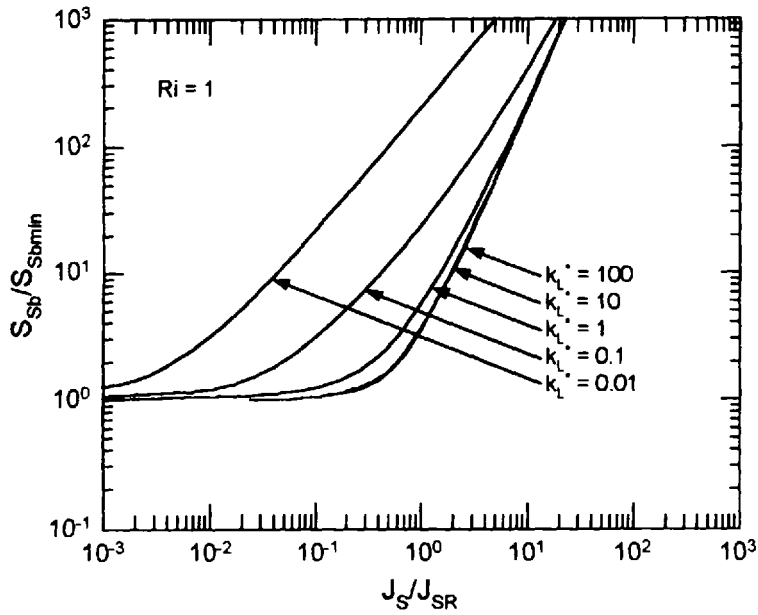


Figure 15.15 Normalized loading curves for Ri = 1. (From M. S. Heath, S. A. Wirtel, B. E. Rittmann, and D. R. Noguera, Closure to discussion of 'Simplified design of biofilm processes using normalized loading curves'. *Research Journal, Water Pollution Control Federation* **63**:91–92, 1991. Copyright © Water Environment Federation; reprinted with permission.)

growing as a biofilm. Two techniques are available, one based on steady-state experiments[35] and the other on transient-state experiments.[29] Space does not allow us to discuss those procedures, but the reader should be aware of their existence, should the need to measure biofilm kinetics arise.

15.2.4 Modeling Transport and Reaction: Limiting-Case Solutions

Although the pseudoanalytical approach to modeling transport and reaction has greatly simplified the computation of the flux into a steady-state biofilm, it still does not lead to closed form solutions to bioreactor mass balances. Many circumstances exist, however, in which direct calculation of the bulk substrate concentration in a biofilm reactor would be advantageous. Consequently, several investigators have proposed simplifying assumptions for limiting cases that allow closed analytical solutions. There are four of them for steady-state biofilms.[31]

Deep Biofilm. As we saw in Section 15.2.3, it is possible to solve directly for the flux into a deep biofilm, which is one in which the substrate concentration at the solid–biofilm interface is zero. Under that condition, the dimensionless flux is given by Eq. 15.29, which was presented earlier.[30,36] Its use results in a different flux for each liquid–biofilm interface substrate concentration.

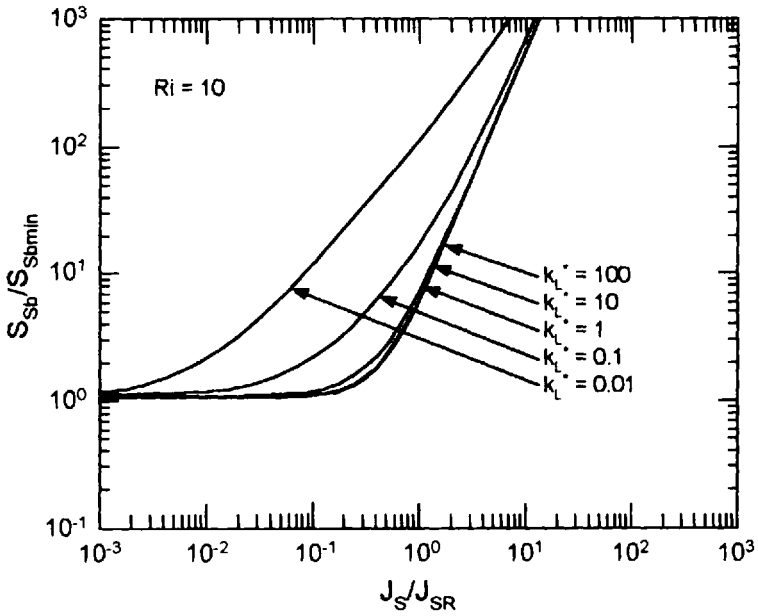


Figure 15.16 Normalized loading curves for Ri = 10. (From M. S. Heath, S. A. Wirtel, B. E. Rittman, and D. R. Noguera, Closure to discussion of "Simplified design of biofilm processes using normalized loading curves." *Research Journal, Water Pollution Control Federation* **63**:91–92, 1991. Copyright © Water Environment Federation; reprinted with permission.)

Fully Penetrated Biofilm. A fully penetrated biofilm is one in which the change in the substrate concentration with depth within the biofilm is negligible. In other words, the entire biofilm contains substrate at almost the same concentration as the liquid–biofilm interface. By assuming that $S_{S,f}$ is equal to S_{S_s} throughout the biofilm, an analytical solution can be obtained.[36]

$$J^*_{S,fp} = \left(\frac{S^*_{S_s}}{1 + S^*_{S_s}}\right) L_f^* \tag{15.43}$$

where L_f^* is given by

$$L_f^* = L_f \left(\frac{\hat{q}_H \cdot X_{B,H}}{D_e \cdot K_S}\right)^{0.5} \tag{15.44}$$

First-Order Biofilm. When the substrate concentration at all points within the biofilm is much less than the half-saturation coefficient, the Monod equation can be approximated as first-order with respect to substrate concentration (Eq. 3.38). This allows the differential equation describing reaction within a steady-state biofilm to be rewritten as:

$$D_e \frac{d^2S_S}{dx^2} - \frac{X_{B,H} \cdot \hat{q}_H \cdot S_S}{K_S} = 0 \tag{15.45}$$

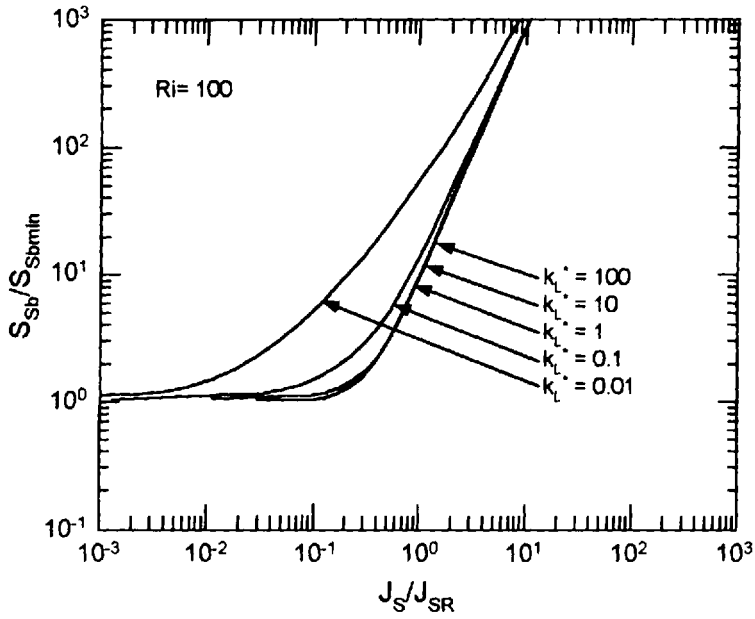


Figure 15.17 Normalized loading curves for Ri = 100. (From M. S. Heath, S. A. Wirtel, B. E. Rittmann, and D. R. Noguera, Closure to discussion of "Simplified design of biofilm processes using normalized loading curves." *Research Journal, Water Pollution Control Federation* **63**:91–92, 1991. Copyright © Water Environment Federation; reprinted with permission.)

For the boundary conditions that

$$\frac{dS_S}{dx} = 0 \text{ at } x = 0 \tag{15.46}$$

and

$$S_S = S_{Ss} \text{ at } X = L_f \tag{15.47}$$

this equation can be solved analytically, giving the following when placed into dimensionless form:[30]

$$J^*_{S,first} = S^*_{Ss} \tanh\left[\left(\frac{1 + Ri}{Ri}\right) J^*_{S,first}\right] \tag{15.48}$$

Zero-Order Biofilm. When the substrate concentration at all points within the biofilm is much greater than the half-saturation coefficient, the Monod equation can be approximated as zero-order with respect to substrate concentration (Eq. 3.37). This allows the differential equation describing reaction within a steady-state biofilm to be rewritten as:

$$D_e \frac{d^2S_S}{dx^2} - X_{B,Hf} \cdot \hat{q}_H = 0 \tag{15.49}$$

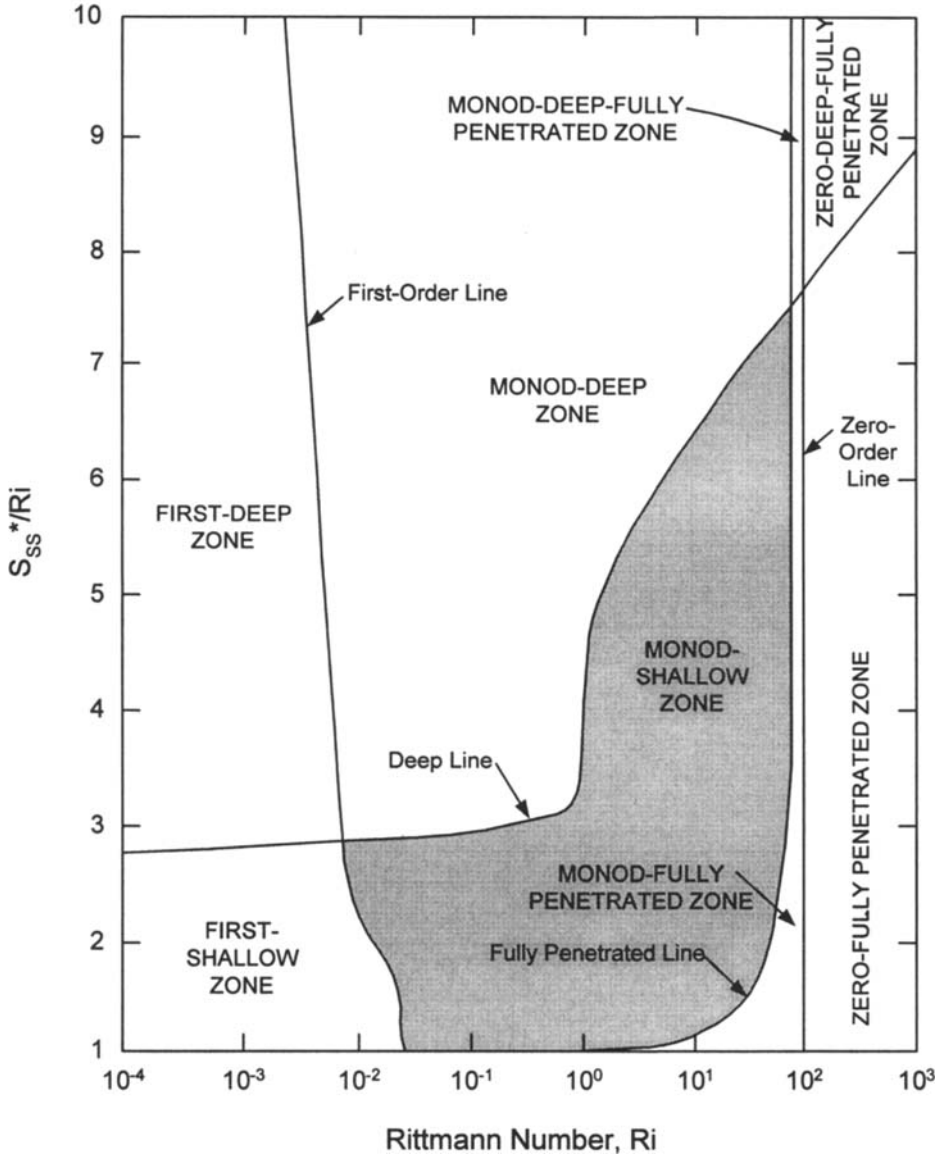


Figure 15.18 Conditions under which a limiting-case solution to the steady-state biofilm model differs from the full pseudoanalytical solution by less than 1.0%. Each unshaded area represents a limiting-case solution. The shaded area, labeled Monod-shallow zone, indicates the conditions under which none of the limiting case solutions are accurate. (From P. B. Sáez and B. E. Rittmann, Error analysis of limiting-case solutions to the steady-state-biofilm model. *Water Research* **24**:1181–1185, 1990. Copyright © Elsevier Science Ltd.; reprinted with permission.)

Solution of this equation with the boundary conditions expressed by Eqs. 15.46 and 15.47 gives the following when placed into dimensionless form:[35]

$$J^*_{S,zero} = L_f^* \tag{15.50}$$

Other Cases. It is theoretically possible for either first-order or zero-order biofilms to be either deep- or fully-penetrated. In such cases, both requirements must be satisfied. A biofilm that is neither first-order nor zero-order has been called a Monod biofilm.[31] A biofilm that is neither deep nor fully-penetrated has been called a shallow biofilm.[31]

Error Analysis. The limiting-case solutions are fine provided the simplifying assumptions are appropriate for the conditions encountered. If they are not, their application can lead to gross errors.[31,36] Thus, one must be sure a limiting-case solution is appropriate before using it. To help in that assessment, Sáez and Rittmann[31] conducted an error analysis of the limiting-case solutions and prepared Figure 15.18 showing conditions under which a limiting-case solution of the steady-state biofilm model differs from the full pseudoanalytical solution by less than 1.0%. Several important points arise from the figure. First, fully penetrated biofilms are difficult to attain, occurring only when the Rittmann number is greater than 10, i.e., only when the relative growth potential is low. This suggests that the fully-penetrated case is of limited utility. First-order biofilms occur when Ri is small, but Monod-biofilms cover a broad range of Ri values. In fact, the Monod-deep zone could be expanded into the first-order-deep zone, since first-order kinetics is just a limiting case of Monod kinetics and the deep-biofilm equation is straightforward. Perhaps the most important point, however, is that the full pseudoanalytical approach must be used over a broad range of conditions, i.e., the Monod-shallow zone, which covers much of the expected range of Ri values.[31]

In summary, while several limiting-case solutions are available in the literature, most are applicable only under very restricted conditions. The exception to this is the case of the deep biofilm. Thus, Eq. 15.29 represents the most useful limiting-case solution. Nevertheless, for the practical range of Ri values, many problems will require use of the full pseudoanalytical solution.

15.3 EFFECTS OF MULTIPLE LIMITING NUTRIENTS

All of the models presented in Section 15.2 consider only a single limiting nutrient, the electron donor. However, in aerobic systems it is quite likely that the concentration of electron acceptor, i.e., oxygen, will drop to sufficiently low values within the biofilm for the microorganisms to become limited by both the electron acceptor and the electron donor. This follows from the fact that for most biofilm systems, the maximum concentration of oxygen in the bulk liquid phase is limited to the saturation concentration associated with air at atmospheric pressure, which is roughly 8 to 10 mg/L. In addition, in some attached growth reactors, the concentration of oxygen in the bulk liquid is likely to be much lower than saturation.

The importance of the dissolved oxygen concentration to the behavior of a biofilm is shown clearly in Figure 15.19.[48] The biofilm depicted was grown with ammonia-N as the sole electron donor, so nitrifying bacteria were the predominant types present. We saw earlier that they have a relatively high half-saturation coefficient for oxygen and thus their activity is influenced strongly by the dissolved oxygen

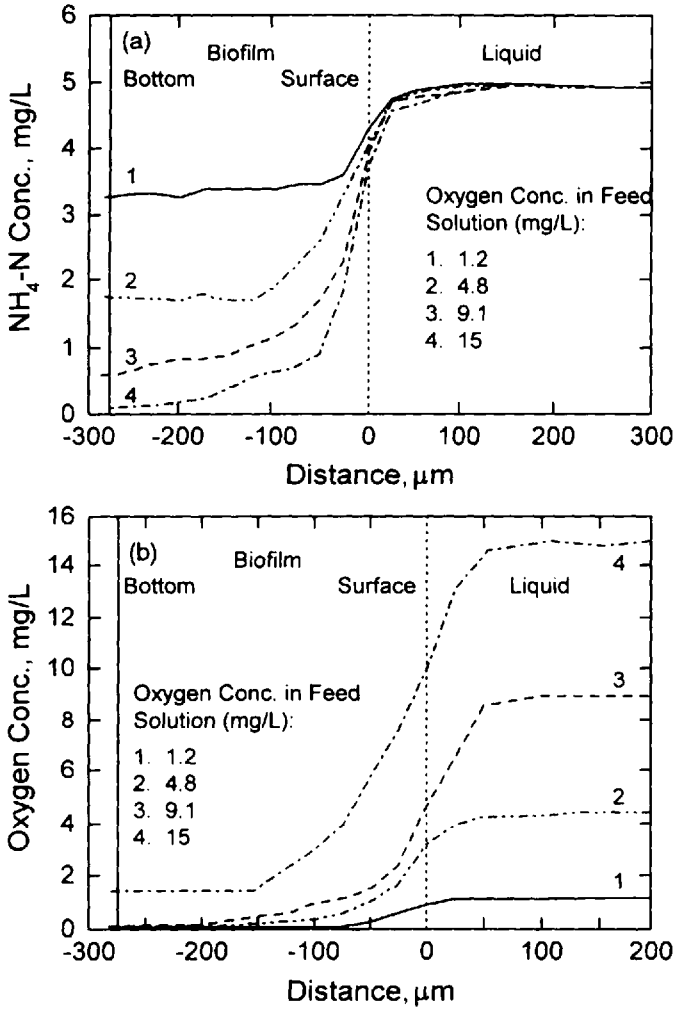


Figure 15.19 Concentration profiles for (a) ammonia-N and (b) dissolved oxygen in nitrifying biofilms subjected to different concentrations of dissolved oxygen in the bulk liquid phase. (Adapted from T. C. Zhang, Y. C. Fu, and P. L. Bishop, Competition in biofilms. *Water Science and Technology* **29**(10/11):263–270, 1994. Copyright © Elsevier Science Ltd.; reprinted with permission.)

concentration. After growth, the biofilm was placed in a test chamber and the profiles of ammonia-N, nitrate-N, and dissolved oxygen concentrations were measured with microelectrodes for different dissolved oxygen concentrations in the bulk liquid phase.[48] The ammonia-N profile is shown in Figure 15.19a and the dissolved oxygen profile is shown in Figure 15.19b. Examination of the figure reveals that when the dissolved oxygen concentration in the bulk liquid was high (15 mg/L), ammonia-N was exhausted before the oxygen. However, as the concentration of dissolved oxygen in the bulk liquid was decreased, oxygen became exhausted at shallower depths in the biofilm, thereby limiting the amount of ammonia-N converted to nitrate-N. This

suggests that for accurate depiction of substrate removal in a biofilm, consideration must be given to the concentration of the electron acceptor as well as the electron donor.

Modeling of dual nutrient limitation requires the use of an interactive double-Monod kinetic expression like Eq. 3.46 in the rate equations for both the electron donor and the electron acceptor, with the two rates being linked stoichiometrically by Eq. 3.34. More importantly, it requires transport of both constituents to be considered. This means that another mass balance equation like Eq. 15.8, with its associated boundary conditions, must be written for the electron acceptor and solved along with the mass balance equation on substrate. The boundary conditions would be analogous to Eqs. 15.9 and 15.10, but would be written in terms of the electron acceptor. Furthermore, the diffusivity and mass transfer coefficient for that equation and its boundary conditions must be for the electron acceptor. Because of the increased number of parameters, this set of equations does not lend itself to the effectiveness factor or the pseudoanalytical approach. Rather, the full equations must be solved numerically for each situation.

Because modeling of systems under dual nutrient limitation is complex, having a way to decide when a biofilm can be considered to be limited by either the electron donor or the electron acceptor alone would be convenient because then the techniques of Section 15.2 could be applied. Unfortunately, this question has only been addressed with the noninteractive dual-substrate model (Eq. 3.47), which is not an accurate representation of the effects of two limiting nutrients.[3] Furthermore, the question is sufficiently complex so as to make the answer nonintuitive. Thus, at this time it is impossible to establish clear guidelines as to when a biofilm can be considered limited by only the electron donor or the electron acceptor. However, it can be stated that when the dimensionless concentration of one is in stoichiometric excess relative to the dimensionless concentration of the other in the bulk liquid, then the one present in least amount can be considered to be a single limiting constituent, allowing the approaches of Section 15.2 to be used. This suggests that only in dilute substrate solutions saturated with dissolved oxygen can the electron donor be considered to be the rate limiting substance in aerobic environments. One important use of biofilm reactors is the treatment of contaminated groundwater and the removal of trace organic compounds from drinking water. Because both of those situations often involve dilute substrate solutions, a single-substrate biofilm model may be accurate enough for them. Furthermore, because anoxic and anaerobic systems often have high electron acceptor concentrations, it is likely that higher electron donor concentrations can be considered rate limiting in them, thereby extending the applicability of single-substrate biofilm models. Nevertheless, because of the benefits associated with the use of the effectiveness factor and pseudoanalytical techniques, it is apparent that the question of when dual-substrate models must be applied needs more systematic study.

15.4 MULTISPECIES BIOFILMS

All of the models we have considered so far in this chapter have considered only a single type of bacteria growing on a single electron donor. Furthermore, they have considered only one type of electron acceptor. However, as we saw when we studied

suspended growth systems, if both organic carbon and ammonia-N are available, then both heterotrophs and autotrophs will grow if the environmental conditions are appropriate. In addition, if an electron donor is present but environmental conditions change from aerobic to anoxic, then facultative heterotrophs can change the nature of their electron acceptor. We have already seen that the oxygen concentration decreases with depth in a biofilm and can approach zero. Thus, if nitrate is present, the potential for denitrification exists in the interior of a biofilm. In other words, just as the potential existed in suspended growth systems for many of the events discussed in Chapter 3, so too does the potential exist in attached growth systems. Consequently, in order for a model to have general utility, it should consider all of these possibilities.

The modeling of multiple events within a biofilm is a good deal more complex than their modeling in suspended growth systems. There are many reasons for this. First is the necessity to consider both transport and reaction simultaneously. We have already seen how that is handled for a single electron donor and electron acceptor provided in the bulk fluid. Extension of those concepts to multiple donors and acceptors in the bulk fluid is not complicated; it just increases the number of differential equations that must be solved. However, it must be recognized that when an electron acceptor such as nitrite-N or nitrate-N is generated within the biofilm, transport can occur in either or both directions from the point of generation, depending on the concentration gradient established. This, too, must be considered in the equations, complicating them somewhat. Another event that complicates the modeling of multispecies biofilms is the competition between the various types of bacteria for the electron acceptor. Autotrophs require molecular oxygen as their electron acceptor and heterotrophs will use it in preference to nitrate-N and nitrite-N when it is present. We saw earlier, however, that heterotrophs have a lower half-saturation coefficient for oxygen than autotrophs do. This means that heterotrophs can lower the oxygen concentration within the biofilm to the point that autotrophs cannot grow. This puts the autotrophs at a disadvantage and limits the region in which they can grow. Perhaps the most important complication arises, however, from the competition for space within the biofilm. In a suspended growth system the biomass is distributed uniformly and is lost from the system in proportion to its concentration. In other words, all of the biomass has the same retention time. This is not true in an attached growth system. Rather, biomass grows outward from a solid support and is removed by detachment at the liquid–biofilm interface. This displacement toward the interface must be considered in any multispecies model. In addition, as we saw in Figure 15.6, the distribution of biomass is not uniform throughout the biofilm. This means that different types of bacteria have different residence times in the biofilm, and the model must be structured to consider this as well.

As one might deduce from the description above, the modeling of multispecies biofilms is among the most complicated activities in the modeling of biochemical operations for wastewater treatment. Consequently, space does not permit us to explore the subject to the same depth that we explored the single species, single substrate models. Rather, we will consider only a few key concepts. Those wishing to develop a greater understanding of the subject should read Chapter 11 in *Biofilms*[18] and any of the papers describing modeling efforts.[22,24,27,39,40,41] Those wishing to use a computer code implementing a general purpose multispecies biofilm model should acquire a copy of AQUASIM.[25]

To understand any of the multispecies biofilm models in the literature, it is important to understand why one type of bacteria can be displaced by another within a biofilm. That is, one must understand competition for space. Consider a control volume within a biofilm as shown in Figure 15.7. When net growth is positive in the control volume, the biomass increases. If the biomass density is constant, biomass must cross the control volume's boundary, giving a biomass flux.[18,39,40] The mass balance on the control volume must incorporate the flux of biomass into it from deeper within the biofilm and the flux of biomass out toward the liquid–biofilm interface. The flux of biomass into the control volume is proportional to the integrated net growth of biomass, i.e., growth minus decay, deeper in the biofilm.[27] Biomass debris will also be generated within the biofilm due to decay (Eq. 3.57) and it will occupy space, also contributing to the flux. This flux causes the biofilm to increase in depth over time until the flux of biomass due to net growth is just balanced by the loss per unit area due to surface detachment. One effect of this flux, which starts at the biofilm-support interface and increases with distance from that interface, is a net migration of particles from the interior of the biofilm to the liquid–biofilm interface. If a particular species of bacteria cannot grow rapidly enough in the biofilm to equal this displacement, it will eventually be lost. Thus, the question of coexistence is one of whether the net growth rate of one species is great enough to allow it to compete with other species for space.

Rittmann and Manem[27] have developed a model for a steady-state biofilm containing i bacterial species that are competing only for space, i.e., they each have their unique electron donor, S_{Si}, and the concentration of the electron acceptor is not rate-limiting. It is an extension of the steady-state biofilm model presented in Section 15.2.3 and requires combining a new mass balance on biomass with transport and reaction equations for each of i substrates. In a multispecies biofilm, the density of any one species (including biomass debris, X_D) is a fraction f_{XBi} of the total biomass density, X_{Bf}:

$$X_{B,i} = f_{XBi}X_{Bf} \tag{15.51}$$

in which $X_{B,i}$ is the density of species i at a point in the biofilm, and X_{Bf} is assumed constant. (For this development, X_D will be considered to be one value of $X_{B,i}$.) The sum of the densities must equal the total density, thus:

$$\sum_{i=1}^{n} f_{XBi} = \sum_{i=1}^{n} \left(\frac{X_{B,i}}{X_{Bf}}\right) = 1 \tag{15.52}$$

in which n is the total number of biomass types. A steady-state mass balance on species i in a fixed control volume of biofilm leads to:[27]

$$\frac{df_{XBi}}{dx} \sum_{i=1}^{n} \int_{0}^{x} (\mu_i - b_i f_D) f_{XBi}\, dx = (\mu_i - b_i f_D) f_{XBi} - f_{XBi} \sum_{i=1}^{n} (\mu_i - b_i f_D) f_{XBi} \tag{15.53}$$

in which the values of f_{XBi} and μ_i not inside the integral are for position x only, and f_D is the fraction of active biomass contributing to biomass debris, as defined in Eq. 3.53. The rate coefficient b_i is the traditional decay coefficient and μ_i is the specific growth rate for the particular type of biomass represented by i. For example, if species 1 is a heterotroph, μ_1 will be a μ_H and if species 2 is an autotroph, μ_2 will

be a μ_A. The boundary condition for Eq. 15.53 is that there is no flux of any species into the attachment surface:

$$J_X = 0 \text{ at } x = 0 \tag{15.54}$$

For a steady-state biofilm, the growth of all species due to substrate utilization is just equal to the loss by detachment and decay. For multispecies, this can be written as:

$$\sum_{i=1}^{n}(J_{Si} \cdot Y_i) = X_{Bf}\left(b_D \cdot L_f + \sum_{i=1}^{n}\int_0^{L_f}(f_{XBi}f_D b_i)\,dx\right) \tag{15.55}$$

This equation can be used in place of Eq. 15.26 when writing a model for multispecies biofilm growth and substrate utilization. It is used in combination with equations like Eq. 15.23 with associated boundary conditions for each of i substrates.

Rittmann and Manem[27] then used the concept of S_{Sbmin} to decide whether coexistence of two species is possible. First, S_{Sbmin} should be calculated for each species using Eq. 15.22. The species whose bulk substrate concentration is closest to its S_{Sbmin} value is the one whose ability to exist in the biofilm is questionable. For this development, that species will be defined as species 2. Two conditions must be met by species 2 for it to coexist with species 1 in the biofilm. First, it must have a net positive specific growth rate somewhere in the biofilm. As a consequence:

$$S_{S2} > \frac{K_{S2} \cdot b_2}{\hat{q}_2 \cdot Y_2 - b_2} \tag{15.56}$$

at some position in the biofilm. Second, it must have a sufficiently fast specific growth rate to allow it to compete with species 1 for space. Because S_{Sb1}/S_{Sbmin1} is greater than S_{Sb2}/S_{Sbmin2}, the most favorable location for species 2 to compete with species 1 is near the attachment surface, where S_{S1} is the lowest. Because $J_X = 0$ at the attachment surface (Eq. 15.54), Eq. 15.53 gives for species 2:

$$f_{XB2}\left[(\mu_2 - b_2) - \sum_{i=1}^{n} f_{XBi}(\mu_i - b_i f_D)\right] = 0 \text{ at } x = 0 \tag{15.57}$$

Because f_{XB2} must be greater than zero for species 2 to be present, the terms inside the brackets in Eq. 15.57 must equal zero. Consequently:

$$\mu_{2as} - b_2 = b_C \tag{15.58}$$

in which

$$b_C = f_{XB1as}(\mu_{1as} - b_1 f_D) + f_{XB2as}(\mu_{2as} - b_2 f_D) \tag{15.59}$$

and the subscript "as" refers to the location at the attachment surface. The term b_C is called the competition coefficient. At the limit of coexistence, f_{XB2as} approaches zero, in which case b_C becomes

$$b_C = f_{XB1as}(\mu_{1as} - f_D b_1) \tag{15.60}$$

The minimum bulk liquid concentration of substrate 2 that will just allow coexistence of the two species can be obtained by noting that it must provide a specific growth rate for species 2 at the attachment surface that will satisfy Eq.15.58. Consequently:

$$S_{Sbmin2C} = \frac{K_{S2}(b_2 + b_{C'})}{\hat{q}_2 Y_2 - (b_2 + b_{C'})} \tag{15.61}$$

where $b_{C'}$ is defined by Eq. 15.60. In other words, one can use a single substrate, single species model to estimate $\mu_{1,as}$, allowing $b_{C'}$ to be quantified. This will allow quantification of $S_{Sbmin2C'}$. If S_{b2} is greater than that value, then the two species will be able to coexist in the biofilm.

Examination of Eq. 15.61 shows that it is similar to Eq. 15.22 for a single species biofilm, except that $b_{C'}$ replaces b_D. This follows from the fact that at the limit of coexistence, species 2 is protected from detachment by species 1, but must grow rapidly enough to compete with it for space. For the special case in which the biofilm is deep for species 1, i.e., $S_{S1} = 0$ at $x > 0$, $f_{XB1,as}$ approaches zero, which makes $b_{C'}$ approach zero.[27] In this case, $S_{Sbmin2C'}$ is:

$$S_{Sbmin2C'} = \frac{K_{S2} \cdot b_2}{\hat{q}_2 Y_2 - b_2} \tag{15.62}$$

which is the same as for a single-species biofilm undergoing no detachment. When the biofilm is fully penetrated with respect to substrate 1, $f_{XB1,as}$ approaches 1, $b_{C'}$ approaches b_D, and $S_{Sbmin2C'}$ is given by the normal equation for S_{Sbmin}, Eq. 15.22.[27] For all cases between a fully penetrated and a deep biofilm, $S_{Sbmin2C'}$ will gradually decrease from the value given by Eq. 15.22 to the value given by Eq. 15.62. Thus, having a deeper biofilm with respect to species 1 protects species 2 from detachment and lowers its effective S_{Sbmin} value.

The above analysis suggests that the more slowly growing species can be protected from loss provided that $b_{C'}$ is less than b_D. The caveat in this analysis is that it assumes that the two species do not compete for the electron acceptor at limiting concentrations. If that occurs, the specific growth rates of the two species within the biofilm will also be influenced by the electron acceptor concentration profile and their relative half-saturation coefficients for it. This greatly complicates the analysis and makes it difficult to come up with a single criterion as was done above. However, Wanner and Gujer[39] have investigated the question of competition between heterotrophs and autotrophs competing for dissolved oxygen being supplied from the bulk liquid and have examined the conditions for coexistence. The interactive double Monod equation, Eq. 3.46, was used to express the effects of the electron donor and the electron acceptor on the specific growth rate of each type of organism. Nitrification was assumed to occur in one step as was done in International Association on Water Quality (IAWQ) activated sludge model (ASM) No. 1. Analysis of the question with the kinetic and stoichiometric coefficients in Table 15.2 gave the results in Figure 15.20. Any combination of bulk liquid biodegradable COD and ammonia-N concentrations that lies in zone H will result in a fully heterotrophic biofilm and no nitrification will occur. This suggests that nitrification can only occur in a biofilm process after the bulk of the organic matter has been removed. Any combination that lies in zone A will result in a fully autotrophic biofilm. This suggests that fully autotrophic biofilms will be rare. Finally, combinations of bulk liquid COD and

Table 15.2 Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used to Generate Figure 15.20[10]

Symbol	Units	Value
$\hat{\mu}_H$	hr^{-1}	0.20
K_S	mg/L as COD	5.0
$K_{O,H}$	mg/L as O_2	0.10
Y_H	mg biomass COD/mg substrate COD	0.40
b_H	hr^{-1}	0.0083
$\hat{\mu}_A$	hr^{-1}	0.040
K_{NH}	mg/L as N	1.0
$K_{O,A}$	mg/L as O_2	0.10
Y_A	mg biomass COD/mg N	0.22
b_A	hr^{-1}	0.0021
D_{eS}	cm^2/hr	0.035
D_{eA}	cm^2/hr	0.062
D_{eO}	cm^2/hr	0.073

ammonia-N concentrations that lie in zone AH will result in a two-species biofilm in which both nitrification and carbon oxidation occur. However, the larger the bulk liquid COD associated with a given ammonia-N concentration, the greater the contributions of the heterotrophs to the biofilm. These effects of competition for space and oxygen have a strong effect on the design of biofilm processes for carbon oxidation and nitrification as we will see in Chapters 19, 20 and 21.

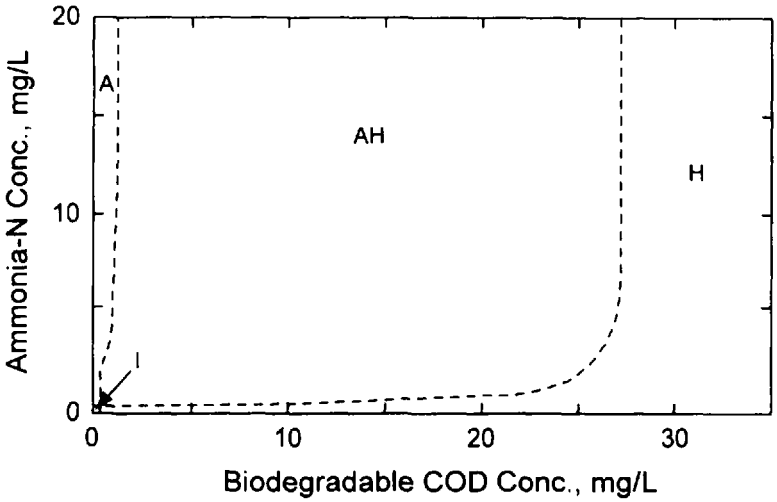


Figure 15.20 Effects of bulk fluid ammonia-N and biodegradable COD concentrations on the coexistence of autotrophic and heterotrophic biomass in a biofilm. The kinetic and stoichiometric coefficients used are given in Table 15.2. (After Wanner and Gujer.[10])

15.5 KEY POINTS

1. Biofilms are nonuniform structures consisting of discrete cell clusters attached to each other and to a solid support with extracellular polymeric material. The spaces between clusters form vertical and horizontal voids, with the vertical voids acting as pores and the horizontal voids acting as channels. Nevertheless, most current mathematical models for biofilm reactors assume that transport within a biofilm is by diffusion alone with a constant diffusion coefficient.
2. In an attached growth bioreactor, the substrate concentration at which the reaction takes place is less than the substrate concentration in the bulk liquid phase because of the resistances encountered in the transport of substrate from the bulk liquid into the biofilm. Therefore, it is necessary to combine physical mass transport with microbial reactions when modeling attached growth reactors.
3. Mass transfer from the bulk liquid phase to the liquid–biofilm interface is considered to occur across a hypothetical stagnant liquid film adjacent to the interface. The rate of mass transfer across that film is proportional to the drop in substrate concentration across it and the proportionality constant, k_l (T^{-1}), is called the mass transfer coefficient.
4. The transport of substrate within a biofilm may be described by Fick's law with the free diffusion coefficient, D_w (L^2T^{-1}), replaced by an effective diffusion coefficient, D_e.
5. The term flux refers the mass flow rate of a substance per unit area. The flux of substrate across the stagnant liquid layer to the liquid–biofilm interface must equal the flux of substrate into the biofilm by diffusion, which must equal the rate of substrate consumption per unit of biofilm planar surface area. Thus, the terms flux and substrate utilization rate are used interchangeably for biofilms.
6. The effectiveness factor, η_e, is defined as the ratio of the actual, observed substrate removal rate per unit of biofilm planar surface area to the theoretical rate that would occur in the absence of mass transfer resistance. Thus, the effectiveness factor is a correction factor, which, when multiplied by the homogeneous reaction rate without mass transfer resistance gives the actual rate in the presence of mass transfer resistance.
7. The overall effectiveness factor, η_{eO}, accounts for the presence of both external and internal mass transfer resistance. Its values must be determined numerically and are commonly presented graphically in terms of a Sherwood number called the Biot number, Bi, the Thiele modulus, ϕ, and a modified Thiele modulus, ϕ_l. They are functions of the bulk liquid substrate concentration. In general, the overall effectiveness factor decreases rapidly with increases in ϕ_l, increases slightly with increases in ϕ, and increases rapidly with increases in Bi.
8. The presence of a biofilm prevents washout of a continuous stirred tank reactor (CSTR). As a result, CSTRs containing biofilms are capable of removing substrate at hydraulic residence times (HRTs) well below those that cause washout in a CSTR without a biofilm.

9. The pseudoanalytical approach to modeling transport and reaction utilizes simple algebraic expressions for the flux of substrate into a biofilm. The simple algebraic expressions are based on an analysis of the results from the numerical solution of the differential equations describing transport and reaction in a biofilm. The availability of simple algebraic equations eliminates the need to repetitiously solve numerically a set of nonlinear differential equations while modeling the performance of a biofilm reactor.
10. A steady-state biofilm is one in which the gains in biofilm mass due to biomass growth are just balanced by the losses in biofilm mass due to the combined effects of microbial decay within the biofilm and detachment by shear at the liquid–biofilm interface. As a consequence, a minimum bulk liquid substrate concentration, S_{sbmin}, is required to drive the growth reactions to balance the losses and concentrations below that value cannot support a steady-state biofilm.
11. The dimensionless minimum bulk liquid substrate concentration, S^*_{sbmin}, is determined solely by parameters that depend on the biomass, substrate, and fluid regime. It represents the ratio of the biomass specific loss rate to the biomass maximum net specific growth rate, giving it special significance for steady-state biofilms. Consequently, it has been named the Rittmann number and given the symbol Ri.
12. Implementation of the pseudoanalytical approach for the analysis and design of biofilm reactors is made easier by the use of normalized loading curves which present a normalized flux as a function of the normalized minimum bulk substrate concentration. Normalization of the plots makes them applicable for a broad range of parameter values.
13. Closed form solutions allowing direct computation of the substrate flux into a biofilm are possible for several limiting cases: a deep biofilm, a fully penetrated biofilm, a first-order biofilm, and a zero-order biofilm. Error analysis has shown that all but the deep biofilm case are applicable only under very restricted conditions.
14. Multispecies biofilm modeling requires that all electron donors and electron acceptors for both heterotrophs and autotrophs be considered, just as they were in the IAWQ activated sludge models. Furthermore, appropriate transport and reaction equations must be incorporated for each component. The complexity involved prevents the effectiveness factor and pseudoanalytical approaches from being used. Rather, a set of differential equations must be solved simultaneously.
15. Because there is a net flux of particulate material from the biofilm support toward the liquid–biofilm interface, more rapidly growing bacteria can displace slowly growing bacteria from a biofilm. However, because bacteria in the interior of the biofilm are protected from surface shear, rapidly growing bacteria can protect slowly growing bacteria, allowing them to survive in a biofilm under conditions where they would not be able to form a biofilm alone.
16. Multispecies modeling results suggest that because of competition for space and dissolved oxygen, nitrifying bacteria cannot survive in biofilms when the bulk organic substrate concentration exceeds approximately 27

mg/L as COD. However, nitrifiers and heterotrophs can coexist in biofilms for concentrations below that value.

15.6 STUDY QUESTIONS

1. Describe the structure of a biofilm as it is now conceived and contrast that conceptualization with the conceptualization used in mathematical models of biofilms.
2. Define the mass transfer coefficient, k_l, and relate it to the thickness of the hypothetical stagnant liquid layer that is assumed to lie between the bulk liquid and the liquid–biofilm interface.
3. Describe the transport of substrate within a biofilm, including in your description an explanation of the effective diffusivity.
4. Draw a sketch depicting the substrate concentration profile within a biofilm. Then tell what the slope of the profile must be at the liquid–biofilm interface as well as at the biofilm-support interface (assuming an impermeable support) and explain why.
5. What is the role of an effectiveness factor?
6. Name and define the three parameters that characterize the overall effectiveness factor for Monod kinetics in a flat biofilm. Describe briefly the effects of these parameters on the value of the overall effectiveness factor.
7. A CSTR with a volume of 1.0 L receives a sterile feed with a substrate concentration of 200 mg/L as COD at a flow rate of 1.0 L/hr. The CSTR contains a biofilm with the kinetic and stoichiometric parameters listed in Table SQ15.1. The system variables are also listed there. Prepare a graph showing the surface area of biofilm required to reduce the effluent substrate concentration to values between 1 and 100 mg/L. Compute at least four data points and then explain why the curve has the shape that it does. Use the effectiveness factor approach.
8. Describe a steady-state biofilm and state the conditions required to maintain it.
9. Explain why S_{Sbmin} has special significance as a parameter in the pseudoanalytical approach to solving the equations quantifying substrate

Table SQ15.1 Kinetic Parameters, Stoichiometric Coefficients, and System Variables for Study Question 7

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD · hr)	0.33
K_S	mg/L as COD	20
Y_{Hobs}	mg biomass COD/mg substrate COD	0.60
$X_{B,H}$	mg biomass COD/cm^3	40
k_l	cm/hr	5.0
D_e	cm^2/hr	0.025
L_f	cm	0.020

flux into a steady-state biofilm. Then explain the significance of the Rittmann number, telling what values greater than and less than 1.0 suggest.

10. A steady-state biofilm, described by the kinetic and stoichiometric coefficients given in Table SQ15.2, exists in an environment where the bulk substrate concentration is maintained at a constant value. Determine how the bulk substrate concentration affects the substrate flux into the biofilm and the biofilm thickness. Calculate at least four points over the range of bulk substrate concentrations from 0.1 to 1.0 mg/L.
11. A synthetic wastewater with a biodegradable COD of 20 mg/L (0.020 mg/cm^3) is flowing at a rate of 1.0 L/hr (1000 cm^3/hr) into a single CSTR containing a biofilm media with a specific surface area of 150 m^2/m^3 (1.50 cm^2/cm^3). The wastewater, the bioreactor, and the associated biofilm have the characteristics listed in Table SQ15.2. What total surface area of biofilm would be required to reduce the biodegradable COD to 0.5 mg/L (0.0005 mg/cm^3)? What bioreactor volume is required to house the media? Determine your answer by the pseudoanalytical approach.
12. Repeat Study Question 11 using the normalized loading curves.
13. Consider the situation presented in Examples 15.2.3.1 and 15.2.3.2. If a synthetic wastewater with the same kinetic characteristics was introduced at a rate of 1.0 L/hr into a chain of three CSTRs containing the same media, what influent COD concentration could be treated to a final effluent concentration, i.e., from bioreactor number three, of 0.5 mg/L of COD if each bioreactor had a volume of 1.0 L?
14. Describe the situations under which each of the limiting-case solutions of the biofilm flux equation (deep, fully-penetrated, first-order, and zero-order biofilms) are applicable.
15. Describe the situations under which it is acceptable to assume that a biofilm is behaving as if it were limited by only the electron donor or the electron acceptor.
16. Explain why a fast growing species of bacteria is able to prevent a slow growing species from growing in a biofilm even when they are not competing for either the electron donor or the electron acceptor.

Table SQ15.2 Kinetic Parameters, Stoichiometric Coefficients, and System Variables for Study Questions 10 and 11

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD · hr)	0.75
K_S	mg/L as COD	15.0
Y_H	mg biomass COD/mg substrate COD	0.54
b_H	hr^{-1}	0.004
$X_{B,H}$	mg biomass COD/cm^3	45
k_l	cm/hr	5.0
D_s	cm^2/hr	0.025
b_D	hr^{-1}	0.003

17. What is the competition coefficient and what is its significance to the determination of whether two species can coexist in a biofilm when they are competing only for space and not for the electron donor or acceptor?
18. Describe the conditions under which autotrophs and heterotrophs can coexist in a biofilm in which they are competing for the dissolved oxygen and explain why they are able to do so.

REFERENCES

1. Atkinson, B., *Biochemical Reactors*, Pion Limited, London, 1974.
2. Atkinson, B. and I. J. Davies, The overall rate of substrate uptake (reaction) by microbial films. Part I. A biological rate equation. *Transactions of Institution of Chemical Engineers* **52**:248–259, 1974.
3. Bae, W. and B. E. Rittmann, A structured model of dual-limitation kinetics. *Biotechnology and Bioengineering* **49**:683–689, 1996.
4. Blanch, H. W. and D. S. Clark, *Biochemical Engineering*, Marcel Dekker, New York, 1996.
5. Cannon, F. S., Discussion of 'Simplified design of biofilm processes using normalized loading curves'. *Research Journal, Water Pollution Control Federation* **63**:90, 1991.
6. Characklis, W. G. and K. C. Marshall, eds, *Biofilms*, Wiley, New York, 1990.
7. Characklis, W. G. and K. C. Marshall, Biofilms: a basis for an interdisciplinary approach. In *Biofilms*, W. G. Characklis and K. C. Marshall, eds. Wiley, New York, pp. 3–15, 1990.
8. Characklis, W. G., M. H. Turakhia, and N. Zelver, Transport and interfacial transfer phenomena. In *Biofilms*, W. G. Characklis and K. C. Marshall, eds. Wiley, New York, pp. 265–340, 1990.
9. Costerton, J. W., Z. Lewandowski, D. E. Caldwell, D. R. Korber, and H. M. Lappin-Scott, Microbial biofilms. *Annual Review of Microbiology* **49**:711–745, 1995.
10. de Beer, D., P. Stoodley, and Z. Lewandowski, Liquid flow in heterogeneous biofilms. *Biotechnology and Bioengineering* **44**:636–641, 1994.
11. de Beer, D., P. Stoodley, and Z. Lewandowski, Liquid flow and mass transport in heterogeneous biofilms. *Water Research* **30**:2761–2765, 1996.
12. de Beer, D., P. Stoodley, F. Roe, and Z. Lewandowski, Effects of biofilm structures on oxygen distribution and mass transport. *Biotechnology and Bioengineering* **43**:1131–1138, 1994.
13. Fink, D. J., T.-Y. Na, and J. S. Schultz, Effectiveness factor calculations for immobilized enzyme catalysts. *Biotechnology and Bioengineering* **15**:879–888, 1973.
14. Finlayson, B. A., *The Method of Weighted Residuals and Variational Principles*, Academic Press, New York, 1972.
15. Gjaltema, A., P. A. M. Arts, M. C. M. van Loosdrecht, J. G. Kuenen, and J. J. Heijnen, Heterogeneity of biofilms in rotating annular reactors: occurrence, structure, and consequences. *Biotechnology and Bioengineering* **44**:194–204, 1994.
16. Grady, C. P. L. Jr., Modeling of biological fixed films—A state-of-the-art review. In *Proceedings, First International Conference on Fixed-Film Biological Processes*, Y. C. Wu, E. D. Smith, R. D. Miller, and E. J. O. Patken, eds. University of Pittsburgh, Pittsburgh, pp. 344–404, 1982.
17. Grady, C. P. L. Jr. and H. C. Lim, *Biological Wastewater Treatment: Theory and Applications*, Marcel Dekker, New York, 1980.
18. Gujer, W. and O. Wanner, Modeling mixed population biofilms. In *Biofilms*, W. G. Characklis and K. C. Marshall, eds. Wiley, New York, pp. 397–443, 1990.

19. Heath, M. S., S. A. Wirtel, and B. E. Rittmann, Simplified design of biofilm processes using normalized loading curves. *Research Journal, Water Pollution Control Federation* **62**:185–192, 1990.
20. Heath, M. S., S. A. Wirtel, B. E. Rittmann, and D. R. Noguera, Closure to discussion of 'Simplified design of biofilm processes using normalized loading curves'. *Research Journal, Water Pollution Control Federation* **63**:91–92, 1991.
21. Jansen, J. and G. H. Kristensen, Fixed film kinetics—Denitrification in fixed films. Report No. 80–59, Department of Sanitary Engineering, Technical University of Denmark, Lyngby, Denmark.
22. Kissel, J. C., P. L. McCarty, and R. L. Street, Numerical simulation of mixed-culture biofilm. *Journal of Environmental Engineering* **110**:393–411, 1984.
23. Lazarova, V. and J. Manem, Biofilm characterization and activity analysis in water and wastewater treatment. *Water Research* **29**:2227–2245, 1995.
24. McCarty, P. L., M. Reinhard, and B. E. Rittmann, Trace organics in groundwater. *Environmental Science and Technology* **15**:40–51, 1981.
25. Reichert, P., AQUASIM, a tool for simulation and data analysis of aquatic systems. *Water Science and Technology* **30**(2):21–30, 1994.
26. Rittmann, B. E., The effect of shear stress on biofilm loss rate. *Biotechnology and Bioengineering* **24**:501–506, 1982.
27. Rittmann, B. E. and J. A. Manem, Development and experimental evaluation of a steady-state, multispecies biofilm model. *Biotechnology and Bioengineering* **39**:914–922, 1992.
28. Rittmann, B. E. and P. L. McCarty, Model of steady-state-biofilm kinetics. *Biotechnology and Bioengineering* **22**:2343–2357, 1980.
29. Rittmann, B. E., L. Crawford, C. K. Tuck, and E. Namkung, *In situ* determination of kinetic parameters for biofilms: isolation and characterization of oligotrophic biofilms. *Biotechnology and Bioengineering* **28**:1753–1760, 1986.
30. Sáez, P. B. and B. E. Rittmann, Improved pseudoanalytical solution for steady-state biofilm kinetics. *Biotechnology and Bioengineering* **32**:379–385, 1988.
31. Sáez, P. B. and B. E. Rittmann, Error analysis of limiting-case solutions to the steady-state-biofilm model. *Water Research* **24**:1181–1185, 1990.
32. Sáez, P. B. and B. E. Rittmann, Accurate pseudoanalytical solution for steady-state biofilms. *Biotechnology and Bioengineering* **39**:790–793, 1992.
33. Siegrist, H. and W. Gujer, Mass transfer mechanisms in a heterotrophic biofilm. *Water Research* **19**:1369–1378, 1985.
34. Stewart, P. S., R. Murga, R. Srinivasan, and D. de Beer, Biofilm structural heterogeneity visualized by three microscopic methods. *Water Research* **29**:2006–2009, 1995.
35. Suidan, M. T. and Y. T. Wang, Unified analysis of biofilm kinetics. *Journal of Environmental Engineering* **111**:634–646, 1985.
36. Suidan, M. T., B. E. Rittmann, and U. K. Traegner, Criteria establishing biofilm-kinetic types. *Water Research* **21**:491–498, 1987.
37. Tartakovsky, B. and S. R. Guiot, Modeling and analysis of layered stationary anaerobic granular biofilms. *Biotechnology and Bioengineering* **54**:122–130, 1997.
38. Trulear, M. G. and W. G. Characklis, Dynamics of biofilm processes. *Journal, Water Pollution Control Federation* **54**:1288–1301, 1982.
39. Wanner, O. and W. Gujer, Competition in biofilms. *Water Science and Technology* **17**(2/3):27–44, 1984.
40. Wanner, O. and W. Gujer, A multispecies biofilm model. *Biotechnology and Bioengineering* **28**:314–328, 1986.
41. Wanner, O. and P. Reichert, Mathematical modeling of mixed-culture biofilms. *Biotechnology and Bioengineering* **49**:172–184, 1996.
42. Weber, W. J. Jr. and DiGiano, F. A., *Process Dynamics in Environmental Systems*, Wiley, New York, 1996.

43. Wilderer, P. A. and W. G. Characklis, Structure and function of biofilms, In *Structure and Function of Biofilms*, W. G. Characklis and P. A. Wilderer, eds. Wiley, New York, pp. 5–17, 1989.
44. Yang, S. and Z. Lewandowski, Measurement of local mass transfer coefficient in biofilms. *Biotechnology and Bioengineering* **48**:737–744, 1995.
45. Zhang, T. C. and P. L. Bishop, Structure, activity and composition of biofilms. *Water Science and Technology*, **29**(7):335–344, 1994.
46. Zhang, T. C. and P. L. Bishop, Experimental determination of the dissolved oxygen boundary layer and mass transfer resistance near the fluid–biofilm interface. *Water Science and Technology* **30**(11):47–58, 1994.
47. Zhang, T. C., Y. C. Fu, and P. L. Bishop, Competition in biofilms. *Water Science and Technology* **29**(10/11):263–270, 1994.
48. Zhang, T. C., Y.-C. Fu, and P. L. Bishop, Competition for substrate and space in biofilms. *Water Environment Research* **67**:992–1003, 1995.
49. Zhang, S. and P. M. Huck, Parameter estimation for biofilm processes in biological water treatment. *Water Research* **30**:456–464, 1996.

This page intentionally left blank

16

Aerobic Growth of Biomass in Packed Towers

Several types of attached growth bioreactors were listed in Table 1.2. Among those, the most widely used is the packed tower, which contains microorganisms growing on an immobile support over which wastewater flows in thin sheets. Most recently installed packed towers use plastic media as the immobile support. Two types are in current use, random packing, which is typically in the form of cylinders approximately 5 cm in diameter and 5 cm long, and bundle media, which consists of sheets formed into self-supporting modules with vertical surfaces. The media is held within a containment structure that provides physical support for random packing but is necessary only to prevent splashing and to minimize wind effects for bundle media. Clarified wastewater is distributed uniformly over the top of the media by the distribution system, which may be either rotary or fixed. Fixed nozzles may discharge either continuously or intermittently, in which case they have hydraulic characteristics similar to rotary distributors. After passing over the media in thin sheets, the treated wastewater is collected in the underdrain system, which is open to the atmosphere to allow free movement of air through the tower. From there, it flows to a gravity settler for removal of biomass.

When wastewater containing organic matter, nitrogen, or other electron donors flows through a packed tower, microorganisms consume the substrates and grow attached to the media as a biofilm. The flow of water over the biofilm imparts a shear force to it, keeping its thickness relatively constant by removing cells generated by consumption of the substrate. The suspended biomass is then removed by gravity settling before discharge of the treated effluent. In some cases, effluent from the tower is recirculated back to the top to allow control of the flow rate through the tower and to influence the concentration of substrate. The recirculated flow is generally clarified effluent, but not always.

Although a packed tower is mechanically simpler than most suspended growth systems, it is more difficult to model for a number of reasons. First, both the electron donor and the electron acceptor must be transported into the biofilm for reaction, as discussed in Chapter 15. This means that the models for substrate removal are much more complicated. Second, although flow is generally assumed to be in thin sheets in most models, in reality, the flow patterns are quite complex over all of the media types. Third, various types of bacteria must compete for nutrients and space in the biofilm, as discussed in Section 15.4, rather than being homogeneously distributed as in a suspended growth culture. Fourth, the biofilm will not be evenly distributed over all of the media; rather, the distribution will depend on the both the flow patterns

of the fluid and the concentration of the substrate in it. This makes the performance of packed towers much more stochastic than the performance of suspended growth systems. However, like most models of suspended growth systems, most packed tower models tend to be deterministic.

As a consequence of the various factors in the preceding paragraph, no consensus models yet exist for biofilm processes that are comparable to International Association on Water Quality (IAWQ) activated sludge model (ASM) Nos. 1 and 2 for suspended growth processes. Models exist that account for transport and reaction in the manner discussed in Chapter 15, but they are generally limited to a single limiting nutrient and consider fluid flow to be in thin sheets with no intermixing. At the other extreme, models exist that seek to account for the complexity in fluid flow through packed towers, but they use limiting case solutions for transport and reactions, thereby limiting their generality, as discussed in Section 15.2.4. As a consequence, it is currently impossible to use mechanistic modeling to investigate the theoretical performance of packed towers with the same degree of confidence that modeling was used in Part II to investigate suspended growth systems. Nevertheless, much can be learned from models that idealize the hydraulic regime as thin film flow, as long as one views their results qualitatively rather than quantitatively. Consequently, we will use such a model herein to learn how certain factors influence packed tower performance and then will consider other important aspects not included in the model.

16.1 MODEL FOR SUBSTRATE REMOVAL IN A PACKED TOWER

16.1.1 Simplifying Assumptions

By making certain simplifying assumptions, the effectiveness factor approach may be used to investigate the theoretical performance of a packed tower, and this has been done by Grady and Lim.[1] Consequently, their approach will be presented here. The first assumption is that the electron acceptor is present in excess so that the electron donor acts as a single limiting nutrient. The second is that only one type of microorganism is present; heterotrophs if the electron donor is organic matter and autotrophs if it is ammonia-N. The results will be presented in terms of heterotrophs removing soluble chemical oxygen demand (COD), but they will be qualitatively similar to those for autotrophs removing ammonia-N in the absence of heterotrophs. The third is that steady-state conditions prevail so that microorganisms are sheared from the surface of the biofilm at a rate equal to their growth, resulting in a biofilm of constant thickness. The fourth assumption is that the thickness of the biofilm is constant throughout the tower. That thickness is an input parameter and is not computed in the model. The fifth assumption is that detached biomass is carried along with the liquid flow and contributes to substrate removal. The sixth assumption is that only clarified effluent is recirculated around the tower and that the clarifier is perfect so that the recirculation contains no biomass. The seventh assumption is that flow is applied uniformly across the entire tower cross-section and that the fluid flows as a thin sheet of constant thickness throughout the tower depth. The final assumption is that flow through the tower conforms to perfect plug–flow characteristics. All of these assumptions have important implications to the output from the

modeling effort and prevent the model from being used as a quantitative design tool. Nevertheless, the output is sufficiently qualitatively accurate to allow the major characteristics of packed towers to be observed and understood.

The effectiveness factor approach developed in Section 15.2.2 considered the biofilm to be flat with a characteristic thickness L_f. Because of the nature of the media in a packed tower, however, the surface of the biofilm is seldom flat. Consequently, at first glance it would appear that the entire effectiveness factor approach would be invalid for a packed tower. Careful investigation of the situation, however, reveals that this is not the case. The thickness of the active biofilm is on the order of 100 μm while the equivalent radius of the support media is greater than 30 mm.[10] Thus, the biofilm may be looked on as an annulus whose thickness is negligibly small in comparison to its total radius so that it may be approximated as a flat plate with the same thickness. In other words, even though the surface of the biofilm is not actually flat, it may be looked on as a flat plate, thereby allowing the mass balance equation for substrate within the biofilm to be written in planar geometry. Even if this flat plate approximation could not be used, however, the overall effectiveness factor could still be used because it has been found that the numerical values of the effectiveness factors for curved shapes like cylinders and spheres are very close to those of flat plates if the proper value of the characteristic length, L_f, is used.[1] The proper value for arbitrary shapes is:

$$L_f = \frac{\text{volume of biofilm}}{\text{external surface area for reactant penetration}} \tag{16.1}$$

Thus, even if the surface cannot be considered to behave exactly like a flat plate, proper choice of L_f will minimize the error, thereby allowing us to account for the effects of both internal and external mass transfer by using the overall effectiveness factor approach with values of η_{eO} as given by Figure 15.9.

16.1.2 External Mass Transfer

The flow patterns within a packed tower are very complex, reflecting interactions among fluid elements flowing over different support surfaces, variations in the cross-sectional area available for flow in random packing, irregularities caused by channeling, and short circuiting due to droplets falling from protrusions in the biofilm.[8] Because of these complex flow patterns it has been necessary to develop empirical correlations for the liquid phase mass transfer coefficient, k_L, within such towers.[15,18] One common form of correlation relates the mass transfer coefficient to the Schmidt number ($Sc = \mu_w/\rho_w D_w$) and the Reynolds number ($Re = v\rho_w d/\mu_w$), where μ_w is the fluid viscosity, ρ_w is its density, v is its bulk fluid velocity past the biofilm, D_w is the diffusivity of the substrate in water, and d is a dimension characterizing the media. Recognizing that the product of the velocity and the density of a fluid is the mass velocity, M, the Reynolds number is often written as Md/μ_w. Using this concept, Wilson and Geankoplis[22] reported the following correlations for mass transfer to liquids in packed beds:

$$k_L = (0.25\ M/\varepsilon\rho_w)Sc^{-0.67}Re^{-0.31} \qquad 55 < Re < 1500 \tag{16.2}$$

$$k_L = (1.09\ M/\varepsilon\rho_w)Sc^{-0.67}Re^{-0.67} \qquad 0.0016 < Re < 55 \tag{16.3}$$

These equations are restricted to $0.35 < \varepsilon < 0.75$, where ε is the void space between media elements as a fraction of the total bed volume, i.e, the porosity. Because the Reynolds number is directly proportional to M, Eq. 16.2 predicts that the mass transfer coefficient increases with $M^{0.69}$, when M is the superficial mass velocity based on the entire bed cross-sectional area normal to the direction of flow. Unfortunately, most plastic media used in packed towers for wastewater treatment have void fractions on the order of 0.95 and thus similar correlations are needed for predicting k_l in such systems. Dimensional analysis, as well as the above equations, suggest the use of a correlation of the form:

$$k_l = aM^b \tag{16.4}$$

where a depends on the properties of the fluid and the media while b depends on the flow range employed. The general utility of such correlations is an open question because mass transfer in biological packed towers is influenced strongly by the presence and characteristics of the biofilm,[5,7,11] which will depend to a large degree on the nature and concentrations of the substrates being degraded. However, it is clear now that mass transfer coefficients measured with clean media do not accurately reflect mass transfer in towers containing biofilms, and the needed studies will not be easy to perform. Nevertheless, the continued refinement of mechanistically based models for attached growth systems requires that they be done. In spite of the weakness in the mass transfer correlation used by Grady and Lim,[4] worthwhile information can be obtained from their simulations and they will be presented herein.

16.1.3 General Model

The schematic diagram of a packed tower is shown in Figure 16.1. A wastewater at flow rate F containing a soluble substrate at concentration S_{s0} flows to the tower where it is intermixed with recirculated effluent from the clarifier containing residual substrate at concentration S_{sc}, thereby diluting the influent concentration to S_{s1}, which is the applied concentration. The value of S_{s1} may be calculated from a mass balance on the mixing point:

$$S_{s1} = \frac{(S_{s0} + \alpha S_{sc})}{1 + \alpha} \tag{16.5}$$

where α is the fraction of the influent flow that is recirculated. The total flow rate through the tower is $F(1 + \alpha)$ and it is applied uniformly and continuously across the entire cross-sectional area, A_c. The cross-sectional area for fluid flow is $f_{Ac}A_c$, where f_{Ac} is the fraction of the tower cross-sectional area occupied by the liquid film. The wastewater is considered to be devoid of biomass, as is the recirculated flow. Thus, no biomass enters the tower. However, because the biofilm in the tower is considered to be at steady-state, biomass is continuously removed from its surface and carried down through the tower to the clarifier. This means that the concentration of biomass in suspension increases as the fluid moves through the tower, giving a concentration X_{Bc} in the flow entering the clarifier. The suspended biomass will contribute to substrate removal, although the attached biofilm will be responsible for most of it.

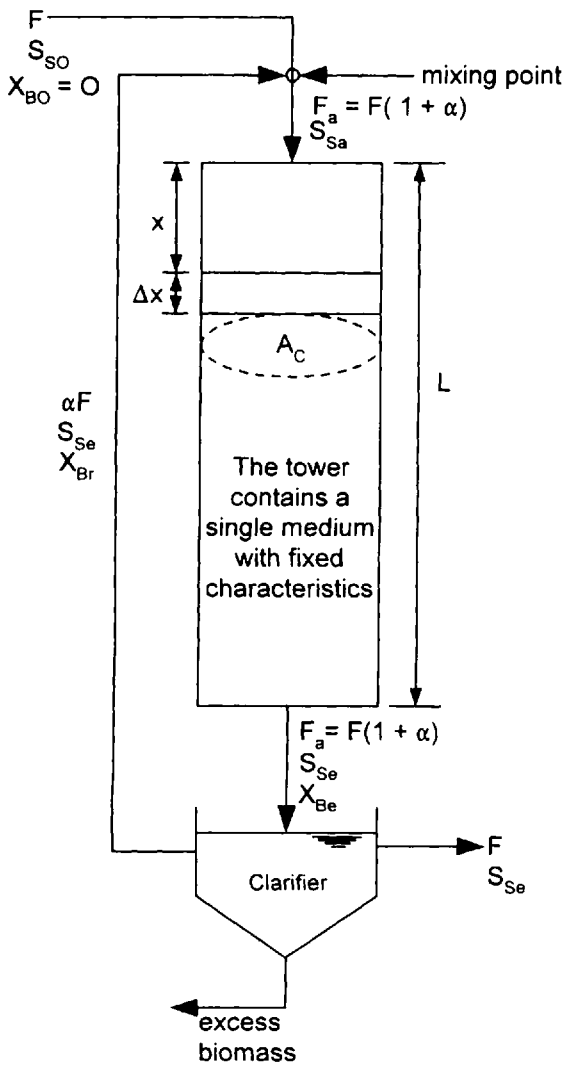


Figure 16.1 Schematic diagram of a packed tower with recirculation of clarified effluent.

Grady and Lim[4] assumed that the flow of fluid through a packed tower behaved in a plug–flow manner, with no vertical intermixing. We now know that intermixing occurs due to the flow of droplets from protrusions from the biofilm,[5] so the assumption of perfect plug–flow is not correct. The implications of this will be discussed later. Under the assumption of perfect plug–flow a steady-state mass balance on substrate can be written over an infinitesimal volume element of the packed tower as shown in Figure 16.1. Application of the usual limiting process and use of the overall effectiveness factor as presented in Section 15.2.2 to account for simultaneous transport and reaction leads to:

$$\frac{dS_{Sb}}{dx} + \eta_{eO}\left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}}\right)\left(\frac{a_S A_c}{F(1+\alpha)}\right) X_{BH} \cdot L_f + \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}}\right)\left(\frac{f_{Ac} A_c}{F(1+\alpha)}\right) Y_{Hobs}(S_{Sa} - S_{Sb}) = 0$$

$$S_{Sb} = S_{Sa} \text{ at } x = 0 \tag{16.6}$$

where a_s is the specific surface area (biofilm surface area per unit volume of tower), and Y_{Hobs} is an observed yield as used in conjunction with Eq. 15.19. The justification for its use was given with that equation. The second term in the equation represents substrate removal by the attached biomass whereas the third term is the removal by the suspended biomass. The performance of an ideal packed tower may be simulated by numerically integrating Eq. 16.6 using the overall effectiveness factor given by Figure 15.9 or an empirical correlation as discussed in Section 15.2.2. When many simulations are to be performed it is more convenient to use an empirical correlation because it can be substituted into Eq. 16.6 prior to integration. A technique must be used, however, that will allow proper choice of the correlation parameters, depending on the values of Bi, ϕ^2, and ϕ_f.

16.2 PERFORMANCE OF AN IDEAL PACKED TOWER

Predictions of ideal tower performance were obtained by Grady and Lim[4] by numerically integrating Eq. 16.6 with a fourth-order Runga–Kutta method using the parameter values in Table 16.1, unless noted otherwise. Equations of the form:

$$\eta_{eO} = c\phi_f^d \tag{16.7}$$

Table 16.1 Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used to Generate Figures 16.2–16.8

Symbol	Units	Value
$\hat{q}_H$	mg substrate COD/(mg biomass COD·hr)	0.26
K_S	mg/L as COD	30
Y_{Hobs}	mg biomass COD/mg substrate COD	0.50
k_f	cm/sec	$1.39[F(1+\alpha)/100\,A_c]^{0.5}$
D_e	cm^2/hr	0.139
X_{BH}	mg biomass COD/cm^3	32
L_f	cm	0.10
ε	—	0.35
a_s	cm^2/cm^3	1.0
f_{Ac}	—	0.35
L	cm	200
A_c	cm^2	200
F	L/hr	20
α	—	0
S_{SO}	mg/L as COD	300
X_{BO}	mg/L as COD	0

were used to allow estimation of the overall effectiveness factor as a function of Bi, ϕ^2, and ϕ_f. For situations in which the curves in Figure 15.9 were nonlinear, the curves were divided into sections within which the error in the use of Eq. 16.7 was minimal. Furthermore, because the value of the Biot number is affected by the feed flow rate, F, the recirculation ratio, α, and the cross-sectional area of the tower, A_c, interpolation formulas were developed to allow estimation of η_{eo} for any value of Bi between 0.1 and 10.0.

When there was no recirculation (α = 0), Eq. 16.6 was integrated directly starting with the known value of the applied substrate concentration, $S_{sa} = S_{so}$. When there was recirculation, however, the applied substrate concentration was not known because the effluent substrate concentration was unknown (see. Eq. 16.5). Therefore, an iterative procedure was used. First, the effluent substrate concentration, S_{se}, was guessed and used in Eq. 16.5 to calculate the applied substrate concentration, S_{sa}, needed to begin the integration of Eq. 16.6. At the end of the integration, the effluent substrate concentration was checked against the guessed value. If they did not agree, the process was repeated using as the next trial value the effluent substrate concentration just calculated. This numerical scheme was found to converge very rapidly, giving a better than 99.9% agreement within a few iterations.

Typical profiles of substrate concentration and overall effectiveness factor as functions of tower depth are shown in Figure 16.2 for a tower without recirculation. There it can be seen that the substrate concentration drops with tower depth in an exponential manner. This decrease in removal rate with depth is typical of packed towers and is due to the dependency of the rate on the substrate concentration through the Monod equation and to the declining effectiveness factor. The effectiveness factor also decreases in an exponential manner with tower depth for the two lower substrate concentrations. When the applied substrate concentration is high, on the other hand, it acts to keep the effectiveness factor large over a greater portion of the tower depth. Only after the substrate concentration has dropped considerably does the effectiveness factor start to decline in an exponential fashion. It should be noted that the curves for the highest feed substrate concentration do not tell the complete story. This is because the model considers only a single limiting nutrient and does not consider the possibility of oxygen limitations. In reality, application of such a high substrate concentration would result in oxygen limitations in the upper reaches of the tower, causing poorer performance than the model indicates.[7,11] In that case, substrate removal would be controlled by the rate of oxygen transfer, causing the same mass of substrate to be removed in each successive section of tower depth. This would cause the concentration to decrease linearly with depth with a smaller slope than shown in the figure.

One assumption in the model of Grady and Lim[4] is that the biofilm thickness is constant throughout the entire tower depth. Rittmann[17] performed simulations with a steady-state biofilm model and showed that the biofilm thickness declines with depth in a perfect plug–flow reactor because of the declining substrate concentration. While this finding does not alter the conclusion that the substrate concentration drops in an exponential manner with depth, it means that substrate removal will not be as great toward the bottom of a tower as Figure 16.2 suggests. Furthermore, for a steady-state situation the lowest attainable substrate concentration will approach S_{sbmin} as given by Eq. 15.22. Substrate concentrations below that value would not be able to sustain a biofilm.

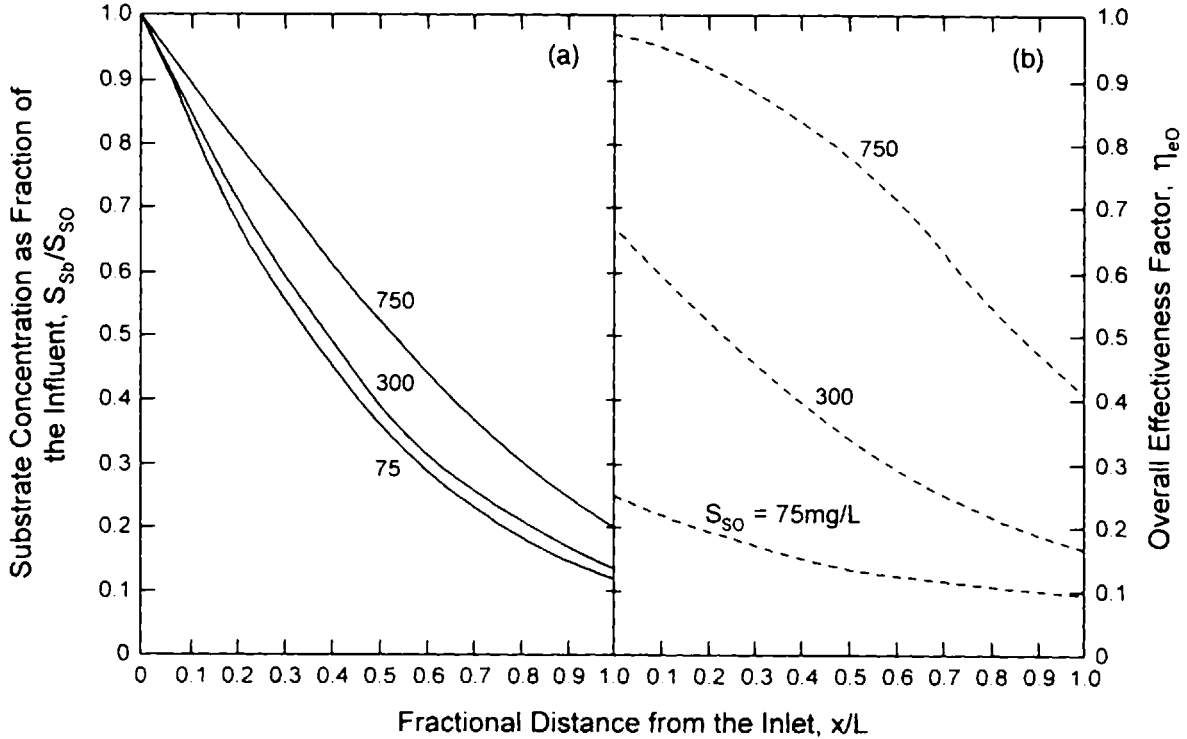


Figure 16.2 Effect of depth on the performance of a packed tower without recirculation receiving influents with different substrate concentrations: (a) effects on substrate concentration profiles; (b) effects on overall effectiveness factor profiles. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 16.1 unless otherwise specified.

Figure 16.3 shows the effect of influent substrate concentration, S_{SO}, on the removal of substrate by a tower without recirculation. The range of substrate concentrations is much greater than could be applied in practice without oxygen limitations or excessive biofilm buildup, but it is presented to illustrate some of the fundamental characteristics of towers. As the feed substrate concentration is increased, the fractional substrate removal decreases although the mass removal rate, $F(S_{SO} - S_{Se})$, increases. Because the tower has been assumed to have a constant and fixed biofilm thickness throughout, the amount of biomass in the tower is fixed. Consequently, the decrease in the fractional substrate removal is a direct result of the intrinsic kinetics of substrate removal for that fixed amount of biomass. At low substrate concentrations the fractional removal is almost constant as the feed substrate concentration is increased. This is because the Monod equation simplifies to a first order reaction with respect to substrate concentration under those conditions. Because mass transport is also first order and the amount of biomass is fixed, the substrate removal rate increases linearly with increases in substrate concentration, giving a constant fractional removal. As the feed substrate concentration is increased further, however, the Monod equation becomes zero order with respect to substrate, which means that the substrate mass removal rate becomes controlled entirely by

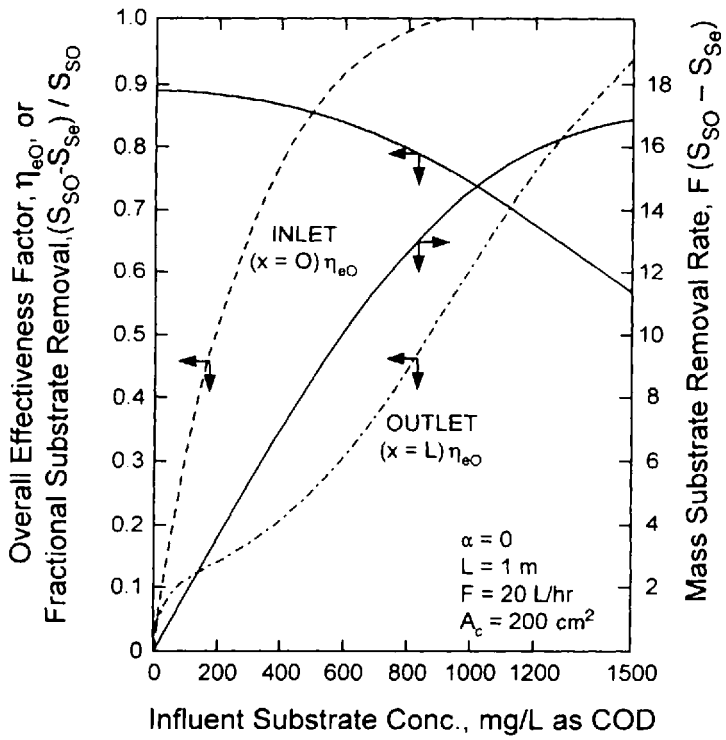


Figure 16.3 Effect of influent substrate concentration on the performance of a packed tower without recirculation. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 16.1 unless otherwise specified.

transport. Ultimately, the system will approach its maximum removal capability and the mass removal rate will reach a limit.

The effects of the feed substrate concentration on the overall effectiveness factor at the inlet and outlet are also shown in Figure 16.3. The inlet effectiveness factor approaches 1.0 at high substrate concentration, showing that transport is not limiting and that the biofilm is fully penetrated with substrate at a sufficiently high concentration to maximize the reaction rate. In addition, it can be seen that the variation in the effectiveness factor over the tower length declines as the feed substrate concentration is increased, showing that this condition prevails throughout much of the tower at higher feed concentrations.

The effects of influent flow rate are shown in Figure 16.4. Actually, as the influent flow rate is increased for a tower of fixed cross-sectional area, the superficial velocity of flow through the tower, which is the flow rate divided by the cross-sectional area, is increased. For biological packed towers this parameter is referred to as the total hydraulic loading, which will be given the acronym THL and the symbol Λ_H. In the general case, it is calculated by:

$$\Lambda_H = \frac{F(1 + \alpha)}{A_c} \tag{16.8}$$

All of the terms in Eq. 16.8 have been defined previously. The total hydraulic loading (THL) is an important design parameter for packed towers because minimum values must be attained to keep all of the media wet. The upper abscissa in Figure 16.4 shows the THL for the case considered. All of the values presented are below the minimum THL required for complete utilization of high-rate bundle media but are consistent with THLs for random media of the type considered in the simulations.

As the flow rate to the tower is increased the fractional substrate removal decreases whereas the mass substrate removal rate increases. This behavior is typical of plug–flow reactors containing biofilms. A higher flow rate brings a higher mass of substrate per unit time into contact with the biofilm. Since the mass of microorganisms in the biofilm is constant, the only way the mass rate of substrate removal can be increased is for the intrinsic specific rate to increase. Because the Monod equation defines the specific removal rate, this requires an increase in the bulk substrate concentration throughout the tower. Furthermore, because the Monod equation is nonlinear and approaches a maximum rate, the bulk substrate concentration must increase proportionally more than the feed concentration to bring about the required increase in the mass removal rate, thereby making the fractional removal decrease. The effects are offset somewhat by the effects of the increased THL on the external mass transfer rate. As indicated by Eq. 16.4, an increase in flow rate over the media will cause the external mass transfer coefficient to increase, thereby increasing the overall effectiveness factor. Because the influent substrate concentration is constant in Figure 16.4, the curve for η_{e0} at the tower inlet reflects these effects. The curve for the outlet, however, reflects changes in substrate concentration as well as changes in flow rate. Nevertheless, the increase in the overall effectiveness factor with its associated increase in the reaction rate at any point in the tower is not large enough to compensate for the required increase in substrate concentration or the decrease in residence time of the fluid in the tower, so the total fractional substrate removal in the tower is decreased. Because of these effects, greater flow rates and THLs gen-

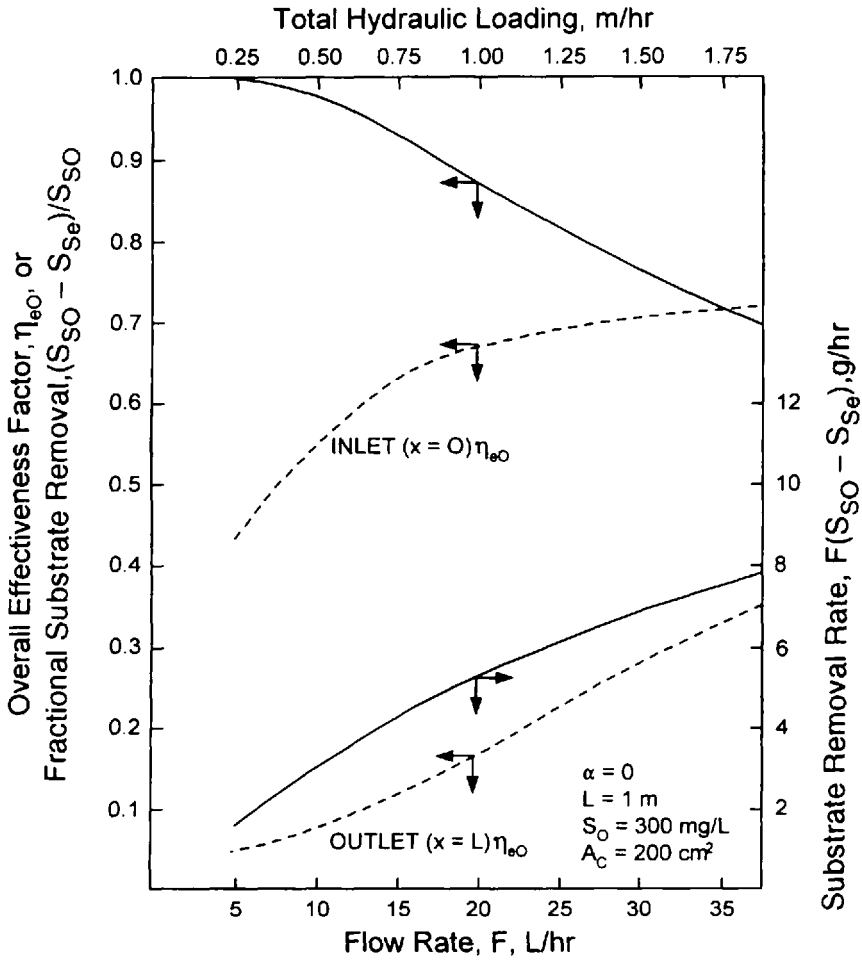


Figure 16.4 Effect of influent flow rate and the associated total hydraulic loading on the performance of a packed tower with fixed cross-sectional area without recirculation. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 16.1 unless otherwise specified.

erally require greater tower depths to achieve a fixed effluent concentration, as shown in Figure 16.5.

Recirculation of clarified effluent has a complicated effect on tower performance. First, depending on the amount, it will reduce the applied substrate concentration by dilution of the feed with the treated effluent, as indicated by the inlet concentrations in Figure 16.6a. It will also result in flatter substrate concentration profiles, which would be expected because a plug–flow reactor behaves more like a continuous stirred tank reactor (CSTR) as the recirculation ratio approaches infinity. Although not considered in the model of Grady and Lim,[4] recirculation also acts to provide a more uniform biofilm thickness throughout a tower.[17] The lower applied substrate concentration reduces the effectiveness factor at the tower inlet, although

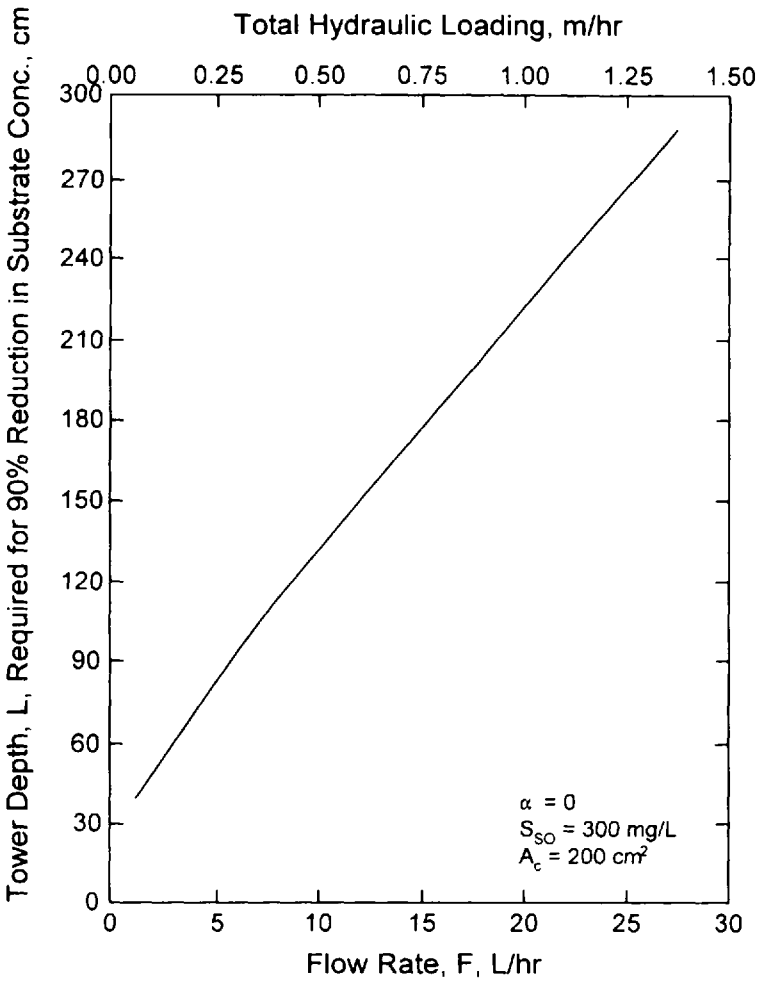


Figure 16.5 Effect of influent flow rate and the associated total hydraulic loading on the depth of packed tower with fixed cross-sectional area required to achieve 90% reduction in substrate concentration in the absence of recirculation. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 16.1 unless otherwise specified.

the impact is ameliorated somewhat by the increased flow rate through the tower, which increases the external mass transfer coefficient. Thus, the effectiveness factor at the inlet does not decrease as much as the applied substrate concentration. Furthermore, as a consequence of the increase in the external mass transfer coefficient, the effectiveness factor in the lower portions of the tower increases as the recirculation ratio is increased, as shown in Figure 16.6b. The decreased applied substrate concentration reduces the reaction rate. When this is coupled with the reduction in the effectiveness factor in the top of the tower, the result is less substrate removal in the upper portions. Furthermore, even though the increased flow rate increases the effectiveness factor in the lower portions of the tower, the increase cannot overcome

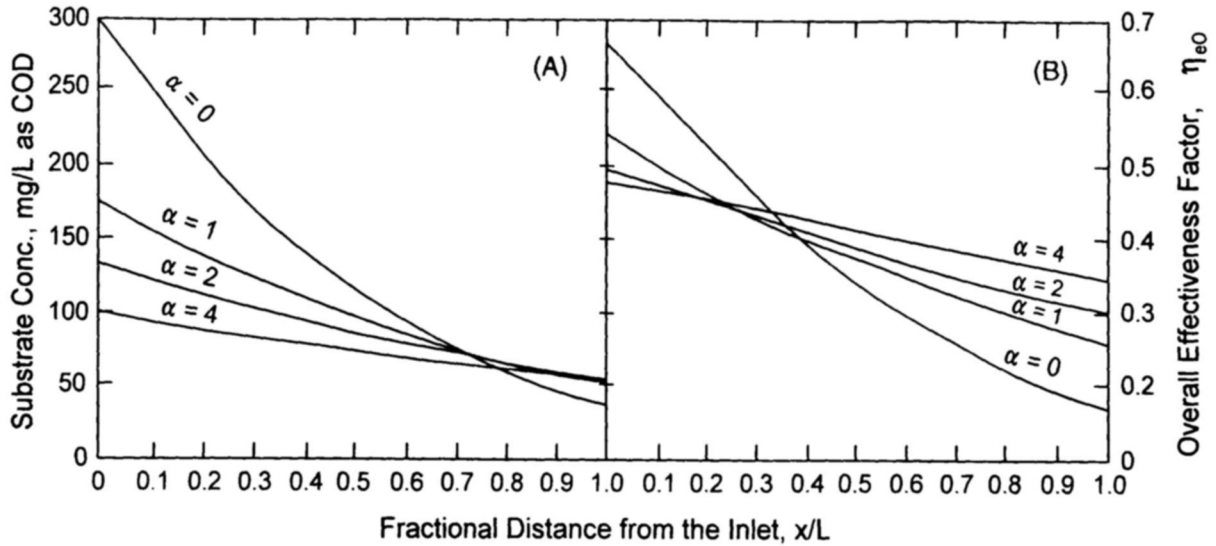


Figure 16.6 Effect of recirculation of effluent from a perfect clarifier on the performance of a packed tower of fixed size: (A) effects on substrate concentration profiles; (B) effects on overall effectiveness factor profiles. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 16.1 unless otherwise specified.

the reduced reaction rate due to dilution of the substrate, with the result that the output substrate concentration is higher from a tower with recirculation, as can be seen in Figure 16.6b.

A more detailed examination of the effect of recirculation on the effluent substrate concentration is presented in Figure 16.7. At low recirculation rates the decrease in reaction rate associated with the lower applied substrate concentration offsets the positive effect of recirculation on the mass transfer coefficient, causing the percent substrate removal to decrease. However, as the recirculation ratio is increased further, the effect of dilution eventually becomes minimal so that it is offset by the increase in the external mass transfer coefficient and the percent substrate removal increases slightly. Although these findings are consistent with that from other models with different assumptions,[12,17] no generalizations should be made about the magnitude of the effect of recirculation because it will depend on the feed flow rate to the tower as well as the mass transfer characteristics of the media. Thus, while recirculation will generally reduce the fractional removal of substrate across a tower, the degree of reduction will be system specific.

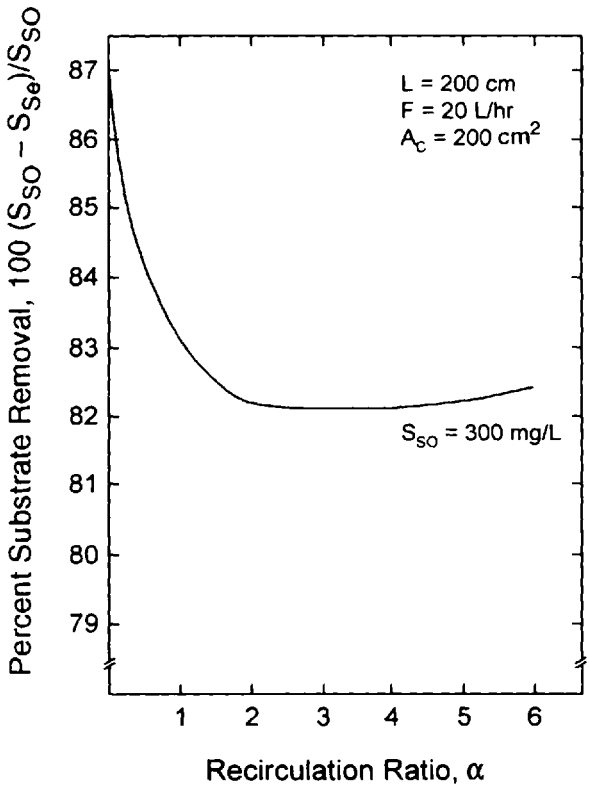


Figure 16.7 Effect of recirculation ratio on the ability of a packed tower of fixed size to remove substrate. The recirculation of effluent is from a perfect clarifier. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 16.1 unless otherwise specified.

Although the model results discussed above show that recirculation of clarified effluent will decrease substrate removal, circumstances exist in which recirculation could increase it. For example, if the feed substrate concentration was so high that oxygen transfer limited substrate removal, recirculation could decrease the problem by reducing the reaction rate and increasing the oxygen transfer rate. Furthermore, the presence of biomass in the recirculated flow can have an impact. The results in Figures 16.6 and 16.7 were obtained by assuming that the settler was perfect so that no biomass was present in the recirculation flow. It is possible, however, that if biomass had been present the reaction term for substrate removal by suspended organisms would have been large enough to make the effluent substrate concentration lower than it was without recirculation.[11] Thus, while it is true that recirculation generally reduces substrate removal through packed towers, one must not conclude that the effects of recirculation are always negative. Rather, each situation must be evaluated on a case by case basis.

During design of a packed tower for a given feed flow rate, an engineer may choose any cross-sectional area that gives a THL that is acceptable for the media under consideration. However, because of the impacts of THL on tower performance, different cross-sectional areas will require towers of different depth. These effects are illustrated in Figure 16.8 for a tower without recirculation. An increase in the cross-sectional area results in an increase in the surface area of biofilm per unit length of tower, and thus it is apparent that an increase in cross-sectional area should decrease the depth of tower required to remove a given fraction of substrate. However, because the increase in cross-sectional area also decreases the THL, the decrease in depth is not in proportion to the increase in the area so that the net result is an increase in tower volume. This happens because the decrease in THL decreases the external mass transfer coefficient, thereby decreasing the overall effectiveness factor. The results in Figure 16.8 suggest that the total media volume will be minimized by choosing a tall, thin tower rather than a short, fat one. Because a similar conclusion has been reached with other models,[9,12,17] as well as experimentally,[16,20] it appears to be general and would be expected to be true for other parameter values as well. It should be recognized, however, that the decrease in tower volume associated with increased tower depth may be less than shown in Figure 16.8 for other parameter values. Consequently, from a practical perspective, the effects of cross-sectional area and tower depth may not be significant in some situations.[12]

The model results presented herein are very useful for understanding the fundamental characteristics of packed towers. The complexity of such models and the difficulty of evaluating the parameters in them have prevented their widespread use in practice. Rather, most design of packed towers is based on empirical, rather than mechanistic, models. Consequently, in Section 16.4 we will briefly introduce such models. First, however, we need to consider several factors not included in the model presented here.

16.3 OTHER FACTORS NOT CONSIDERED IN MODEL

The model presented in Section 16.3 assumed that a packed tower acts as a perfect plug–flow reactor with no intermixing of fluid elements as the wastewater undergoing treatment flows downward through it. This is a simplification. In reality, two

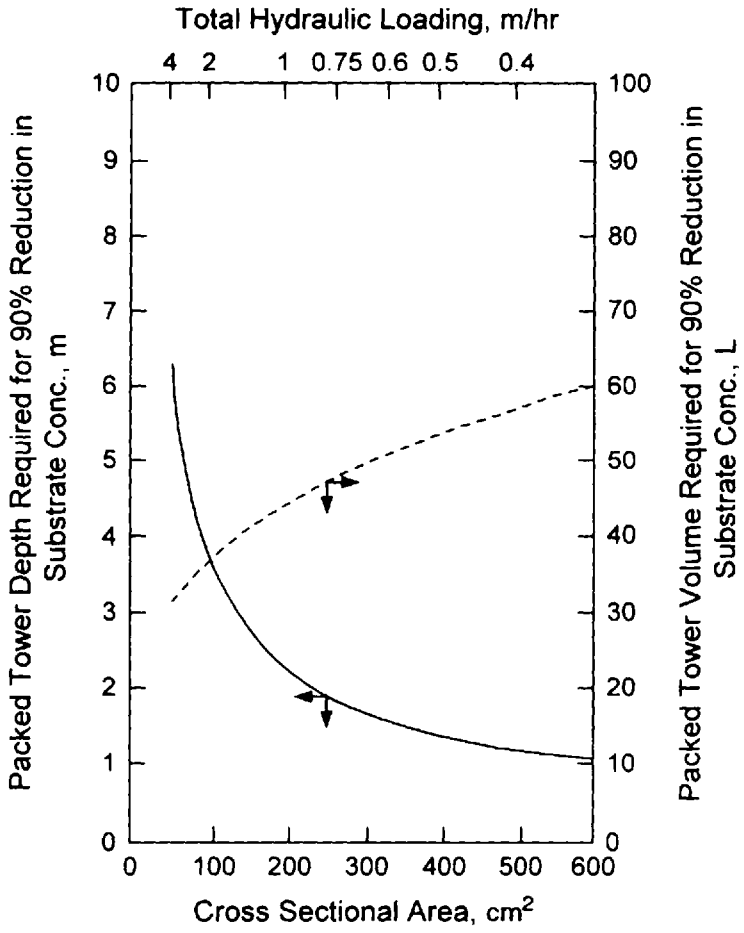


Figure 16.8 Effect of cross-sectional area and the associated total hydraulic loading at a fixed flow rate on the depth and volume of media in a packed tower without recirculation required to achieve a 90% reduction in substrate concentration. The values of the kinetic parameters, stoichiometric coefficients, and system variables are given in Table 16.1 unless otherwise specified.

factors act to cause significant degrees of intermixing, the nature of the biofilm and the type of media used for packing.

From observations of bundle media with a clear plastic front section, Hinton and Stensel[5] concluded that while sheet flow is the predominant type in that media, droplet flow is also important and leads to significant fluid intermixing. Droplet flow results from protrusions in the biofilm. Although the cause of protrusions is unknown, they act as sites of droplet formation, allowing droplets to bypass portions of the media before intersecting it and the sheet flow again. The points where the droplets hit are highly turbulent, resulting in significant intermixing. Thus, while the majority of the flow moves without intermixing, this behavior is frequently interrupted by fluid that has bypassed treatment, causing fluid with a higher substrate concentration

to be mixed back in. The net result is less substrate removal than predicted by models like the one in the preceding section.

Logan et al.[14] have attempted to incorporate the effects of media geometry into a model for packed towers. They maintain that different types of media have different flow patterns in them and that the type of media will influence the degree of intermixing that occurs. Some types of media will establish long flow paths over which the fluid flows in laminar flow. This allows the concentration gradient depicted in Figure 15.4 to be established and maintained over long distances. In contrast, other media have intersections of flow paths. Mixing occurs at those intersections, destroying the concentration gradient and establishing a new average concentration from which a new gradient is established. That gradient will then be maintained until the next mixing point. Modeling results have shown that these effects can cause differences in performance, with the media exhibiting intermixing giving better substrate removal.[14] Nevertheless, it should be pointed out that the improvement shown by intermixing is uncertain, particularly in light of the effects discussed in the preceding paragraph, which are likely to be highly stochastic. Thus, it is apparent that much more study of packed tower hydraulics is required before accurate mechanistic models can be written.

Another factor not considered in the model of Grady and Lim[4] is oxygen transfer and the potential for oxygen limitations in the biofilm. The implications of this were discussed earlier, but it would be worthwhile considering oxygen transfer further here. Oxygen transfer in biofilm systems cannot be studied accurately in the absence of biological activity and this has hampered development of oxygen transfer relationships.[7,11] There are two reasons for this. First, the biofilm itself influences the fluid flow patterns over the media, thereby preventing studies done on clean media from being representative of media containing biofilms.[5] Second, because the biological activity maintains a low oxygen concentration in the bulk liquid, oxygen transfer continues in situations where the water flowing over clean media would have become saturated. Furthermore, because of the stochastic nature of fluid flow over packed tower media, 10 to 25% of the total biofilm surface area can be devoid of a covering bulk liquid film at any instant.[5] This suggests that the oxygen concentration gradient into the biofilm can change drastically in a random manner, making models that assume a fixed gradient inaccurate. In spite of this complex situation, both theoretical[11] and experimental[7] studies have reached similar conclusions about the maximum oxygen transfer rate for bundle media, which is approximately 650 mg $O_2/(m^2 \cdot hr)$. Translation of this into a maximum substrate removal rate without oxygen limitations depends on the true growth yield (see Eq. 3.34), the kinetics of biodegradation, and the relative diffusivities of oxygen and the substrate, but it seems unlikely that removal rates in excess of 1400 mg COD/$(m^2 \cdot hr)$ could be handled without oxygen limitation and the potential problems associated with it. This suggests that at typical THLs, the maximum applied substrate concentration that would avoid oxygen transfer limitations is around 250 mg COD/L.[12]

Finally, the model of Grady and Lim[4] was limited to one type of bacteria growing on a single limiting electron donor. However, as we have seen throughout this book, both heterotrophic and autotrophic bacteria will grow simultaneously in the same culture when possible, each with its own electron donor, but competing for the electron acceptor. In addition, in packed towers, as in any biofilm process, the two types of bacteria must compete for space, as discussed in Section 15.4. This

competition has a significant effect on packed towers and causes carbon oxidation to occur in the upper portion of the tower, followed by nitrification in the lower portion. Reexamination of Figure 15.20 makes the reason for this behavior clear. For most wastewaters, the relative concentrations of COD and ammonia-N in the feed to a packed tower lie in the region marked H in that figure. Consequently, the heterotrophs displace the autotrophs in the upper portions of the tower and only COD removal will occur. As the fluid passes downward through the tower, the COD concentration is reduced relative to the ammonia-N concentration, resulting in a zone where both types of bacteria can exist, allowing simultaneous carbon oxidation and nitrification. Finally, further down the tower the concentrations of the two constituents will lie in region A of Figure 15.20 and nitrification will be the predominate reaction. While this sequential pattern is a common observation in practice, its existence can be influenced somewhat by recirculation, which reduces the concentration gradients in the tower, allowing simultaneous carbon oxidation and nitrification to occur over a greater length of the tower in some situations.

16.4 OTHER PACKED TOWER MODELS

It is apparent by now that a packed tower is a very involved biochemical operation that is difficult to model theoretically. Consequently, empirical and semiempirical approaches have been used to model it and a number of design equations have been developed. The NRC[16] and Gollar–Gotaas[3] equations are strictly empirical because they are based only on the application of regression analysis techniques to experimental observations. Consequently, they are valid only in the limited region over which the original data were available and can only be used for interpolation within that region. They cannot be used for extrapolation. Semiempirical models are also primarily interpolative, but can be successfully employed for extrapolation if the extrapolations do not extend them beyond the range over which their simplifying assumptions are valid. Finally, a truly mechanistic model can be used for extrapolation to new conditions, but care must still be exercised to be sure that extrapolations do not violate the basic assumptions of the model. In order to recognize the assumptions inherent in them we shall review some semiempirical and mechanistic models.

16.4.1 Velz Model

The model of Velz[21] states that the substrate concentration decreases in an exponential manner with depth, which is a characteristic we observed in Figure 16.2:

$$\frac{S_{Se}}{S_{Sa}} = \exp[-KL] \tag{16.9}$$

where K is a constant. Comparison of Eq. 16.9 with Eq. 16.6 shows that the Velz equation is limited to the situation in which the removal of substrate by suspended microorganisms is negligible, the substrate consumption rate is first order with respect to the substrate concentration ($S_{Sb} << K_S$), and the overall effectiveness factor is constant throughout the tower. Under these circumstances Eq. 16.6 can be integrated to yield:

$$\frac{S_{Se}}{S_{Sa}} = \exp\{-[(\hat{q}_H/K_S)a_vA_c\eta_{eO}X_{B,H}L_f/F(1 + \alpha)]L\} \tag{16.10}$$

from which we see that:

$$K = (\hat{q}_H/K_S)a_vA_c\eta_{eO}X_{B,H}L_f/F(1 + \alpha) = (\hat{q}_Ha_v\eta_{eO}X_{B,H}L_f)/(K_S\Lambda_H) \tag{16.11}$$

Thus, the Velz equation is limited to one THL rate.

16.4.2 Eckenfelder Model

Eckenfelder[2] has explicitly accounted for the effects of flow rate:

$$\frac{S_{Se}}{S_{Sa}} = \exp\left\{-K_1a_v^{1-m}[A_c/F(1 + \alpha)]^nL\right\} \tag{16.12}$$

where m and n are parameters that are dependent on the specific media used and K_1 is a rate coefficient. This equation is a modification of the Velz equation in which the effluent concentration is allowed to depend explicitly on F, A_c, α, and a_v. Therefore, we may look upon it as a limiting case of Eq. 16.6 in which the removal of substrate by suspended microorganisms is negligible, the substrate consumption rate is first order with respect to substrate ($S_{Sb} << K_S$), and the overall effectiveness factor remains constant throughout the tower but varies with the THL, $F(1 + \alpha)/A_c$. The last point becomes clear if we rearrange Eq. 16.12:

$$\frac{S_{Se}}{S_{Sa}} = \exp\left\{-K_1a_v^{1-m}[A_c/F(1 + \alpha)][A_c/F(1 + \alpha)]^{n-1}L\right\} \tag{16.13}$$

and compare it to Eq. 16.10. Thus, if η_{eO} is proportional to $[A_c/F(1 + \alpha)]^{n-1}$ they are equivalent for a given media. According to Liptak,[10] most media have an n value of 0.7 to 0.8. Consequently, a 28 fold increase in THL would result in a 1.95 to 2.72-fold increase in $[A_c/F(1 + \alpha)]^{n-1}$. Likewise, for the values of Bi, ϕ^2, and ϕ_f likely to be found in a packed tower, a 28-fold increase in THL will increase the overall effectiveness factor by a factor of 2.5. Thus, it appears that the changes in $[A_c/F(1 + \alpha)]^{n-1}$ in the Eckenfelder model are similar to the changes in the overall effectiveness factor caused by flow.

16.4.3 Kornegay Model

The differential equation used by Kornegay[9] is:

$$\frac{dS_{Sb}}{dx} = -K_2[A_c/F(1 + \alpha)]\left(\frac{S_{Sb}}{K_g + S_{Sb}}\right)$$

$$S_{Sb} = S_{Sa} \text{ at } x = 0 \tag{16.14}$$

where K_2 depends on the substrate and type of media, and K_g is a pseudo half-saturation coefficient which decreases asymptotically to K_S as the THL becomes very large. Equation 16.14 may be rewritten as:

$$\frac{dS_{Sb}}{dx} = -K_2[A_c/F(1 + \alpha)]\left(\frac{S_{Sb}}{K_S + S_{Sb}}\right)\left(\frac{K_S + S_{Sb}}{K_g + S_{Sb}}\right) \tag{16.15}$$

where K_S is the intrinsic half-saturation coefficient which remains invariant for a particular microorganism and substrate. If there is no substrate removal by suspended microorganisms, Eq. 16.6 reduces to:

$$\frac{dS_{Sb}}{dx} = -K_1[A_c/F(1 + \alpha)] \left(\frac{S_{Sb}}{K_S + S_{Sb}}\right) \eta_{eO}$$

$$S_{Sb} = S_{S,a} \text{ at } x = 0 \tag{16.16}$$

in which K_1 depends on the substrate and the type of media. Comparison of Eq. 16.15 with Eq. 16.16 reveals that they are equivalent for a given media if:

$$\eta_{eO} = \frac{K_S + S_{Sb}}{K_e + S_{Sb}} \tag{16.17}$$

Let us examine the requirement on η_{eO} in more detail. Inspection of Eq. 16.17 reveals that the overall effectiveness factor: approaches unity as the substrate concentration becomes very large; approaches unity when the THL becomes very large so that K_e approaches K_S; decreases as the substrate concentration decreases, approaching an asymptotic value of K_S/K_e as the substrate concentration approaches zero; and decreases as the THL is decreased, causing K_e to increase. These trends are consistent with those predicted from Figure 15.9. Thus, it appears that the Kornegay model can be interpreted as a limiting case of Eq. 16.6 in which substrate removal by suspended microorganisms is negligible and the overall effectiveness factor is given by Eq. 16.17.

16.4.4 Schroeder Model

Schroeder's[19] model states that:

$$\frac{dS_{Sb}}{dx} = -\left(\frac{\hat{q}_H S_{Sb}}{K_S + S_{Sb}}\right)\left(\frac{a_c A_c X_{B,H} L_f}{F(1 + \alpha)}\right) \eta_{el}$$

$$S_{Sb} = S_{S,a} \text{ at } x = 0 \tag{16.18}$$

where η_{el} is the internal effectiveness factor. Comparison of Eq. 16.18 with Eq. 16.6 reveals that the Schroeder equation is a limiting case in which both the removal of substrate by suspended microorganisms and the external mass transfer resistance are assumed to be negligible. The assumption of negligible external mass transfer resistance may only be valid when the THL is very high. Furthermore, Schroeder assumed that the internal effectiveness factor is directly proportional to the bulk substrate concentration. Figures 16.2 and 16.6 suggest, however, that this assumption is a poor one.

16.4.5 Logan et al. Model

Logan et al.[14] developed a model that attempts to account for the hydrodynamic properties of different packed tower media. The model assumes that the fluid flows in thin films that is generally laminar, allowing a parabolic velocity profile to be established. Transport of a single limiting substrate is assumed to occur across that liquid film in the absence of reaction in the film, i.e., there is no substrate removal by suspended microorganisms. The flux into the biofilm is given by an empirical

expression that assumes first order kinetics and Brownian collisions between the substrate molecules and the cells.[14] Although the rationale for the flux equation is not entirely clear, the model essentially assumes that substrate removal is limited by transport to the biofilm surface and not by the biofilm kinetics.[13] In other words, only external mass transport is assumed to be limiting, which is consistent with the experimental observations of others.[13] The unique feature of this model is that it considers the hydraulic characteristics of the media. In some types of media, the fluid is assumed to flow in thin films throughout the tower depth, allowing stable velocity and concentration profiles to be established. In others, however, intermixing is assumed to occur at regular intervals because of the geometry of the media, disrupting the velocity and concentration profiles at that point, thereby requiring them to be reestablished from the average concentration after intermixing. Although simultaneous transport of substrate and oxygen is not considered and only a single substrate is assumed to limit reaction, the capability is provided to allow computation of the maximum possible oxygen transport rate in the system. A computer code for the model suitable for microcomputers can be downloaded from the World Wide Web. Links to it can be found in the website of the Association of Environmental Engineering Professors, http://bigmac.civil.mtu.edu/aeep.html.

16.4.6 Hinton and Stensel Model

Hinton and Stensel[6] developed a model that considers transport of both substrate and oxygen into a biofilm, as well the effects of hydrodynamics in the tower. Consumption of substrate within the biofilm is conceptualized as a one-dimensional diffusion mass transport process with dual substrate limited kinetics. Rather than Monod kinetics, however, the model uses Blackman kinetics, which treats the reaction as being either first-order or zero-order with respect to the substrate and oxygen concentrations. The biofilm is assumed to be deep, with the active depth being determined by exhaustion of either the substrate or oxygen. The hydraulic characteristics of the tower are described as laminar liquid film flow which is interrupted at regular intervals by falling liquid drops. The interruptions provide mixing so that a uniform average concentration is established at the start of each laminar flow zone. Overall tower performance is determined by repeatedly solving the model equations for the short laminar flow sections and combining the results of many sections. The average section length is similar to the distance between mixing zones in the model of Logan et al.,[14] so although the models are conceptually different, their portrayals of the system hydrodynamics are similar in effect. Substrate transport is all through the laminar liquid film, but oxygen transport occurs both through the liquid film and through direct contact of the biofilm with the bioreactor gas phase when flow is shifted between paths. Predictions from the model were in agreement with experimental observations in the lab and in pilot plants.[6]

16.5 KEY POINTS

1. A packed tower may be modeled by assuming plug–flow and allowing the substrate to be consumed by both dispersed and attached microorganisms. However, to more accurately reflect reality, intermixing should oc-

cur at regular intervals to account for falling liquid droplets and intersecting flows within the media. Both internal and external mass transfer must be accounted for and this may be done by the effectiveness factor approach if only a single limiting nutrient is considered.

2. The substrate concentration drops exponentially with depth in packed towers either with or without recirculation of clarified effluent.
3. When there is no recirculation and when the feed substrate concentration is high, the effectiveness factor remains large over a portion of the tower depth and then drops exponentially thereafter with the tower depth.
4. As the feed substrate concentration to a packed tower is increased, the fractional substrate removal decreases, although the mass removal rate increases almost linearly until it approaches its maximum value. The effectiveness factor also increases and its variation over the tower depth declines.
5. As the feed flow rate to a packed tower is increased, the fractional substrate removal decreases whereas the mass substrate removal rate increases. The effectiveness factor increases with the flow rate but not sufficiently to offset the effect of a higher velocity so that faster flow rates generally require greater tower depths to achieve a fixed effluent concentration.
6. Higher recirculation ratios result in flatter substrate and effectiveness faster profiles in a packed tower. The percent substrate removal generally decreases as the recirculation ratio is increased, with the result that larger towers are required to achieve the same fractional substrate removal when recirculation is used.
7. An increase in the cross-sectional area of a packed tower results in a decrease in the depth required to remove a given fraction of substrate, but not in proportion to the increase in the area. Thus, the media volume required to achieve a given effluent substrate concentration increases as the cross-sectional area is increased.
8. Oxygen transfer can limit substrate removal in a packed tower. The maximum oxygen transfer rate is approximately 650 mg/(m^2·hr) and loadings that require more oxygen than that will not achieve expected levels of substrate removal. At typical THLs, the maximum allowable applied substrate concentration to avoid that limit is around 250 mg COD/L.
9. Because of competition for space and oxygen between heterotrophic and autotrophic bacteria, carbon oxidation generally occurs at the top of packed towers and nitrification near the bottom.
10. A number of semiempirical and mechanistic models are available for modeling the performance of biological packed towers in practice. Many are limiting case approximations of the model presented in Eq. 16.6.

16.6 STUDY QUESTIONS

1. Describe the assumptions inherent in the packed tower model of Grady and Lim,[4] telling how they affect the generalities about packed tower performance arrived at from simulations.

2. Describe the packed tower model of Grady and Lim,[4] giving its general characteristics.
3. Draw sketches showing typical substrate concentration and effectiveness factor profiles within a packed tower without recirculation and explain why the curves have the shapes that they do.
4. Draw a sketch showing the effect of the influent substrate concentration on the fractional substrate removal and the mass removal rate of substrate in a packed tower. Also explain why the mass removal rate curve has the shape that it does.
5. Draw a sketch showing the effect of the influent flow rate on the fractional substrate removal and the mass removal rate of substrate in a packed tower of fixed size. Also explain why the curves have the shapes that they do.
6. Describe and explain the effects of recirculation of clarified effluent on the fractional substrate removal in a packed tower as predicted by the model of Grady and Lim.[4] Also describe situations in which the effects could be different from the model predictions and explain why.
7. Draw a sketch showing the effect of packed tower cross-sectional area on the depth and volume of media needed to achieve a particular fractional substrate removal, and explain the effect.
8. Describe the factors that can prevent a packed tower from behaving as a perfect plug-flow reactor and explain their impact on tower performance.
9. Explain why nitrification typically follows carbon oxidation in a packed tower.

REFERENCES

1. Aris, R., On shape factors for irregular particles. I. The steady state problem. Diffusion and reaction. *Chemical Engineering Science* **6**:262–268, 1957.
2. Eckenfelder, W. W. Jr., Trickling filter design and performance. *Journal of the Sanitary Engineering Division, ASCE* **87**(SA6):33–40, 1961.
3. Galler, W. S. and H. B. Gotaas, Analysis of biological filter variables. *Journal of the Sanitary Engineering Division, ASCE* **90**(SA6):59–79, 1964.
4. Grady, C. P. L. Jr. and H. C. Lim, *Biological Wastewater Treatment: Theory and Applications*, Marcel Dekker, New York, 1980.
5. Hinton, S. W. and H. D. Stensel, Experimental observations of trickling filter hydraulics. *Water Research* **25**:1389–1398, 1991.
6. Hinton, S. W. and H. D. Stensel, A mechanistic model of substrate uptake and oxygen consumption in trickling filters. *Proceedings of the Water Environment Federation 66th Annual Conference and Exposition*, Volume 1, *Research*, pp. 295–306, 1993.
7. Hinton, S. W. and H. D. Stensel, Oxygen utilization of trickling filter biofilms. *Journal of Environmental Engineering* **120**:1284–1297, 1994.
8. Kehrberger, G. J. and A. W. Busch, The effects of recirculation on the performance of trickling filter models. *Proceedings of the 24th Industrial Waste Conference*, Purdue University Engineering Extension Series No. 138, 37–52, 1969.
9. Kornegay, B. H., Modelling and simulation of fixed film biological reactors for carbonaceous waste treatment. In *Mathematical Modelling for Water Pollution Control Processes*, T. M. Keinath and M. Wanielista, eds. Ann Arbor Science, Ann Arbor, Michigan, pp. 271–315, 1975.

10. Liptak, B. G., *Environmental Engineer's Handbook*, Volume I, Chilton Book, Randor, Pennsylvania, 1974.
11. Logan, B. E., Oxygen transfer in trickling filters. *Journal of Environmental Engineering* **119**:1059–1076, 1993.
12. Logan, B. E., S. W. Hermanowicz, and D. S. Parker, Engineering implications of a new trickling filter model. *Journal, Water Pollution Control Federation* **59**:1017–1028, 1987.
13. Logan, B. E., S. W. Hermanowicz, and D. S. Parker, Authors' response to discussion of 'A fundamental model for trickling filter process design' and 'Engineering implications of a new trickling filter model'. *Journal, Water Pollution Control Federation* **61**:364–366, 1989.
14. Logan, B. E., S. W. Hermanowicz, and D. S. Parker, A fundamental model for trickling filter process design. *Journal, Water Pollution Control Federation* **59**:1029–1042, 1987.
15. McCabe, W. L., J. C. Smith, and P. Harriott, *Unit Operations of Chemical Engineering*, Fifth Edition, McGraw-Hill, New York, 1993.
16. NRC Subcommittee Report, Sewage treatment at military installations, Chapter V. Trickling filters. *Sewage Works Journal* **18**:897–982, 1946.
17. Rittmann, B. E., Comparative performance of biofilm reactor types. *Biotechnology and Bioengineering* **24**:1341–1370, 1982.
18. Satterfield, C. N., *Mass Transfer in Heterogeneous Catalysis*, MIT Press, Cambridge, Massachusetts, pp. 79–128, 1970.
19. Schroeder, E. D., *Water and Wastewater Treatment*, McGraw-Hill, New York, p. 294, 1977.
20. Sorrels, J. H. and P. S. A. Zeller, Two-stage trickling filter performance. *Sewage and Industrial Waste* **28**:943–954, 1956.
21. Velz, C. J., A basic law for the performance of biological filters. *Sewage Works Journal* **20**:607–617, 1948.
22. Wilson, E. J. and C. J. Geankoplis, Liquid mass transfer at very low Reynolds numbers in packed beds. *Industrial and Engineering Chemistry Fundamentals* **5**:9–14, 1966.

17

Aerobic Growth of Heterotrophs in Rotating Disc Reactors

As shown in Table 1.2, one type of attached growth bioreactor is the rotating disc reactor (RDR). In it, closely spaced discs are mounted on a common horizontal shaft placed very near to or touching the liquid surface in a long narrow tank. The shaft is rotated at constant speed, thereby allowing any point on a disc to be alternately submerged and exposed to the atmosphere. When water containing organic matter, nitrogen, and other nutrients flows through the bioreactor, microorganisms consume the substrates and grow attached to the disc as a biofilm. The rotating action imparts a shear force to the biofilm, keeping its thickness relatively constant by removing the cells generated by consumption of the substrate. The turbulence generated by the rotation transfers oxygen to the bulk liquid and keeps the sloughed microorganisms in suspension so they can be carried out in the effluent. The most common arrangement of the discs is with the shaft perpendicular to the direction of liquid flow, as shown in Figure 17.1. Under those circumstances, the turbulence is sufficient to make the substrate concentration uniform throughout the tank. In other words, for all practical purposes the tank can be considered to be completely mixed and can be modeled as such. We saw in Chapter 7 that bioreactors arranged in series perform better than a single bioreactor of similar total volume. Because of this and because of the modular nature of RDRs, most applications use a series of bioreactors. Consequently, the performance of an RDR system can be modeled as a series of continuous stirred tank reactors (CSTRs) containing biofilms.

Just as in a packed tower, when the concentration of organic matter is high, nitrifying bacteria are unable to compete with the heterotrophic bacteria for space in the biofilm and thus the main reaction is carbon oxidation. However, once the concentration of organic matter has been reduced, nitrifiers can compete effectively for space and nitrification becomes significant. As a consequence, in systems containing a series of RDRs, carbon oxidation is the predominant reaction in the first few bioreactors while nitrification is more important in the later stages. In some cases, wastewater from which the organic matter has been removed in an upstream biochemical operation is applied to an RDR specifically for the oxidation of ammonia-N to nitrate-N, in which case the biofilm is composed almost entirely of nitrifying bacteria. Regardless of the nature of the reactions, however, the configuration of the RDR gives it characteristics that differ from that of a packed tower. In order to gain an appreciation for the differences and similarities of the two attached growth bioreactor types we will consider only growth of heterotrophic bacteria with a single limiting organic substrate using the model of Grady and Lim.[3] The reader should recognize, however, that nitrifying biofilms will behave in a similar manner.

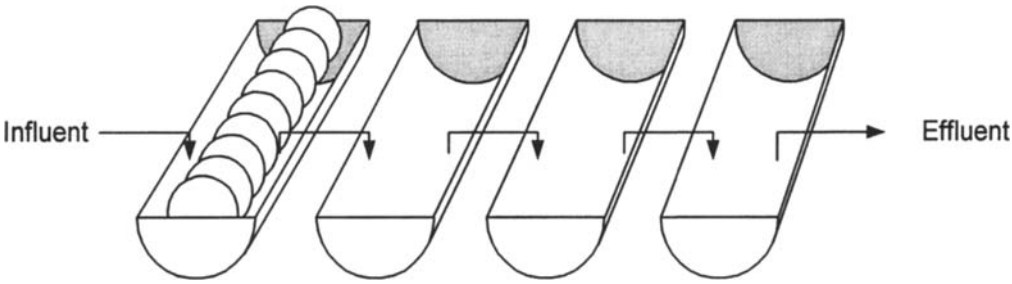


Figure 17.1 Schematic diagram of an RDR system.

17.1 MODEL FOR SUBSTRATE REMOVAL IN A SINGLE ROTATING DISC REACTOR

Because of the complex nature of an RDR, certain simplifying assumptions must be made to model it. The first is that steady-state conditions prevail so that microorganisms are sheared from the surface of the biofilm at a rate equal to their growth. Thus, the biofilm behaves as a steady-state biofilm of known thickness. That thickness is an input parameter and is not computed in the model. The second assumption is that the turbulence in the bioreactor fluid is sufficient to keep the detached biomass in suspension so that it can be washed out with the effluent. The third is that both the attached and detached microorganisms contribute to substrate removal. The fourth is that oxygen and other nutrients are present in excess so that the organic substrate is the growth limiting nutrient. In other words, dual limitation by both the electron donor and the electron acceptor in the biofilm is not considered. The fifth is that the thickness of the liquid film is uniform over the aerated sector of the disc. The final assumption is that the substrate concentration in the liquid film on the aerated sector depends only on the circumferential angle θ and not on the radial position. In other words, the liquid film in the aerated sector is treated as a plug flow reactor on top of the biofilm.

The effectiveness factor approach to modeling transport and reaction was used by Grady and Lim.[3] Consequently, they incorporated the effects of both internal and external mass transfer resistances into an overall effectiveness factor as discussed in Section 15.2.2. Although the surface of a disc is generally not flat, the diameter of the undulations is large in comparison to the thickness of the active biofilm so that the effectiveness factor for a flat biofilm should be applicable. However, because of the differences between the submerged and the aerated sectors of the disc, separate effectiveness factors are required for each.

17.1.1 External Mass Transfer

As shown in Figure 17.2, each disc can be divided into two sectors, submerged and aerated. Because the biofilm is attached to the disc, it moves through the bulk fluid in the submerged sector, thereby making the external mass transfer coefficient, $k_{l,s}$, a function of the rotational speed, ω. As a point on the surface of the disc leaves the submerged sector and enters the aerated sector, a thin film of liquid adheres to

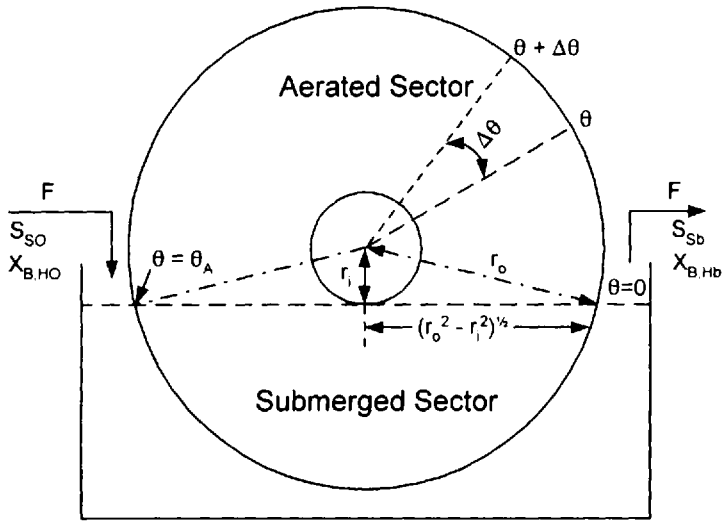


Figure 17.2 Schematic diagram of a single disc in an RDR system.

it and is carried along with it. Although this film can be assumed to have no motion relative to the biofilm on the disc, its thickness is a function of the rotational speed of the disc and consequently, the mass transfer coefficient for the substrate within it, $k_{l,s}$, also depends upon the speed.

Mass transfer from a fluid in laminar flow to the surface of a submerged rotating disc was analyzed by von Karman and given by Levich:[10]

$$k_{l,s} = 1.55\, D_w^{0.667}(\mu_w/\rho_w)^{-0.167}(\omega/r)^{0.50} \tag{17.1}$$

where $k_{l,s}$ is the submerged external mass transfer coefficient, r is the radius of the disc, D_w is the diffusivity of the substrate in water, ρ_w is the fluid density, μ_w is the fluid viscosity, and ω is the rotational speed of the rotating disc. Equation 17.1 indicates that the external mass transfer coefficient will increase with the square root of the rotational speed. In practice, however, both the proportionality constant and the power on the rotational speed may be different due to deviations from the assumptions made in deriving the equation. Nevertheless, it is possible to obtain correlations of the form:

$$k_{l,s} = e \cdot \omega^f \tag{17.2}$$

where e and f are coefficients with e depending on the physical properties of the fluid and radius of the disc.

In the aerated sector, the entrained fluid forms a stagnant layer on top of the biofilm. Thus, it will be assumed that the external mass transfer coefficient in the aerated sector, $k_{l,a}$, is equal to the diffusivity divided by the thickness of the stagnant liquid film as given by Eq. 15.3. Thus, quantification of that mass transfer coefficient requires knowledge of the liquid film thickness. The thickness of the liquid film entrained on a flat plate withdrawn vertically from a quiescent liquid has been analyzed by Landau and Levich[9] and found to be:

$$L_w = \alpha v^{0.667} \tag{17.3}$$

where L_w is the film thickness, v is the withdrawal velocity, and α is a parameter that is dependent on the fluid properties. Since the withdrawal velocity of a point on a rotating disc depends on its radial position, some average velocity should be used, such as:

$$v = \pi\omega\left(\frac{r_o + r_i}{2}\right) \tag{17.4}$$

where r_o and r_i are the outer and inner radii of the submerged sector, respectively, as shown in Figure 17.2. This suggests that the average film thickness is given by:

$$L_w = \alpha_1 \cdot \omega^{0.667} \tag{17.5}$$

where α_1 is a coefficient whose value depends on the fluid properties, the size of the disc, and its degree of submergence. Hartman[7] has reported a stagnant liquid film thickness of 40 μm for a smooth rotating disc and Eq. 17.5 predicts a thickness of 60 μm under the conditions studied. Grieves,[4] on the other hand, reported that the thickness of the stagnant liquid layer on top of a rotating biofilm ranged from 50 to 200 μm and was not reproducible. Such variability is due to the surface film depicted in Figure 15.1. Therefore, it is necessary to add an arbitrary amount to the thickness predicted by Eq. 17.5 to account for the retention of fluid by the surface biofilm. Hence, a more appropriate form might be:

$$L_w = \alpha_2 + \alpha_3 \cdot \omega^{\alpha_4} \tag{17.6}$$

where α_2, α_3, and α_4 must be determined experimentally. This expression for the stagnant film thickness can be substituted into Eq. 15.3 to obtain an expression depicting the effect of rotational speed on the external mass transfer coefficient for the aerated sector, $k_{L,a}$:

$$k_{L,a} = \frac{D_w}{\alpha_2 + \alpha_3\omega^{\alpha_4}} \tag{17.7}$$

It is important to recognize that $k_{L,s}$ and $k_{L,a}$ are influenced differently by the rotational speed of the disc. As Eq. 17.2 shows, the external mass transfer coefficient in the submerged sector will increase as the rotational speed is increased. However, as seen with Eq. 17.7, the external mass transfer coefficient in the aerated sector will decrease.

Once the external mass transfer coefficients have been estimated for each of the sectors, the overall effectiveness factors can be determined from Figure 15.9. They will not be the same, in part because of the differences between $k_{L,s}$ and $k_{L,a}$. Consequently, $\eta_{eO,s}$ and $\eta_{eO,a}$ are used to denote the overall effectiveness factors for the submerged and aerated sectors, respectively. The substrate consumption rates per unit area of biofilm for each of the sectors are given by Eq. 15.16 with the appropriate values for the overall effectiveness factors.

17.1.2 General Model

First let us consider the case of a single bioreactor as shown in Figure 17.2. Taking the liquid volume in the tank as the control volume, V, we see that the substrate is brought in by two streams: the influent flow, F, and the liquid film on the aerated sector of each disc, F_L. Two streams also comprise the output: the effluent flow, F,

and the liquid film entrained by the rotating discs, F_L. Substrate is removed by the biofilm within the submerged sector and by suspended microorganisms. Thus, the steady-state mass balance equation on substrate in the tank is:

$$F \cdot S_{SO} + F_L \cdot S_{SLR} - F \cdot S_{Sb} - F_L \cdot S_{Sb} - \eta_{eOS} \cdot X_{B,Hb} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \right) A_s - X_{B,Hb} \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \right) V = 0 \tag{17.8}$$

which yields:

$$\frac{1}{\tau} = \eta_{eOS} \cdot X_{B,Hb} \cdot L_f \left[\frac{\hat{q}_H \cdot S_{Sb}}{(K_S + S_{Sb})(S_{SO} - S_{Sb})} \right] \frac{A_s}{V} + Y_{Hobs} \left(\frac{\hat{q}_H \cdot S_{Sb}}{K_S + S_{Sb}} \right) - \frac{F_L}{V} \left(\frac{S_{SLR} - S_{Sb}}{S_{SO} - S_{Sb}} \right) \tag{17.9}$$

when Eq. 15.19 is substituted for $X_{B,Hb}$. In these equations η_{eOS} is the overall effectiveness factor for the biofilm in the submerged sector, S_{Sb} is the substrate concentration in the tank and in the entrained liquid film at the point of entrainment, S_{SLR} is the substrate concentration in the entrained liquid film at the point of return, and A_s is the total surface area in the submerged sector. The other terms retain their usual meanings from Chapter 15. The volumetric flow rate of the entrained liquid film is given by:

$$F_L = 2N\pi(r_o^2 - r_i^2)L_s\omega \tag{17.10}$$

where N is the number of discs in the bioreactor and therefore takes on integer values only. If we ignore the edges of the discs the total surface area in the submerged sector is (See Figure 17.2):

$$A_s = 2N \left[\frac{r_o^2(2\pi - \theta_A)}{2} - r_i(r_o^2 - r_i^2)^{0.5} \right] = N(2\pi r_o^2) \left\{ \cos^{-1} \left(\frac{r_i}{r_o \cdot \pi} \right) - \frac{r_i}{r_o \cdot \pi} \left[1 - \left(\frac{r_i}{r_o} \right)^2 \right]^{0.5} \right\} \tag{17.11}$$

where $2\pi r_o^2$ is the total surface area of one disc and the expression in the braces, {}, represents the fractional submergence of the disc. Equation 17.11 can be substituted into either Eq. 17.8 or Eq. 17.9, thereby making the performance of the RDR a function of the number, size, and degree of submergence of the discs.

Equations 17.8 and 17.9 both contain a term for the concentration of substrate in the liquid film being returned to the tank by the rotation of the disc. Thus, before either of them can be solved, information must be available about that concentration. As the disc rotates through the aerated sector the microorganisms in the biofilm remove substrate from the entrained liquid film, thereby making S_{SLR} less than S_{Sb}. Furthermore, since the entrained film acts like a plug–flow reactor, S_{SLR} will be dependent on S_{Sb}. This dependency can be shown by writing a steady-state mass balance equation for substrate in a differential element of the entrained film as shown in Figure 17.2. When performing the balance the assumption is made that the substrate concentration at any point on the aerated sector of the disc, S_{SL}, is a function

of only θ and is independent of the radial position of the point. Ignoring the removal of substrate by organisms suspended in the liquid film, the mass balance is:

$$\left(\frac{F_l}{2N}\right) S_{sl}\Big|_{\theta} - \left(\frac{F_l}{2N}\right) S_{sl}\Big|_{\theta+\Delta\theta} - \eta_{eO_2} \cdot X_{BHB} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{sb}}{K_s + S_{sb}}\right) \frac{(r_o^2 - r_i^2)\Delta\theta}{2} = 0 \tag{17.12}$$

where η_{eO_2} is the overall effectiveness factor for the biofilm in the aerated sector. The usual limiting process leads to:

$$\frac{dS_{sl}}{d\theta} = -\eta_{eO_2} \cdot X_{BHB} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{sb}}{K_s + S_{sb}}\right) \frac{(r_o^2 - r_i^2)N}{F_l} \tag{17.13}$$

Substitution of Eq. 17.10 for F_l gives:

$$\frac{dS_{sl}}{d\theta} = -\eta_{eO_2} \cdot X_{BHB} \cdot L_f \left(\frac{\hat{q}_H \cdot S_{sb}}{K_s + S_{sb}}\right)\left(\frac{1}{2\pi L_\alpha \omega}\right) \tag{17.14}$$

Because of the assumption that the substrate concentration in the liquid film is a function of θ alone, the boundary condition can be approximated by:

$$S_{sl} \approx S_{sb} \text{ at } \theta = 0 \tag{17.15}$$

and the concentration in the returning liquid can be approximated by:

$$S_{slR} \approx S_{sl} \text{ at } \theta = \theta_A \tag{17.16}$$

Thus, Eqs. 17.9 and 17.14 are the mass balance equations that describe the removal of substrate by the RDR and they must be solved simultaneously. For a given set of reactor conditions the unknowns are S_{slR}, S_{sb}, and η_{eO_2} in Eq. 17.9 and S_{sb} and η_{eO_2} in Eq. 17.14. As we saw in Section 15.2.2, however, η_{eO_2} and η_{eO_2} are functions of S_{sb} as well as the physical and operational characteristics of the bioreactor. The only difference is that η_{eO_2} depends on $k_{L\alpha}$ while η_{eO_2} depends on k_{LS}. Thus, the value of η_{eO_2} is expected to decrease with increasing rotational speed while that of η_{eO_2} is expected to increase.

Because η_{eO_2} is a function of S_{sb}, Bi, and ϕ^2, Eq. 17.14 cannot be integrated analytically. Rather, it must be integrated numerically as discussed in Section 15.2.2. Equation 17.9 is a nonlinear algebraic equation and thus a numerical technique is applicable to it also. It must be recognized, however, that these two equations are coupled by the boundary conditions, Eqs. 17.15 and 17.16.

If the kinetic and mass transfer parameters are known, Eqs. 17.9 and 17.14 can be used to calculate the substrate concentration remaining in a bioreactor with known characteristics, although an iterative approach must be used. Because η_{eO_2} and η_{eO_2} are functions of S_{sb}, it is convenient to iterate on it. A value of S_{sb} is assumed and used to estimate η_{eO_2}, which is then used in Eq. 17.14 to obtain the value of S_{sl} at the point of return. The assumed value is also used to estimate η_{eO_2}, which is then used along with S_{slR} to calculate a new value for S_{sb} with Eq. 17.9. The iteration ends when the calculated value of S_{sb} is the same as the assumed one.

17.2 PERFORMANCE OF A SINGLE ROTATING DISC REACTOR

Predictions of the performance of a single, ideal rotating disc reactor were obtained by numerically solving a slightly simplified form of the equations given above using the parameter values given in Table 17.1.[3] To allow the use of Figure 15.9 in the numerical techniques, equations were used to allow estimation of η_{eOs} as a function of Bi, ϕ^2, and ϕ_f. The forms of those equations are also given in Table 17.1. Because the substrate concentration in the aerated sector is usually small, the substrate removal rate was approximated as a first order reaction, which allowed η_{eOa} to be approximated by:

$$\eta_{eOa} = \frac{1}{\frac{\hat{q}_H}{K_S k_{La}} + \frac{\phi}{\tanh \phi}} \tag{17.17}$$

using Eq. 17.7 for k_{La} and the expression given in Table 17.1 for L_∞.

The effects of flow rate on the removal of substrate in an RDR with a fixed number of discs rotating at a fixed speed are shown in Figure 17.3. As would be anticipated from Figures 15.10 and 15.11 in which the effects of the hydraulic residence time (HRT) on the performance of a CSTR were presented, increases in flow-rate (which cause a decrease in the HRT) cause an increase in the effluent

Table 17.1 Kinetic Parameters, Stoichiometric Coefficients, and System Variables Used to Generate Figures 17.3–17.11

Symbol	Units	Value
$\hat{q}_h$	mg substrate COD/(mg biomass COD·hr)	2.0
K_S	mg/L as COD	50
Y_{Hobs}	mg biomass COD/mg substrate COD	0.50
k_{La}	cm/sec	$0.1863\ (\omega/r_{t,})^{0.5}$
D_c	cm^2/hr	0.02
L_∞	cm	$2.144 \times 10^{-3}[\omega(r_{t,} + r_t)/2]^{0.667}$
X_{BAH}	mg biomass COD/cm^3	50
L_f	cm	0.01
V	cm^3	3.0×10^6
N	—	44
r_o	cm	150
r_i	cm	30
ω	rev/min	2.0
F	cm^3/hr	3.0×10^6
S_{so}	mg/L as COD	100
η_{eOs}		
at Bi = 1	$0.66\phi_f^{1.864}$	
at Bi = 5	$0.1987\phi_f^{1.869} + 0.832\phi_f^{1.3681}$; $1.65 < \phi_f < 10$	
	$0.1987\phi_f^{1.869} + 0.7045\phi_f^{1.0362}$; $1.25 < \phi_f < 1.65$	
at Bi = 10	$1.1904\phi_f^{1.3681}$	

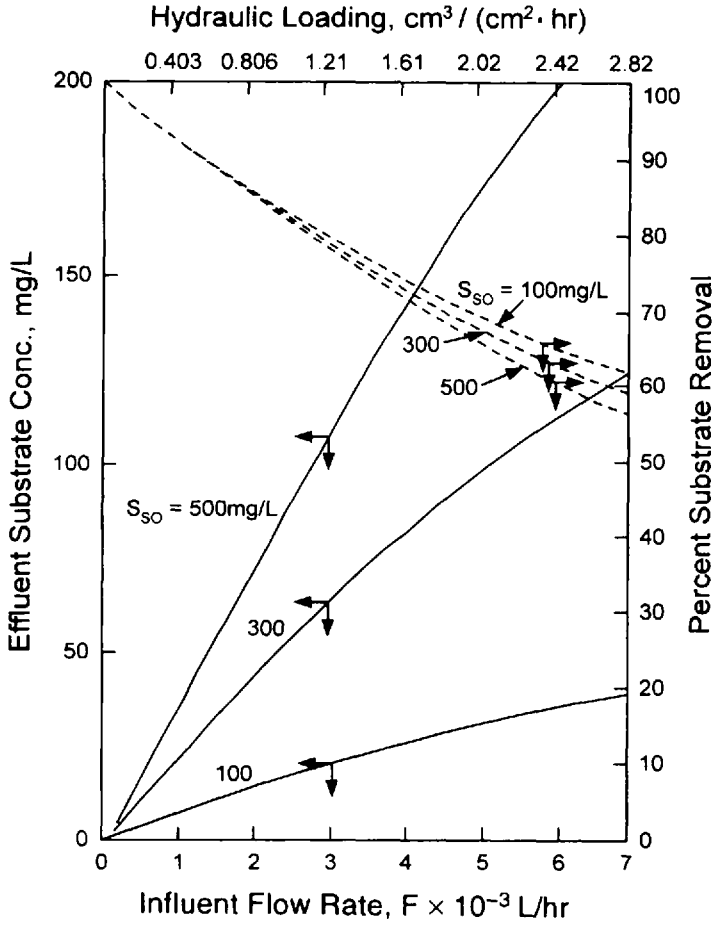


Figure 17.3 Effects of influent flow rate and influent substrate concentration on the performance of a single RDR.

substrate concentration and a decrease in the percent removal. As with a packed tower, this follows from the fact that the mass of biomass in the reactor is constant, so that the rate of application of substrate per unit of biomass is increased as the flow-rate is increased. Consequently, a higher substrate concentration must surround the biomass to drive the removal reactions at a faster rate. In an RDR the hydraulic loading is a commonly used indicator of performance. It is defined as the flow per unit of total wetted surface area and is shown across the top of Figure 17.3 to give an indication of its effect on performance.

The effects of influent substrate concentration are also shown in Figure 17.3. Examination of the curves shows that the effluent substrate concentration will increase as the influent substrate concentration is increased. As was noted for increases in flow-rate, increases in the influent substrate concentration increase the mass of organic matter applied per unit of biomass. As a consequence, the substrate concen-

tration surrounding the biomass must increase to drive the removal reactions at a faster rate, thereby increasing the effluent substrate concentration. It is interesting to note, however, that the percentage substrate removal is relatively independent of the influent substrate concentration at low hydraulic loadings. This is due to the fact that at low hydraulic loadings, the effluent substrate concentrations are low relative to the value of the half-saturation coefficient (K_S = 50 mg/L as COD) so that the Monod kinetics reduce to first-order kinetics, thereby making the percentage substrate removal independent of the influent substrate concentration. At higher hydraulic loadings, the effluent substrate concentrations are significant relative to the half-saturation coefficient so that Monod kinetics prevail. Consequently, the percentage substrate removal decreases as the influent substrate concentration is increased, just as it does in other attached growth bioreactors.

Figure 17.4 shows the effect of rotational speed on the performance of an RDR. As the rotational speed is increased, the percent substrate removal increases up to an upper limit characterized by the other system parameters. Several factors interact to cause this response. First, as shown in Eq. 17.1, the external mass transfer coefficient in the submerged sector increases as the rotational speed of the discs is increased, thereby increasing the overall effectiveness factor. Consequently, substrate is consumed more rapidly by the submerged biofilm as the rotational speed is increased. The events occurring in the aerated sector are more complicated, however. Examination of Eq. 17.3 reveals that an increase in rotational speed will cause the thickness of the liquid film on the aerated sector to increase, which will make the mass transfer coefficient in the liquid film, $k_{l,a}$, decrease (see Eq. 17.7). The decrease

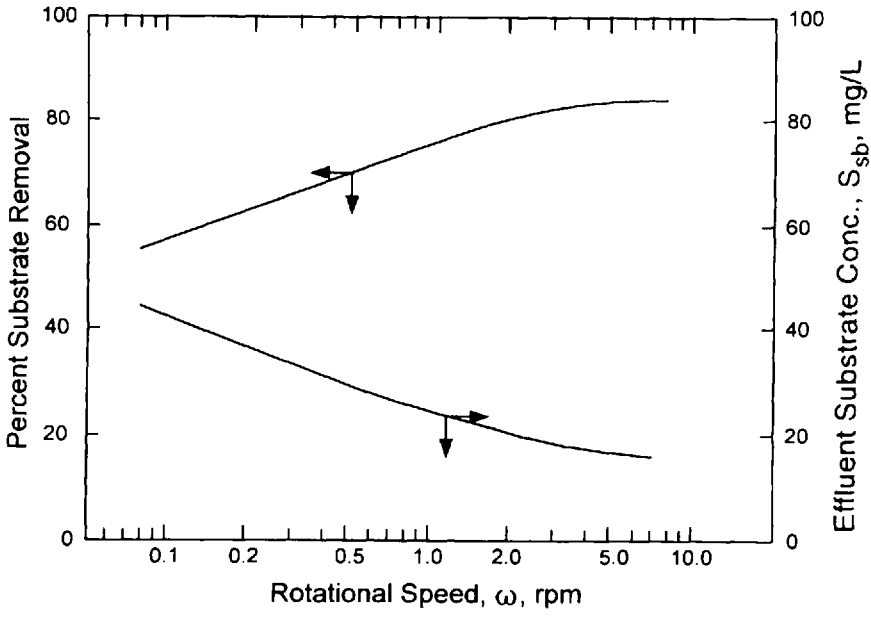


Figure 17.4 Effects of rotational speed on the performance of a single RDR.

in $k_{l,a}$ will decrease the overall effectiveness factor in the aerated sector, thereby reducing the rate of substrate removal and increasing the substrate concentration in the liquid returning to the bulk fluid from the aerated sector. The increase in the thickness of the liquid film, on the other hand, causes the volume of fluid carried with the discs into the aerated sector to increase as the rotational speed is increased. As a consequence, even though the concentration of substrate returning to the bulk liquid increases as the rotational speed is increased, the mass of substrate removed per unit time by the aerated sector increases slightly. Hence, the rate of substrate removal by both the submerged and aerated sectors increases as the rotational speed is increased, with the result that the effluent substrate concentration will decrease, as shown in Figure 17.4. There it can be seen that the effect is more significant when the rotational speed is small, but decreases as the speed is increased. Apparently, at low rotational speeds the resistance to mass transfer in the submerged sector is quite important so that a decrease in it makes a significant improvement in bioreactor performance. Beyond a rotational speed of 2 rpm, however, the mass transfer resistance is not as important and hence the performance does not improve significantly with further increases in rotational speed. It should be noted that very high rotational speeds can cause excessive shear of biomass from the discs, thereby decreasing

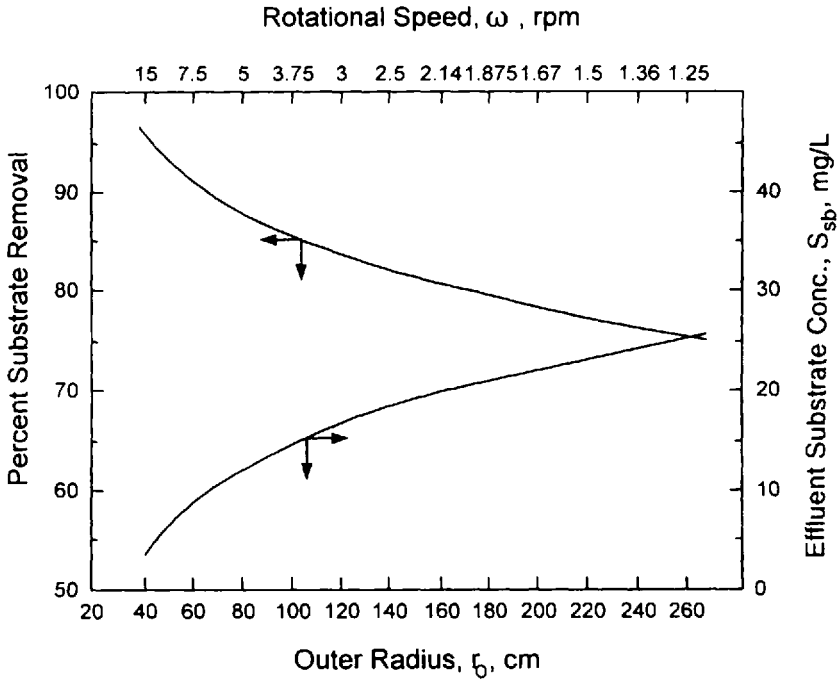


Figure 17.5 Effect of disc size on the performance of a single RDR with constant hydraulic loading and constant peripheral velocity.

performance. This possible effect is not reflected in the model, which assumes a constant value for the biofilm thickness, L_f.

Scale-up of RDRs is complicated because of the many factors involved. One approach that found use early in the development of RDRs was to maintain a constant peripheral velocity while maintaining the same hydraulic loading, probably as a result of recommendations based on lightly loaded systems.[2] To investigate the efficacy of this practice, simulations were performed to investigate the effect of disc size while maintaining the peripheral velocity and the hydraulic loading constant as the disc diameter was increased, and the results are presented in Figure 17.5. Maintenance of a constant peripheral velocity required that the rotational speed be decreased as the disc diameter was increased, so the rotational speed corresponding to each disc size is shown on the upper abscissa. Examination of Figure 17.5 shows clearly that the scale-up strategy is not effective since the percent substrate removal decreases as the disc size is increased, an observation that has been made in practice as well.[13] This is a result of two effects. First, because the rotational speed is decreased to maintain a constant peripheral velocity as the disc size is increased, the external mass transfer coefficient in the submerged sector is decreased. This has the effect of decreasing the overall effectiveness factor for that sector, reducing the substrate removal rate. Second, the decrease in rotational speed reduces the volume of liquid carried through the aerated sector relative to the influent flow rate, which decreases the mass of substrate removed there, in spite of the increase in the external mass transfer coefficient in the aerated sector associated with the decrease in the liquid film thickness, L_A. Consequently, the loss in performance associated with an increase in disc size is primarily due to the effect of rotational speed on the rates of mass transfer in both sectors and the movement of liquid through the aerated sector. In addition to the problems discussed above, several other effects complicate the problem of scale-up.[11] As a result, and because no suitable scale-up strategy has been found, it is recommended that pilot studies be performed with full-scale discs.

The effect of the number of discs on the performance of an RDR is shown in Figure 17.6. An increase in the number of discs causes a corresponding increase in the submerged biofilm area (and associated quantity of biomass) and in the volume of fluid carried with the discs into the aerated sector. Consequently, the substrate removal rates in both the submerged and the aerated sectors will increase with an increase in the number of discs, causing a reduction in the effluent substrate concentration and an increase in the percent removal.

Figure 17.7 shows the effect of fractional submergence of the discs on substrate removal in an RDR with a fixed number of discs rotating at a constant speed. By reducing the inner radius, r_i, the discs can be submerged to any fraction up to 0.5. An increase in the fractional submergence increases the submerged area, A_s, and allows more microorganisms to grow on a disc of fixed size. Consequently, it causes the substrate removal rate in the submerged sector to increase. It also increases the volume of fluid carried with the rotating discs into the aerated sector, thereby causing the substrate removal rate in that sector to increase as well. Therefore, the percent substrate removal will increase as the degree of submergence is increased. Although not reflected in the model, submergence in excess of 0.5 will decrease the rate of oxygen transfer in the system, thereby hurting bioreactor performance.

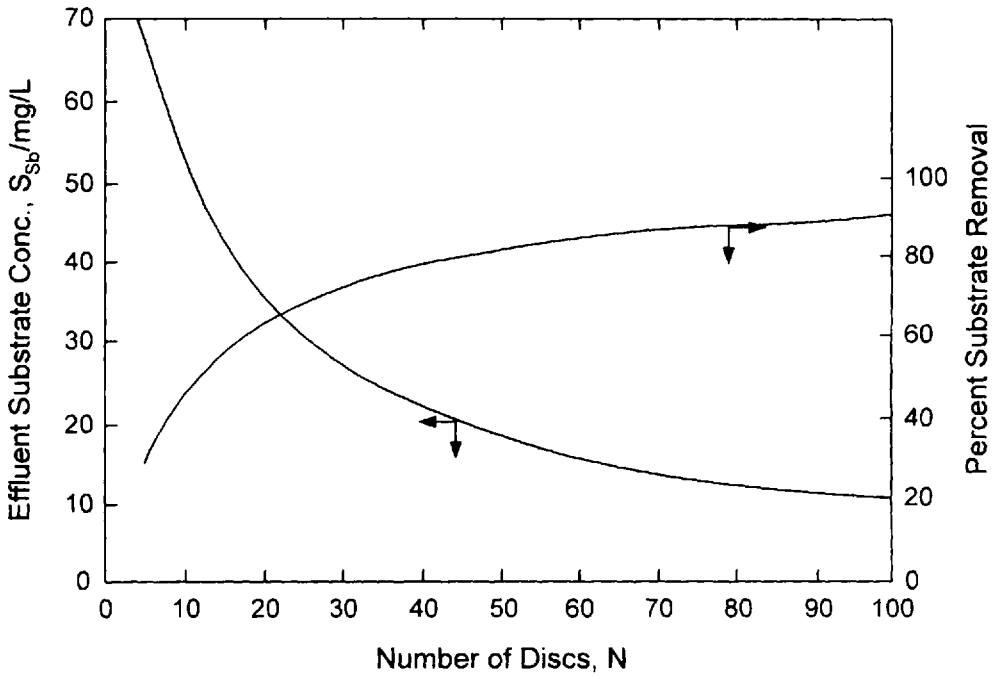


Figure 17.6 Effect of the number of discs on the performance of a single RDR.

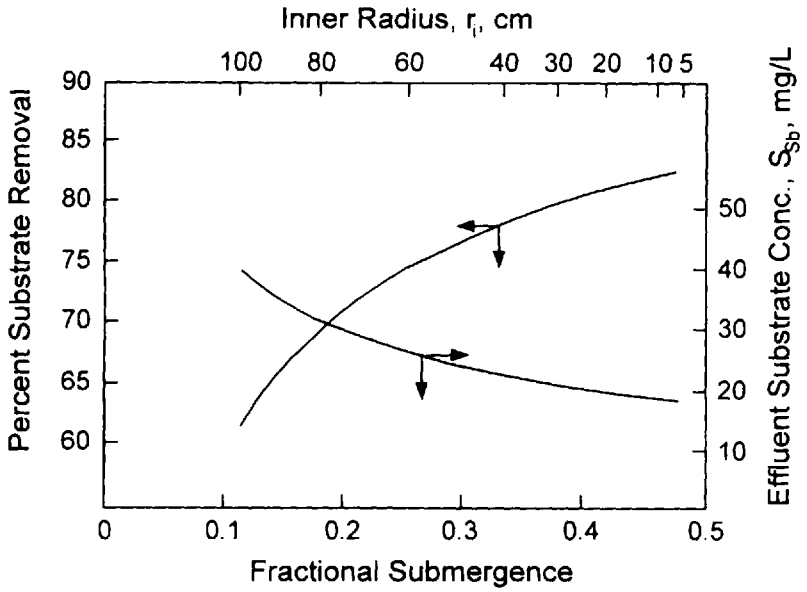


Figure 17.7 Effect of the degree of submergence of the discs on the performance of a single RDR.

17.3 MODEL FOR SUBSTRATE REMOVAL IN A CHAIN OF ROTATING DISC REACTORS

Because of the advantages associated with staged systems, in most situations, a number of RDRs will be used in series to achieve the desired effluent substrate concentration in the most economical manner. Such a system can be modeled as a number of CSTRs in series and mass balance equations can be written for each in turn. Therefore, for a system of M RDRs in series, there are M equations similar to Eqs. 17.8 and 17.14. The only difference is that the bioreactors after the first one will receive biomass from the previous one and that input must be incorporated into the mass balance equations. If all necessary parameter values are given and it is desired to know the effluent substrate concentration from the Mth RDR, we would solve the equations sequentially, starting with the first bioreactor and working our way down the chain.

17.4 PERFORMANCE OF A CHAIN OF ROTATING DISC REACTORS

Using the approach outlined in the preceding section, it is possible to investigate the performance of a number of RDRs in series. Since the effects of the individual parameters will be the same for a chain of bioreactors as for a single one, we will limit this discussion to the effects of the influent characteristics on the substrate removal profile through a six-reactor chain. For convenience, we will again assume that the reaction in the aerated sector is first order with respect to substrate, although it should be recognized that this assumption may be weak for the first one or two reactors in the chain. The effect would be for the model to over-estimate the substrate removal in those reactors. The parameter values are given in Table 17.1 and the influent conditions are given in the figures.

Figure 17.8 shows the profiles of substrate concentration along the bioreactor chain for three different influent substrate concentrations. There it can be seen that the majority of the substrate is removed in the first two reactors and that the others contribute relatively little. Furthermore, as would be expected, more removal occurs in the later stages as the influent concentration is increased. This is a direct result of the behavior observed for a single stage. Even though this is a stagewise configuration, it would be instructive to make a semilog plot of substrate concentration as a function of stage number as shown in Figure 17.9. Examination of this figure reveals that the substrate concentration decreases in an almost exponential manner as the liquid moves from stage to stage, giving almost straight lines in the semilog plot. This behavior is to be expected because the Monod equation behaves in a first-order manner when the substrate concentration is low relative to the half-saturation coefficient, and first order reactions cause an exponential decline in substrate concentration in a stagewise system. This particular behavior makes it easy to predict the effect of additional stages on the performance of a chain of RDRs. Furthermore, it suggests that if the performance of a single stage were known, a reasonable prediction of the performance of a chain could also be made. Another interesting observation from Figure 17.9 is that the lines for different influent sub-

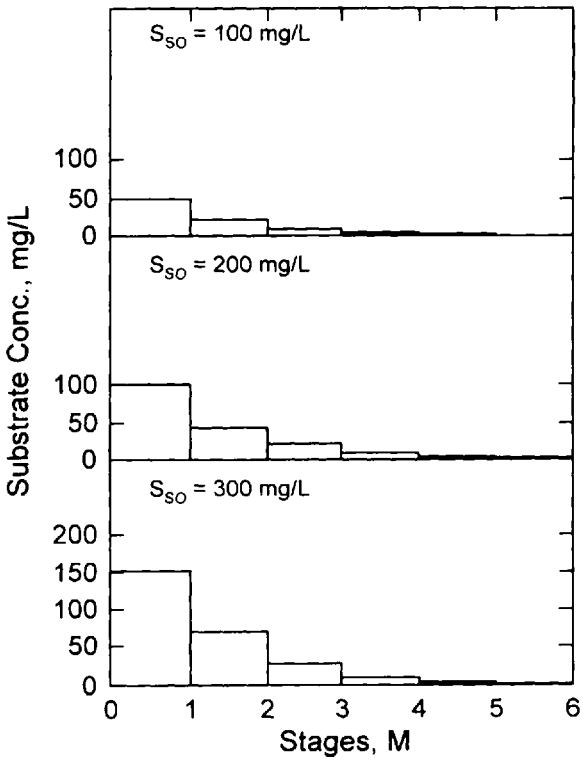


Figure 17.8 Effect of influent substrate concentration on the substrate concentration profile through a chain of six RDRs.

strate concentrations are almost parallel. This suggests that data from one substrate concentration can be extrapolated with reasonable accuracy to another, provided that care is taken to ensure that the range over which the extrapolation is made is reasonable.

The effects of hydraulic loading on the substrate concentration profile within a chain of RDRs is shown in Figure 17.10 where the hydraulic loading decreases as the number of discs, N, is increased. As would be anticipated from the previous discussion, an increase in the hydraulic loading shifts the removal of substrate to reactors further down the chain. Figure 17.11 shows the same information plotted as a semilog function of the reactor number. Examination of that figure reveals that the linearity of the plot is best at lower hydraulic loading rates. This suggests that the extrapolation procedure in the preceding paragraph will work best at lower hydraulic loadings. It is also apparent from the figure that the slope of the semilog plot is a function of the hydraulic loading. This suggests that a plot of that slope versus the hydraulic loading would provide a way to interpolate between loadings.

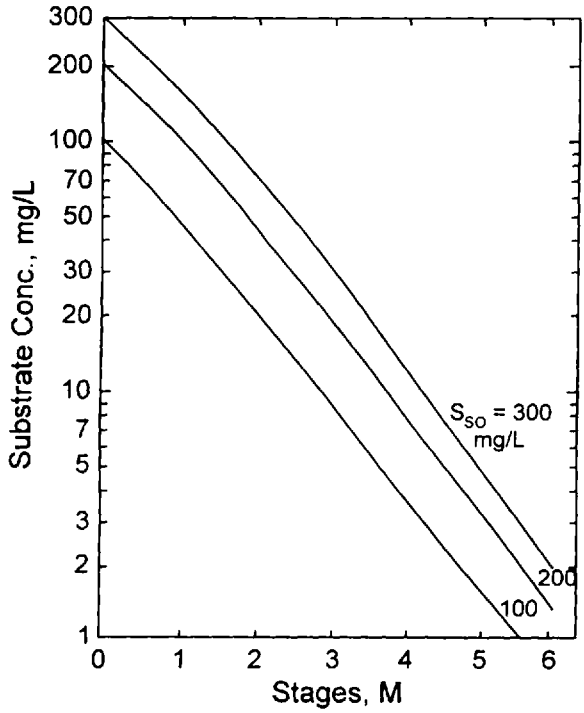


Figure 17.9 Semi-log plot of the effect of influent substrate concentration on the substrate concentration profile through a chain of six RDRs.

17.5 OTHER ROTATING-DISC REACTOR MODELS

Several other RDR models are available in the literature. Although no model is widely accepted for design purposes, it would be instructive to review briefly the key features of these models in comparison to the model presented above. A detailed analysis of several RDR models has been presented by Spengel and Dzombak,[11] and readers interested in more information on this subject are urged to consult their work.

17.5.1 Kornegay Model

The Kornegay[8] model ignores the two distinct sectors, aerated and submerged, and assumes that the entire biofilm is exposed to the substrate concentration in the bulk liquid. It also assumes that the rate of substrate removal by suspended cells is negligible. The effects of mass transfer are accounted for by assuming that the Monod half-saturation coefficient is a function of the rotational speed of the discs. This is equivalent to making the overall effectiveness factor dependent on the substrate concentration and the half-saturation coefficient. Thus, Kornegay's[8] model can be viewed as a special case of the one presented above in which $L_w = 0$, $\eta_{e(S)}$ is determined by S_{sb} and K_S, and the area of the submerged sector is equal to the area of disc covered by biofilm.

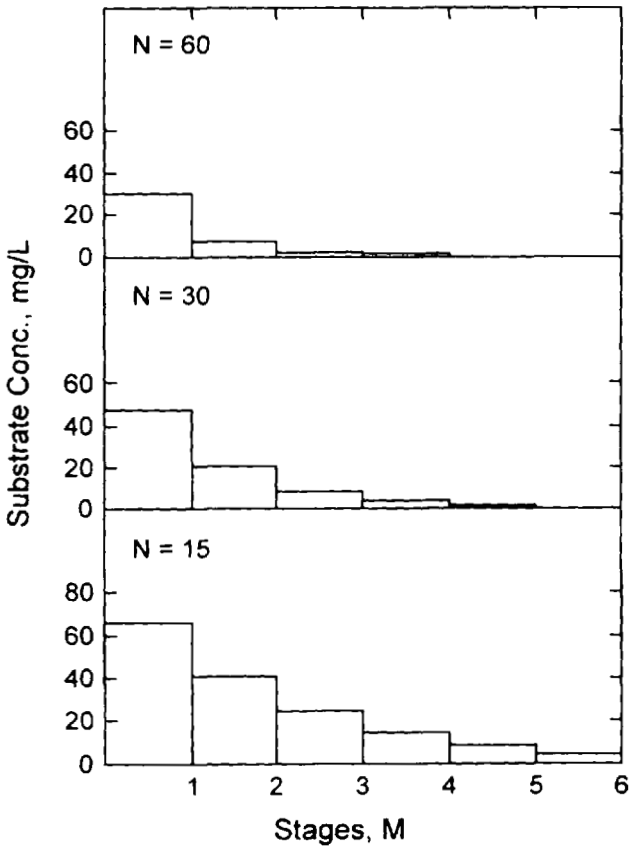


Figure 17.10 Effect of the number of discs per stage (N) on the substrate concentration profile through a chain of six RDRs. The hydraulic loading is inversely proportional to N.

17.5.2 Model of Hansford et al.

The steady-state model of Hansford et al.[6] has many similarities to the one of Grady and Lim[1] used herein. It recognizes that there are two sectors, submerged and aerated, and that a rotating disc carries a liquid film from the tank into the aerated sector as it turns. Their model was developed by writing four substrate mass balance equations: (1) for the liquid film in the aerated sector, (2) for the biofilm in the aerated sector, (3) for the liquid in the tank, and (4) for the biofilm in the submerged sector. The following assumptions were made: the liquid in the tank is completely mixed so that the substrate concentration is uniform; the liquid film and the biofilm in the aerated sector are completely mixed in both the axial and radial directions so that the substrate concentration is a function of only the angular position, θ; the substrate concentration in the biofilm on the submerged sector is uniform with respect to radial, axial, and angular position; the thickness of the liquid film in the aerated sector is constant and independent of the rotational speed; and the suspended biomass in the tank consumes no substrate. The major difference between this model and one of Grady and Lim[1] is that it assumes that the substrate concentration in the biofilm is

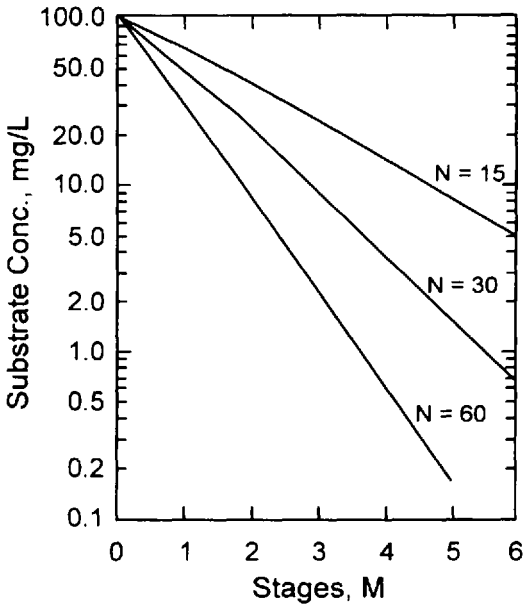


Figure 17.11 Semi-log plot of the effect of the number of discs per stage (N) on the substrate concentration profile through a chain of six RDRs. The hydraulic loading is inversely proportional to N.

independent of depth. This is equivalent to saying that there is no mass transfer resistance within the biofilm or that the thickness of the biofilm is so small that the reaction takes place only at its surface. In either case, the internal effectiveness factor is unity. Furthermore, this model assumes that the reaction takes place only in the biofilm rather than in both the biofilm and bulk liquid. Finally, the thickness of the liquid film in the aerated sector is assumed to be constant, whereas the model used herein allows variations due to rotational speed. In spite of these differences, the trends predicted by the model of Hansford et al.[6] are qualitatively similar to those in this chapter, as well as to trends observed in the field.

17.5.3 Model of Famularo et al.

The RDR model of Famularo et al.[1] incorporates the transport of both oxygen and substrate up to and into the biofilm. The basic components of the model, i.e., the mass transfer effects, reaction rates, etc., are very similar to the ones in the model of Grady and Lim,[3] although the effectiveness factor approach is not used. Rather, they use a finite difference procedure to solve the nonlinear coupled differential equations. A minor difference is that they modeled the aerated sector as four mixed regions in series, whereas a plug-flow model was used by Grady and Lim.[3] The differences in the results caused by this are likely to be slight, and indeed, the trends predicted by the model are in agreement with the trends presented in this chapter. The major benefit of the model of Famularo et al.[1] is that the inclusion of oxygen transfer effects allows prediction of the loading conditions under which oxygen limitations are likely to occur.

17.5.4 Model of Watanabe

Watanabe[12] has developed a model for carbon oxidation, nitrification, and denitrification in an RDR that employs a number of empirical relationships. While a direct solution technique is used, the kinetics are simplified, with zero order intrinsic kinetics being assumed for both nitrification and denitrification, giving overall half-order reaction rates when transport is considered. In addition, an empirical relationship is employed to estimate the distribution of heterotrophic and autotrophic biomass in the biofilm. The model also assumes that the oxygen flux into the biofilm is independent of the composition of the microbial community, so that the use of oxygen by the two bacterial types will depend upon their distribution in the biofilm.

17.5.5 Model of Gujer and Boller

Gujer and Boller[5] have also developed an RDR model that incorporates carbon oxidation, nitrification, and denitrification. The reactions included are aerobic growth of heterotrophs, anoxic growth of heterotrophs, and aerobic growth of nitrifiers. Decay of both heterotrophs and autotrophs is by the lysis:regrowth approach, with the lost biomass going directly to readily biodegradable substrate since slowly biodegradable substrate is not included. Surface shear of the biofilm is included, allowing the biofilm thickness to be an output from the model. Dual limitation by both electron donor and acceptor is also included. Because of that, the direct solution approach discussed in Section 15.3 is employed. The RDR is modeled as a series of CSTRs with reaction occurring only in the submerged sector. The model was able to simulate the effect of organic substrate biodegradation on nitrification and the results were consistent with the concepts presented in Section 15.4, with nitrification occurring in the later stages after the bulk of the organic matter had been utilized. Although the authors caution that the model is not suitable for design, it represents a significant step in the modeling of all possible events in an RDR, although it does not include the effects of disc rotational speed.

17.5.6 Model of Spengel and Dzombak

Spengel and Dzombak[11] built on the work of Famularo et al.[1] to develop a very complete model that considers both carbon oxidation and nitrification with competition for oxygen in the biofilm. Their substrate utilization rate terms are double Monod expressions (Eq. 3.46), considering both the electron donor and the electron acceptor (oxygen). However, like the model of Grady and Lim,[3] the biofilm thickness and density have to be provided as input. Two simplifications are made to facilitate their solution. First, the disc is divided into four sectors, two in the submerged region and two in the aerated region, and the substrate removal rates are calculated for each sector. Second, the biofilm is divided into four layers, with each being considered as a mixed zone. In that way substrate transport into the biofilm can be considered easily. Spengel and Dzombak[11] were able to calibrate the model to mimic well the results from pilot tests, thereby facilitating the interpretation of those tests. Although several issues still need resolution, it is possible that the model will be useful to assist in scale-up.

17.6 KEY POINTS

1. The performance of a rotating disc reactor (RDR) system can be modeled as a series of continuous stirred tank reactors (CSTRs) containing biofilms.
2. In an RDR there are two external mass transfer coefficients that need to be evaluated, one in the submerged sector, $k_{l,s}$, and the other in the aerated sector, $k_{l,a}$. The former increases with the rotational speed of the rotating disc in accordance with a power law, while the latter decreases in accordance with an inverse power law.
3. A general model for substrate removal in a single RDR is obtained by writing a steady-state mass balance on substrate around the liquid volume in the tank (CSTR) and a steady-state mass balance on substrate around an infinitesimal liquid volume on the aerated sector of the rotating disc plug–flow reactor (PFR).
4. Increases in flow-rate cause increases in the effluent substrate concentration from an RDR.
5. The effluent substrate concentration increases as the influent concentration to an RDR is increased. At low flow-rates, the percent substrate removal is relatively independent of the influent substrate concentration, while at high flow-rates it decreases as the influent substrate concentration is increased.
6. As the rotational speed of an RDR is increased, the percent substrate removal increases up to an upper limit characterized by the system parameters.
7. The effluent substrate concentration from an RDR increases as the disc size is increased, provided that the bioreactor is operated at constant hydraulic loading and constant peripheral velocity.
8. An increase in the number of discs in an RDR results in an increase in the substrate removal rate and a decrease in the effluent substrate concentration.
9. An increase in the fractional submergence of the discs results in an increase in the substrate removal rate in an RDR.
10. Rotating disc reactors are usually run in series in order to economically achieve a desired effluent substrate concentration.
11. In a chain of RDRs, the majority of the substrate is removed in the first few reactors and the others contribute relatively little. At higher substrate concentrations, more removal occurs in the later stages. The substrate concentration decreases in an almost exponential manner as the liquid moves from stage to stage.
12. An increase in hydraulic loading shifts the removal of substrate to reactors further down the chain.

17.7 STUDY QUESTIONS

1. Explain why and how the external mass transfer coefficient for the liquid in the aerated sector of an RDR is influenced by the rotational speed.

2. Explain why there is a difference between the overall effectiveness factors for the aerated and submerged sectors of an RDR.
3. Describe in general terms the approach that must be taken to model the performance of an RDR.
4. Describe the effects of influent flow rate, influent substrate concentration, and rotational speed on the performance of a single RDR and explain why they occur.
5. Explain why the percent substrate removal is independent of the influent substrate concentration when the hydraulic loading is low, but is dependent on it when the hydraulic loading is high.
6. Explain why it is not possible to scale-up an RDR by assuming that a full-scale unit will have the same effluent quality as a pilot-scale unit with the same hydraulic loading and peripheral velocity.
7. Explain why RDRs are usually run in series.
8. Explain why it is common for the substrate concentration to drop in an exponential fashion from stage to stage in an RDR system with multiple stages in series.

REFERENCES

1. Famularo, J., J. A. Mueller, and T. Mulligan, Application of mass transfer to rotating biological contactors. *Journal, Water Pollution Control Federation* **50**:653–671, 1978.
2. Friedman, A. A., L. E. Robbins, and R. C. Woods, Effect of disc rotational speed on biological contactor efficiency. *Journal, Water Pollution Control Federation* **51** 2678–2690, 1979.
3. Grady, C. P. L. Jr. and H. C. Lim, A conceptual model of RBC performance. *Proceedings, First National Symposium/Workshop on Rotating Biological Contactor Technology*, E. D. Smith, R. D. Miller, and Y. C. Wu, eds. University of Pittsburgh, Pittsburgh, pp. 829–859, 1980.
4. Grieves, C. G., Dynamic and steady state models for the rotating biological disc reactor. Ph.D. Dissertation, Clemson University, Clemson, South Carolina, 1972.
5. Gujer, W. and M. Boller, A mathematical model for rotating biological contactors. *Water Science and Technology* **22**(1/2):53–73, 1990.
6. Hansford, G. S., J. F. Andrews, C. G. Grieves, and A. D. Carr, A steady state model for the rotating biological disc reactor. *Water Research* **12**:855–868, 1978.
7. Hartmann, H., Undersuchung uber die biologische Reinigung von Abwasser mit Hilfe von tauchtop Korperanlagen. *Stuttgarter Berichte zur Siedlungswasserwirtschaft Kommissionsverlag*, Band 9, R., Oldenbourg, Munich, 1960.
8. Kornegay, B. H., Modeling and simulation of fixed film biological reactors for carbonaceous waste treatment. In *Mathematical Modeling for Water Pollution Control Processes*, T. M. Keinath and M. Wanielista, eds. Ann Arbor Science, Ann Arbor, Michigan, pp. 271–315, 1975.
9. Landau, L. D. and V. G. Levich, Dragging of a liquid by a moving plate. *Acta Physicochimica, U.R.S.S.* **17**:42–54, 1942.
10. Levich, V. G., *Physicochemical Hydrodynamics*, Prentice-Hall, Englewood Cliffs, New Jersey, p. 69, 1962.

11. Spengel, D. B. and D. A. Dzombak, Biokinetic modeling and scale-up considerations for rotating biological contactors. *Water Environment Research* **64**:223–235, 1992.
12. Watanabe, Y., Mathematical modelling of nitrification and denitrification in rotating biological contactors. In *Mathematical Models in Biological Wastewater Treatment*, S. E. Jørgensen and M. F. Gromiec, eds. Elsevier, New York, pp. 419–471, 1985.
13. Wilson, R. W., K. L. Murphy, and J. P. Stephenson, Scaleup in rotating biological contactor design. *Journal, Water Pollution Control Federation* **52**:610–621, 1980.

This page intentionally left blank

18
Fluidized Bed Biological Reactors

The biofilm reactors that we have considered in the past two chapters both come under the heading of fixed media bioreactors in which the media stays in one position as the wastewater flows past. As wastewater flows down through a packed tower, the biofilm grows attached to a solid support that is either stacked or dumped into place, removing the soluble substrate as it does so. Careful selection of the hydraulic and organic loadings is required to ensure that sufficient shear occurs relative to growth to prevent the interstitial spaces in the media from plugging, causing the system to fail. In addition, any suspended solids applied to the tower must be sufficiently small to allow them to pass through those spaces without being trapped, because entrapment would lead to plugging. Likewise, in a rotating disk reactor, the rotational speed must be selected to ensure sufficient shear to prevent the biofilm from bridging the spaces between the disks, thereby blocking contact of the wastewater with the biofilm. Furthermore, as with the packed tower, care must be exercised concerning the admission of suspended solids.

In contrast to the fixed media bioreactors, the bioreactors considered in this chapter come under the broad heading of mobile bed bioreactors.[15] Such bioreactors include all biofilm systems with continuously moving media, whether that movement is induced by high air or water velocities or by mechanical stirring. Moving media provide several distinct advantages: they allow better control of biofilm thickness, have superior mass transfer characteristics, are not subject to clogging, and provide very high surface areas for biofilm development while maintaining low pressure drops.[15] As a consequence, they are rapidly gaining acceptance, even though they are relatively new. In this chapter we focus on the theoretical performance of one type of mobile bed bioreactor, the fluidized bed biological reactor (FBBR), because it is one of the more popular ones and has some distinctive characteristics of which environmental engineers should be aware. In Chapter 21 we consider its design, as well as that of some other mobile bed bioreactors.

18.1 DESCRIPTION OF FLUIDIZED BED BIOLOGICAL REACTOR

18.1.1 General Characteristics

A fluidized bed bioreactor, Figure 18.1,[28] is one in which the biofilm grows attached to small carrier particles that remain suspended in the fluid, i.e., fluidized, by the drag forces associated with the upward flow of water. The term bioparticle generally denotes a biofilm covered carrier particle, although in some cases (e.g., granules)

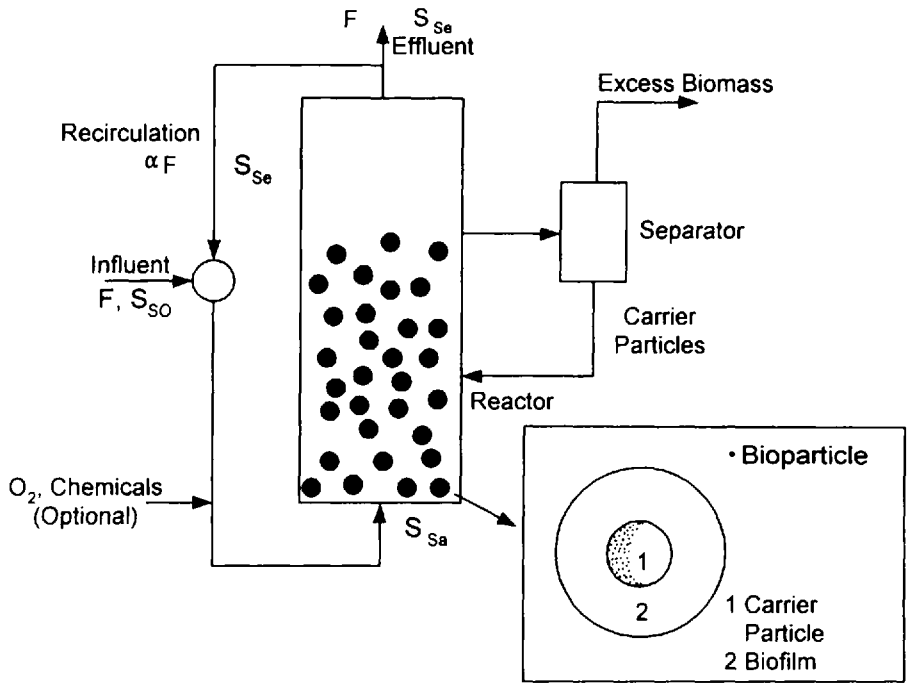


Figure 18.1 Schematic Diagram of an FBBR.

bioparticles develop without the presence of a carrier particle. Most FBBRs are two-phase systems, containing only water and bioparticles, and if oxygen is required it is dissolved in the recirculation flow prior to its return to the reactor. However, recent advances in system design have allowed the incorporation of a gas phase, thereby allowing oxygen transfer to occur directly in the bioreactor.[15] Although the popularity of three-phase systems is increasing, they are considerably more complicated to model than two-phase systems, and thus, our discussion is limited to the latter. It should be noted that the designation of two-phase and three-phase is made with regard to reactants. Some so-called two-phase systems, such as denitrifying and methanogenic systems, actually have a gas phase in them because of the gas produced by the biological reactions. Nevertheless, they can generally be considered to be two phase for modeling purposes as long as the gas flow rate is small relative to the liquid flow rate.[12] This restriction may not be met for systems that are very heavily loaded, however, and care should be exercised in applying two-phase models to such FBBRs.

Because the bioparticles are retained in the reactor, the effluent from an FBBR often contains a sufficiently low suspended solids concentration to allow its discharge without clarification. Maintenance of the appropriate velocity to achieve the desired degree of suspension usually requires recirculation of bioreactor effluent. As biomass grows, the bioparticles become larger, causing the bed to expand in height. To prevent uncontrolled bed expansion, leading to loss of the bioparticles in the effluent, they are usually removed in a systematic manner to maintain a desired bed height. If the bioparticles contain a carrier particle, the excess biomass is removed in a

separator, allowing the carrier particles to be returned. In this way a constant quantity of biomass can be kept in the system while maintaining an effluent low in suspended solids.

In general, FBBRs can be divided into two categories, depending on the nature of the bioparticles.[1] Tower bioreactors are those in which the bioparticles are composed entirely of biomass, without a carrier particle at the center, whereas supported-film bioreactors are those in which the biomass grows as a film on a carrier particle like sand, anthracite, or activated carbon. The sorptive properties of activated carbon provide distinct advantages in some cases, but complicate the analysis of FBBRs using it as the carrier particles. Consequently, its use is not discussed in this chapter. Rather, the reader should consult other sources for more information.[31] Distinction between the two types of FBBRs is necessary because the presence or absence of a carrier particle has a strong influence on the way the bioparticles behave as they grow larger, as we will see later. The upflow anaerobic sludge blanket (UASB) bioreactor is an important example of a tower bioreactor. In it the bioparticles grow as small spherical granules containing the complex microbial community associated with methanogenic systems. Most other FBBRs are supported-film bioreactors, and in fact, the two terms are usually used synonymously, as we will do here. Our discussion here is limited to supported-film bioreactors.

The main advantage of FBBRs over other attached growth bioreactors is that the small size of the carrier particles provides a very large specific surface area for biomass growth. Whereas the media in packed towers and rotating disk bioreactors have specific surface areas on the order of 100 m^2/m^3 (see Table 19.2 and Section 20.1.1), the specific surface area provided by typical carrier particles in an FBBR is on the order of 1,000 to 3,000.[20,28] This allows the maintenance of very high biomass concentrations, ranging from 15,000 mg/L in aerobic FBBRs to 40,000 mg/L in anoxic ones.[20,28] This, in turn, allows very short hydraulic residence times (HRTs) to be used, often on the order of minutes. While packed towers could theoretically contain media of similar size, the downward flow of liquid would cause excessive pressure drop and be prone to clogging as biomass growth occurred. It would also be subject to plugging through entrapment of suspended solids that might enter in the influent. Such solids can pass through FBBRs, however, because of the open structure of the fluidized bed. More information about the characteristics of these unique bioreactors can be found elsewhere.[3,7,17]

18.1.2 Nature of the Biofilm

In Chapter 15, we discussed the concept of a steady-state biofilm. In an FBBR, development of a steady-state biofilm would require the excess biomass to be continually removed from the bioparticle surface and carried away in the effluent. It would also require the bed height to be sufficiently large to accommodate the quantity of biomass associated with the steady-state biofilm. Neither of these requirements is particularly desirable in practice. As noted above, one advantage of the FBBR is that it can be operated in a way that eliminates the need for a final clarifier. This would not be possible if biomass were constantly being sheared from the bioparticles in the FBBR. Furthermore, the thickness associated with a steady-state biofilm is likely to exceed the active thickness of the biofilm,[1] i.e., the depth to which reactants penetrate. This means that the bed height and volume associated with a steady-state

biofilm would be greater than that required to achieve the desired effluent quality with a fully active biofilm. Consequently, one characteristic of most FBBRs is the continual wastage of biomass from the top of the bed to maintain a constant bed height less than that associated with a steady-state biofilm. Another complicating factor is that wastage occurs from the top of the bed because that is where the bioparticles with the thickest biofilm reside. After the biomass on the removed bioparticles has been reduced by subjecting them to surface shear forces, the carrier particles are returned to the bed where they again serve as a support for biomass growth. Initially, the carrier particles fall to the bottom of the bed, but they migrate upward as biofilm builds up on them. (The reason for this behavior is explained later.) Consequently, the biofilm thickness on any individual bioparticle is continually changing, which differs from the assumptions associated with steady-state biofilms. Nevertheless, many models of FBBRs assume the existence of a steady-state biofilm to reduce computational complexity. Although such models are very useful for understanding the major factors influencing the behavior of FBBRs, it should be recognized that they differ significantly from the characteristics of most operating FBBRs.

One interesting attribute of FBBRs is that the dry density of the biofilm on a bioparticle depends on the thickness of that biofilm.[9,20,25,28] Biofilm dry density is defined as the attached dry biomass per unit wet biofilm volume and is the same as the biomass concentration in a biofilm, $X_{B,III}$, used in Chapter 15. However, the term dry density, and its associated symbol, ρ_{fd}, is used here to make clear the distinction between the amount of biomass per unit volume of biofilm and the amount per unit volume of bioreactor, X_B. The exact relationship between density and thickness varies from study to study, but Figure 18.2[28] shows one that has been used in FBBR modeling. An important characteristic is that dry densities are very high (on the order of

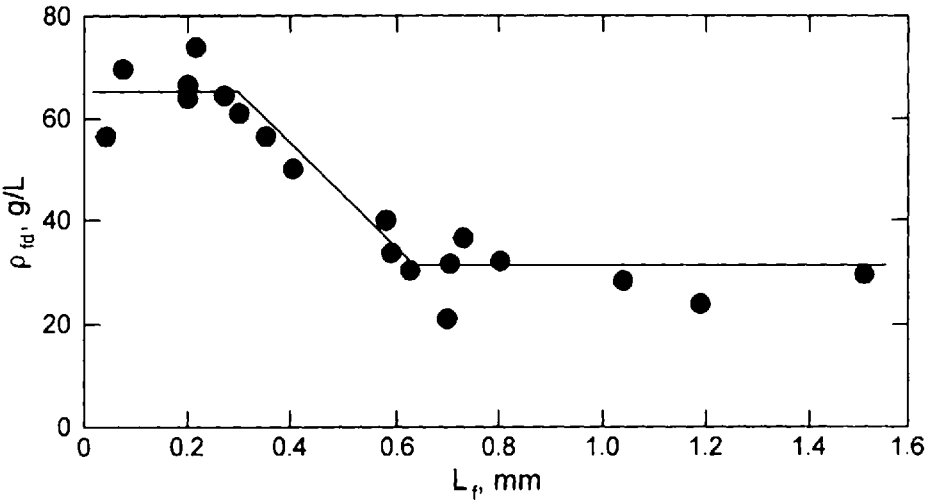


Figure 18.2 Effect of biofilm thickness, L_f, on the dry density, ρ_{fd}, of a denitrifying biofilm. (From W. K. Shieh and J. D. Keenan, Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology* **33**:131–169, 1986. Copyright © Springer-Verlag New York, Inc.; reprinted with permission.)

70 g/L) for thin biofilms (on the order of 200 μm or less), but decrease markedly as the biofilm grows thicker. Although the reason for this behavior is poorly understood, it is probably related to the activity of the biofilm. Thin biofilms tend to be fully penetrated by both electron donor and electron acceptor, whereas the interior of thick biofilms is devoid of one or both of these substances. The resulting environment leads to microbial reactions that can decrease the quantity of viable microbial cells, thereby decreasing the density. Another possibility is that the hydrodynamic conditions that lead to thin biofilms bring about a morphological change in the biofilm that make it denser.[6] Regardless of the mechanism, however, it is clear from this behavior that thicker biofilms do not necessarily lead to more active biomass. This is one reason that FBBRs are commonly operated to give thin biofilms.

The hydrodynamic conditions required to give a thin biofilm are fairly complex and somewhat counterintuitive. Thinner biofilms develop in systems that have higher first-order detachment rate coefficients, b_{1}. While at first glance it might appear that b_{1} will increase whenever the upward velocity of the fluid (superficial velocity) is increased, this is not the case. Generally, the superficial velocities used in FBBRs result in low Reynolds numbers (<10), which means that surface shear is likely to be small.[28] Lower superficial velocities, however, result in a smaller degree of bed expansion, which means that there is a higher probability of collisions among particles. The attrition caused by those collisions has a larger effect on the detachment coefficient than the fluid velocity past the biofilm surface.[12,28] Consequently, higher values of b_{1} have been observed at lower superficial velocities.[11] Empirical models are available that account for the various factors influencing the detachment rate coefficient.[6]

18.2 FLUIDIZATION

It is apparent from the preceding that the hydrodynamic conditions in a fluidized bed influence the biological characteristics of the bioreactor in important ways. In addition, they also define its physical characteristics, which influences its performance. Consequently, it is important to have a clear understanding of fluidization.

18.2.1 Fluidization of Clean Media[8,16]

Consider a cylindrical vessel like that in Figure 18.1 in which water is flowing upward through a bed of small, biomass-free, spherical, carrier particles of equal diameter. As the superficial upflow velocity of the water (equivalent to the total hydraulic loading in a packed tower, defined in Eq. 16.8) is increased, the pressure drop through the bed, i.e., the frictional forces acting on the particles, will increase, but the height of the bed will remain constant as shown in Figure 18.3. Ultimately a point will be reached (A) at which the pressure drop through the bed just counterbalances the force of gravity on the particles (taking into consideration the buoyant force of the displaced fluid) and the particles begin to move. As the superficial velocity is increased further, the particles will move apart so that the frictional forces associated with the local velocity past the particles continue to counterbalance the force of gravity on the particles. As a consequence, the pressure drop through the bed remains constant while the porosity (ε) of the bed increases, causing the bed

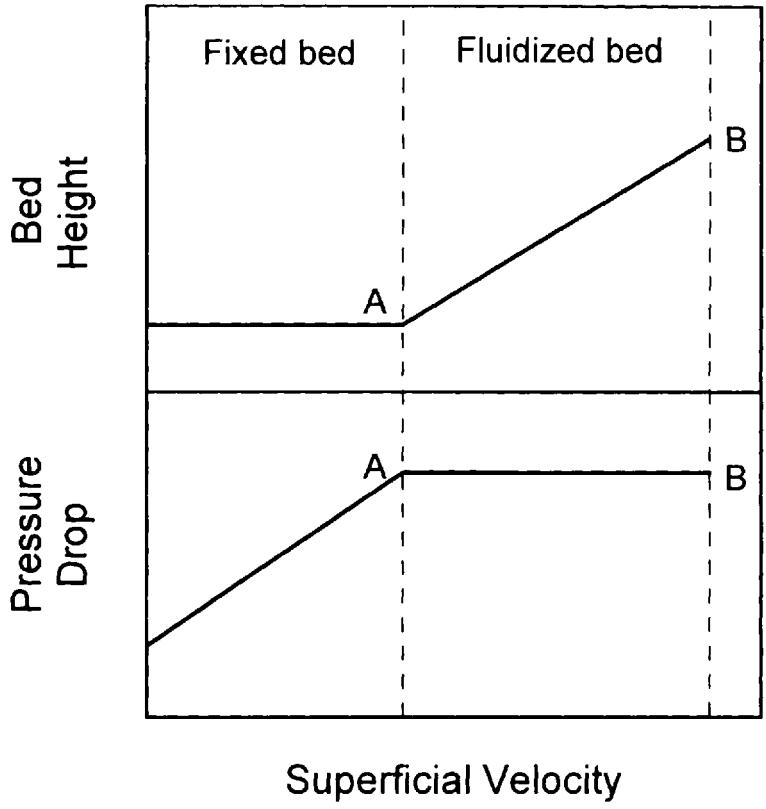


Figure 18.3 Idealized effect of superficial velocity on the pressure drop and bed height of a fluidized bed.

height to increase. Any particle that rises above the top of the bed due to transient nonuniformities in flow will encounter a local upward velocity that is less than the particle's terminal settling velocity and the particle will fall back into the bed. Ultimately, a point (B) will be reached at which the local fluid velocity around the particles is equal to their terminal settling velocity. If the superficial velocity is increased beyond that point the upward drag forces on the particles will exceed the downward gravitational forces and the particles will be carried away and the bed will cease to exist. It is then in continuous fluidization.

Consideration of Figure 18.3 suggests that two velocities are critical to defining the operating range for a fluidized bed, those associated with points A and B. The velocity associated with point A is called the minimum fluidization velocity. The formula for its computation can be derived by equating the pressure drop as given by the Ergun equation to the weight of the bed per unit area of cross-section, allowing for the buoyant force of the displaced water. This results in a quadratic equation for the minimum fluidization velocity, v_{mf}:

$$\frac{1.75\,\rho_w v_{mf}^2}{d_p}\frac{1}{\varepsilon_M^3} + \frac{150\,\mu_w v_{mf}}{d_p^2}\frac{1-\varepsilon_M}{\varepsilon_M^3} - g(\rho_p - \rho_w) = 0 \qquad (18.1)$$

in which d_p is the diameter of the particle, ρ_p is its density, ρ_w is the density of water, μ_w is the viscosity of water, g is gravitational acceleration, and ε_M is the minimum porosity at incipient fluidization, i.e, point A. For roughly spherical particles, ε_M is generally between 0.40 and 0.45, decreasing slightly with increasing particle size. The velocity associated with point B is the terminal settling velocity of the carrier particles, v_t. If it is exceeded the particles are carried away in continuous fluidization and the bed is destroyed. The equation for its computation is derived by equating the drag force on the particle to the gravitational force minus the buoyant force. The result is:

$$v_t = \left[\frac{4g(\rho_p - \rho_w)d_p}{3C_D\rho_w}\right]^{0.5} \tag{18.2}$$

in which C_D is the drag coefficient. The value of C_D depends on the Reynolds number and correlations are available to relate the two for spherical particles as well as for other shapes. The appropriate velocity to use in computation of the Reynolds number is the terminal settling velocity of the particle. The result is usually called the terminal Reynolds number, Re_t. Because the terminal settling velocity depends on the Reynolds number (through the drag coefficient) and the Reynolds number depends on the velocity, an iterative procedure may be required to compute the terminal settling velocity, depending on the nature of the C_D vs. Re_t relationship. Techniques are available for defining ranges of Reynolds numbers over which direct solutions may be possible and the reader should consult other sources to learn more about them.[8,16]

The height that a fluidized bed of clean carrier particles attains (H_{Bp}) depends directly on the porosity that results from the applied superficial velocity. This follows from the fact that the mass of particles in the bed is constant. Therefore:

$$H_{Bp} = H_{Rp}\left(\frac{1 - \varepsilon_R}{1 - \varepsilon}\right) \tag{18.3}$$

where H_{Rp} is a reference bed height and ε_R is the porosity associated with it. Some use the minimum bed height immediately prior to fluidization as H_{Rp}, in which case, ε_R will be ε_M.[8] Others[21,29] avoid the need to know ε_M by using as H_{Rp} the height that would be occupied by the carrier particles if they formed a solid block with mass equal to the total mass of carrier particles present, in which case ε_R would be zero and H_{Rp} would be given by:

$$H_{Rp} = \frac{\mathbf{M}_p}{\rho_p A_c} \tag{18.4}$$

where $\mathbf{M}_p$ is the mass of carrier particles and A_c is the cross-sectional area of the FBBR. Substitution of Eq. 18.4 into Eq. 18.3 gives:

$$H_{Bp} = \frac{\mathbf{M}_p}{\rho_p A_c(1 - \varepsilon)} \tag{18.5}$$

which can be used to calculate the height of clean carrier particles in a fluidized bed. However, regardless of which definition of H_{Rp} is used, prediction of the bed height associated with a given superficial velocity requires prediction of the porosity, ε. This can be done by using the Richardson–Zaki equation:

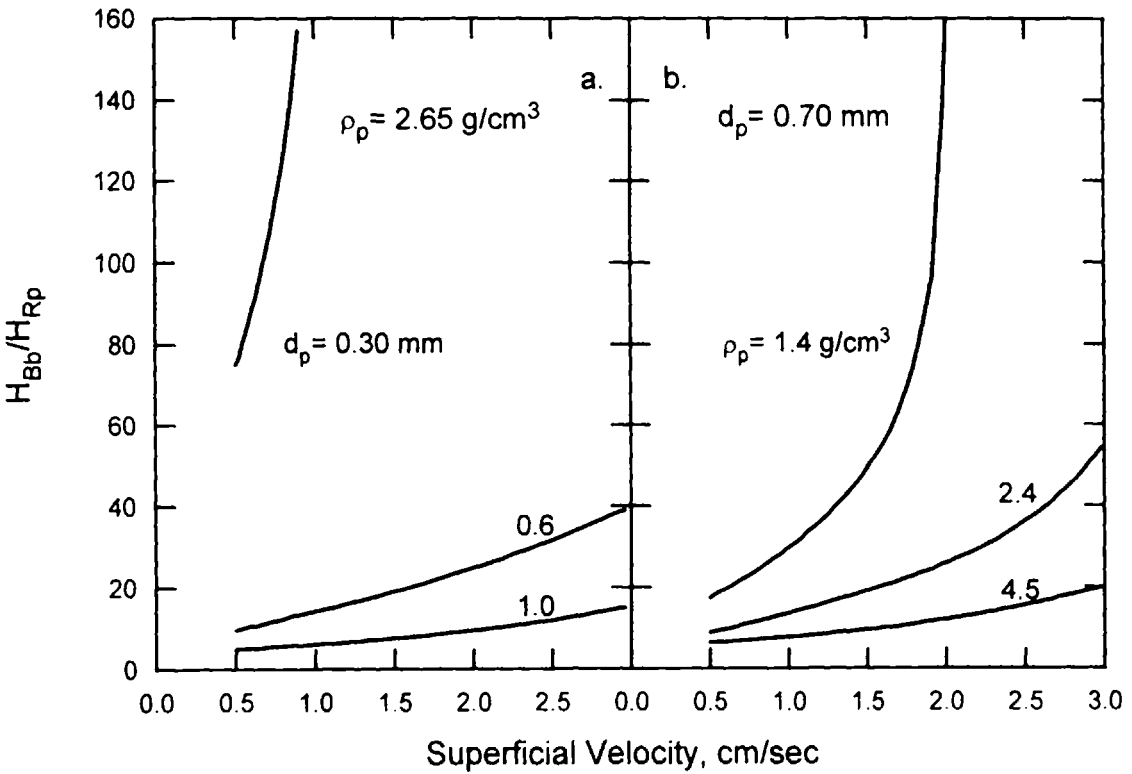


Figure 18.4 Effect of superficial velocity, v, carrier particle diameter, d_p, and carrier particle density, ρ_p, on the bed height of a fluidized bed, H_{Bb}, relative to the reference height of the carrier particles, H_{Rp}. The carrier particles are covered with a biofilm with a thickness of 100 μm and a dry density of 65 g/L. (From W. K. Shieh and J. D. Keenan, Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology* **33**:131–169, 1986. Copyright © Springer-Verlag New York, Inc.; reprinted with permission.)

$$\frac{v}{v_t} = \varepsilon^n \tag{18.6}$$

in which v is the applied superficial velocity and n is a coefficient. The value of n can be correlated with the Galileo number or with the Reynolds number calculated on the basis of the terminal settling velocity. It typically takes on values between 2 and 5 for clean particles, with larger values being associated with smaller Galileo or Reynolds numbers, i.e., smaller particles. Several correlations are available.

Selection of the carrier particle for an FBBR is an important consideration because it has several influences, as we will see shortly. One important consideration is the stability of the bed to fluctuations in flow rate; that is, to fluctuations in superficial velocity. If the expanded bed height is overly sensitive to such fluctuations, small variations in flow rate might carry particles out in the effluent, thereby destroying the bed. Figure 18.4 illustrates how carrier particle size and density affect bed height over the range of superficial velocities commonly encountered.[28] The curves were obtained by simulation and the conditions employed are indicated in the figure. Those curves illustrate that the stability of a bed increases with increasing carrier particle size and density. Common silica sand has a density of around 2.65 g/cm^3, which provides reasonable stability over a broad range of superficial velocities.

18.2.2 Effects of Biomass on Fluidization

The growth of biofilm on the carrier particles changes their fluidization characteristics. This is due to three things. First, growth of the biofilm will change the size of the particle. Second, unless the carrier particle has a density equivalent to the wet density of the biofilm, growth of the biofilm will change the overall effective density of the particle. Third, the surface properties of the biofilm will differ from those of the clean particle, thereby changing the relationship between the drag coefficient, C_D, and the Reynolds number. In addition, growth of the biofilm may change the sphericity of the particle, but that effect has been found to be small,[24] and is not considered further here. Rather, we assume spherical particles.

Terminal Settling Velocity. Examination of Eq. 18.2 shows that the terminal settling velocity of a particle depends on its diameter, its density, and the drag coefficient. Since all of those characteristics are altered by growth of a biofilm, it becomes clear that biofilm growth changes the terminal settling velocity. The influence of biofilm thickness, L_f, on bioparticle diameter, d_b, is very straightforward:

$$d_b = d_p + 2L_f \tag{18.7}$$

The influence of the biofilm growth on the overall effective density of the bioparticle, ρ_b, depends on the density of the carrier particle, ρ_p, and the wet density of the biofilm, ρ_{fw}, as well as the relative volumes occupied by the carrier particle and the biofilm.[12,22,30] Since the volume of a sphere is proportional to its diameter cubed, the relationship is:[30]

$$\rho_b = \rho_p \left(\frac{d_p}{d_b}\right)^3 + \rho_{fw}\left[1 - \left(\frac{d_p}{d_b}\right)^3\right] \tag{18.8}$$

The biofilm wet density is related to its dry density, ρ_{fd}, and the weight fraction moisture content of the biofilm, P':[30]

$$\rho_{fw} = \frac{\rho_{fd}}{1 - P'} \tag{18.9}$$

The value of moisture content has been found to be approximately 0.93 over a broad range of biofilm thicknesses.[30] We saw earlier that the biofilm dry density depends on its thickness, and the biofilm wet density also varies with biofilm thickness. Thus, it is not surprising that a number of investigators have reported different values for the wet biofilm density.[22] Nevertheless, a value of 1.1 g/cm^3 has been assumed to be typical of biomass.[22]

The influence of biofilm growth on the drag coefficient has been studied by several investigators.[10,19,22,24] All correlate the drag coefficient to the terminal Reynolds number, Re_t, using an expression of the type:

$$C_D = aRe_t^{\ b} \tag{18.10}$$

Figure 18.5[24] shows three relationships and compares them to the relationship of Schiller et al. (referenced in 24) for clean spherical particles. The equations are given in Table 18.1. Two things are evident from the figure. First, the growth of a biofilm increases the drag coefficient relative to that of a clean particle with equivalent terminal Reynolds number, i.e., equivalent diameter and density. Second, the rela-

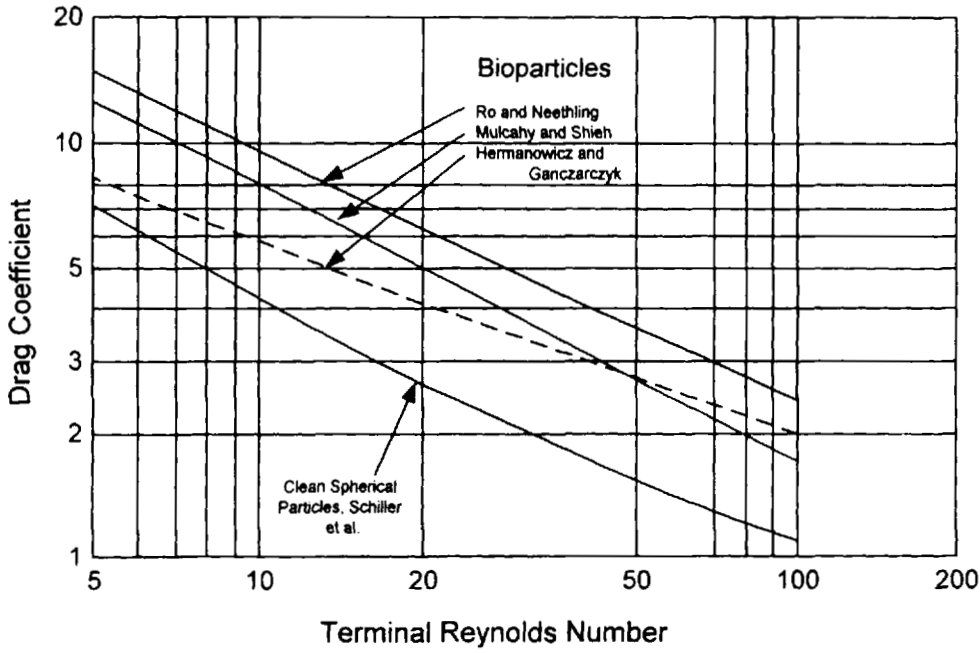


Figure 18.5 Effect of terminal Reynolds number, Re_t, on the drag coefficient, C_D, of bioparticles. (From K. S. Ro and J. B. Neethling, Terminal settling characteristics of bioparticles. *Research Journal, Water Pollution Control Federation* **62**:901–906, 1990. Copyright © Water Environment Federation; reprinted with permission.)

Table 18.1 Equations Depicting the Relationships Between the Drag Coefficient and the Terminal Settling Velocity of Bioparticles Shown in Figure 18.5

Equation	Source
$C_D = 17.1\ Re_t^{-0.47}$	Hermanowicz and Ganczarczyk[10]
$C_D = 36.66\ Re_t^{-0.667}$	Mulcahy and Shieh[19]
$C_D = 24\ Re_t^{-1.0} + 21.55\ Re_t^{-0.515}$	Ro and Neethling[24]

tionships found by the three studies on biofilms are all different, suggesting that the influence of biofilm growth on C_D may be case specific.

Figure 18.6[24] shows the effect of biofilm growth on the terminal settling velocity of bioparticles in which sand (ρ_p = 2.65 g/cm^3) with a diameter of 0.5 mm serves as the carrier particle. The values were calculated with Eq. 18.2 using the bioparticle density, ρ_b, from Eq. 18.8 in place of ρ_p and the bioparticle diameter, d_b,

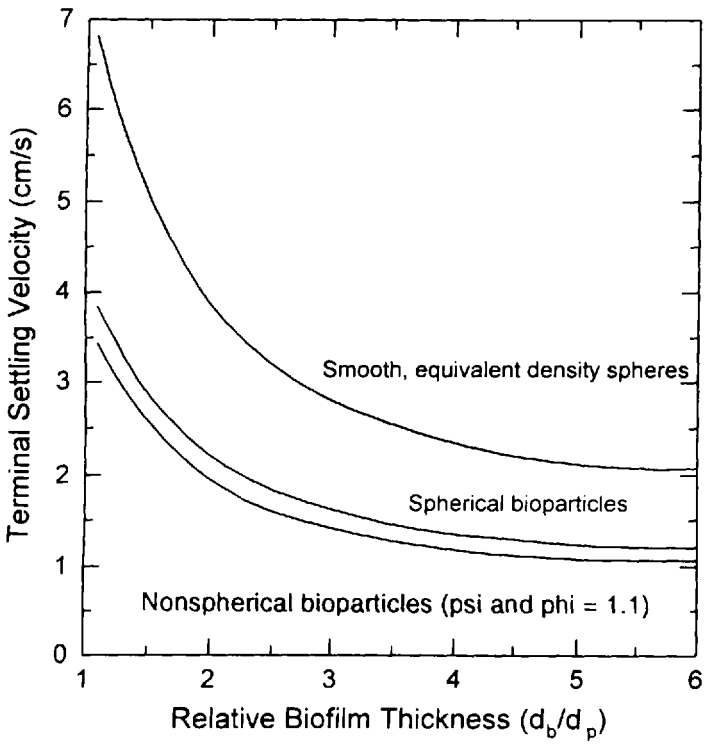


Figure 18.6 Effect of relative biofilm thickness, d_b/d_p, on the terminal settling velocity of spherical and nonspherical bioparticles with sand carrier particles with a diameter, d_p, of 0.5 mm. For comparison, the terminal settling velocity of smooth, equivalent density spheres is also shown. (From K. S. Ro and J. B. Neethling, Terminal settling characteristics of bioparticles. *Research Journal, Water Pollution Control Federation* **62**:901–906, 1990. Copyright © Water Environment Federation; reprinted with permission.)

from Eq. 18.7 in place of d_p. The value of C_t was computed from the correlation of Ro and Neethling[24] shown in Figure 18.5. Three important points are evident in the figure. First, the terminal settling velocity of the bioparticles decreases as the biofilm thickness increases. Since terminal settling velocity is directly proportional to the diameter of a particle (see Eq. 18.2), the decrease in terminal settling velocity associated with an increase in biofilm thickness is due to the decrease in the effective density of the bioparticle (see Eq. 18.8). Second, as the biofilm thickness increases, a point is eventually reached at which further increases have little effect. In that region the effects of increases in diameter are approximately equal to the effects of decreases in density. Third, the settling velocity of a bioparticle is always lower than that of a smooth sphere of equivalent diameter and density. This is due to the effect of the biofilm on the drag coefficient. The latter point is true for a wide range of carrier particle sizes and densities, as well as for a broad range of biofilm thicknesses, with the effect that the settling velocity of a bioparticle is always between 55 and 60% of the velocity of an equivalent density smooth sphere of the same diameter.[24]

The effects of particle density are shown in Figure 18.7[22] for a case in which the growth of the biofilm has no effect on the relationship between the drag coefficient and the terminal Reynolds number. In other words, it assumes that clean carrier particles have the same surface characteristics as those with biofilm. There it can be seen that biofilm growth can increase the settling velocity of carrier particles of low density. In fact, the counteracting effects of the changes in density and diameter can make the settling velocity of bioparticles containing low density carrier particles change in complex ways as they grow larger, particularly when the effects on the drag coefficient are also considered. This can have a significant effect on the migration of bioparticles in FBBRs.

Bed Porosity and Expansion. Because growth of a biofilm changes the terminal settling velocity of a particle, it also changes its fluidization properties. One effect is on the porosity associated with a given superficial velocity. According to the Richardson–Zaki equation (Eq. 18.6) if the superficial velocity is held constant and the terminal settling velocity of a particle is changed, the porosity of the bed will change. This, is turn will change the height of the fluidized bed, as indicated by Eq. 18.3. Another effect is on the reference bed height. Particles with a larger diameter occupy more space. Thus, the reference bed height will be larger, which will cause the expanded bed height to increase as well. Because the volume of a sphere is proportional to its diameter cubed, the value of the reference bed height for bioparticles, H_{Rb}, can be related to the mass of carrier particles present by:[21,29]

$$H_{Rb} = \frac{\mathbf{M}_p}{\rho_p A_c} \left(\frac{d_b}{d_p}\right)^3 \tag{18.11}$$

This equation can be substituted into Eq. 18.3 to give the height of a fluidized bed containing bioparticles:

$$H_{Bb} = \frac{\mathbf{M}_p}{\rho_p A_c (1 - \varepsilon)} \left(\frac{d_b}{d_p}\right)^3 \tag{18.12}$$

The Richardson–Zaki equation (Eq. 18.6) has been shown to be applicable to bioparticles, although the correlation between the coefficient n and the Reynolds or Galileo number is different from that for clean particles.[19,27,33] One that works well

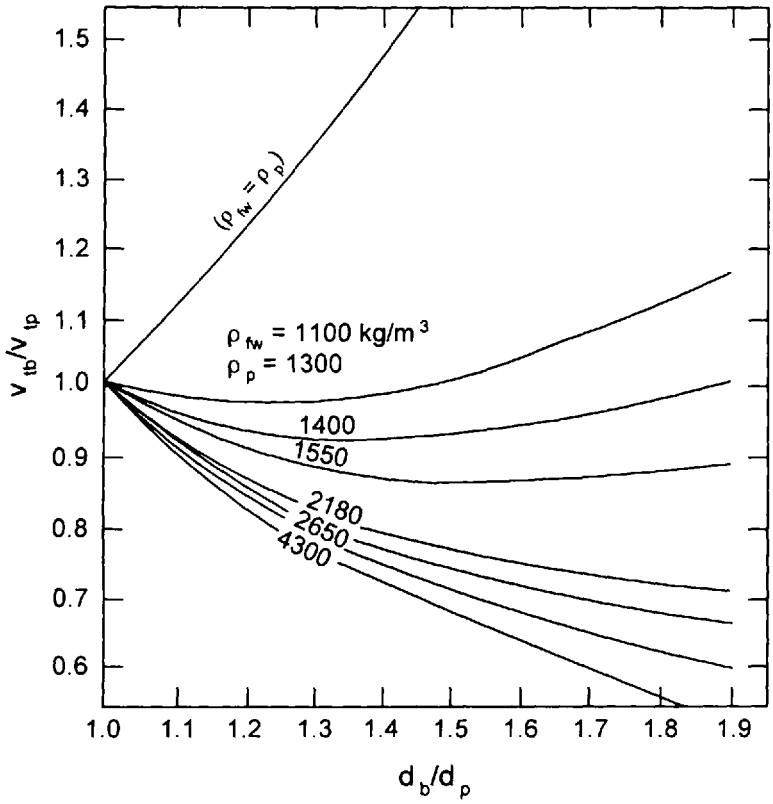


Figure 18.7 Effect of relative biofilm thickness, d_b/d_p, and carrier particle density, ρ_p, on the terminal settling velocity of a bioparticle, v_{tb}, relative to the terminal settling velocity of its carrier particle, v_{tp}. (From J. Myška and J. Švec, The distributive properties of a fluidized bed with biomass. *Water Research* **28**:1653–1658, 1994. Copyright © Elsevier Science Ltd.; reprinted with permission.)

for a broad range of particle sizes and densities is that of Shieh and Chen,[27] which was developed from the data of Mulcahy and LaMotta:[18]

$$n = 47.36\ Ga^{0.2576} \qquad 1{,}000 < Ga < 15{,}000 \tag{18.13}$$

where Ga is the Galileo number, given by:

$$Ga = \frac{d_b^3 \rho_w (\rho_b - \rho_w) g}{\mu_w^2} \tag{18.14}$$

The approach above has been used to demonstrate through modeling the effect of biofilm thickness on the degree of expansion of an FBBR containing sand with a diameter of 0.4 mm as the carrier particle.[28] The results are shown in Figure 18.8 in which the expanded bed height has been normalized relative to H_{Rp} as computed with Eq. 18.4. There it can be seen that even thin biofilms have a strong effect on the height of a fluidized bed. Consequently, during design, careful consideration must

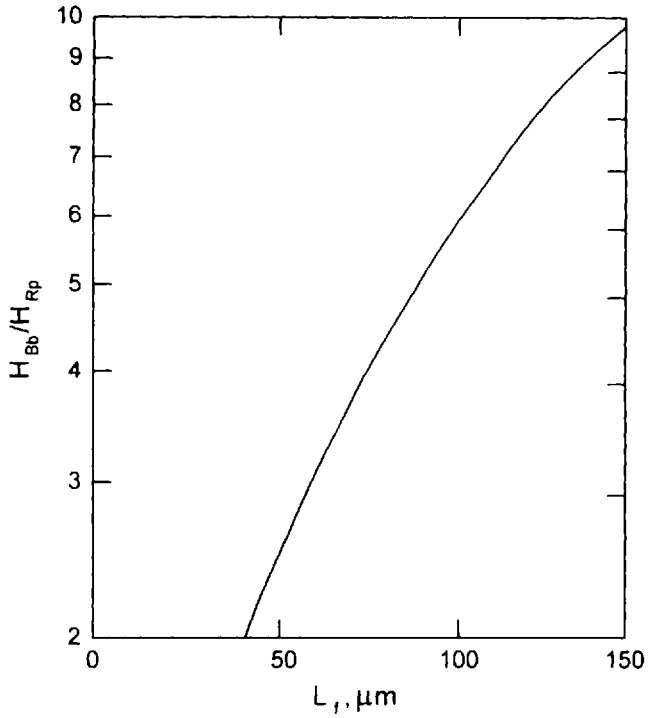


Figure 18.8 Effect of biofilm thickness, L_f, on the height of a fluidized bed, H_{fb}, relative to the reference height of the carrier particles, H_{Rp}. The following conditions were assumed: d_p = 0.4 mm, ρ_p = 2.65 g/cm^3, v = 1 cm/sec, ρ_{bf} = 65 g/L, P' = 0.93. (From W. K. Shieh and J. D. Keenan, Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology* **33**:131–169, 1986. Copyright © Springer-Verlag New York, Inc.; reprinted with permission.)

be given to the configuration of an FBBR to ensure that it is capable of containing the desired amount of media once a biofilm of the desired thickness has developed.

Solids Mixing. The movement of particles in a fluidized bed is a very complex subject that is incompletely understood.[12] In fact, the circumstances and assumptions associated with an analysis of mixing strongly influence the conclusions reached. Nevertheless, it is important to understand the basic forces at work in a fluidized bed as biofilm grows.

Andrews[1] has presented a very thorough analysis of the factors influencing solids mixing. First, it must be recognized that there are two counteracting tendencies affecting particle movement. One is the tendency of fluidized particles to move randomly, which is a disordering tendency. The other is caused by the development of particles of different size due to biofilm growth. If the terminal settling velocities of the various particles are not all the same, the bed tends to stratify, with rapidly settling particles near the bottom and slowly settling ones near the top. This is an ordering tendency, but whether such a tendency is stable depends on the density of the carrier particles.

Bioparticles containing carrier particles of low density, similar to the wet density of the biofilm, tend to stratify because the density doesn't change significantly as the biofilm grows. Only the diameter changes. The same is true for bioparticles without carrier particles, such as UASB granules. In that case, larger particles have a higher settling velocity, causing them to move to the bottom of the bed, where they are exposed to more substrate, causing them to grow even larger. Conversely, smaller particles move to the top, where they are exposed to less substrate, which causes the biofilm to grow more slowly, or even decrease in size because of decay and surface shear. This leads to stratification of the bed, with the possible development of bioparticle sizes well in excess of the optimal, thereby increasing the quantity of inactive biomass in the bed. Under such a situation, biomass wastage should be done from the bottom of the bed.

Bioparticles containing carrier particles of high density, on the other hand, tend to form well mixed beds, although a degree of stratification can be induced. With high-density carrier particles, the settling velocity of the bioparticles decreases as the thickness of the biofilm increases. As a consequence, larger bioparticles tend to move to the top of the bed. Once there, however, they receive less substrate, which causes them to decrease in size, thereby allowing them to move downward into a region of higher substrate concentration. Bioparticles with thin biofilms, on the other hand, move toward the bottom of the bed, where they are exposed to high substrate concentrations, causing more rapid growth and an increase in size. The resulting situation is unstable, inducing motion within the bed, leading ultimately to a relatively uniform bioparticle size throughout the bed.

Control of bed height in an FBBR requires continual wastage of biomass. Otherwise, nonoptimal sized bioparticles develop and the bed height becomes very large.[1] Common practice is to waste biomass from the top of a bed containing high-density carrier particles. This induces stratification in the bed because the size of any individual bioparticle is continually changing, preventing the development of a steady-state biofilm. By continually wasting large bioparticles from the top and returning clean carrier particles, which then migrate to the bottom where they exposed to high substrate concentrations, the bed is maintained in a dynamic state. Consequently, stratification of such beds is a common occurrence.[2,23] The above analysis is based on the assumption of a uniform carrier particle size. If there are significant differences in carrier particle size, the bed tends to stratify based on carrier particle size rather than bioparticle size.[2] As a consequence, larger support particles tend to stay at the bottom where they accumulate biofilm beyond the optimum thickness, while smaller carrier particles migrate to the top from where they can be ineffectually cycled through the biomass wastage device. Consequently, it is important for FBBRs to have a uniform particle size.

18.2.3 Relationship Between Fluidization and Biomass Quantity

It is clear from the preceding that there is a complex relationship between the fluidization regime imposed on an FBBR and the quantity of biomass that may be present. Consequently, it is difficult to intuitively reason out the relationship. Luckily, however, it is a straightforward task to calculate the biofilm thickness that would be associated with a given fluidization regime, provided that sufficient substrate is sup-

plied to maintain that biofilm. If that thickness can be assumed to be representative of the average thickness that could be maintained in an FBBR with a given fluidization regime, then the biomass concentration can be calculated.[21,28-30]

An iterative approach must be used to calculate the biofilm thickness that can be maintained in an FBBR. Figure 18.9 summarizes an approach based on that of

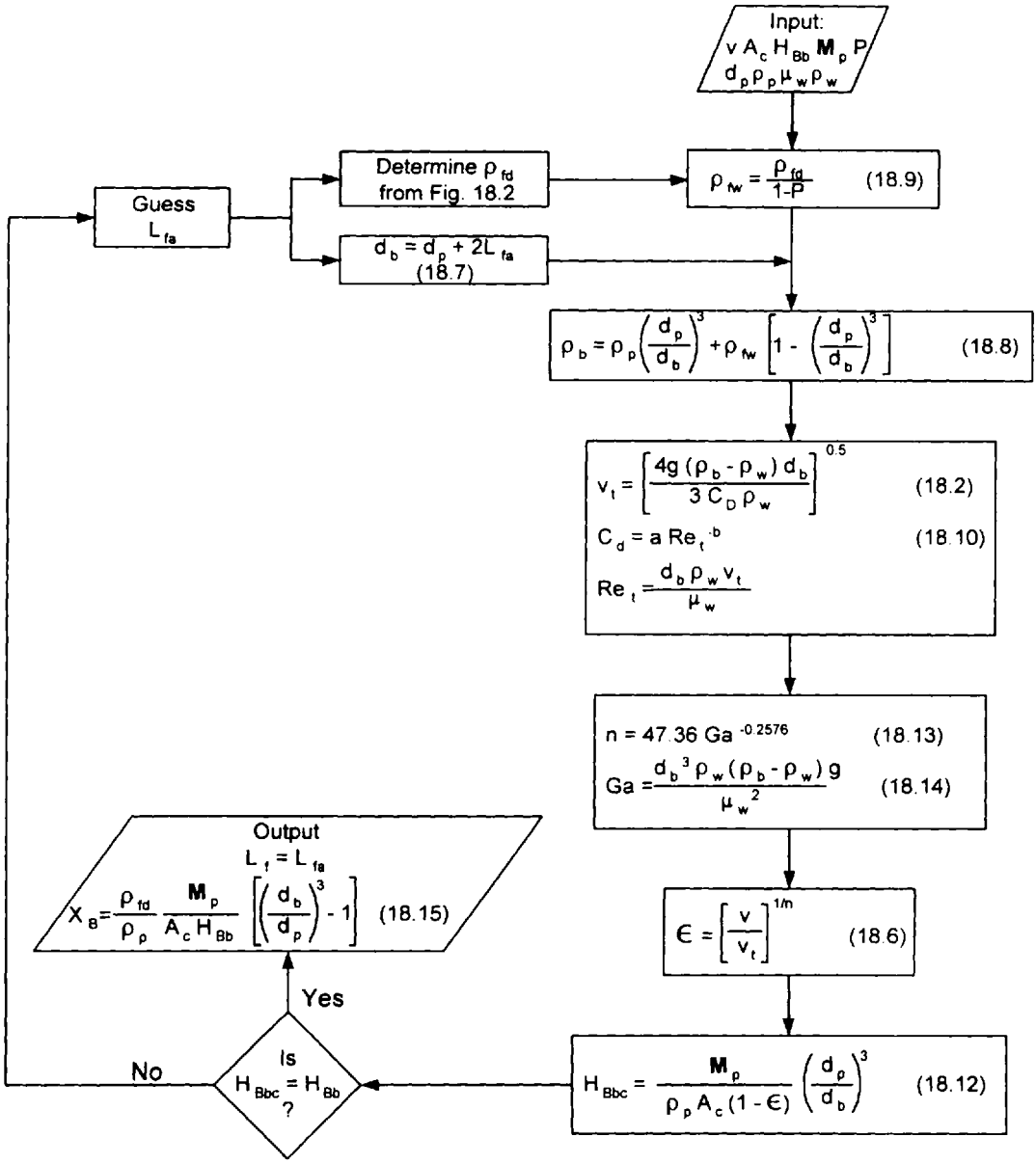


Figure 18.9 Algorithm for computation of the biofilm thickness associated with a fluidized bed.

Shieh and Keenan.[28] First, the characteristics of the FBBR must be established, including the desired superficial velocity, v, the FBBR cross-sectional area, A_c, the desired fluidized bed height, H_{Bb}, the mass of carrier particles, $\mathbf{M}_p$, their diameter, d_p, and their density, ρ_p. In addition, the properties of the fluid such as its density and viscosity must be established, as should the biofilm moisture content, P′. The computation begins by assuming a biofilm thickness. The assumed value is given the symbol L_{fa} to denote it as an assumed value. The biofilm dry density associated with the assumed biofilm thickness can be determined from information such as that in Figure 18.2 or an appropriate empirical equation, allowing the biofilm wet density to be calculated with Eq. 18.9. The bioparticle diameter can be calculated with Eq. 18.7, and that, in turn, can be used to calculate the bioparticle density with Eq. 18.8. The terminal settling velocity of the bioparticle can then be calculated with Eq. 18.2 (substituting d_b for d_p and ρ_b for ρ_p) using a relationship for C_D such as one of the ones in Figure 18.5 as expressed with Eq. 18.10. The coefficient n in the Richardson–Zaki equation can then be estimated with Eq. 18.13, allowing the porosity of the fluidized bed to be calculated with Eq. 18.6. It can then be used to calculate the bed height. The calculated value is denoted as H_{Bbc}. The value of H_{Bbc} is then compared to the desired bed height used to begin the computations. If they are equal, then the assumed biofilm thickness is correct and can be taken as the true thickness, L_f. If $H_{Bbc} > H_{Bb}$, then the assumed biofilm thickness is too large and a new smaller value should be assumed for repeating the computations. Conversely, if $H_{Bbc} < H_{Bb}$, a larger biofilm thickness should be assumed. Finally, once the correct biofilm thickness has been found, the concentration of biomass per unit volume of fluidized bed, X_B, can be calculated with:

$$X_B = \frac{\rho_{fd}}{\rho_p} \frac{\mathbf{M}_p}{A_c H_{Bb}} \left[\left(\frac{d_b}{d_p} \right)^3 - 1 \right] \tag{18.15}$$

The procedure illustrated in Figure 18.9 can be used to investigate the effect of biofilm thickness on the biomass concentration in an FBBR. The result of such an exercise is shown in Figure 18.10[28] for sand as the carrier particle with a diameter of 0.4 mm. The biomass dry density was assumed to be constant with a value of 65 g/L, which is consistent with Figure 18.2 over the range of biofilm thicknesses considered. Examination of the figure reveals that a biofilm thickness of 100 μm maximizes the concentration of biomass. Beyond that thickness, the increase in biomass associated with a thicker biofilm is offset by the reduction in the number of particles per unit volume due to increased bed expansion. If the objective was to maximize biomass concentration, then a biofilm thickness of 100 μm should be chosen. It should be recognized, however, that such a strategy may not optimize overall FBBR performance. To do that, consideration must also be given to the effectiveness of the biofilm.[28] We examine that question below.

It should also be recognized that Eq. 18.15 simply determines the biomass concentration and biofilm thickness that could be maintained by the hydrodynamic conditions in the bed. It does not tell whether they can be supported by the substrate loading on the bioreactor or whether a desired effluent substrate concentration can be achieved. One way to estimate whether a desired film thickness can be supported is to compare it to a steady-state biofilm. Because we wish to maintain the biofilm in a dynamic state, the biofilm thickness in the FBBR must be equal to or less than the steady-state biofilm thickness. To calculate the steady-state biofilm thickness,

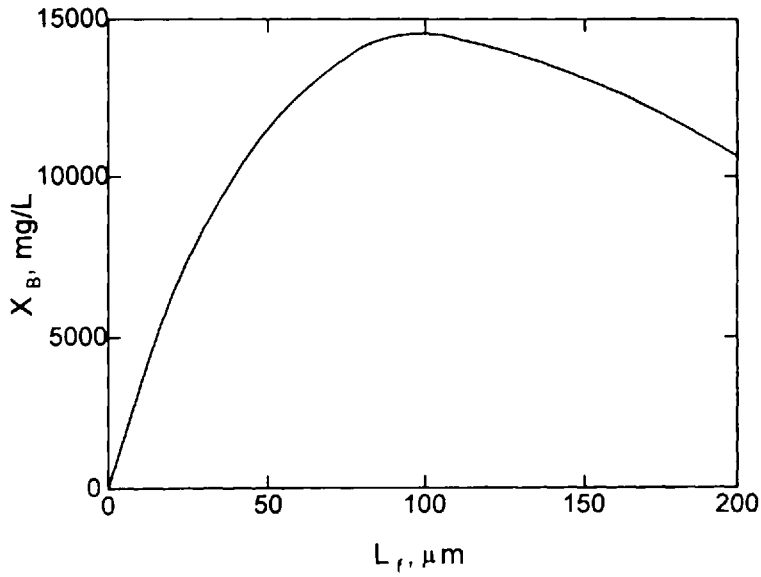


Figure 18.10 Effect of biofilm thickness, L_f, on the biomass concentration, X_B, in an FBBR. The following conditions were assumed: d_p = 0.4 mm, ρ_p = 2.65 g/cm^3, v = 1 cm/sec, ρ_{Bd} = 65 g/L, P′ = 0.93. (From W. K. Shieh and J. D. Keenan, Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology* **33**:131–169, 1986. Copyright © Springer-Verlag New York, Inc.; reprinted with permission.)

assume that the FBBR is completely mixed and that all bioparticles are exposed to a substrate concentration that is the average of the applied (considering recirculation) and the desired effluent substrate concentrations. Knowing that concentration, the steady-state biofilm thickness can be determined using the pseudoanalytical approach, as illustrated in Example 15.2.3.1. As long as the calculated biofilm thickness is less than the steady-state value, it is possible to support that thickness. Whether the computed biomass concentration can achieve the desired effluent concentration requires application of an FBBR model incorporating the assumed mass of carrier particles supporting the desired biofilm thickness.

18.3 MODELING FLUIDIZED BED BIOLOGICAL REACTORS

The modeling of FBBR performance requires the integration of information from several sources.[21] First, a biofilm model must be available that gives the rate of substrate conversion by individual bioparticles. Such a model can be developed using one of the approaches presented in Chapter 15. The only major difference is the incorporation of spherical coordinates to account for the shape of the bioparticles. Second, a model must be available that accounts for the effects of fluidization on the biofilm thickness and the number of bioparticles per unit of fluidized bed volume that can be maintained by the hydrodynamic conditions imposed. Such a model was

presented in Figure 18.9. Finally, one must have available an overall bed model that links the biofilm and fluidization models to yield substrate concentration as a function of axial position within the FBBR. The structure of this component model depends on the fluid regime within the FBBR.

Several FBBR models are available in the literature,[1,2,12,14,20,21,23,26,28,29,32] although many are variations of the same basic model. Most use an effectiveness factor approach for modeling transport and reaction within the biofilm and thus consider only a single limiting nutrient. In addition, most assume a uniform carrier particle size and consider the biofilm thickness to be uniform throughout the bed. In spite of these simplifications, the modeling of FBBRs is the most complex of all of the biochemical unit operations because of the interactions between the biofilm and fluidization submodels. Space does not allow us to consider all of the features of the models. Rather, only the major points are given and the reader is encouraged to consult the references for more details.

18.3.1 Biofilm Submodel

The biofilm submodel must consider simultaneous reaction and transport. Consequently, correlations must be available for relating the liquid phase mass transfer coefficient, k_l, to the hydraulic conditions in the bioreactor. Several have been proposed.[2,12,28] Shieh and Keenan[28] recommend the use of a correlation developed for fluidized beds from experimental data collected at Reynolds numbers within the range common to FBBR operation:

$$k_l = \frac{0.81}{\varepsilon}\left[\frac{D_w^{1.333} v \rho_w^{0.333}}{\mu_w^{0.333} d_b}\right] \tag{18.16}$$

This correlation reveals that a typical value of k_l is 0.01 cm/s for FBBR conditions, which some believe is sufficiently high to allow external mass transfer resistance to be ignored.[21] Consequently, this is frequently done to simplify computations.[1,14,21,28] The most exact approach, however, would be to consider both internal and external mass transfer resistance as was done in Chapter 15.

The effectiveness factor approach is the most common method of handling simultaneous reaction and transport in the biofilm, although all of the approaches in Chapter 15 have been used. While models[12] are available that use effectiveness factors for intrinsic Monod kinetics through the use of relationships like that in Figure 15.9, more assume the limiting case of either zero order ($S_{sb} >> K_s$) or first order ($S_{sb} < K_s$) intrinsic kinetics. For zero order kinetics in the absence of external mass transfer resistance, the effectiveness factor is defined as the biofilm volume containing substrate divided by the total biofilm volume.[21] Consequently, when the bioparticle is fully penetrated with substrate, the effectiveness factor has a value of 1.0. For first order kinetics in the absence of external mass transfer resistance, the effectiveness factor has been defined as the substrate flux into the spherical bioparticle divided by the intrinsic rate when all of the biofilm is surrounded by substrate at the bulk substrate concentration.[21]

For both first and zero order kinetics in the absence of external mass transfer resistance, the effectiveness factor can be correlated with an appropriate Thiele mod-

ulus. Furthermore, the correlations can be reduced to single curves for each type of kinetics by using an appropriate characteristic biofilm thickness, L_{fc}, which is defined as:

$$L_{fc} = \frac{\text{biofilm volume}}{\text{biofilm exterior surface area}} \tag{18.17}$$

$$L_{fc} = \frac{d_b^3 - d_p^3}{6d_b^2} \tag{18.18}$$

For zero order kinetics, the zero order effectiveness factor, η_{eZ}, can be correlated with a modified zero order Thiele modulus, ϕ_{Zm}, defined as:[28]

$$\phi_{Zm} = L_{fc}\left(\frac{\rho_{bd}\hat{q}_H}{D_e S_{sb}}\right)^{0.5} \tag{18.19}$$

The correlation is shown in Figure 18.11, and is described well by the following empirical equation:[28]

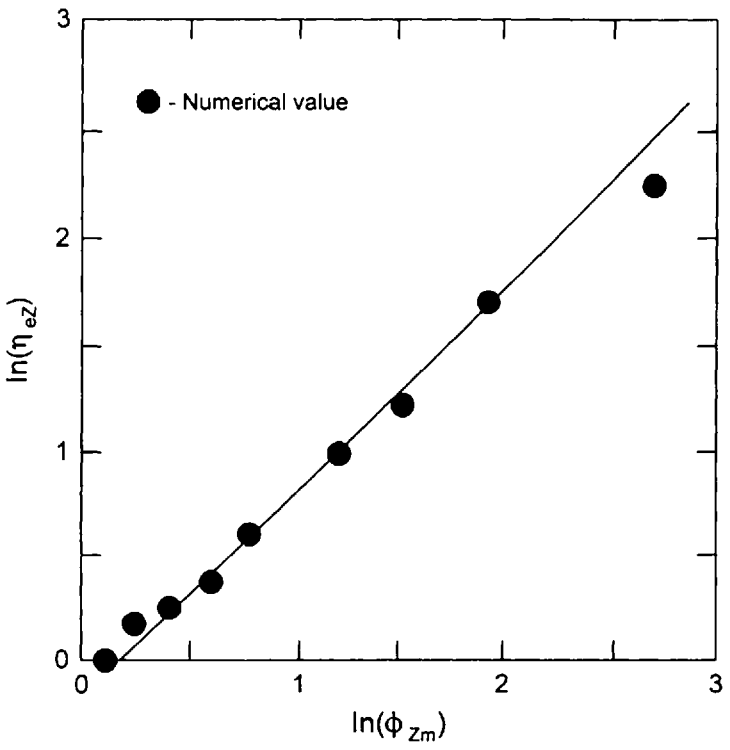


Figure 18.11 Relationship between the bioparticle zero-order effectiveness factor, η_{eZ}, and the modified zero-order Theile modulus, ϕ_{Zm}. (From W. K. Shieh and J. D. Keenan, Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology* **33**:131–169, 1986. Copyright © Springer-Verlag New York, Inc.; reprinted with permission.)

$$\eta_{cZ} = \frac{1.2712}{\phi_{Zm}} \tag{18.20}$$

For first order kinetics, the first order effectiveness factor, η_{c1}, can be correlated with a modified first order Thiele modulus, ϕ_{1m}, defined by Eq. 15.12 except for the use of the characteristic film thickness, L_{fc}:[28]

$$\phi_{1m} = L_{fc}\left(\frac{\rho_{iH}\cdot\hat{q}_H}{K_S\cdot D_e}\right)^{0.5} \tag{18.21}$$

(Recall that ρ_{iH} and $X_{B,H}$ are the same.) It should be noted that $\hat{q}_H/K_S$ is equivalent to k_c, the mean reaction rate coefficient defined Eq. 3.45. The correlation between η_{c1} and ϕ_{1m} is shown in Figure 18.12, and is described well by the first order, non-spherical form for homogeneous reaction media proposed by Aris:[5]

$$\eta_{c1} = \frac{\coth(3\phi_{1m})}{\phi_{1m}} - \frac{1}{3\,\phi_{1m}^2} \tag{18.22}$$

The points in the figures were computed for a variety of bioparticle and carrier particle sizes, thereby demonstrating that the characteristic film thickness works well as a normalizing factor.

Because the characteristic biofilm thickness serves as a normalizing factor that allows effectiveness factor correlations developed for planar coordinates to be used with spherical particles, it should be possible to use the general correlation for Monod kinetics shown in Figure 15.9 by using an appropriately modified Thiele modulus. This would allow external mass transfer resistance to be handled with little additional effort. Therefore, depending on the kinetic and mass transfer characteristics of the system, Eq. 18.20, Eq. 18.22, or Figure 15.9 can be used to calculate the substrate removal rate by bioparticles surrounded by substrate at a given concentration. This information can then be used to calculate system performance in the same manner as used in Chapters 15–17.

Another approach used in Chapter 15 for determining substrate removal rates by biofilms is the pseudoanalytical approach with a steady-state biofilm. The assumption of a steady-state biofilm is consistent with the situation encountered in packed towers and rotating disk reactors, but is inconsistent with FBBRs from which bioparticles are constantly wasted, as discussed previously. Thus, while FBBR models are available that assume steady-state biofilms, one must question their relevance to most operating FBBRs. As a consequence, steady-state biofilm FBBR models are not discussed here.

18.3.2 Fluidization Submodel

The effect of the fluidization regime on the thickness of biofilm that can be maintained on carrier particles of a given size and density constrained within a bed of fixed height was discussed in Section 18.2.3. Figure 18.9, shown there, presents the algorithm for calculating that biofilm thickness, and as such, represents a fluidization submodel that can be used. No further discussion of it is needed here. It should be noted, however, that such a model assumes a uniform biofilm thickness throughout the FBBR, which may not conform to reality for some FBBRs, as pointed out previously. Other models[1,23] are capable of handling variations in particle size within

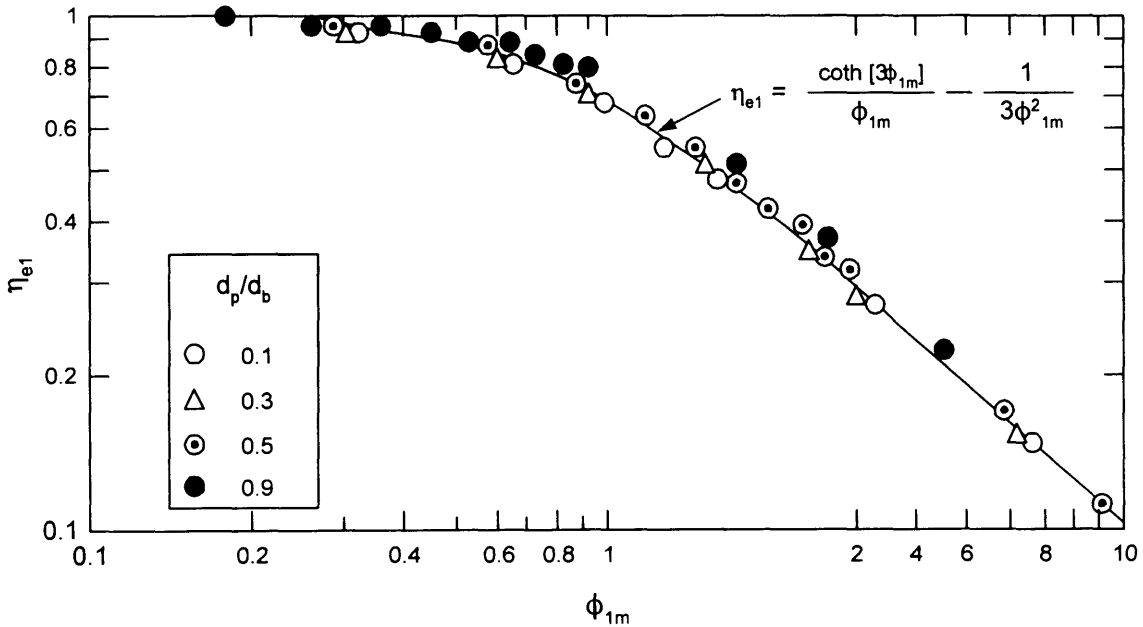


Figure 18.12 Relationship between the bioparticle first-order effectiveness factor, η_{e1}, and the modified first-order Theile modulus, ϕ_{1m}. (From W. K. Shieh and J. D. Keenan, Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology* **33**:131–169, 1986. Copyright © Springer-Verlag New York, Inc.; reprinted with permission.)

the bed, but because of space constraints and their added complexity, they are not discussed here. Rather, the reader is referred to the cited papers.

18.3.3 Reactor Flow Submodel

The reactor flow submodel must link the biofilm and fluidization submodels to allow computation of the performance of an FBBR. The type of model that should be used depends on the hydraulic regime in the FBBR. The nature of that regime is determined primarily by the degree of substrate utilization across the bed and the amount of effluent recirculated to maintain the appropriate fluidization velocity and to provide the required amount of electron acceptor. If the recirculation ratio is high and the influent substrate concentration is low, so that the change in substrate concentration across the bed (after dilution of the influent by the recirculation flow) is small, then the bed can be considered to behave as if it were a completely mixed reactor. The validity of this approach can be checked easily by comparing the diluted influent concentration as calculated with Eq. 16.5 to the assumed effluent concentration. On the other hand, if the recirculation ratio is low (e.g., <2) or the degree of dilution is small, then the FBBR must be treated as a plug–flow reactor, either with or without axial dispersion, or as a series of continuous stirred tank reactors (CSTRs). However, consideration need not be given to changes in the degree of axial dispersion from point to point in an FBBR due to differences in the porosity because that level of complexity cannot be justified.[1] Examples of all of these approaches can be found in the literature, depending on the situation being modeled. The equations used are typical of these various flow regimes as discussed in previous chapters and thus are not presented here.

Basically the approach to FBBR modeling is iterative, with the number of loops depending on the reactor flow submodel. Only a completely mixed FBBR (both bioparticles and liquid) is considered to describe the procedure, but the concepts can be extended to other flow regimes, which usually require more iterative loops. First, the characteristics of the FBBR must be established, including the desired superficial velocity, v, the FBBR cross-sectional area, A_c, the desired fluidized bed height, H_{fb}, the mass of carrier particles, $\mathbf{M}_p$, their diameter, d_p, and their density, ρ_p. The biofilm thickness that can be maintained by these conditions can then be computed using the procedure in Figure 18.9. That thickness determines the bioparticle diameter, which defines the characteristic biofilm thickness, L_{fc}, which is used to determine the effectiveness factor in the biofilm submodel. If the effectiveness factor expression includes the bulk substrate concentration, then one must be assumed. It is equivalent to the effluent substrate concentration for a completely mixed FBBR. The biofilm submodel is then used in the reactor flow submodel to compute the output substrate concentration. This is a direct computation for a completely mixed FBBR, but an iterative procedure is required for a plug–flow or tanks-in-series flow regime. If the computed concentration is different from the assumed value, then a new value must be assumed and the procedure repeated until the computed effluent concentration agrees with the value assumed. This is the effluent substrate concentration from the FBBR. The entire procedure can be repeated for different initial conditions, i.e., v, $\mathbf{M}_p$, d_p, or ρ_p, thereby relating performance to those conditions. This allows identification of the conditions giving an effluent concentration equal to or less than some desired value.

18.4 THEORETICAL PERFORMANCE OF FLUIDIZED BED BIOLOGICAL REACTORS

We saw in Section 18.2 that for a given superficial velocity the expansion of a fluidized bed depends on the size and density of the carrier particles, as well as on the thickness of the biofilm. In addition, for a given degree of expansion (porosity), the number of particles per unit bed volume also depends on those factors. Consequently, the biomass concentration in the bed is influenced by them as well. The effect of biofilm thickness on the biomass concentration was illustrated in Figure 18.10, and similar figures could be generated illustrating that values that maximize the biomass concentrations exist for the carrier particle size and density also.[28,29] Thus, from consideration of the effects of fluidization alone, it can be seen that complex interactions exist among the factors that influence FBBR performance. Fluidization effects do not tell the whole story, however. Because of the need for transport of reactants into the biofilm, not all of the biomass has the same activity. That is why the effectiveness factor is less than 1.0. Furthermore, the effectiveness factor depends on the size of the bioparticle and the thickness of the biofilm, as reflected in the modified Thiele moduli as used in Figures 18.11 and 18.12. This suggests that the combination of bioparticle characteristics that maximizes the ability of the FBBR to remove substrate is different from that which maximizes biomass concentration.[28,29] Because each situation is unique and complex, mathematical models are required for their analysis and several have been developed that integrate submodels of the type discussed in the preceding section.[1,2,4,12,13,21,23,26,28,29]

The theoretical performance of FBBRs can be examined with those models. The result from one such exercise is shown in Figure 18.13.[29] It shows the effect of particle diameter and biofilm thickness on the time required for 90% removal of substrate by biomass with an intrinsic zero-order reaction in an FBBR containing a fixed mass of carrier particles operated with a fixed superficial velocity. The flow regime in the FBBR was characterized as plug–flow. Thus, the required reaction time corresponds to the fractional height in the bed at which 90% of the substrate is removed. In other words, it corresponds to a required bed height and media mass. Examination of the figure reveals that the optimal (smallest) reaction time (and therefore bed size) is associated with moderately thin biofilms growing on small carrier particles. In fact, others have shown that for a given carrier particle size, the bioparticle effectiveness factor is maximized when the biofilm thickness is slightly less than the thickness at which all substrate would be exhausted.[2,4] Consequently, the optimum biofilm thickness depends on the diffusivity of the substrate in the biofilm and the biodegradation kinetics. The benefit of small carrier particles derives directly from the fact that for a given mass of carrier particles, the surface area for biofilm growth increases as the carrier particle diameter decreases. Nevertheless, the curvature associated with the optimal region in Figure 18.13 is relatively shallow in both dimensions, suggesting that the designer has some latitude in selecting a carrier particle size and the desired biofilm thickness, i.e., fluidization conditions. Similar conclusions regarding the relative effects of carrier particle size and biofilm thickness have also been reached with a model assuming first order intrinsic kinetics.[2,4] Thus, they can be considered to be general.

The above information considers selection of the optimal carrier particle size and biofilm thickness. Once they have been fixed and the mass of carrier particles

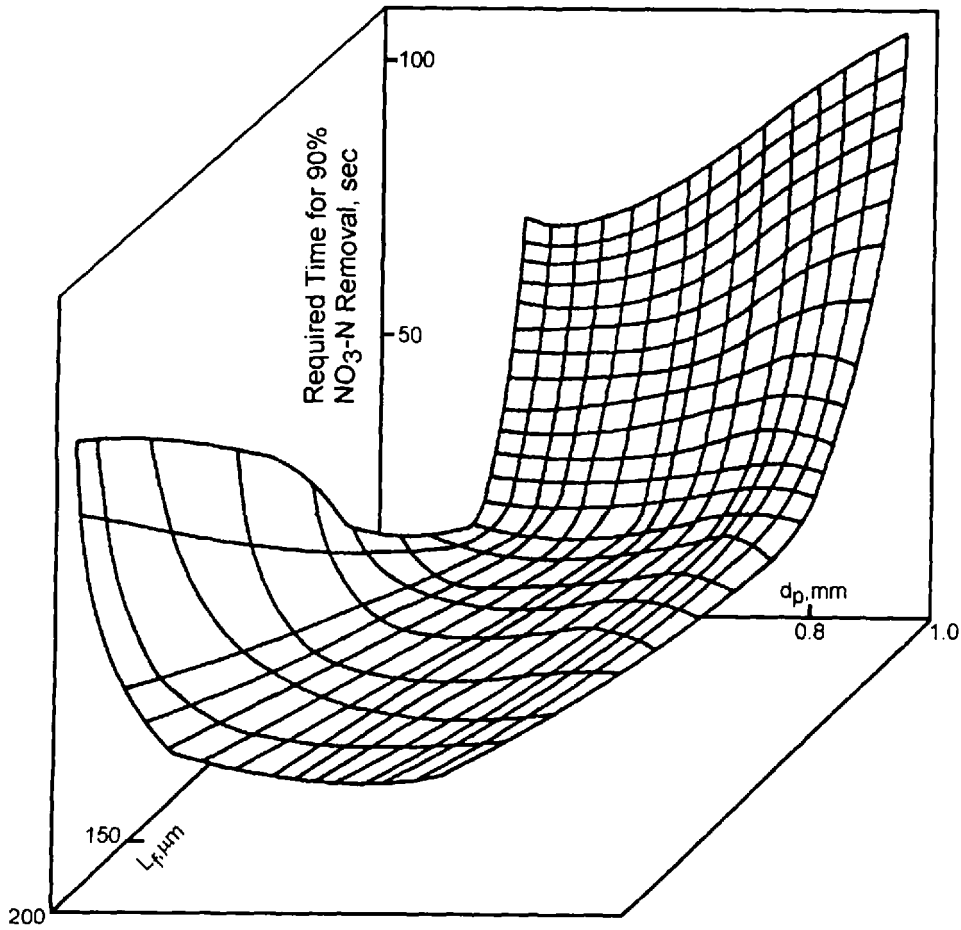


Figure 18.13 Combined effects of carrier particle diameter, d_p, and biofilm thickness, L_f, on the time required to remove 90% of the influent nitrate in a denitrifying FBBR. (From W. K. Shieh, L. T. Mulcahy, and E. J. LaMotta, Mathematical model of the fluidized bed biofilm reactor. *Enzyme and Microbial Technology* **4**:269–275, 1982. Copyright © Elsevier Science Ltd.; reprinted with permission.)

and porosity have been selected, then the required superficial velocity (and associated bed diameter) and bed height become fixed. A question then arises about the performance of the FBBR if the influent flow rate or substrate concentration changes. Some understanding of the response can be obtained by considering what would happen to the quantity of biomass in the system, which depends on how the FBBR is operated. Consider the case in which biomass wastage is practiced to maintain a fixed bed height and the recirculation rate is adjusted to maintain a constant superficial velocity through the bed. If the influent flow rate or concentration was increased while maintaining the same mass input rate of substrate and the same superficial velocity, then the impact on system performance would be minimal because the mass input of substrate per unit of biomass would stay about the same. On the other hand,

if the mass input rate of substrate increased, the output substrate concentration would increase by a proportionally greater amount. Because of the increased input, the biomass would be exposed to higher substrate concentrations, which would cause it to grow faster, leading to thicker biofilms. This would cause the degree of expansion in the bed to increase, which would reduce the number of carrier particles associated with the fixed bed height. Even though each carrier particle left in the system would have a thicker biofilm, the mass of biomass in the system would decrease because the increased biomass on each carrier particle would not compensate for the loss of carrier particles. Furthermore, the effectiveness of each carrier particle would be decreased because of the increased film thickness. On the other hand, if the recirculation rate was decreased to maintain the same mass of carrier particles within the prescribed bed height, the increase in the output substrate concentration would not be as great because more biomass could be maintained in the system. Likewise, if the bed height was allowed to expand to accommodate the increased mass input rate of substrate, there would be little impact on performance.

The above suggests that an FBBR can be thought of somewhat like a suspended growth system. If a suspended growth CSTR is operated at a fixed solids retention time (SRT), the mass of biomass in the system increases in proportion to an increase in the influent mass flow rate of substrate, and the effluent concentration remains the same. On the other hand, if it is operated at a fixed mixed liquor suspended solids (MLSS) concentration, the process loading factor (U) increases (the SRT decreases), and the effluent concentration increases. The FBBR acts similarly. If it is operated in a manner that allows the mass of biomass to increase, the impact of an increase in the mass input rate is minimal. Conversely, if the operational practice results in the same or less biomass, performance suffers.

The concept of SRT in a fluidized bed is a helpful one, but one must recognize that the situation is more complex than in a suspended growth bioreactor because of the mass transfer limitations in biofilms.[9] Long SRTs can lead to thick biofilms, which have a lower effectiveness factor. Thus, FBBRs with long SRTs can have lower volumetric removal rates. Conversely, at short SRTs, even though the biofilms are thin, the amount of biomass may be insufficient to get good removal. In other words, the fact that there is an optimal biofilm thickness associated with a given particle size means that there is also an optimum SRT for a given situation.

18.5 SIZING A FLUIDIZED BED BIOLOGICAL REACTOR

The sizing of an FBBR proceeds in a logical and straight forward manner, utilizing the information presented earlier in this chapter. As with all other biological processes, the parameters in the model must be specified, as must the influent flow rate and concentration, and the desired substrate removal across the system. Shieh and Keenan[28] have presented procedures for estimating the needed parameters. The sizing of the FBBR entails choosing a porosity, the carrier particle, the optimal biofilm thickness, the superficial velocity and the associated recirculation and bioreactor cross-section, and the bed height.[2]

Andrews[2] advocates using the smallest porosity that prevents the particles from agglomerating or having a collision frequency that would cause excessive shear. He states that a porosity of 0.60 is a reasonable compromise for the minimum porosity in the fluidized bed, which would be at the base. If the porosity is fixed, the next decision is to select the nature of the carrier particles. This is an important decision because everything else follows from it. We saw in Figure 18.4 that carrier particle size and density have an important impact on the stability of the bed. Light, small particles form a bed that is susceptible to large fluctuations in expanded bed height by small variations in superficial velocity. Sand (ρ_p = 2.65) with a particle diameter of around 0.5–0.6 mm is commonly used because it is readily available and offers good stability.

Having chosen the porosity and the carrier particle, it is now possible to choose the biofilm thickness that maximizes the average volumetric reaction rate in the system, which is equivalent to the product of the effectiveness factor times the biomass concentration. Using the average substrate concentration across the tower, the effectiveness factor can be calculated as a function of biofilm thickness using the characteristic biofilm thickness, L_c, given by Eq. 18.18 and the appropriate modified Thiele modulus. If it is necessary to consider external mass transfer resistance in determining the effectiveness factor, then the superficial velocity associated with each biofilm thickness has to be computed for use in the appropriate mass transfer coefficient correlation. The biomass concentration can be calculated with Eq. 18.23, which was derived by substituting Eq. 18.12 into Eq. 18.15:

$$X_B = \rho_{bd}(1 - \varepsilon)\left[1 - \left(\frac{d_p}{d_b}\right)^3\right] \tag{18.23}$$

It should be recalled that the dry density of the biofilm is a function of its thickness, as shown in Figure 18.2. Consequently, an appropriate correlation should be used with Eq. 18.23. The optimal biofilm thickness is obtained by plotting the product of the effectiveness factor times X_B as a function of the biofilm thickness and selecting the value that maximizes the product.

Once the optimum biofilm thickness has been chosen, the superficial velocity required to achieve the desired porosity can be calculated with Eq. 18.6. The terminal settling velocity of the bioparticle can be calculated with Eq. 18.2 after replacement of ρ_p with ρ_b and d_p with d_b. The value of the coefficient n can be calculated with Eqs. 18.13 and 18.14. At this point, a check should be made to ensure that the required superficial velocity does not exceed the terminal settling velocity of the clean carrier particle or any bioparticle desired in the bioreactor. If it does, and there is no error in the computations, then the chosen carrier particle is not feasible for the desired biofilm thickness and another must be selected.

Having selected the superficial velocity, it is now possible to determine the required cross-sectional area and associated height for the FBBR. The superficial velocity in an FBBR is equivalent to the total hydraulic loading in a packed tower, which was given by Eq. 16.8. Thus:

$$v = \Lambda_H = \frac{F(1 + \alpha)}{A_c} \tag{18.24}$$

Since the superficial velocity is known, A_c can be calculated after the recirculation ratio, α, has been chosen. Several factors go into the selection of α.[28] When an aerobic two-phase FBBR is being used, oxygen is provided by dissolving it in the recirculation stream. When high-purity oxygen is used for the supply, approximately 60 mg/L of oxygen can be dissolved in wastewater and used in the FBBR with less than 1% loss.[28] Thus, the amount of recirculation required can be calculated from a mass balance on chemical oxygen demand (COD) across the FBBR. Recirculation can also be provided to maintain a constant superficial velocity across the FBBR when the influent flow is variable. Often, however, it is best to set the influent equal to the highest expected flow rate across the system and to add to it the amount of recirculation required to transfer the needed oxygen, and to use those values in Eq. 18.24 to calculate the cross-sectional area.

Finally, after selection of α and A_c, the tower height, H_{fb}, can be calculated. This requires use of the appropriate reactor flow submodel as described in Section 18.3.3. If the recirculation flow is small, then a model for plug-flow or plug-flow with dispersion would be appropriate. Alternatively, a tanks-in-series model could be used as well. On the other hand, if the recirculation rate is high, it might be possible to treat the entire bioreactor as a CSTR. In addition, if the bed is likely to be stratified with a variety of bioparticle sizes, that expectation can be incorporated into the reactor flow submodel. The computational procedure involved depends on the type of model employed. The goal, however, is to determine the residence time or expanded bed height required to achieve the required effluent substrate concentration. Once H_{fb} is known, then the mass of carrier particles, M_p, can be calculated with Eq. 18.12.

18.6 KEY POINTS

1. A fluidized bed bioreactor (FBBR) is one in which the biofilm grows attached to small carrier particles that remain suspended in the fluid, i.e., fluidized, by the drag forces associated with the upward flow of water. It has several advantages over other attached growth bioreactors: better control of biofilm thickness, superior mass transfer characteristics, less tendency to clog, very high surface areas for biofilm development, and low pressure drops.
2. One attribute of FBBRs is that the dry density of the biofilm that develops on a bioparticle depends on the thickness of the biofilm, with thinner films exhibiting higher density. Consequently, thicker biofilms do not necessarily lead to more active biomass.
3. Two superficial velocities are critical to defining the operating range for a fluidized bed, those associated with points A and B in Figure 18.3. The velocity associated with point A is called the minimum fluidization velocity and is the velocity at which the particles just begin to move apart. The velocity associated with point B is the terminal settling velocity of

the carrier particles. If it is exceeded the particles are carried away in continuous fluidization and the bed is destroyed.

4. The growth of biofilm on carrier particles changes their terminal settling velocity. This is due to three things. First, growth of the biofilm changes the size of the particle. Second, unless the carrier particle has a density equivalent to the wet density of the biofilm, growth of the biofilm changes the overall effective density of the particle. Third, the surface properties of the biofilm differ from those of the clean particle, thereby changing the relationship between the drag coefficient, C_D, and the Reynolds number.
5. Because growth of a biofilm changes the terminal settling velocity of a particle, it also changes its fluidization properties. As a consequence, the height of a fluidized bed increases as the thickness of the biofilm increases.
6. Bioparticles containing uniform diameter carrier particles of low density, similar to the wet density of the biofilm, tend to stratify because the bioparticle density doesn't change as the biofilm grows. Under such a situation, biomass wastage should be done from the bottom of the bed. Conversely, bioparticles containing uniform diameter carrier particles of high density tend to form well mixed beds because the density of the bioparticle decreases as the biofilm grows. Biomass wastage from these beds should be done from the top.
7. The biofilm thickness that can be maintained by the hydrodynamic conditions in an FBBR is controlled by the superficial upflow velocity imposed on the bioreactor and the desired fluidized bed height. It must be calculated iteratively by the procedure illustrated in Figure 18.9.
8. Modeling the performance of an FBBR requires linkage of a biofilm submodel with fluidization and reactor flow submodels. The procedure is iterative for most situations.
9. For a given mass of carrier particles with a given density, there is a combination of biofilm thickness and carrier particle diameter that maximizes the conversion rate of substrate per unit bioreactor volume. The biofilm thickness associated with that optimum is usually slightly less than the biofilm depth at which all substrate would be exhausted, and is usually small.
10. When sizing an FBBR, a porosity around 0.60 should be chosen because it prevents the particles from agglomerating or having a collision frequency that would cause excessive shear. Sand (ρ_p = 2.65) with a particle diameter of around 0.5–0.6 mm is commonly used as a carrier particle because it is readily available and offers good stability of bed height against changes in superficial velocity.

18.7 STUDY QUESTIONS

1. Describe the general characteristics of an FBBR, including its advantages over other attached growth processes, and differentiate between a tower bioreactor and a supported-film bioreactor.

2. Explain why thin biofilms are often more desirable than thick ones.
3. Describe what happens to the pressure drop and the porosity of a bed of small, spherical particles as the superficial upflow velocity through it is increased and explain why those events occur.
4. Spherical sand (ρ_p = 2.65 g/cm^3) particles with a diameter of 0.6 mm are to be fluidized by water at a temperature of 20°C. Determine the minimum velocity for fluidization and the terminal settling velocity of the particles. Assume that the minimum porosity at fluidization is 0.45. State the relationship that you chose to use to determine the drag coefficient from the terminal Reynolds number and justify your choice.
5. Describe how growth of a biofilm influences the terminal settling velocity of a bioparticle containing sand as the carrier particle. Also explain why the effect occurs.
6. Rework Study Question 4, but assume that a biofilm with a thickness of 0.10 mm has grown on the carrier particle. Assume that the weight fraction moisture content of the biofilm is 0.93 and that its dry density varies with the biofilm thickness as shown in Figure 18.2.
7. An FBBR has a diameter of 7.5 cm and contains 1.0 kg of spherical sand (ρ_p = 2.65 g/cm^3) carrier particles with a diameter of 0.6 mm. Determine the bed height when a biofilm with a thickness of 0.10 mm has grown on the particles and they are fluidized with a superficial velocity of 2.5 cm/sec. Assume that the weight fraction moisture content of the biofilm is 0.93 and that its dry density varies with the biofilm thickness as shown in Figure 18.2.
8. Explain how and why the density of the carrier particle influences the degree of solids mixing that occurs in an FBBR.
9. An FBBR has a diameter of 7.5 cm and contains 1.0 kg of spherical sand (ρ_p = 2.65 g/cm^3) carrier particles with a diameter of 0.6 mm. Determine the biofilm thickness that could be carried on the bioparticles if the bed height is maintained at 1.0 m while the bed is being fluidized with a superficial velocity of 1.5 cm/sec. Assume that the weight fraction moisture content of the biofilm is 0.93 and that its dry density varies with the biofilm thickness as shown in Figure 18.2.
10. Explain the significance of the characteristic biofilm thickness as defined by Eqs. 18.17 and 18.18.
11. Prepare a flow diagram showing the steps that must be followed to calculate the effluent substrate concentration from a completely-mixed FBBR with intrinsic zero-order kinetics.
12. Repeat Study Question 11 for intrinsic first-order kinetics.
13. Prepare a flow diagram showing the steps that must be followed to calculate the effluent substrate concentration from an FBBR that has a plug–flow liquid phase, but a uniform bioparticle size. Assume intrinsic first-order kinetics.
14. Explain why the biofilm thickness that maximizes the quantity of biomass in an FBBR may not be the same as the biofilm thickness that maximizes the conversion rate of substrate per unit bioreactor volume.
15. Prepare a flow diagram showing the steps that must be followed in sizing an FBBR to achieve a desired effluent concentration.

REFERENCES

1. Andrews, G. F., Fluidized-bed fermenters: a steady-state analysis. *Biotechnology and Bioengineering* **24**:2013–2030, 1982.
2. Andrews, G. F., Selecting particles for fluidized-bed bioreactors with flocculent biomass. *Biotechnology Progress* **2**:16–22, 1986.
3. Andrews, G. F., Fluidized-bed bioreactors. *Biotechnology and Genetic Engineering Reviews* **6**:151–178, 1988.
4. Andrews, G. F. and J. Przezdziecki, Design of fluidized-bed fermentors. *Biotechnology and Bioengineering* **28**:802–810, 1986.
5. Aris, R., *Elementary Chemical Reactor Analysis*, Prentice-Hall, Englewood Cliffs, New Jersey, 1969.
6. Chang, H. T., B. E. Rittmann, D. Amar, R. Heim, O. Ehlinger and Y. Lesty, Biofilm detachment mechanisms in a liquid-fluidized bed. *Biotechnology and Bioengineering* **38**: 499–506, 1991.
7. Cooper, P. F. and B. Atkinson, eds., *Biological Fluidized Bed Treatment of Water and Wastewater*, Ellis Horwood Publishers, Chichester, England, 1981.
8. Coulson, J. M., J. F. Richardson, J. R. Backhurst, and J. H. Harker, *Chemical Engineering, Volume Two, Unit Operations*, Third Edition, Pergamon Press, Oxford, 1978.
9. Hermanowicz, S. W. and Y.-W. Cheng, Biological fluidized bed reactor: hydrodynamics, biomass distribution and performance. *Water Science and Technology* **22**:(1/2):193–202, 1990.
10. Hermanowicz, S. W. and J. J. Ganczarczyk, Some fluidization characteristics of biological beds. *Biotechnology and Bioengineering* **25**:1321–1330, 1983.
11. Hermanowicz, S. W. and J. J. Ganczarczyk, Dynamics of nitrification in a biological fluidized bed reactor. *Water Science and Technology* **17**(2/3):351–366, 1984.
12. Hermanowicz, S. W. and J. J. Ganczarczyk, Mathematical modeling of biological packed and fluidized bed reactors. In *Mathematical Models in Biological Waste Water Treatment*, S. E. Jorgensen and M. J. Gromiec, eds. Elsevier, pp. 473–524, 1985.
13. Kargi, F. and J. K. Park, Optimal biofilm thickness for fluidized-bed biofilm reactors. *Journal of Chemical Technology and Biotechnology* **32**:744–748, 1982.
14. Kim, B. R., Approximate solution for a fluidized-bed biofilm model. *Water Research* **26**:1271–1275, 1992.
15. Lazarova, V. and J. Manem, Advances in biofilm aerobic reactors ensuring effective biofilm activity control. *Water Science and Technology* **29**(10/11):319–327, 1994.
16. McCabe, W. L., J. C. Smith and P. Harriott, *Unit Operations of Chemical Engineering*, Fifth Edition, McGraw-Hill, New York, 1993.
17. Mishra, P. N. and P. M. Sutton, Biological fluidized beds for water and wastewater treatment: a state-of-the-art review. *Biodeterioration and Biodegradation* **8**:340–357, 1991.
18. Mulcahy, L. T. and E. J. LaMotta, *Mathematical Model of the Fluidized Bed Biofilm Reactor*, Report No. 58–78–2, Department of Civil Engineering, University of Massachusetts at Amherst, 1978.
19. Mulcahy, L. T. and W. K. Shieh, Fluidization and reactor biomass characteristics of the denitrification fluidized bed biofilm reactor. *Water Research* **21**:451–458, 1987.
20. Mulcahy, L. T., W. K. Shieh and E. J. LaMotta, Kinetic model of biological denitrification in a fluidized bed biofilm reactor (FBBR). *Water Science and Technology* **12**(6): 143–157, 1980.
21. Mulcahy, L. T., W. K. Shieh and E. J. LaMotta, Simplified mathematical models for a fluidized bed biofilm reactor. *Water - 1980, AIChE Symposium Series*, **77**(209):273–285, 1981.
22. Myška, J. and J. Švec, The distributive properties of a fluidized bed with biomass. *Water Research* **28**:1653–1658, 1994.

23. Nieuwstad, T. J., Modeling, optimization and design of fluidized beds for biological denitrification. *Water Science and Technology* **17**(2/3):367–383, 1984.
24. Ro, K. S. and J. B. Neethling, Terminal settling characteristics of bioparticles. *Research Journal, Water Pollution Control Federation* **62**:901–906, 1990.
25. Ro, K. S. and J. B. Neethling, Biofilm density for biological fluidized beds. *Research Journal, Water Pollution Control Federation* **63**:815–818, 1991.
26. Shieh, W. K., Suggested kinetic model for the fluidized-bed biofilm reactor. *Biotechnology and Bioengineering* **22**:667–676, 1980.
27. Shieh, W. K. and C.-Y. Chen, Biomass hold-up correlations for a fluidised bed biofilm reactor. *Chemical Engineering Research and Design* **62**:133–136, 1984.
28. Shieh, W. K. and J. D. Keenan, Fluidized bed biofilm reactor for wastewater treatment. *Advances in Biochemical Engineering/Biotechnology* **33**:131–169, 1986.
29. Shieh, W. K., L. T. Mulcahy and E. J. LaMotta, Mathematical model of the fluidized bed biofilm reactor. *Enzyme and Microbial Technology* **4**:269–275, 1982.
30. Shieh, W. K., P. M. Sutton and P. Kos, Predicting reactor biomass concentration in a fluidized-bed system. *Journal, Water Pollution Control Federation* **53**:1574–1584, 1981.
31. Sutton, P. M. and P. N. Mishra, Activated carbon based biological fluidized beds for contaminated water and wastewater treatment: a state-of-the-art review. *Water Science and Technology* **29**(10/11):309–317, 1994.
32. Tang, W.-T. and L.-S. Fan, Steady-state phenol degradation in a draft tube, gas–liquid–solid fluidized-bed bioreactor. *AIChE Journal* **33**:239–249, 1987.
33. Thomas, C. R. and J. G. Yates, Expansion index for biological fluidized beds. *Chemical Engineering Research and Design* **63**:67–70, 1985.

Part V
Applications: Attached Growth Reactors

Part IV presents the fundamental principles of ideal attached growth reactors and their application to packed towers, rotating disk reactors, and fluidized bed reactors. In Part V, those principles are applied to the practical design and operation of a variety of attached growth reactors. Chapter 19 addresses the design of trickling filters, the principal packed tower used in practice. Chapter 20 addresses rotating biological contactors, the principal rotating disk reactor used in practice. Both have been widely used in practice for removal of biodegradable organic matter, combined carbon oxidation and nitrification, and separate stage nitrification. Finally, Chapter 21 addresses a variety of submerged fixed film reactors that have undergone various degrees of development and application in practice. These include downflow and upflow packed bed reactors, fluidized bed reactors, and combined suspended and attached growth reactors. They have been used for removal of biodegradable organic matter, combined carbon oxidation and nitrification, separate stage nitrification, and denitrification using both the carbon in the wastewater itself and supplemental carbon (such as methanol). As in Part III, several process design approaches are presented, representing diverse degrees of sophistication and information requirements. The reader is referred to Chapter 9 for discussions of the iterative nature of biological process design and the need for a variety of design procedures.

This page intentionally left blank

19
Trickling Filter

The term trickling filter represents an array of attached growth biochemical operations in which wastewater is applied to fixed media in an air-filled packed tower. Treatment of the wastewater is accomplished by microorganisms growing attached to the media, of which there are several types. Trickling filters are aerobic and are used to oxidize biodegradable organic matter, forming biomass. The produced biomass sloughs from the media and is separated from the treated wastewater in a downstream clarifier. Trickling filters are also used to oxidize ammonia-N to nitrate-N. Nitrification can occur either in a trickling filter that is being used for oxidation of organic matter, a process called combined carbon oxidation and nitrification, or in a trickling filter receiving wastewater that has previously been treated to remove organic matter, a process called separate stage nitrification. Both are discussed in this chapter; the theoretical performance of packed towers is discussed in Chapter 16.

19.1 PROCESS DESCRIPTION

The trickling filter was one of the first biochemical operations developed for the treatment of municipal and industrial wastewaters. Initial experimentation with the use of gravel beds for wastewater treatment occurred at the Lawrence Experiment Station in Massachusetts in 1889.[62] This was followed by research in England in the late 1890s and early 1900s, research in the United States in the early 1900s, and initial full-scale applications in the United States in the late 1900s and early 1910s. The popularity of the trickling filter increased throughout the first half of the twentieth century until, in the 1950s, it was the most popular biochemical operation in the United States. Then, during the 1960s and 1970s, the popularity of the activated sludge process increased due to better economic and performance characteristics relative to other processes available at that time. Since then, however, new media designs, process configurations, and increased understanding of biofilm processes have resulted in improved trickling filter economics and performance, causing a resurgence in use.[61,62] As a result, the trickling filter is a viable option for many carbon oxidation and nitrification applications.

19.1.1 General Description

Figure 19.1 presents a schematic diagram of a trickling filter. A typical trickling filter consists of five major components: (1) the media bed, (2) the containment structure,

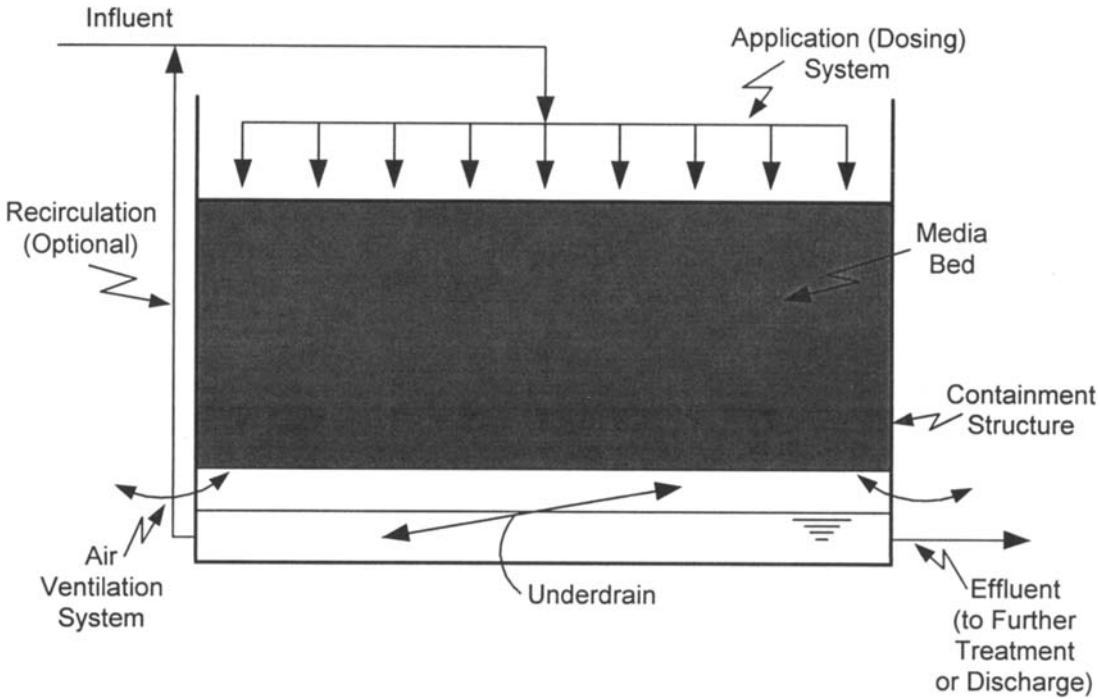


Figure 19.1 Schematic diagram of a trickling filter.

(3) the wastewater application (or dosing) system, (4) the underdrain system, and (5) the ventilation system.[61,62] The media bed provides the surface upon which the microorganisms grow. Media options consist of rock, wood, and synthetic plastic of various types and configurations.

The containment structure retains the media and applied wastewater and controls the effects of wind. Some media, such as rock and random plastic, are not self-supporting and, in these instances, the containment structure must also support the media. The containment structure is often constructed of concrete, either poured in place or precast panels. Other materials such as wood, fiberglass, and coated steel have also been used, particularly when the media is self-supporting.

The application system uniformly applies the wastewater to the media bed. Uniform application is necessary to ensure wetting of all of the media. The application system is also used to control dosing frequency, which affects process performance.

The underdrain system has two functions. The first is to collect the treated effluent for conveyance to further treatment or to discharge. The second is to provide a plenum to allow air passage through the open media bed, thereby providing the oxygen required for aerobic metabolism. Clay or concrete underdrain blocks are often used for rock media trickling filters because of the weight that must be supported. Many types of underdrain systems, such as concrete piers, wood stringers, and reinforced fiberglass grating, are used with other media.

Oxygen required to meet the metabolic needs of the microorganisms is provided by the vertical flow of air through the media. As discussed in Section 19.2.5, ventilation to provide that air can be supplied either by natural draft or by mechanical means. In natural draft systems, the difference in density between air inside and outside the trickling filter causes air within the trickling filter to either rise or sink. This results in a continuous flow of air through the media. Density differences arise because air within the trickling filter quickly becomes saturated with water vapor and reaches the temperature of the applied wastewater. Consequently, the magnitude of the density difference depends on the temperature and humidity of the ambient air. One disadvantage of natural draft ventilation is that neutral density conditions can occur, resulting in the absence of air-flow through the trickling filter and the development of anaerobic conditions. In forced draft ventilation systems, the air is applied to the trickling filter by mechanical means. In all cases, air must be uniformly distributed across the media to ensure that oxygen is provided to the entire bioreactor.

As indicated in Figure 19.1, trickling filter effluent may be recirculated and mixed with the influent wastewater prior to its application to the trickling filter. Recirculation dilutes the influent wastewater and also allows separation of the hydraulic and organic loadings to the unit. Recirculation is an essential process component in some applications. The need for recirculation and the various recirculation configurations are discussed later.

The influent to a trickling filter must generally be pretreated to remove nonbiodegradable particulate matter such as plastics, rags, and stringy material. Materials of this type can easily plug the distributor and the media, leading to unequal flow distribution and poor performance. Debris removal by coarse screens is not generally acceptable for trickling filter applications, but adequate removal can be accomplished using fine screens (generally 1 mm opening or less) or primary clarifiers. Primary clarifiers are used most often.

The media provides a surface for the growth of microorganisms and the mechanism for retaining the microorganisms in the unit. Organic matter removal and nitrification occur by the same mechanisms as in any other aerobic biochemical operation. Soluble organic matter diffuses into the biofilm located on the media surface and is used as a carbon and energy source by heterotrophic bacteria. Colloidal and particulate organic matter also diffuse into the biofilm where they are first removed by sorption and entrapment. They are subsequently hydrolyzed into soluble organic matter by the action of extracellular enzymes. The soluble organic matter is then metabolized by the heterotrophic bacteria contained within the biofilm. Ammonia-N also diffuses into the biofilm where part is used by the heterotrophs for biomass synthesis and the remainder is oxidized to nitrate-N by nitrifying bacteria. The removal of organic matter and nitrification result in the production of additional biomass and increased biofilm thickness. When the biofilm reaches a thickness that can no longer be supported on the media, the excess sloughs off and passes into the treated effluent. Chapter 15 describes the role of diffusion in controlling the metabolic processes occurring within the biofilm.

Trickling filter effluents are usually treated in clarifiers to remove the produced biomass, although this may not be necessary in some separate stage nitrification applications because of the low yield of nitrifying bacteria. For example, consider a separate stage nitrification application in which 20 mg/L of ammonia-N is being oxidized. Since the yield coefficient for nitrifying bacteria is approximately 0.15 mg TSS/mg ammonia-N oxidized, only about 3 mg/L of nitrifying bacteria will be produced. This increase in suspended solids concentration may not be significant in relation to plant effluent suspended solids limits, and thus a liquid–solids separation device may not be required downstream of the trickling filter.[57]

As discussed in Chapter 16, the liquid flow pattern through a trickling filter may generally be thought of as plug–flow with dispersion. Because of this flow pattern and because the microorganisms are fixed on the media, variations in the composition of the biomass often exist along the depth of the trickling filter. This is in contrast to the activated sludge process where biomass recycle results in a uniform biomass composition throughout the bioreactor. The variation in biomass composition through the depth of a trickling filter can have significant impacts on process performance. For example, carbon oxidation typically occurs in the upper portion of combined carbon oxidation and nitrification systems, while nitrification occurs in the lower portion,[59] as illustrated in Figure 19.2. This is due to competition between heterotrophic and autotrophic bacteria for space within the biofilm, as discussed in Chapter 15. In the upper levels of a trickling filter, both organic matter and ammonia-N concentrations are relatively high and will, generally, not limit the specific growth rate of either the heterotrophic or nitrifying bacteria. Under these conditions the heterotrophic bacteria can grow faster than the nitrifying bacteria and out compete them for space within the biofilm. As the wastewater flows down through the trickling filter, organic matter is removed, ultimately causing its concentration to limit the specific growth rate of the heterotrophic bacteria. Because the ammonia-N concentration is still high, a point is reached at which the specific growth rate of the nitrifying bacteria exceeds the specific growth rate of the heterotrophic bacteria. Under these conditions the nitrifying bacteria can effectively compete with the heterotrophic bacteria for space and will become established in the biofilm. As discussed

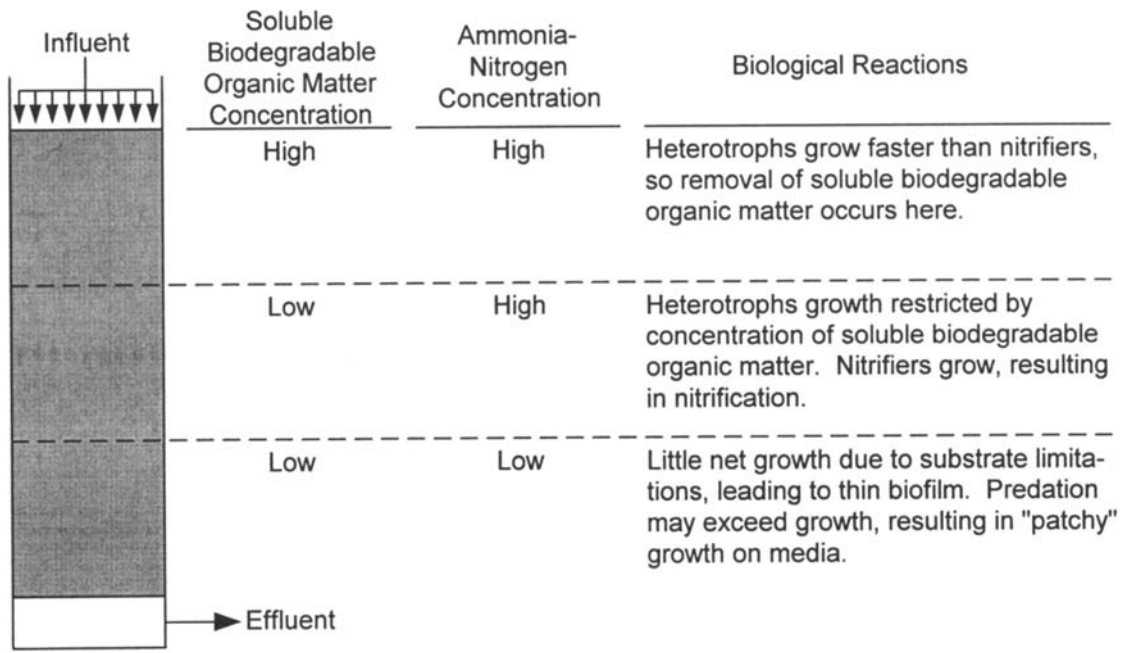


Figure 19.2 Representation of the biological reactions occurring at various depths in a trickling filter accomplishing combined carbon oxidation and nitrification.

in Chapter 15, the soluble biodegradable organic matter concentration must be reduced to about 20 mg/L as COD before this can occur.[41,59]

The plug–flow nature of the trickling filter can also result in reduced growth and biomass accumulation in the lower portion of the tower, leading to patchy growth as indicated in Figure 19.2.[6,44] This occurs because of the low yield of nitrifying bacteria and the presence of predators that consume trickling filter biomass. The reduced biofilm thickness can result in a diminished wastewater treatment capacity in the lower portion of the tower, which can be particularly important when the loading increases, such as during diurnal high flow events.

Although trickling filters are generally thought of as aerobic processes, in most cases the biofilm is relatively thick and exceeds the depth of oxygen penetration.[64] Consequently, the biofilm consists of an outer aerobic layer and an inner anoxic/anaerobic layer. This affects process performance in many significant ways. For example, the occurrence of a zone of low dissolved oxygen (DO) concentration within the biofilm allows denitrification to occur, although the extent is limited. The removal of organic matter typically occurs in the upper levels of the trickling filter, whereas nitrification occurs in the lower levels. As a consequence, the concentration of organic matter is low in the region where nitrate-N is produced. However, some denitrification can occur when treated effluent containing nitrate-N is recirculated to the process influent.[41]

19.1.2 Process Options

Trickling filter process options vary with the treatment objective, the media type, and the nature of the other unit operations in the process train.

Treatment Objectives. Trickling filters are used to treat a wide variety of wastewaters to achieve various treatment objectives. Consequently, as indicated in Table 19.1, those two factors can be used to characterize the trickling filter process. Because the degree of treatment is often determined by the process organic loading rate, it can be used as a quantitative indicator of the degree of treatment. The total organic loading is the mass flow rate of biodegradable organic matter in the influent wastewater (excluding recirculation) divided by the media volume, V_M. It is given the acronym TOL and the symbol A_S, and is calculated as:

Table 19.1 Trickling Filter Process Applications

Application/objective	Influent wastewater	TOL kg $BOD_5/(m^3 \cdot day)$
Roughing	Screened wastewater or primary effluent	1.5–3.5
Carbon oxidation	Screened wastewater or primary effluent	0.7–1.5
Combined carbon oxidation and nitrification	Screened wastewater or primary effluent	<1.0
Separate stage nitrification	Secondary effluent	NA[a]

[a]Not applicable.

$$\Lambda_S = \frac{F(S_{SO} + X_{SO})}{V_M} \tag{19.1}$$

The concentration of biodegradable organic matter is typically expressed as either five-day biochemical oxygen demand (BOD_5) or as chemical oxygen demand (COD), making the units for TOL kg $BOD_5/(m^3 \cdot day)$ or kg $COD/(m^3 \cdot day)$.

Influent wastewater containing both organic matter and ammonia-N can be treated in roughing, carbon oxidation, or combined carbon oxidation and nitrification applications. The difference between them is the degree of treatment, as indicated by the TOL.[61,62] A relatively high TOL is used in a roughing filter, resulting in residual organic matter and suspended solids concentrations that generally exceed the conventional definition of secondary treatment (30 mg/L each of BOD_5 and total suspended solids [TSS]). Roughing applications are generally used to lower the concentration of organic matter prior to further biological treatment. A somewhat lower TOL is used in a carbon oxidation application so that relatively complete removal of organic matter (approaching the definition of secondary treatment) is achieved. Finally, an even lower TOL is used to achieve combined carbon oxidation and nitrification. Carbon oxidation is relatively complete in the upper portion of the trickling filter, thereby allowing the growth of nitrifying bacteria in the lower portion, as described above.

Separate stage nitrification differs in that a stream that is relatively low in biodegradable organic matter and suspended solids (BOD_5 and TSS generally less than 30 mg/L), but with a significant ammonia-N concentration, is applied to the trickling filter.[6,15,44,57,61,62] As a consequence, the microbial population that develops in the biofilm is enriched in nitrifying bacteria. In this instance the TOL is not an issue. Instead, the ammonia loading and other operational factors determine the degree of ammonia removal. For example, the nitrification efficiency might be correlated with the total ammonia-N loading (TAL) with units of kg NH_3-$N/(m^3 \cdot day)$ and symbol Λ_{NH}. Alternatively, there may be instances in which it is advantageous to express nitrification performance in terms of the loading of total Kjeldahl nitrogen (TKN) on the process. In that case, one might speak of a total nitrogen loading (TNL) with units of kg $N/(m^3 \cdot day)$ and symbol Λ_N.

Media Type. Many types of media have been used in trickling filters, but in general, they can be divided into rock media and high-rate media.[11,61,62] Table 19.2 summarizes the characteristics of various rock and high-rate media, while Figure 19.3 presents photographs of several typical media.

An ideal rock media consists of rounded river rock of relatively uniform size and shape.[34,61,62] Typical rock media are approximately 5 cm in diameter, although larger and smaller rock can be used. Rounded river rock is durable, and its rounded nature and uniform size minimize bed consolidation and plugging. Irregular materials such as slag have also been used, but they exhibit a greater potential for plugging due to the entrapment of biomass by the irregular surfaces and openings. Media attrition will also occur due to freeze–thaw cycles, resulting in the production and accumulation of fines that contribute to plugging. This is particularly true for slag media due to their rough surfaces which allow water to intrude into the media.

As indicated in Table 19.2, the primary characteristics of rock trickling filter media are high unit weight, low specific surface area, and low void space. These impose significant design and operational constraints. The higher the unit weight,

Table 19.2 Characteristics of Trickling Filter Media

Media type	Description	Size cm	Unit weight kg/m^3	Specific surface area m^2/m^3	Void space %
Rock	Rounded river rock of uniform size (desirable) or more irregular slag (less desirable)	2.5–7.5 5–10	1,500 1,600	60 45	50 60
High-rate					
Bundle	Sheets of PVC formed in various configurations and fastened together to form bundles with openings of various sizes and orientations	61 × 61 × 122 61 × 61 × 122	30–80 65–95	88–105 140–150	>95 >95
Vertical, semi-corrugated (VSC)					
Vertical, fully corrugated (VFC)					
60° crossflow (XF)					
Random (RA)	Irregular shapes manufactured of extruded plastic elements	2–10 (varies for individual products)	30–65 50–80	80–115 140–165	>94 >94
Horizontal (HO)	Wooden pallets, typically constructed of redwood or pressure treated lumber	122 × 122 × 4.8	165	46	>90

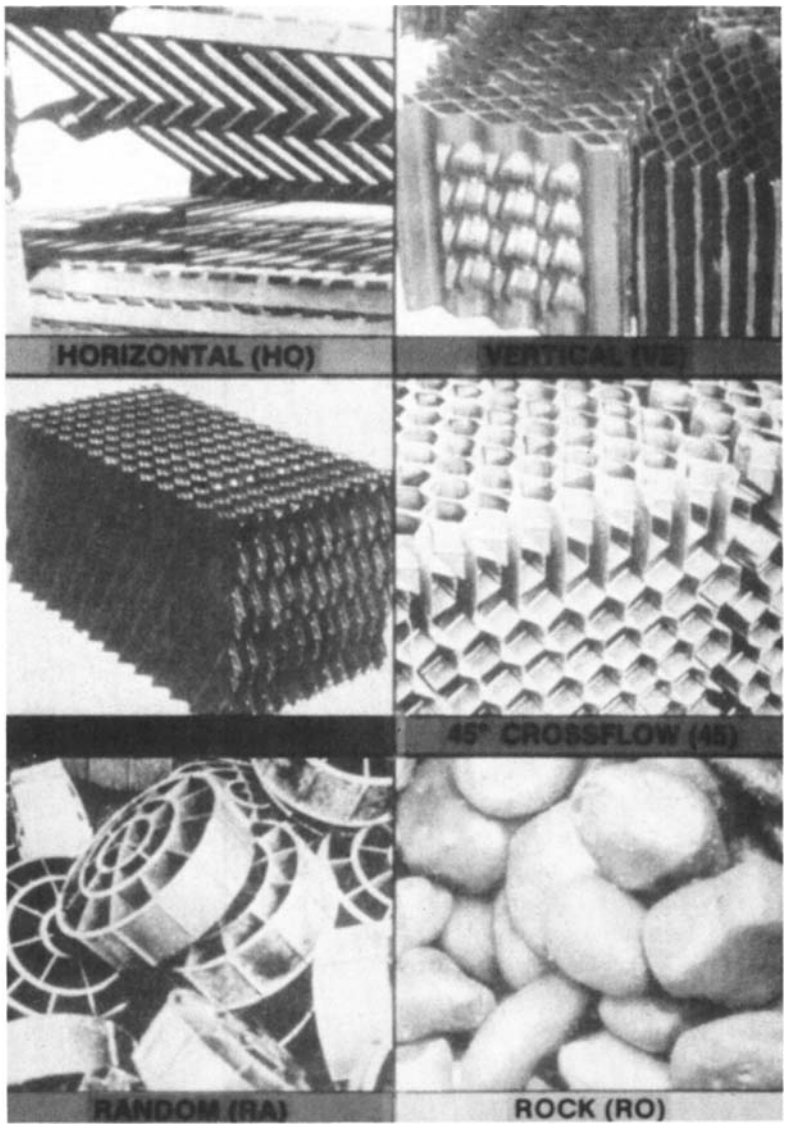


Figure 19.3 Typical trickling filter media. (From Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992. Copyright © Water Environment Federation; reprinted with permission.)

the greater the structural requirements and the shallower the media depth must be to avoid media crushing. For rock media trickling filters, media depths are typically on the order of 2 m, which restricts treatment efficiency to some extent. Some have been constructed to greater depths using special construction techniques, but this is not conventional practice. The relatively low specific surface area and void space restrict the organic loadings that can be applied. The low specific surface area limits

the capacity for biofilm growth and consequently, the treatment capability. Similarly, the low void volume limits the space available for the passage of air, water, and sloughed biomass through the media, thereby increasing the potential for media plugging at higher organic loadings. As a consequence, rock media have typically been used at low to moderate TOLs, generally in the range of 0.5 to 1.5 kg $BOD_5/(m^3 \cdot day)$. More significantly, the shallow media depth and low organic loadings result in low wastewater hydraulic application rates. Although recirculation can be used to provide increased total hydraulic loadings (THLs, see Eq. 16.8), typical design practice has been to limit recirculation ratios to minimize energy requirements. As a consequence, THL to rock media is typically on the order of 0.5 m/hr, which is significantly less than that used with high-rate media. The adverse impacts of these low THLs on process performance are discussed later.

As illustrated in Table 19.2, high-rate media are characterized by significantly lower unit weight, higher specific surface area, and greater void space than rock media. As a consequence, media depths are typically greater (generally 5 to 7 m), and a wider range of TOLs can be used (up to 3.5 kg $BOD_5/(m^3 \cdot day)$). Because of the greater media depth, the cross-sectional area of a high-rate media trickling filter will be significantly less than that of a comparably sized rock media trickling filter. A THL of 1.8 m/hr is typically considered necessary to fully wet and completely utilize high-rate trickling filter media.

The effects of media type on trickling filter performance are discussed more fully in Section 19.2.6, but a brief summary is provided here. Data suggest that the performance of rock and high-rate media trickling filters are similar when they are loaded at relatively low organic loading rates, less than about 1 kg $BOD_5/(m^3 \cdot day)$. In contrast, at higher TOLs the performance of trickling filters using high-rate media is superior. Other operating characteristics of rock and high-rate media trickling filters tend to be somewhat different. For example, excess biomass production rates may be somewhat less in rock media trickling filters than in high-rate media trickling filters.[21] This is thought to result from the greater biomass retention characteristics of rock media, resulting in increased degradation (probably anaerobic) of the retained biomass. The biomass retention characteristics may also result in variations in the settling characteristics of the produced biosolids and in the clarity of the settled effluent.

Plastic high-rate trickling filter media are manufactured with a variety of specific surface areas. Variations in the specific surface area are accomplished by modifications in the media configuration and dimensions. Consider plastic sheet (or bundle) media as an example. Increased specific surface area is accomplished by manufacturing media sheets with slightly smaller indentations, which allows media sheets to be placed closer together. Media with a specific surface area of approximately 100 m^2/m^3 is typically used when screened wastewater or primary clarifier effluent is being treated, whereas media with a specific surface area of approximately 140 m^2/m^3 is typically used when secondary effluent is being treated. The more open, low specific surface area media is needed to avoid plugging problems associated with the relatively high biomass production rates experienced when screened wastewater or primary clarifier effluent is being treated. Since the biomass production rate is lower when secondary effluent is being treated, media with a higher specific surface area media can be used in that application.

As indicated in Table 19.2, the specific surface area of wood media is relatively low. However, experience indicates that the application of return activated sludge to a trickling filter containing this media results in significant interstitial biomass growth, which increases its effective treatment capacity.[10,22] Side-by-side testing indicates that its efficiency will be equivalent to that of bundle media when RAS is recycled to the wood media. No similar improvement in performance is observed for the other media. Process configurations that recycle RAS to an upstream trickling filter are discussed in the next section.

Coupled Trickling Filter/Activated Sludge Systems. Coupled trickling filter/activated sludge (TF/AS) systems use an upstream trickling filter combined with a downstream suspended growth biochemical operation to accomplish overall wastewater treatment.[10,22] Figure 19.4 illustrates a typical system. Such systems are referred to as containing coupled processes because no liquid–solids separation device is provided between the trickling filter and the suspended growth bioreactor. As a consequence, biomass that grows on the trickling filter sloughs off and passes directly into the suspended growth bioreactor where it is enmeshed into and becomes part of the suspended biomass. A significant portion of the biomass contained in the suspended growth bioreactor (generally 60% to 90%) is originally grown in the trickling filter.

Trade-offs exist relative to the sizes of the two biochemical operations. If a relatively small trickling filter is used, then a larger suspended growth bioreactor must be used to accomplish a specified treatment goal. Conversely, if a relatively large trickling filter is used, the treatment goal can be accomplished using a smaller

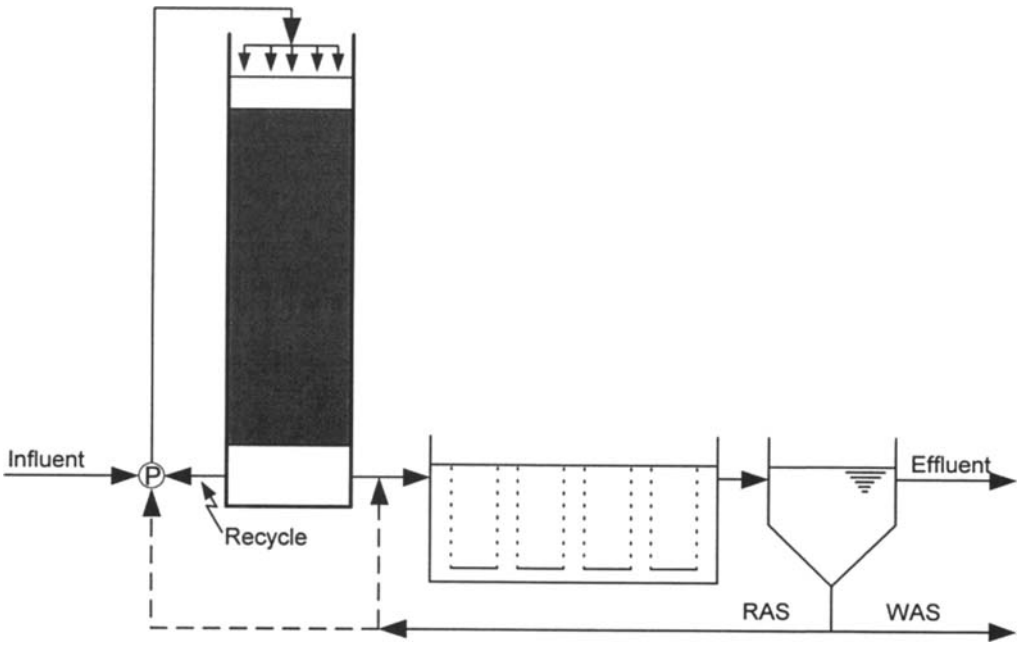


Figure 19.4 Schematic diagram of the coupled trickling filter/activated sludge (TF/AS) process.

Table 19.3 Coupled TF/AS System Options

Process option	Unit size		RAS recycle destination
	Trickling filter	Suspended growth bioreactor	
Trickling filter/solids contact (TF/SC)	Large	Small	Suspended growth bioreactor
Activated biofilter (ABF)	Large	None	Trickling filter
Roughing filter/ activated sludge (RF/AS)	Small	Large	Suspended growth bioreactor
Biofilter/activated sludge (BF/AS)	Small	Large	Trickling filter

suspended growth bioreactor. This trade-off provides one of the criteria for classifying coupled TF/AS systems. Another criterion is the location for the return activated sludge (RAS); either around the entire system, i.e., to the trickling filter, or around only the suspended growth bioreactor. Table 19.3 uses these two criteria to classify coupled TF/AS systems.

Both the trickling filter/solids contact (TF/SC) and the activated biofilter (ABF) processes use relatively large trickling filters and small suspended growth bioreactors. In both of these processes, removal of organic matter is achieved in the trickling filter. The suspended growth bioreactor is used primarily to flocculate and enmesh fine suspended solids contained in the trickling filter effluent and to prepare the biosolids for efficient removal in the secondary clarifier. In the TF/SC process the RAS is recycled to the suspended growth bioreactor, which is referred to as a solids contact basin. In the ABF process, no distinct suspended growth bioreactor is provided. Rather, underflow from the clarifier is recycled to the trickling filter and the suspended biomass is developed and contained only in the recirculating fluid mass.

The roughing filter/activated sludge (RF/AS) and biofilter/activated sludge (BF/AS) processes use small trickling filters. As a consequence, the removal of organic matter in the trickling filter is incomplete and the suspended growth bioreactor removes the remaining organic matter. In the RF/AS process, the RAS is recycled only to the suspended growth bioreactor, whereas in the BF/AS process it is recycled to the trickling filter. In some instances (particularly the treatment of high concentration, readily biodegradable wastewaters such as from the food processing industry), recycle of the RAS to the trickling filter can improve sludge settling characteristics. This is thought to be related to a "selector effect" that is achieved by contacting the RAS with high concentrations of organic matter in the highly aerated biofilter environment.[21,22] The use of selectors to control sludge settleability in activated sludge systems is discussed in Section 10.2.1.

19.1.3 Comparison of Process Options

The trickling filter process is a stable, reliable process that is capable of providing economical wastewater treatment. Energy requirements are typically lower than in

suspended growth systems, which was one reason for the renewed popularity of trickling filters in the 1980s, a time when energy costs were escalating rapidly. The process is also relatively simple to operate. Capital costs can be high compared to other biochemical operations, and process operation cannot be adjusted as easily in response to loading and/or performance variations. The biggest drawback of trickling filters is that their performance may not meet current discharge standards. However, the performance of coupled trickling filter/activated sludge systems will generally equal that of suspended growth systems. The simplicity, stability, and low energy requirements for the trickling filter process have made it a popular option for carbon oxidation and nitrification. Primary clarification is typically provided prior to a trickling filter to minimize the debris loading.

Table 19.4 summarizes the benefits and drawbacks of the various trickling filter options. Roughing applications can provide very economical removal of organic matter, particularly from high-strength wastewaters. Further treatment of the roughing filter effluent (in addition to secondary clarification) is typically required prior to final discharge, but the size of the downstream treatment system is reduced.

Carbon oxidation applications offer the advantages of favorable economics, simple design and operation, and well known process and facility design procedures. Process performance is consistent and reliable, but may not meet the stringent standards now typically required.

Combined carbon oxidation and nitrification and separate stage nitrification applications are simple, both to design and to operate. However, experience with them is somewhat limited and performance relationships are still evolving. It is expected that this drawback will be eliminated in the future as more direct experience accumulates with these options.

A comparison of rock and high-rate media indicates substantial advantages for high-rate media. The only drawback associated with high-rate media is that media collapses can occur as a result of improper application or manufacturing. This suggests that experience is necessary in its application. Due to the associated benefits, nearly all new trickling filters are constructed using high-rate media. However, a large number of rock media trickling filters exist and many are providing effective and economical service.

The coupled trickling filter/activated sludge process was developed to take advantage of the energy efficiency and stability of the trickling filter while also achieving the excellent effluent quality obtained with the activated sludge process. Experience indicates that this objective is typically achieved. Performance differences between the four primary TF/AS options are relatively minor. An economic tradeoff exists between the more energy-efficient TF/SC and ABF processes and the less capital intensive RF/AS and BF/AS processes. Multiple modes of operation are often incorporated into full-scale facilities so that selection of a single coupled process option is not necessary.

19.1.4 Typical Applications

The trickling filter process is widely accepted and used for the aerobic biological treatment of wastewaters. It is used for both the removal of organic matter and nitrification. Its long history has resulted in a large number of operating installations. Many of the older installations use rock media, and a large number of successful

Table 19.4 Trickling Filter Process Comparison

Process	Benefits	Drawbacks
Treatment objectives		
Roughing	Economical, particularly for high-strength wastewaters Simple to design and operate Process and facility design well known	Further treatment typically required prior to discharge Generally requires secondary clarification
Carbon oxidation	Economical Simple to design and operate Process and facility design well known	Performance is consistent, but may not reliably meet stringent performance standards Generally requires secondary clarification
Combined carbon oxidation and nitrification	Simple to design and operate	Relatively new process option, performance relationships not well characterized Limited operator flexibility
Separate stage nitrification	Simple to design and operate	Relatively new process option, performance relationships not well characterized
Media type		
Rock	Large number of existing applications Quite effective at low to moderate organic loading rates	Relatively expensive due to structural constraints Not applicable for high loading applications Odor potential
High-rate	Economical Applicable to a wide range of process loadings and applications Process and facility design well known	Media collapses have occurred due to improper application and/or manufacturing

Coupled trickling filter/ activated sludge (TF/AS)		
Trickling filter/solids contact (TF/SC)	Stable, reliable performance Simple to design and operate Low energy Process and facility design well known	Moderate capital cost
Activated biofilter (ABF)	Simple to design and operate Low energy Low capital cost Process and facility design well known	Process performance variable, except at low loading
Roughing filter/activated sludge (RF/AS)	Stable, reliable performance Simple to design and operate Low capital cost Process and facility design well known	Moderate energy cost
Biofilter/activated sludge (BF/AS)	Stable, reliable performance Simple to design and operate Low capital cost Process and facility design well known Improved sludge settling characteristics in some applications	Moderate energy cost

installations of various size exist. Installations constructed in the last 20 years generally use high-rate media, either plastic sheet, random, or horizontal. Plastic sheet media is currently the most popular due to its availability, cost, and good performance characteristics. Horizontal media is seldom used in new installations today due to its higher costs relative to plastic media. However, many existing installations exist because of its popularity during the 1970s.

The use of high-rate media trickling filters for separate stage nitrification has been actively considered for more than two decades, and some facilities have been in operation for a number of years.[47,55,57,61,62] Interest in this application has increased recently as practitioners have identified potential cost and operational advantages. Given the level of interest currently expressed in this technology, it is reasonable to expect more applications to be constructed.

Roughing filters are often used to pretreat industrial wastewaters containing high concentrations of readily biodegradable organic matter. The roughing filter effluent (after secondary clarification) typically receives further treatment, either in another biological treatment system located at the industrial site or in a municipal wastewater treatment system. Roughing filters have also been used to pretreat mixtures of municipal and industrial wastewater containing high concentrations of organic matter.

Use of the trickling filter for carbon oxidation has declined since the 1950s due to its general inability to meet stringent discharge standards. The coupled TF/AS systems were developed in response to this performance short-fall and have been quite successful in meeting stringent discharge standards.[10,22,33,36] Such systems can be designed for carbon removal applications alone or to accomplish combined carbon oxidation and nitrification. They have generally proven effective in achieving this goal while also retaining the basic operating characteristics of the trickling filter process.

While the trickling filter process has been used widely, it still must be considered to be a developing process. This is because of the recent process and mechanical developments, which have improved performance relative to that experienced historically. Continued development of high-rate trickling filter media has resulted in good performance and very competitive media costs. Improved understanding of trickling filter hydraulics has led to revised hydraulic loading and wetting regimes that provide better control of biofilm thickness, thereby producing a biofilm that is thinner, more aerobic, and more active. Experience with these improved facilities is accumulating, and is very encouraging. Recent experience with nitrification in trickling filter applications has caused increased interest in these options. The ongoing development of coupled TF/AS process options has resulted in improved control over effluent suspended solids concentrations, thereby allowing trickling filter based processes to produce effluent quality that rivals that of the activated sludge process.

The RF/AS and BF/AS processes have proven to be quite effective for treating higher strength industrial or industrial/municipal wastewaters. The trickling filter provides process stability and control of filamentous microorganisms, while the suspended growth bioreactor allows an effluent of excellent quality to be produced. In addition, the use of relatively small trickling filters results in a system with moderate capital and operating costs. As noted above, experience indicates that recycle of the RAS to the trickling filter (thereby converting it to a biofilter) generally results in the most complete control of solids settling characteristics. The ABF process has

received relatively little use by itself, but many systems incorporate the flexibility to operate in this mode and it is used effectively to reduce energy costs during periods of lower process loading.

While basic information on the combined carbon oxidation/nitrification and separate stage nitrification processes has been available for nearly 20 years, significant interest in these processes has developed only recently.[14,41,52,56] They offer the potential for reduced energy costs, good process stability, and favorable economics.

Trickling filters have not proven to be as adaptable to nutrient removal as suspended growth bioreactors. While phosphorus removal in coupled TF/AS processes has been demonstrated, their removal capability is hindered by the oxidation of organic matter in the trickling filter.[51] Biological nitrogen removal is also problematic for the same reason. This must be considered when selecting a trickling filter based option for a new or expanded wastewater treatment facility.

19.2 FACTORS AFFECTING PERFORMANCE

Over its long history of use, a large data base has been assembled describing the factors affecting the performance of the trickling filter process. Unfortunately, in many cases the data are contradictory and incomplete. This arises largely because of the interrelation between trickling filter design and operational parameters. It also arises because our understanding of the trickling filter has evolved throughout its history and continues to evolve today. Our current understanding allows recognition that effects once thought to be significant are really artifacts of past design and operational practices and are not fundamental process variables. This section discusses the primary factors that affect trickling filter performance.

19.2.1 Process Loading

The performance of any biochemical operation is affected by the process loading, i.e., the amount of substrate applied per unit time per unit mass of biomass. In suspended growth systems the process loading is expressed as the solids retention time (SRT) or the process loading factor, i.e., F/M ratio. Both have physical meaning in a suspended growth system because the biomass is well mixed and can be characterized by a single parameter. In addition, the operational parameters necessary to calculate them are easily measured. This is not true for attached growth systems such as trickling filters. The biomass cannot be characterized by a single parameter because it is not uniformly distributed through the bioreactor. Furthermore, it is not possible to easily determine the biomass concentration within a trickling filter, thereby making it impossible to calculate either an SRT or a process loading factor. While some values for the biomass concentration in a trickling filter have been reported,[21] no consensus exists as to the appropriateness of this approach. As a consequence, other measures of process loading must be used.

As discussed in Chapter 15, substrate removal in biofilm processes is expressed by the substrate flux, J_S, which is the mass of substrate per unit time transported into and consumed by the biofilm per unit biofilm planar surface area. Logically, the biofilm process loading can be expressed in the same fashion, as the mass of substrate applied per unit time per unit of total planar biofilm area. For organic substrate, the

result, referred to as the surface organic loading (SOL), λ_S, is expressed in units of kg substrate/(m^2·day). The SOL is also sometimes referred to as the applied flux. Another approach to expressing the loading on a trickling filter is the TOL, defined by Eq. 19.1 and typically expressed in units of kg substrate/(m^3·day). The SOL and the TOL, Λ_S, are related by the specific surface area of the media, a_s:

$$\lambda_S = \frac{\Lambda_S}{a_s} = \frac{F(S_{SO} + X_{SO})}{A_s} \tag{19.2}$$

where A_s is the wetted surface area of media available for biofilm growth. Since the quantity of biomass in a trickling filter is proportional to both the media area and the media volume, both the TOL and the SOL are analogous to the process loading factor for suspended growth bioreactors. This analogy is incomplete, however, due to the fact that the biomass composition is uniform in a suspended growth system while it varies with bioreactor depth in a trickling filter, as discussed above. However, it is a useful analogy if not carried too far.

Theoretically the SOL is superior to the TOL as an expression of biofilm process loading. This is because, as discussed in Chapter 15, biofilm reactions are generally limited by mass transfer and the overall reaction rate is a function of the biofilm surface area. An inherent assumption in the concept of the SOL is that the biofilm surface area is proportional to the media surface area. A second assumption is that the substrate is dissolved and diffuses into the biofilm. While these assumptions are good ones for many biofilm processes, for some trickling filter applications they are not. This occurs for several reasons, including media plugging and inadequate wetting, the presence of colloidal organic matter in wastewater, and the occurrence of multiple substrate limitations. The result is that, for some trickling filter applications, performance correlates better with the TOL than with the SOL.

The concept of surface loading is not limited to the removal of organic substrate. Rather, it can also be applied to nitrification. For example, in a trickling filter performing only nitrification, the nitrification efficiency might be correlated with the surface ammonia-N loading (SAL) with units of kg NH_3-N/(m^2·day) and symbol λ_{SH}. Alternatively, one may wish to express nitrification performance in terms of the loading of TKN on the process. In that case, one might speak of a surface nitrogen loading (SNL) with units of kg N/(m^2·day) and symbol λ_S. These surface loadings are related to the corresponding total volumetric loadings in the same way that the SOL is related to the TOL, i.e., by Eq. 19.2.

Plugging and channeling of flow in a trickling filter results in incomplete wetting and utilization of the media surface area provided, thereby making the biofilm surface area less than the media surface area. Plugging and channeling especially occur when a trickling filter is used for carbon oxidation. This is because of high biomass production rates, which cause excess biomass to accumulate in the media. The resulting incomplete media wetting and partial media plugging produce variable and incomplete utilization of the media area provided.[27] Although media with higher specific surface areas are theoretically better, beyond a point they will not result in greater biofilm surface area because the smaller openings associated with the media will cause increased plugging and reduced wetting of the available media surface area. In short, more media surface area does not necessarily mean more treatment capacity. The situation is further complicated by the fact that, in many wastewaters, a significant portion of the biodegradable organic matter is present in either a sus-

pended or colloidal form. This organic matter is removed initially by flocculation and entrapment, just as in suspended growth systems. It is then hydrolyzed and the soluble organic matter degraded. It is not yet clear what controls the removal rate of such organic matter in a biofilm reactor. Moreover, its retention within the trickling filter can result in additional channeling of flow.

As discussed in Section 19.1.2 and illustrated in Table 19.1, the trickling filter TOL can be correlated with its treatment objectives. Performance relationships can also be presented graphically, as illustrated in Figure 19.5 where a typical relationship between the soluble BOD_5 removal efficiency of a trickling filter and its TOL is presented. The nitrification performance of trickling filters accomplishing combined carbon oxidation and nitrification can also be correlated with the TOL, as illustrated in Figure 19.6.[14] The performance data presented in this figure are also correlated with the SOL in the upper axis (called areal loading).

In trickling filters accomplishing either carbon oxidation or combined carbon oxidation and nitrification, it is likely that oxygen is the limiting substance. As discussed in Sections 15.3 and 19.1.1, when both biodegradable organic matter and ammonia-N are present, heterotrophs and autotrophs compete for space in the aerobic portion of the biofilm, with autotrophs being excluded until the concentration of biodegradable organic matter drops below about 20 mg/L as COD. Therefore, if the rate of oxygen transfer into the biofilm is limiting, in the upper portion of the trickling filter, the oxidation rate of biodegradable organic matter will be limited by the rate of oxygen transfer, whereas in the lower portion of the trickling filter oxygen transfer will limit the oxidation rate of ammonia-N. If both substrates are expressed

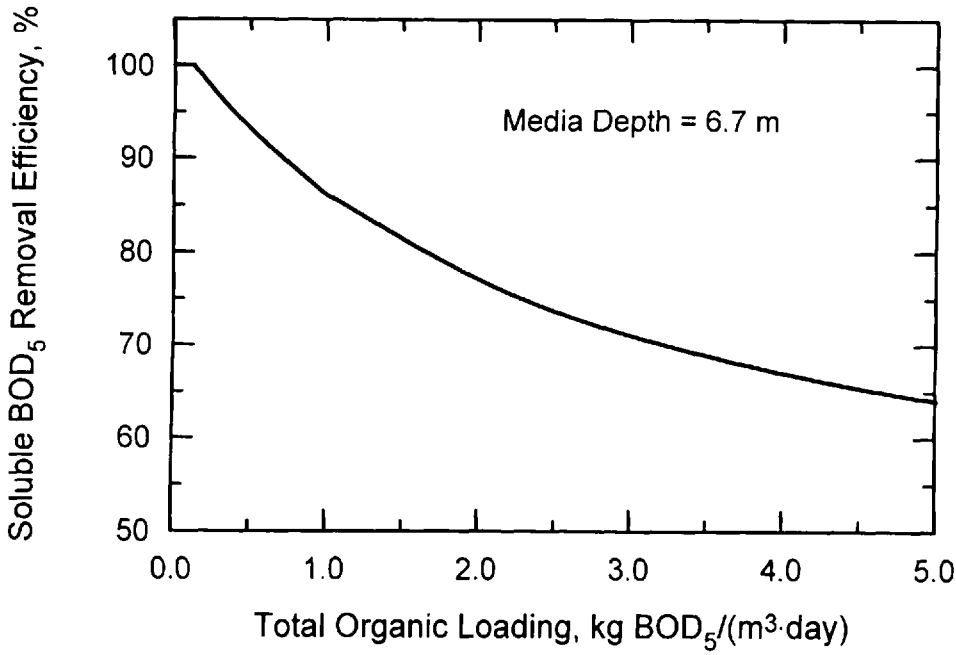


Figure 19.5 Typical relationship between total organic loading (TOL) and soluble BOD_5 removal efficiency for a trickling filter containing high-rate media.

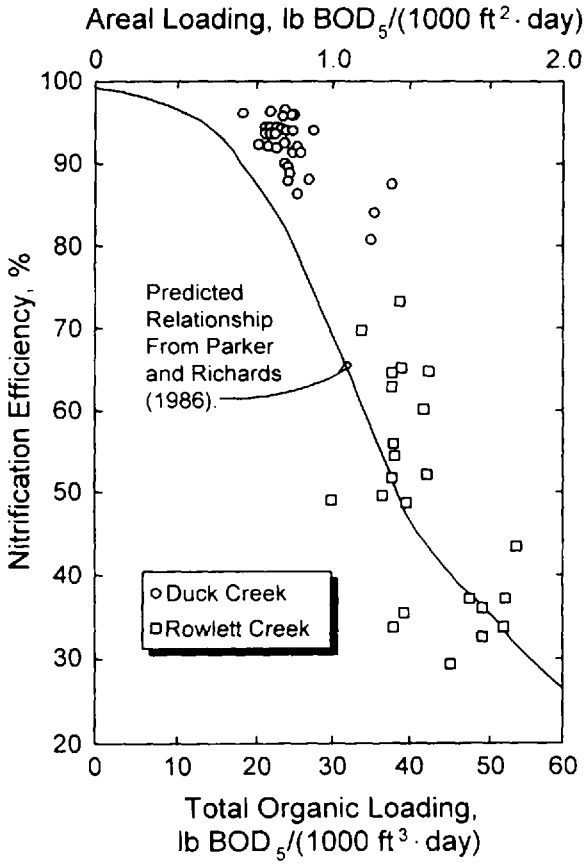


Figure 19.6 Effect of total organic loading on the nitrification efficiency in a trickling filter accomplishing combined carbon oxidation and nitrification. (From G. T. Daigger, L. E. Norton, R. S. Watson, D. Crawford, and R. B. Sieger, Process and kinetic analysis of nitrification in coupled trickling filter/activated sludge processes. *Water Environment Research* **65** 750–758, 1993. Copyright © Water Environment Federation; reprinted with permission.)

in oxygen units, it should be possible to correlate the rate at which oxygen demand is being satisfied with the trickling filter process loading. This has been done, and the rate per unit of media volume is called the oxidation rate, A_{OR}.[13] It is calculated using the following equation:

$$A_{OR} = \frac{F(S_{SO} + 4.57\ S_{NO})}{V_M} \tag{19.3}$$

where S_{SO} is the influent organic matter concentration expressed in BOD_5 units and S_{NO} is the nitrate-N concentration in the trickling filter effluent. At typical process loadings, the BOD_5 is generally equal to the oxygen required to oxidize the biodegradable organic matter applied, which is one reason that S_{SO} is expressed in those units. For trickling filters accomplishing combined carbon oxidation and nitrification, the oxidation rate has been observed to be relatively constant at a value between 0.5

and 1.0 kg O_2 demand/(m^3·day) until the ammonia-N concentration drops below about 3 to 5 mg/L, at which point it becomes rate limiting, causing the rate to drop.[27]

The nature of the biofilm that develops in separate stage nitrification applications is different from that which develops in carbon oxidation applications. This is because of the low yields associated with nitrifying bacteria. In fact, the biofilm that develops in these applications conforms much more closely to the theoretical assumptions upon which the models presented in Chapter 15 and Sections 16.1 and 16.2 are based. Consequently, for these applications media plugging and flow channeling are less of an issue and the SOL can be used effectively to characterize the process loading. Although a number of relationships are available in the literature,[61,62] the one developed by Parker and co-workers[3,43,44] conforms most closely to the theory of Chapter 15 and is presented in Section 19.3.2.

Hydraulic loading rates can also affect trickling filter performance. However, as discussed in Section 16.2, the theoretical performance of a trickling filter is relatively insensitive to the THL as long as the organic loading is fixed. Practical experience confirms this prediction, once a minimum THL is achieved.[61,62] These results have been interpreted as indicating that a minimum THL is necessary to achieve complete wetting and utilization of the trickling filter media. Once this minimum THL has been achieved, further performance improvements are not observed when the hydraulic loading is increased further since the media is fully wet and active. For trickling filters using high-rate media, the minimum THL is approximately 1.8 m/hr. The minimum THL for trickling filters using rock media is not as well defined. However, THL's used in practice typically range from 0.35 to 0.75 m/hr. As discussed below, for economic reasons, most trickling filters are operated with THLs near the minimum for the media used, and thus little information is available about maximum allowable THLs. For high-rate media, the maximum THL is most likely caused by excessive removal of the biofilm. Estimates of such limits can be obtained from the concept of Spülkraft, which is discussed in Section 19.2.7.

19.2.2 Recirculation

The term recirculation refers to the return of trickling filter effluent, either prior to or following secondary clarification, back to the trickling filter influent. The primary purpose of recirculation is to uncouple the hydraulic and organic loading rates. The trickling filter media volume is selected to give the desired organic loading rate, expressed as either the SOL or the TOL. A depth is then selected, fixing the cross-sectional area of the media. The THL that will be achieved is then calculated and compared to the required minimum. If the THL is less than the minimum, either recirculation flow must be provided to make the THL equal or exceed the minimum, or a greater depth must be selected, thereby decreasing the cross-sectional area. The effects of media depth are discussed in the next section. In most cases, increased recirculation will require increased pumping, thereby increasing operating costs. For this reason, recirculation in excess of that required to achieve the required minimum THL is not desirable. Moreover, as discussed Sections 16.2 and 19.2.1, increased recirculation beyond the minimum required to achieve good wetting generally does not result in improved treatment. Thus, recirculation beyond that required to attain the minimum THL is not generally used unless other factors require it.

Another purpose for recirculation applies when a high-strength wastewater is being treated. As discussed in Section 16.3, oxygen limitations may be expected in trickling filters whenever the concentration of biodegradable organic matter exceeds about 250 mg/L as COD. This prediction is generally confirmed by practice.[61,62] Treatment performance may not necessarily be adversely impacted, but odors can be produced by the development of anaerobic conditions within the biofilm. When the concentration of biodegradable organic matter in the influent wastewater exceeds 250 mg/L as COD, it can be diluted by recirculating treated effluent. The quantity of recirculation required to dilute the applied biodegradable COD to 250 mg/L is calculated and compared to the recirculation required to achieve the minimum THL. The greater of the two is then provided.

Several recirculation options have been developed and tested over the long history of the trickling filter process.[61,62] They vary primarily in the source of the recirculation flow, the location to which it is recirculated, and, if trickling filters in series are used, whether recirculation occurs around each individually or around both. Sources of recirculation flow include both clarified and unclarified trickling filter effluent. Recirculation destinations include the trickling filter influent and the influent to an upstream primary clarifier. Recirculation of final clarifier effluent to the influent of a primary clarifier adds treated, oxygenated flow to it, which can reduce odor emissions. However, the cost of such a practice is high since the sizes of both the primary and secondary clarifiers must be increased to accommodate the recirculated flow. Recirculation of clarified effluent directly to the trickling filter has a similar drawback since the secondary clarifier must be sized to accommodate the recirculation flow, but with no compensating benefits. Consequently, the recirculation option illustrated in Figure 19.1, direct recirculation of unclarified trickling filter effluent to the trickling filter influent, is the one most frequently used today. In some instances this approach also results in a simplified pumping system as influent wastewater and recycled effluent can be combined in a single pumping system for application to the trickling filter. This option also allows sloughed biomass to be recirculated to the trickling filter, although the resulting impact on treatment efficiency is small due to the short HRT of the trickling filter.

19.2.3 Media Depth

Trickling filters have been constructed with a wide range of media depths. Depths are generally around 2 m for rock media, with a typical range of 1 to 2.5 m. Depths up to 12 m have been used for high-rate media, but a maximum depth of about 6.7 m is often used because of media structural considerations. Like recirculation, differences of opinion have existed over the years concerning the impact of media depth on trickling filter performance.[61,62] For example, a strict interpretation of the Velz, Eckenfelder, and Kornegay models discussed in Sections 16.4.1, 16.4.2, and 16.4.3, respectively, indicate that, for a constant media volume, trickling filter performance improves as the media depth is increased. Models such as those have provided the basis for constructing trickling filters with relatively large media depths (over 10 m). However, these predictions have been demonstrated to be incorrect. In contrast, the model of Logan et al.,[30] discussed in Section 16.4.5, indicates that, for a constant media volume, trickling filter media depth has only a modest affect on trickling filter performance.

The available data are generally consistent with the predictions of the model of Logan et al.[30] and indicate that the impact of media depth on performance is relatively small for a constant TOL. Modest performance improvement is obtained by increasing media depths up to about 4 m, but performance improvements for further increases are negligible. However, maintenance of the minimum THL, as discussed in the previous two sections, is critical to maintaining consistent trickling filter performance over a wide range of media depths. This issue arises because of practical considerations. For a constant trickling filter media volume, as the media depth is decreased the media cross-sectional area (A_c) increases. Thus, to maintain a constant THL as the cross-sectional area is increased, the recirculation flow rate must be increased. The interactions between TOL, media depth, and THL are illustrated by the following example.

Example 19.2.3.1

A trickling filter utilizing high-rate media is to be sized with a TOL of 1 kg $BOD_5/(m^3 \cdot day)$ to treat a wastewater with a flow rate of 5,000 m^3/day and a BOD_5 concentration of 150 mg/L. The minimum acceptable THL for the media is 1.8 m/hr.

a. What media volume is required?
The required media volume can be calculated with a rearranged form of Eq. 19.1:

$$V_M = \frac{(5{,}000)(150)}{(1.0)(1000)} = 750 \text{ m}^3$$

b. What would the THL be for a media depth of 5 m if no recirculation is used?
At a media depth of 5 m, the trickling filter cross-sectional area is:

$$A_c = \frac{750}{5} = 150 \text{ m}^2$$

The THL can be calculated with Eq. 16.8 using a recirculation ratio of zero:

$$A_H = \frac{5{,}000(1 + 0)}{150} = 33.3 \text{ m/day} = 1.39 \text{ m/hr}$$

This is less than the minimum value of 1.8 m/hr. Consequently, the media provided will not be fully utilized and such a design would not be acceptable.

c. What recirculation rate would be required to attain the minimum THL for a media depth of 5 m?
The required THL is 1.8 m/hr = 43.2 m/day. The recirculation ratio can be calculated with a rearranged form of Eq. 16.8:

$$\alpha = \frac{(43.2)(150)}{5{,}000} - 1 = 0.30$$

Thus, the required recirculation rate is (0.3)(5,000) = 1,500 m^3/day.

d. What media depth would be required to achieve a THL of 1.8 m/hr (= 43.2 m/day) without recirculation?

The required cross-sectional area can be calculated with a rearranged form of Eq. 16.8:

$$A = \frac{5{,}000(1 + 0)}{43.2} = 116 \text{ m}^2$$

To provide a total media volume of 750 m^3, the depth must be:

$$L = \frac{750}{116} = 6.5 \text{ m}$$

The choice between the two alternative designs would have to be made on the basis of economics and other such factors since they should both give similar performance.

This example illustrates the relationship between THL and media depth for a fixed media volume. Although it is recognized today that a minimum THL must be maintained to fully utilize the media and that recirculation can be used to achieve that THL, analysis of historical studies suggests that this requirement was not widely recognized in the past.[30] As a consequence, data were collected from trickling filters with various media depths and with various THLs that were below the minimum value.[7,8,25,61,62] As expected, these data indicated that trickling filters with greater media depths, and consequently greater THLs, performed significantly better than trickling filters with smaller depths. Today we recognize that this effect was largely attributable to the variation in THL and that media depth itself exerted only a minimal influence.

19.2.4 Temperature

Temperature is another factor whose impact on trickling filter performance has historically been poorly understood. Ample full-scale evidence exists demonstrating that the performance of a trickling filter can decline significantly during periods of cold weather.[5] Based on such observations, it was concluded long ago that the trickling filter process is relatively temperature sensitive. With an improved understanding of biofilm processes, however, it was realized that, in many instances, substrate removal is controlled more by mass transfer than by biological reaction.[38] Moreover, since temperature effects on mass transfer are often modest, the effect of temperature on trickling filter performance should also be relatively modest. Further analysis suggests that the observed significant effect of cold weather operation on trickling filter performance is often a result of severe temperature drops due to the physical configuration of the system. When such drops occur, the biological reaction is severely retarded, and performance does indeed suffer. Therefore, the key to good performance is to limit temperature changes to the range over which temperature has little effect and improved design concepts can be used to do that. Of particular importance are physical design concepts to minimize heat loss during cold weather operation.

Heat loss occurs from trickling filters during cold weather by a variety of mechanisms. Conductive heat losses through the walls of the trickling filter are generally minor because the residence time of the fluid within the trickling filter is so short. Rather, the two most important heat loss mechanisms are wind effects and increased ventilation during cold weather. Experience indicates that significant heat

loss can occur as the influent wastewater flows out of the distribution system, through the air, and onto the media. Several methods are available to reduce heat loss by this mechanism. They include:

- Construct deep trickling filters with small cross-sectional areas. The reduced cross-sectional area increases the THL and reduces the need for recirculation, which increases the number of exposures of the wastewater to the cooling influence of flowing through the distributor.
- Extend the side wall upward so that it is 1.5 to 2 m above the distributor. This reduces the cooling effect of wind on the flow being applied to the trickling filter.
- Covering the trickling filter.

Several full-scale installations exist in which retrofitting wind screens to the top of the trickling filters or adding covers has significantly improved cold weather performance.

Ventilation control is also quite important during cold weather to minimize heat loss.[61,62] As discussed above, temperature differences between the wastewater and the ambient air cause the density of the air inside the trickling filter to differ from the density of the air outside, inducing air to flow through the trickling filter. Although wastewater temperatures vary seasonally, their variation is typically much less than the variation in ambient air temperatures, resulting in much greater temperature differentials in the winter. For example, wastewater temperatures during the summer may reach 25°C to 30°C, which are comparable to the air temperature. In contrast, in colder climates influent wastewater temperatures seldom drop below 10°C in winter while ambient air temperatures can reach −10°C to −20°C. Consequently, there is an increased density difference between the air inside and outside the trickling filter in winter, causing increased air flow through the trickling filter. Consequently, the cooling effect increases as the ambient air temperature decreases.

For trickling filters with natural draft ventilation, the effect of the increased temperature difference during winter operation can be controlled by providing adjustable dampers on the air inlets. During cold weather operation the dampers are throttled to restrict air flow through the trickling filter, thereby reducing cooling affects. During warm weather the dampers are opened to increase the air flow. Covering the trickling filter provides an even greater opportunity to control air flow, as long as adjustable dampers are included in the design. Yet another approach is forced draft ventilation, which makes the air flow independent of the ambient air temperature.

19.2.5 Ventilation

The resistance to air flow through a properly designed trickling filter is quite low, and thus only a small motive force is required to induce it. Natural draft ventilation operates very effectively as long as the temperature and humidity differences between the air inside and outside the trickling filter are sufficiently large to generate the needed force. Unfortunately, instances occur in which the densities of air inside and outside the trickling filter are the same.[5,50,53] Neglect for a minute the effect of humidity differences on air density and consider only temperature differences. Wastewater temperatures are relatively constant throughout a typical day, but ambient air

temperatures vary significantly. Consequently, during many days it is quite likely that the ambient air temperature will equal the wastewater temperature at least twice, as illustrated in Figure 19.7. This is particularly true during periods of moderate to warm air temperature. When this occurs, no density difference exists between the air inside and outside the trickling filter and no motive force exists to move air. The fact that both temperature and humidity affect air density complicates the phenomenon somewhat, but does not change its basic nature. As a result, during certain periods little or no air will flow through tricking filters using natural draft ventilation, negatively impacting aerobic treatment. This situation can be tolerated in many instances, particularly when incomplete treatment can be accepted on a temporary basis and when odor production is not a significant problem. In other instances, more positive control of air flow is required. This requires forced draft ventilation.

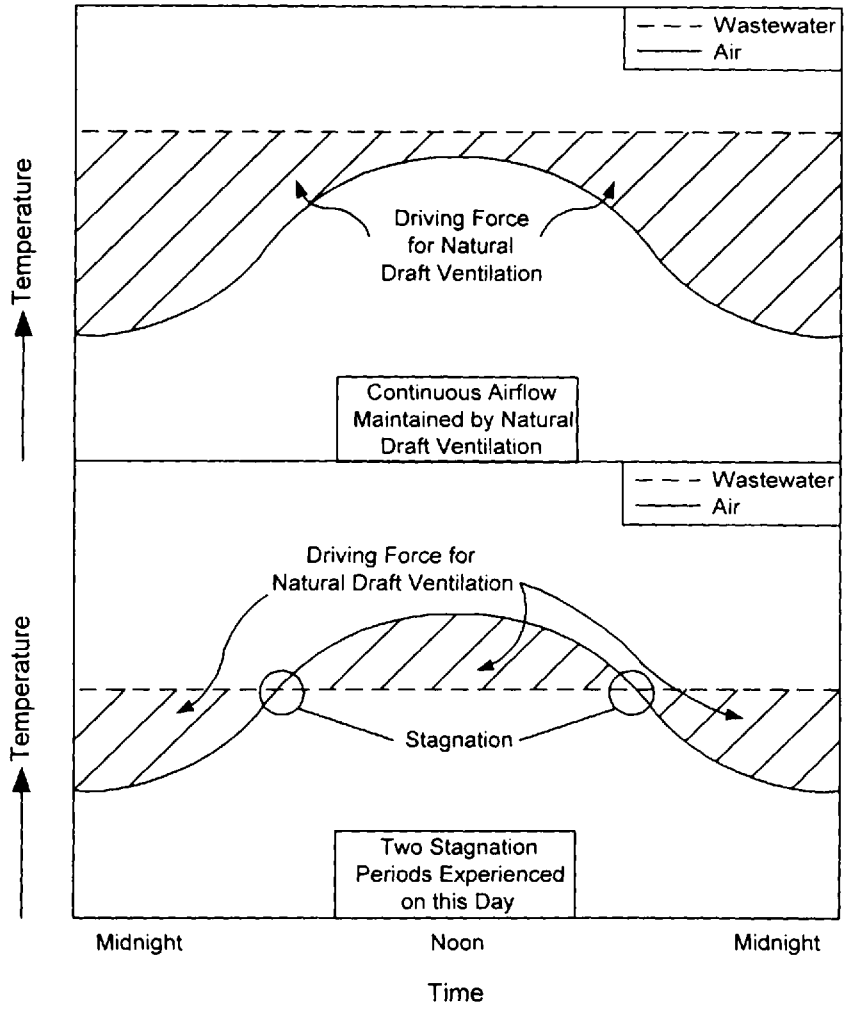


Figure 19.7 Effect of relative temperatures of air and wastewater on natural draft ventilation.

Natural draft ventilation requires the provision of sufficient open area to allow air to flow into the tricking filter, through the media, and out. For trickling filters without covers, little restricts the flow of air at the top. In contrast, air will not enter or exit the bottom of the trickling filter unless openings with sufficient area are provided to allow air to enter the underdrain system. The clear opening between the top of the water surface and the bottom of the trickling filter media must also be sufficient. Consequently, the design of a natural draft ventilation system involves providing sufficient vent area and free distance between the water surface in the underdrain and the bottom of the media. Because the draft is generated uniformly across the entire cross-sectional area of the trickling filter by the difference in air density, uniform distribution of the air occurs naturally.

Forced draft ventilation systems overcome the difficulties associated with natural draft systems by providing positive movement of air through the trickling filter with a fan. In contrast to natural draft ventilation systems, forced draft systems require a means to distribute the air uniformly across the cross sectional area of the trickling filter. This is usually accomplished by connecting a piping network with multiple openings to the ventilation fan, as described in Section 19.3.5.

Natural draft ventilation systems have been used with most full-scale trickling filters. Consequently, our performance data base is for such systems. It is likely, however, that forced draft ventilation systems will improve the performance of and reduce the production of odors from trickling filters. While little full-scale experience is currently available to confirm this hypothesis, it should become available in the future.

19.2.6 Media Type

The types of trickling filter media in common use today are discussed in Section 19.1.2, compared in Table 19.2, and illustrated in Figure 19.3. There are several opinions about the relative merits of these media.[11,39,46,48] Harrison and Daigger[11] conducted a comprehensive investigation of several of the media operated over a wide range of organic loading rates. Their results indicated that all of the tested media gave significant removal of organic matter, but that some did better, both statistically and practically. Interestingly, some media provided superior performance at low TOLs, whereas others provided superior performance at high TOLs. Good performance was provided by rock media at low TOLs but not at high TOLs. At low TOLs, the best performance was provided by 60° crossflow (XF) media. In contrast, fully corrugated vertical media (VFC) provided the best performance at high TOLs. Performance of both the random (RA) and horizontal (HO) media was inferior to that of the bundle media at all TOLs. The superior performance of XF media at low TOLs has been confirmed by several other researchers, but the superior performance of VFC media at high organic loading rates has been contested by others. While results such as these are significant, it must be recognized that nearly all media testing has been conducted at pilot scale. It has been suggested that pilot scale results may not accurately reflect full-scale performance due to the difficulty of simulating full-scale wastewater distribution systems in pilot-scale facilities.[61,62]

Operational differences have also been noted among the various trickling filter media. Some of the more significant ones are as follows:[4,11,61,62]

- A relatively thick and biologically diverse biofilm typically develops on rock media. It can be several mm thick, and it consists of a rich array of bacteria, Eucarya such as protozoa and rotifers, and macroinvertibrates such as worms and fly larvae. These thick biofilms result in a high degree of anaerobic activity. It is thought by some that this causes lower net biomass production rates from rock media trickling filters. The low void volume and irregular pathways through rock media may make it more susceptible to plugging than high-rate media.
- 60° crossflow media possesses good flow redistribution characteristics, particularly in comparison to vertical media. This may be one of the reasons that superior performance is observed for crossflow media at low TOLs. It may also result in increased plugging potential, particularly at high TOLs.
- Random media possesses good flow redistribution characteristics. However, localized dry spots and areas of ponding have been observed, suggesting areas of incomplete utilization. Anaerobic activity may develop within the ponded areas, contributing to odors. The dry areas also provide locations for the growth of nuisance organisms, particularly Psychoda flies.
- Relatively thin biofilms develop on horizontal wood media when applied in conventional trickling filter applications. When RAS is applied in an ABF or BF/AS application, however, a significant amount of interstitial growth develops, which also contributes to organic matter removal.

These differences in operational characteristics should be considered when selecting a media for a particular application.

19.2.7 Distributor Configuration

Several wastewater distribution systems have been used, including fixed nozzles with or without periodic dosing and rotary distributors with or without speed control. Experience indicates that the distributor type significantly affects the hydraulic flow pattern and the biofilm thickness within the trickling filter. Both of these factors significantly affect trickling filter performance.

Figure 19.8 illustrates the two types of distributors typically used in trickling filters, rotary and fixed nozzle. Generally, rotary distributors give better performance than fixed nozzle distributors.[61,62] This occurs for a variety of reasons. One is more uniform flow distribution. Flow is applied quite uniformly to the section of a trickling filter over which a rotary distributor is passing. Since the distributor passes repetitively over the trickling filter, on an average basis wastewater is applied uniformly over its surface. Theoretically, uniform distribution of wastewater can also be obtained with a fixed nozzle distributor. However, experience indicates that uniform distribution is very difficult, if not impossible, to achieve. Fixed nozzles must emit a fine spray to provide uniform application over the trickling filter. This generally requires relatively small openings. However, since most wastewaters contain some particulate matter, small openings are prone to plugging, which disrupts flow distribution. Consequently, fixed nozzle designs generally represent a compromise between flow distribution and plugging potential. The poor operating characteristics of fixed nozzle distributors can be mitigated to a certain extent by using higher recirculation flows to increase the applied THL. Hydraulic loading rates used with fixed

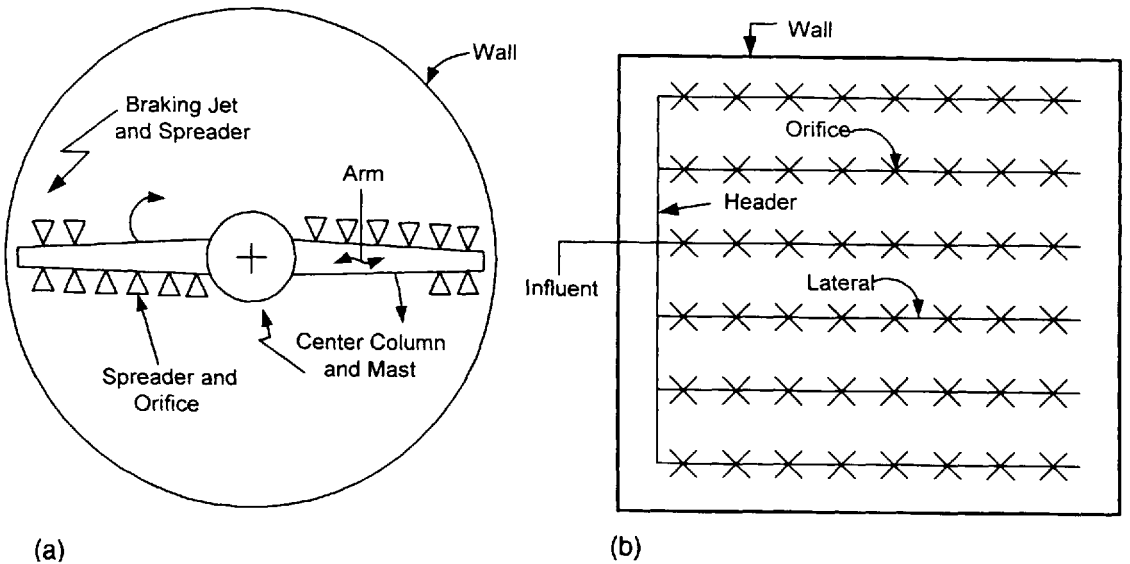


Figure 19.8 Schematic diagrams of (a) rotary and (b) fixed nozzle distributors.

nozzle distributors are often two to three times higher than those used with rotary distributors. The minimum THL for high-rate media of 1.8 m/hr discussed above is that required for rotary distributors. For fixed nozzle distributors, the minimum THL is on the order of 4 to 6 m/hr. Even with these provisions, significant plugging potential still exists for many fixed nozzle distributor designs.

Another advantage of rotary distributors is that the instantaneous hydraulic loading rate that occurs as the distributor passes over a section of the trickling filter is significantly higher than the average THL. This high instantaneous loading rate can produce a flushing effect that removes excess biomass from the trickling filter, thereby maintaining a thinner, more active biofilm. This does not occur when fixed nozzle distributors are used.

The advantages of periodic dosing achieved with rotary distributors in circular trickling filters can also be attained with reciprocating distributors in square or rectangular trickling filters. While rotary distributors have proven to be quite reliable from a mechanical perspective, much less experience exists with reciprocating distributors. As a consequence, most trickling filters are circular units with rotary distributors.

A concept being tested and applied in full-scale applications is Spülkraft, SK, which is the depth of water in mm applied to the trickling filter in one passage of the distributor arm.[45,61,62] For a rotary distributor, the Spülkraft is calculated as follows:

$$SK = \frac{A_H(1{,}000)}{N_a \cdot \omega_d(60)} \tag{19.4}$$

where A_H is the THL in m/hr, N_a is the number of arms on the distributor, and ω_d is the distributor rotational speed in rev/min. It has been suggested that a low SK be used for normal operation and a high value be used for periodic flushing to remove excess biomass and maintain a thin, active biofilm. Review of historical information[23,24,25,32,54] supports this concept. Early results from the application of the concept are quite promising, but greater full-scale experience is required to test and refine it.[1,2,61,62] Table 19.5 presents SK values that are suggested based on current experience and judgment.[61,62]

Most rotary distributors use hydraulic propulsion. The energy of the influent flow is discharged in a horizontal direction, causing the arm to rotate in the opposite direction. The rotational speed can be controlled by discharging some of the influent

Table 19.5 Suggested Spülkraft Values[61,62]

TOL	Spülkraft, SK, mm/pass	
kg $BOD_5/(m^3 \cdot day)$	Design	Flushing
0.25	10–100	≥200
0.50	15–150	≥200
1.00	30–200	≥300
2.00	40–250	≥400
3.00	60–300	≥600
4.00	80–400	≥800

from the opposite side of the distributor arm, but the range that can be achieved is limited. A relatively new development is the power driven distributor, which uses an electric motor to control rotational speed.[61,62] Significantly greater control of rotational speed, and consequently greater control of Spülkraft, is provided by this system.

19.2.8 Wastewater Characteristics

As with all biochemical operations, the characteristics of the wastewater being treated affect the performance of a trickling filter. The more easily biodegradable the wastewater, the higher the organic loading that can be applied while still achieving acceptable effluent quality. There are limits of course, because ultimately, the rate of oxygen transfer will control the rate at which organic matter can be removed and oxidized.

Just as in the activated sludge process, various wastewater components are removed by a variety of mechanisms. Readily biodegradable substrate is removed by diffusion through the biofilm to microorganisms which biodegrade it. Slowly biodegradable substrates are initially removed by flocculation and entrapment mechanisms, just like in the activated sludge process. They are then hydrolyzed by extracellular enzymes before they are biodegraded. The presence of particulate material will adversely impact the capacity of a trickling filter to remove soluble organic matter because when particulate matter is incorporated into the biofilm, it displaces active biomass from the aerobic active biofilm layer.[49] This is also analogous to the activated sludge process where particulate organic matter accumulates in the mixed liquor suspended solids (MLSS).

In contrast to the activated sludge process, the COD/TKN ratio of the wastewater significantly affects tricking filter nitrification efficiency.[61,62] This occurs because of the spatial distribution of microorganisms within the trickling filter, as discussed in Section 9.1.1 and illustrated in Figure 19.2.

19.2.9 Effluent Total Suspended Solids

The overall performance of a trickling filter system depends not only on the removal of soluble and particulate organic matter, but also on the concentration of suspended solids in the system effluent, and this has been the factor that has historically limited system performance. Relatively complete removal of soluble organic matter can be obtained at appropriate loadings, but the concentration of suspended solid may not be low enough to meet performance objectives. In many instances, trickling filter effluents contain finely divided, colloidal suspended solids that settle poorly in secondary clarifiers. This characteristic led to development of coupled trickling filter/activated sludge processes. Coupling a suspended growth bioreactor with a trickling filter introduced a suspended biomass capable of flocculating and removing the colloidal suspended solids contained in the trickling filter effluent.

The concentration of suspended solids in the effluent from a trickling filter system is dependent on the settleability of the suspended solids. Many factors affect that settleability, including the TOL, the characteristics of the wastewater (particularly the proportion of suspended and/or colloidal matter), and the THL. Biological flocculation of suspended solids in the trickling filter effluent generally improves as

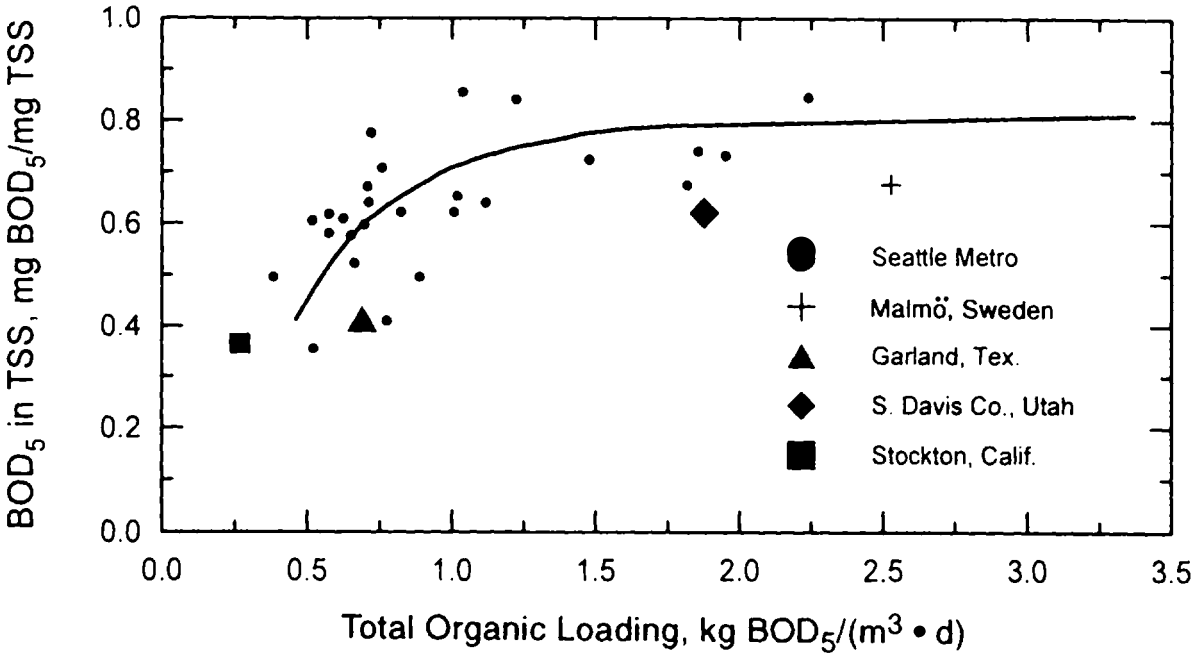


Figure 19.9 Effect of total organic loading (TOL) on the BOD_5 of suspended solids leaving a trickling filter. (From Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992. Copyright © Water Environment Federation; reprinted with permission.)

the TOL is reduced. Poorer flocculation is generally observed from roughing filter applications than for carbon oxidation applications. The best flocculation is obtained with combined carbon oxidation and nitrification applications. This is consistent with experience with the activated sludge process wherein reduced process loading, i.e., higher SRT, results in increased flocculation, as discussed in Section 10.2.1. It has recently been suggested that improved flocculation can be obtained by using hydraulic loadings which maintain a thin, active biofilm. A thin biofilm can be maintained by proper control of the SK factor, as discussed in Section 19.2.7. While more full-scale experience is required to fully test this hypothesis, initial results are promising.

The biodegradable organic matter in the effluent from a trickling filter system is made up of both soluble and particulate material. The concentration of particulate biodegradable organic matter, in turn, is a function of the suspended solids concentration and its biodegradability. The biodegradability of trickling filter effluent suspended solids, as reflected by the BOD_5, is a function of the TOL, as illustrated in Figure 19.9.[61,62] Consequently, the higher the TOL, the lower the suspended solids concentration must be in the system effluent to meet treatment criteria.

Example 19.2.9.1

A trickling filter receiving a wastewater with a total BOD_5 of 150 mg/L achieves 90% removal of soluble BOD_5 by being operated at a TOL of 0.75 kg BOD_5/($m^3 \cdot$ day). What will be the BOD_5 of the clarified effluent if it contains 30 mg/L of suspended solids? Assume that the soluble BOD_5 concentration of the influent wastewater is 65% of the total BOD_5.

a. First calculate the soluble BOD_5 in the effluent.
The soluble BOD_5 in the influent wastewater is (0.65)(150) = 97.5 mg/L. Since 90% of it is removed, the soluble BOD_5 concentration in the trickling filter effluent is 9.75 mg/L.

b. Next estimate the BOD_5 due to particulate matter in the effluent.
Entering Figure 19.9 with a TOL of 0.75 kg BOD_5/($m^3 \cdot$ day), it can be seen that the BOD_5/TSS ratio is 0.5. Since the suspended solids concentration in the effluent is 30 mg/L, the particulate BOD_5 is (0.5)(30) = 15 mg/L.

c. The total BOD_5 is just the sum of the soluble and particulate BOD_5 values.
Total BOD_5 = 9.75 + 15 = 24.75 mg/L.

19.3 PROCESS DESIGN

Although considerable progress has been made in our understanding of substrate transport and microbial growth in biofilms, as discussed in Chapter 15, the application of that understanding to trickling filter design has been slow. This is primarily because trickling filters are considerably more complicated than the conceptual models used to describe biofilms. For example, we saw earlier in this chapter that flow *through* a trickling filter is both intermittent and highly irregular, whereas biofilm models commonly assume steady flow at a constant rate. Thus, while mechanistic

models such as those in Section 16.1 have helped us develop a better understanding of trickling filter performance, they have not found application in practice. Rather, the design of trickling filters is based primarily on empirical correlations and simplified models depicting pilot- and full-scale system performance. Furthermore, because BOD_5 has been used historically to measure the concentration of biodegradable organic matter in those systems, many design relationships are typically presented in terms of it without regard to whether the organic matter is soluble or particulate. Finally, many relationships express the performance of the trickling filter system, i.e., the bioreactor plus the settler that follows it. Although this approach has had a negative impact on our ability to isolate and understand the fundamental processes involved, it has worked satisfactorily when used by experienced designers. Thus, it represents the current state-of-the-art and is the approach that we use here. It is likely, however, that as more basic data are collected from large-scale systems that better approaches will evolve.

This section reviews several procedures for sizing trickling filters for a variety of applications, criteria for sizing the ventilation system, and approaches for sizing coupled TF/AS processes. When sizing a trickling filter, consideration must be given to the nature of the media being used. Trickling filters containing rock and random pack high-rate media can be built to any depth consistent with the structural constraints of the media, as discussed earlier in this chapter. Bundle high-rate media, on the other hand, is modular, and economics dictate that the depth conform to an integer number of modules. One module of such media typically is 0.61 m high, so the depth should be in increments of 0.61 m. Furthermore, as discussed in Section 19.2.3, the maximum unsupported depth to which bundle media can be stacked is about 6.7 m, or eleven modules high. These constraints must be considered when selecting the depth of a trickling filter.

19.3.1 Sizing Trickling Filters with "Black-Box" Correlations

Historically, "black-box" correlations, i.e., those simply correlating output with input, of performance data from full-scale trickling filter applications have been used to size rock trickling filters. As discussed in Section 16.4, two such correlations are those developed by the National Research Council (NRC)[37] and by Galler and Gotaas.[18] As also discussed in that section, great care should be exercised when black-box correlations are used. The data set upon which they are based should be reviewed carefully by the design engineer to ensure that it is similar to the proposed application. If significant differences exist, then the equation should not be used.

The NRC correlation was based on performance data from 34 rock media trickling filters operating at military installations.[37] It reflects the impact of TOL and recirculation on performance. The influent BOD_5 concentrations at the plants were relatively high (generally 200 mg/L or greater). Consequently, the correlation may not accurately reflect the performance of rock media trickling filters treating lower strength wastewater. The Galler–Gotaas[18] correlation is based on 322 observations at typical municipal wastewater treatment plants. It recognizes the importance of organic loading, hydraulic loading, recirculation, media depth, and temperature on system performance. Both the NRC and the Galler–Gotaas correlation equations were developed on the basis of the BOD_5 concentration in the process influent and

the secondary clarifier effluent. Consequently, they implicitly incorporate the impact of both the trickling filter and the secondary clarifier. When using either, it must be assumed that the secondary clarifier is properly sized and operated. Because both black-box correlations are used for sizing rock media trickling filters and few are being designed today, neither is presented here. However, they are presented in standard design manuals.[61,62]

19.3.2 Sizing Trickling Filters with Loading Factor Relationships

The process loading factor approach to trickling filter design reflects the fact that trickling filter performance is generally correlated with the TOL or the SOL, as discussed in Section 19.2.1. The steps required when using this approach include: selection of an appropriate TOL or SOL based on experience and/or on performance correlations such as those shown in Figures 19.5 and 19.6; use of the selected relationship to calculate the required media volume; and use of typical media depths and hydraulic loadings, as described in Sections 19.2.1, 19.2.2, and 19.2.3, to dimension the trickling filter. The following examples illustrate the use of the loading factor relationships to design a roughing filter, a carbon oxidation application, and a trickling filter achieving combined carbon oxidation and nitrification. The first example considers a roughing filter.

Example 19.3.2.1

Size a trickling filter for a roughing filter application in which 75% of the soluble BOD_5 is to be removed. The wastewater flow rate is 5,000 m^3/day, and the total BOD_5 concentration is 150 mg/L. Bundle media is to be used with a minimum THL of 1.8 m/hr. Assume that the relationship between soluble BOD_5 removal and TOL shown in Figure 19.5 is applicable.

a. What TOL should be used to achieve the treatment objective?
From Figure 19.5, a TOL of 2.5 kg $BOD_5/(m^3 \cdot day)$ will achieve the treatment objective.

b. What media volume is required?
The media volume can be calculated with a rearranged form of Eq. 19.1:

$$V_M = \frac{(5{,}000)(150)}{(2.5)(1{,}000)} = 300 \text{ m}^3$$

c. What media depth is required to maintain the minimum allowable THL of 1.8 m/hr with no recirculation?
The media depth is just the volume divided by the cross-sectional area. The area can be obtained from the definition of the THL, given by Eq. 16.8. For $\alpha = 0$ and $A_H = 1.8$ m/hr (= 43.2 m/day):

$$A_c = \frac{5{,}000}{43.2} = 115.7 \text{ m}^2$$

$$L = \frac{300}{115.7} = 2.6 \text{ m}$$

While this is acceptable, as discussed in Section 19.2.3, performance is generally reduced for media depths less than about 4 m. Thus, a greater depth should be used, which is acceptable since that will give a smaller cross-sectional area, thereby giving a greater THL. The depth of a typical sheet media bundle is 0.61 m. If we take 4 m as being the shortest acceptable depth and recognize that an integer number of bundles must be used, then the media depth should be at least 4.27 m, corresponding to 7 layers of media.

d. What THL would result if the trickling filter were 4.27 m high?

$$A_c = \frac{300}{4.27} = 70.4 \text{ m}^2$$

$$\Lambda_H = \frac{5{,}000}{70.4} = 71 \text{ m/day} = 2.96 \text{ m/hr}$$

This is an acceptable THL.

This example illustrates that, for roughing filter applications, the media depth may be established at minimum design values. It also illustrates that relatively high THL rates are sometimes used.

The following example uses the process loading factor approach to size a trickling filter for relatively complete removal of biodegradable organic matter, such as for secondary treatment.

Example 19.3.2.2

Size a trickling filter for 90% removal of soluble BOD_5. As in Example 19.3.2.1, the wastewater flow rate is 5,000 m^3/day, the total BOD_5 concentration is 150 mg/L, and bundle media with a minimum THL of 1.8 m/hr is to be used. Assume that the relationship between soluble BOD_5 removal and TOL shown in Figure 19.5 is applicable.

a. What media volume is required?
From Figure 19.5 it can be seen that a TOL of 0.75 kg $BOD_5/(m^3 \cdot day)$ must be used to achieve 90% removal of soluble BOD_5. The media volume can be calculated with a rearranged form of Eq. 19.1:

$$V_M = \frac{(5{,}000)(150)}{(0.75)(1{,}000)} = 1{,}000 \text{ m}^3$$

b. If a media depth of 6.7 m is used, which is the maximum unsupported depth of plastic sheet bundle media and corresponds to 11 layers of bundles, what recirculation flow rate is required to maintain a THL of 1.8 m/hr (43.2 m/day)?
The recirculation ratio, α, can be calculated with Eq. 16.8 after calculating the cross-sectional area of the trickling filter.

$$A_c = \frac{1{,}000}{6.7} = 149.2 \text{ m}^2$$

$$\alpha = \frac{(43.2)(149.2)}{5{,}000} - 1 = 0.29$$

Thus, the recirculation flow rate is (0.29)(5,000) = 1,445 m^3/day, making the total flow rate to the trickling filter 6,445 m^3/day.

This example illustrates that even though the trickling filter media depth is established at the maximum safe media depth from a structural perspective, a small amount of recirculation may still be required to maintain the minimum required THL. This often occurs when relatively high removal of biodegradable organic matter is needed.

The next example uses the surface loading approach to size a trickling filter for combined carbon oxidation and nitrification.

Example 19.3.2.3

Size a trickling filter to accomplish combined carbon oxidation and nitrification of the wastewater used in the preceding two examples. The treatment goal is to remove more than 85% of the influent ammonia-N. Bundle media with a specific surface area of 100 m^2/m^3, a minimum THL of 1.8 m/hr, and a maximum depth of 6.7 m is to be used.

a. What would an appropriate TOL be?
Figure 19.6 shows that an SOL of 0.002 kg $BOD_5/(m^2 \cdot day)$ is required to achieve a nitrification efficiency of greater than 85%. Equation19.2 can be used to convert this to a TOL for media with a specific surface area of 100 m^2/m^3:

$$A_S = (100)(0.002) = 0.2 \text{ kg } BOD_5/(m^3 \cdot day)$$

b. Calculate the trickling filter media volume.

$$V_M = \frac{(5{,}000)(150)}{(0.2)(1{,}000)} = 3{,}750 \text{ m}^3$$

The media volume could also have been calculated directly from the SOL and the media specific area. At a BOD_5 loading of 750 kg BOD_5/day and an SOL of 0.002 kg $BOD_5/(m^2 \cdot day)$, a total media surface area of 375,000 m^2 would be needed. Since the specific surface area of the media is 100 m^2/m^3, the total media volume is 3,750 m^3.

c. What recirculation flow rate is required to maintain a THL of 1.8 m/hr (43.2 m/day) at a media depth of 6.7 m?
For this application, the cross-sectional area is:

$$A_c = \frac{3{,}750}{6.7} = 560 \text{ m}^2$$

This can be used in Eq. 16.8 to calculate the recirculation ratio, α:

$$\alpha = \frac{(43.2)(560)}{5{,}000} - 1 = 3.84$$

Thus, the recirculation flow rate must be (3.84)(5,000) = 19,190 m^3/day.

This example illustrates that a large media volume may be needed when a high level of treatment is required. This can result in a high recirculation requirement.

Table E19.1 compares the trickling filter sizes required for the applications considered in Examples 19.3.2.1 through 19.3.2.3. The roughing application (Example 19.3.2.1) achieves partial removal of biodegradable organic matter and requires a relatively small trickling filter of moderate depth, operating with a relatively high THL (3.0 m/hr) without recirculation. Removal of biodegradable organic matter to the level of secondary treatment (Example 19.3.2.2) requires a larger trickling filter with a greater depth, operating at the minimum THL (1.8 m/hr) with a modest amount of recirculation. Finally, combined carbon oxidation and nitrification (Example 19.3.2.3) requires an even larger trickling filter, although at the same depth, operating at the minimum THL (1.8 m/hr) with a large amount of recirculation. Capital costs increase as the degree of treatment is increased because larger trickling filters must be constructed. Operating costs also increase because the influent must be pumped to a higher elevation and because more recirculation flow must be pumped.

Next consider the design of a trickling filter for separate stage nitrification. As discussed in Section 19.2.1, separate stage nitrification conforms more closely than other trickling filter applications to the theoretical models of Chapter 15 and Sections 16.1 and 16.2. Consequently, it is best characterized by the surface loading approach. The design methodology proposed by Parker and co-workers[3,43,44] provides the basis for the approach discussed here.

In the upper regions of a trickling filter accomplishing separate stage nitrification the rate of ammonia-N oxidation will be controlled by the rate of oxygen transfer into the biofilm. This occurs because the ammonia-N concentration in the liquid phase is relatively high whereas the dissolved oxygen concentration is limited to the solubility of oxygen in water (in the range of 7 to 12 mg/L for typical wastewater temperatures). Because approximately 4.3 mg of O_2 are required for the oxidation of one mg of ammonia-N, and because the half-saturation coefficient for oxygen is relatively high, oxygen will be the limiting reactant for ammonia-N concentrations above about 3 to 5 mg-N/L. However, in the lower regions of the trickling filter the rate of ammonia-N oxidation will become limited by the ammonia-N concentration. This situation can be characterized by the following expression:

$$J_{NH} = J_{NH,max}\left(\frac{S_{NH}}{K_{s,NH} + S_{NH}}\right) \tag{19.5}$$

Table E19.1 Comparison of Results for Examples 19.3.2.1 – 19.3.2.3

Application	TOL, kg BOD_5/m^3·day	Depth, m	Surface area, m^2	Flow rate, m^3/day: Influent	Flow rate, m^3/day: Recir.
Roughing	2.50	4.27	70.4[a]	5,000	0
Secondary treatment	0.75	6.70	149.2[b]	5,000	1,445
Carbon oxidation and nitrification	0.20	6.70	560.0[b]	5,000	19,190

[a] THL = 3.0 m/hr.
[b] THL = 1.8 m/hr.

where J_{NH} is the ammonia-N flux into the biofilm (g NH_3-N/(m^2 · day)), $J_{NH,max}$ is the maximum flux when ammonia-N is not limiting, S_{NH} is the ammonia-N concentration in the bulk liquid, and $K_{g,NH}$ is a pseudo half-saturation coefficient similar to that employed by Kornegay[29] to account for both mass transport and the degree of saturation on the reaction rate in a trickling filter (See Section 16.4.3). The value of the maximum flux is determined by the rate of oxygen transfer into the biofilm.

Several methods are available for obtaining the parameters in Eq. 19.5. One is through the analysis of operating data from similar, full-scale applications. Another is to operate a pilot plant and to analyze the resulting performance data. Parker and co-workers[3,43,44] have proposed the use of the trickling filter oxygen transfer model of Logan et al.,[31] along with corrections for temperature, the suspended solids concentration in the trickling filter influent, and the THL. With this approach $K_{g,NH}$ must still be selected independently.

Use of the surface loading approach is simply an extension of the approach presented above, but with consideration of the effects of the bulk liquid ammonia-N concentration on the ammonia-N flux into the biofilm. Because the flux equation is approximate, it is typically applied by dividing the trickling filter into a series of incremental elements within which the ammonia-N concentration, and hence the flux, can be considered to be constant. First, an ammonia-N concentration is selected above which the flux can be considered to be constant. The media volume required to reduce the influent concentration to that value is then determined by assuming that the flux is $J_{NH,max}$ and using the approach illustrated in Example 19.3.2.3. The remainder of the trickling filter is then broken into elements within which the concentration change is small, allowing the volume of each increment to be calculated for the flux corresponding to the average concentration within the increment. This is continued down the trickling filter until the desired effluent concentration is attained. The total media volume required is just the sum of the incremental media volumes. The media depth and THL are then specified and the trickling filter dimensions are selected using the same approaches used in Examples 19.3.2.1 through 19.3.2.3. As discussed in Section 19.1.2, higher density media (specific surface area typically 140 m^2/m^3) are usually used for separate stage nitrification applications to increase the volumetric treatment efficiency. This can be done because the risk of plugging and channeling is less with this application.

Example 19.3.2.4

A separate stage nitrification trickling filter is to be sized to treat a high-quality secondary effluent with a flow rate of 5,000 m^3/day and an ammonia-N concentration of 20 mg-N/L. The required effluent ammonia-N concentration is 1 mg-N/L. Pilot testing resulted in the selection of a value for $J_{NH,max}$ of 1.5 g NH_3-N/(m^2 · day) and a value for $K_{g,NH}$ of 1.5 mg-N/L for design purposes. Since most of the biodegradable organic matter has been removed, thereby reducing the chances of plugging, media with a specific surface area of 140 m^2/m^3 will be used to reduce the size of the unit. In addition, a minimum THL of 1.8 m/hr will be maintained.

a. Assuming that the flux will be constant at $J_{NH,max}$ for ammonia-N concentrations of 5 mg-N/L or higher, what media volume is needed to reduce the concentration from 20 mg-N/L to 5 mg-N/L?

The mass removal rate of ammonia-N in the first increment is just the flow rate times the required concentration change, or (5,000)(20 − 5) = 75,000 g NH_3-N/day. Since the maximum flux is 1.5 g NH_3-N/(m^2 · day), the required media surface area is 75,000 ÷ 1.5 = 50,000 m^2. If the specific surface area of the media is 140 m^2/m^3, the required media volume is 50,000 ÷ 140 = 357 m^3.

b. What additional media volume is required to reduce the ammonia-N concentration to 1 mg-N/L, the desired effluent value?
For this analysis, divide the remainder of the trickling filter into zones within which the bulk liquid ammonia-N concentration is reduced by 0.5 mg-N/L, i.e., from 5 to 4.5 mg-N/L, and so on. Use the average ammonia-N concentration within each zone to calculate the flux of ammonia-N into the biofilm in that zone, and then use that flux to calculate the required media volume for the zone. For illustrative purposes consider a reduction in the ammonia-N concentration from 5 to 4.5 mg-N/L. The average ammonia-N concentration for this zone will be 4.75 mg-N/L, and the average flux of ammonia-N into the biofilm will be (Eq. 19.5):

$$J_{NH} = 1.5\left(\frac{4.75}{1.5 + 4.75}\right) = 1.14 \text{ g } NH_3\text{-N/(m}^2 \cdot \text{day)}$$

Then, the media volume will be:

$$V_M = \frac{(5{,}000)(5 - 4.5)}{(1.14)(140)} = 15.7 \text{ m}^3$$

Continue this calculation for each of the zones and tabulate the results. They are shown in Table E19.2. Adding the volumes of the individual zones gives the total media volume required to reduce the ammonia-N concentration from 5 to 1 mg-N/L. It is 152.3 m^3. This compares to a volume of 357 m^3 to reduce the ammonia-N concentration by 15 mg-N/L, from 20 to 5 mg-N/L.

c. What total media volume would be required to reduce the ammonia-N concentration from 20 to 1 mg-N/L?
It is the sum of the two media volumes calculated above, or 357 + 152 = 509 m^3.

Table E19.2 Computation of the Media Volume Required to Reduce the Concentration of Ammonia-N from 5.0 to 1.0 mg/l in Example 19.3.2.4

Zone mg-N/L	$S_{NH,Ave}$ mg-N/L	$J_{NH,Ave}$ g NH_3-N/(m^2 · day)	Media volume m^3
4.5–5.0	4.75	1.14	15.7
4.0–4.5	4.25	1.11	16.1
3.5–4.0	3.75	1.07	16.7
3.0–3.5	3.25	1.03	17.3
2.5–3.0	2.75	0.97	18.4
2.0–2.5	2.25	0.90	19.8
1.5–2.0	1.75	0.81	22.0
1.0–1.5	1.25	0.68	26.3

d. What media depth would be required if no recirculation was used?
The media depth is just the volume divided by the cross-sectional area. The area can be obtained from the definition of the THL, given by Eq. 16.8. For $\alpha = 0$ and $A_H = 1.8$ m/hr (= 43.2 m/day):

$$A_c = \frac{5{,}000}{43.2} = 115.7 \text{ m}^2$$

$$L = \frac{509}{115.7} = 4.4 \text{ m}$$

At 0.61 m/module, this is depth is equivalent to 7.2 modules. Since whole modules must be used, eight will be required. This corresponds to a total media depth of 4.9 m. This will result in a modest increase in the THL, which is acceptable. The diameter of the trickling filter would be 23 m.

19.3.3 Sizing Trickling Filters with the Modified Velz/Germain Equation

An empirical model that has been referred to in the literature as either the modified Velz or the Germain equation has been used extensively to size trickling filters for removal of organic matter.[19,58,61,62] The Velz equation is described in Section 16.4.1. The development of the Velz/Germain equation assumes first order removal of substrate with respect to both the substrate concentration and the biomass concentration. It also assumes that the biomass concentration is constant throughout the trickling filter and that the trickling filter behaves as a plug–flow reactor with a residence time, τ. Based on these assumptions, the concentration of soluble substrate in the effluent, S_{Se}, is related to the concentration of soluble substrate actually applied to the top of the trickling filter, S_{Sa}, by:

$$\frac{S_{Se}}{S_{Sa}} = \exp(-k \cdot X_{B,Hf} \cdot \tau) \tag{19.6}$$

where k is a reaction rate coefficient and $X_{B,Hf}$ is the biomass concentration. However, empirically it was found that the HRT is related to the trickling filter THL and the media depth, L, according to:

$$\tau = \frac{C \cdot L}{A_H^n} \tag{19.7}$$

where C and n are empirical coefficients. Combining these two equations and combining k, C, and $X_{B,Hf}$ into a new coefficient K, called the treatability coefficient, gives:

$$\frac{S_{Se}}{S_{Sa}} = \exp\left(\frac{-K \cdot L}{A_H^n}\right) \tag{19.8}$$

When recirculation is not used, the applied soluble substrate concentration is just the concentration in the process influent wastewater. However, when recirculation is used, the influent is diluted by the recirculated effluent, making the applied substrate concentration less than the concentration in the untreated wastewater. In that case, the applied substrate concentration is given by Eq. 16.5. Combining Eqs. 19.8, 16.5,

and 16.8, which defines the THL, gives the most common form of the modified Velz/Germain equation:

$$\frac{S_{Se}}{S_{SO}} = \left\{ (1 + \alpha)\exp\left[\frac{K \cdot L}{\left(\frac{F(1 + \alpha)}{A_c} \right)^n} \right] - \alpha \right\}^1 \tag{19.9}$$

This equation is dimensional, with the value of K depending on the value of n and the units used for flow rate, F, cross-sectional area, A_c, and depth, L. Consequently, when a value of K is taken from the literature, particular attention must be paid to the units used and those units must be retained. Since K is a reaction rate coefficient, its value is temperature dependent. That dependency is typically expressed by Eq. 3.95, with the temperature coefficient θ set to a value of 1.035. Sometimes Eq. 19.9 is made explicitly dependent on temperature by substituting Eq. 3.95 directly into it. Finally, in many applications of the Velz/Germain equation, the term F/A_c is referred to as the unit wastewater application rate.

As discussed in Section 16.4.2, in the Eckenfelder[16,17] model the treatability coefficient, K, is expressed as $K_f a_c$ where K_f is another treatability coefficient and a_c is the media specific surface area. As discussed previously in this chapter, this may be appropriate for applications where channeling of flow and media plugging are not likely to be important, but is inappropriate when they are. The Eckenfelder equation has received less use than the modified Velz/Germain equation and, consequently, the empirical data base from which to select appropriate model coefficients is much more limited. This reduces its practical utility. However, when sufficient data are available it may be a superior model with which to correlate full-scale or pilot-scale data.

Although the modified Velz/Germain and the Eckenfelder equations were developed theoretically, it is now recognized that their theoretical basis is flawed. Consequently, they must be viewed as empirical models which are best used to correlate performance data. Nevertheless, because the modified Velz/Germain equation has been widely used, a significant empirical database exists, and it has been used to estimate a variety of trickling filter performance characteristics.[61,62] Like the NRC[37] and Galler–Gotaas[18] equations, it has been used to characterize the removal of organic matter by a trickling filter and secondary clarifier system. In these instances, the performance of the secondary clarifier is implicitly incorporated into the treatability coefficient. It has also been used to characterize the removal of organic matter across just the trickling filter itself. Furthermore, in some of those cases the relationship between the concentration of soluble organic substrate in the influent and effluent from the trickling filter is characterized, whereas in others the relationship between the total concentration of biodegradable organic matter in the influent and the concentration of soluble organic matter in the effluent is characterized. Due to this variation in usage, it is important that the basis for treatability coefficients taken from the literature be identified.

The available data[61,62] indicate that the value of n in the modified Velz/Germain equation varies with the trickling filter media type and application, ranging from 0.3 to 0.7. However, it is a standard practice by some to use a value of 0.5 for all situations so that the relative treatability of various wastewaters can be assessed by

the numerical value of the treatability coefficient. Although this approach results in some sacrifice in precision, it does allow easy comparison of the relative performance of various trickling filter applications.

Experience indicates that a number of adjustments must be made to allow the modified Velz/Germain equation to accurately characterize the performance of full-scale trickling filter applications. Adjustments must be made for the depth of the trickling filter media and for the influent waste strength. A correction can also be made if the THL is below the minimum required value of 1.8 m/hr. The need for these adjustments is related to the fact that this model is not fundamental in nature. As discussed above, the best use of the modified Velz/Germain equation is to correlate full-scale or pilot-scale performance data. It can be used to interpolate within a particular data set, but caution should be exercised if performance estimates are to be extrapolated beyond the existing data base. Detailed descriptions of the use of the modified Velz/Germain equation to predict the performance of fullscale trickling filter applications are presented elsewhere.[61,62]

Examination of Eq. 19.9 reveals that it contains eight terms: (1) the wastewater flow rate, F, (2) the wastewater soluble substrate concentration, S_{SO}, β) the desired effluent soluble substrate concentration, S_{Se}, (4) the treatability coefficient, K, (5) the exponent n, (6) the media depth, L, (7) the cross-sectional area, A, and (8) the recirculation ratio, α. In a design setting, F, S_{SO}, S_{Se}, K, and n are all known, leaving three unknowns. Since there is one equation and three unknowns, two will be free design variables and can be chosen at will, allowing the other to be calculated. One approach is to choose the recirculation ratio and the cross-sectional area to set the THL within the allowable range, allowing the media depth, L, to be calculated. Another is to specify the depth, typically at the maximum unsupported depth for the particular media, and then investigate the effects of the recirculation ratio on the area required to maintain the minimum THL. Since total systems costs depend on the media volume used, the amount of recirculation pumped, and the height to which the wastewater and recirculation flow must be pumped, the opportunity exists for the designer to seek an optimal design that minimizes the system costs.

Example 19.3.3.1 illustrates use of the modified Velz/Germain equation to design a trickling filter for the removal of biodegradable organic matter

---

Example 19.3.3.1

Consider the wastewater that was used in the examples of Section 19.3.2, for which 65% of the BOD_5 is soluble. Use the modified Velz/Germain equation to size a trickling filter to produce an effluent with a soluble BOD_5 concentration of 10 mg/L, which is essentially the same effluent quality achieved in Example 19.3.2.2. Assume a value for K of 0.4 $(m/hr)^{0.5}/m$ (which is equivalent to the value of 0.075 $gpm^{0.5}/ft$ often reported in the literature[61,62]) and a value for n of 0.5. The temperature is 20°C. The characteristics of the media are such that the THL must to be maintained at a value of 1.8 m/hr or greater, the depth of one bundle is 0.61 m, and the unsupported depth must be no greater than 1.7 m (11 bundles).

a. What is the influent soluble BOD_5 concentration?
   Since the total BOD_5 in the influent is 150 mg/L and 65% is soluble, the soluble BOD_5 concentration is 97.5 mg/L.

b. What media volume is required if no recirculation is used and the cross-sectional area is selected to maintain a THL of 1.8 m/hr (43.2 m/day)?
The required cross-sectional area can be calculated from the definition of THL, given by Eq. 16.8:

$$A_c = \frac{5{,}000}{43.2} = 115.7 \text{ m}^2$$

The media depth can be calculated with Eq. 19.9 after substituting all of the known values. To be consistent with the units of K, F should be expressed in units of m^3/hr. Thus, F = 208 m^3/hr.

$$\frac{10}{97.5} = \left\{(1 + 0)\exp\left[\frac{(0.4)L}{\left(\frac{208(1 + 0)}{115.7}\right)^{0.5}}\right] - 0\right\}^1$$

Taking the inverse of each side gives:

$$9.75 = \exp[0.298\ L]$$

$$L = \frac{\ln(9.75)}{0.298} = 7.64 \text{ m}$$

This exceeds the typical maximum value of 6.7 m. Furthermore, this represents 12.5 bundles, suggesting that the depth would have to be increased to 13 bundles (7.93 m), exceeding the maximum by two bundles. Thus, such a design would be questionable and the designer would have to carefully investigate the media characteristics before adopting it. If it were used, the total media volume would be (115.7) (7.93) = 917 m^3.

c. If the depth is limited to 6.7 m and the THL is maintained at the minimum value of 1.8 m/hr, what recirculation rate and cross-sectional area are required?
Again, Eq. 19.9 must be used. Recognizing that the term $F(1 + \alpha)/A_c$ is just the THL, which has a value of 1.8 m/hr, gives:

$$\frac{10}{97.5} = \left\{(1 + \alpha)\exp\left[\frac{(0.4)(6.7)}{(1.8)^{0.5}}\right] - \alpha\right\}^1$$

Simplifying and taking the inverse of both sides gives:

$$9.75 = (1 + \alpha)(7.37) - \alpha$$

Solving for α gives:

$$\alpha = 0.37$$

Thus, the recirculation rate is (0.37)(5,000) = 1,850 m^3/day.
For this α, the trickling filter cross-sectional area is:

$$A_c = \frac{(5{,}000)(1 + 0.37)}{43.2} = 159 \text{ m}^2$$

The total media volume is (159)(6.7) = 1,065 m^3, which is quite similar to the volume for Example 19.3.2.2, which had similar design criteria. Comparing this result to the deeper trickling filter sized in part b, this represents an increase in volume of 1,065 − 917 = 148 m^3, or 16%.

The above example illustrates one of the controversial aspects of the modified Velz/Germain equation. It predicts that the required media volume decreases as the media depth increases, which is in contrast with actual experience as discussed in Section 19.2.3. As discussed previously in this section, the value of K is often adjusted for media depth. The correction typically used is:[61,62]

$$K_2 = K_1(L_1/L_2)^{0.5} \tag{19.10}$$

where the subscripts represent two different depths. This further illustrates the empirical nature of the modified Velz/Germain equation and the fact that care must be exercised in its use. Nevertheless, the extensive database available makes the modified Velz/Germain equation useful for design purposes.

19.3.4 The Model of Logan et al.

The model of Logan et al.,[30,31] discussed in Section 16.4.5, represents a new generation of more fundamental trickling filter models. It is based on mass transfer principles for soluble biodegradable organic matter and oxygen. It also accounts for the multi-component nature of most wastewaters by dividing the organic material into five categories, each with different diffusional characteristics. The removal of each category is modeled by assuming the presence of a thick biofilm, i.e., one where the biofilm thickness exceeds that required for aerobic metabolism, and laminar flow over an inclined surface. Substrate removal occurs by transport in the thin liquid film that flows over the media and into the biofilm. The effects of media configuration on performance are predicted based on differences in flow patterns through the various media and their impact on substrate transport in the liquid film. This model provides no coupling between the removal of organic material in the trickling filter and oxygen transfer. Rather, it limits the maximum substrate flux to a value consistent with the oxygen transfer capability of the system.

The model of Logan et al.[30,31] is mentioned to alert the reader to its existence and to the fact that research is ongoing to develop more fundamental trickling filter design models. As discussed previously, this model has been used successfully to predict the oxygen transfer rate, and hence the zero order flux of ammonia-N, in separate stage nitrifying trickling filters. Although some day fundamental models will be used routinely to characterize the performance of full-scale trickling filters, insufficient experience currently exists with the model of Logan et al.[30,31] to conclude that it can be broadly used for the design of full-scale trickling filters.[26]

19.3.5 Ventilation System

Proper design of the ventilation system is necessary to maintain aerobic conditions within a trickling filter. The pressure drop for air flowing through a trickling filter is typically low, often less than 1 mm of water per m of media depth. Due to this low pressure drop, only a small motive force is necessary to cause air to flow. For forced draft ventilation systems the low pressure drop results in relatively low power requirements for the air supply fans, generally on the order of a few kW even for large trickling filters. The head on the ventilation fan will generally be less than 1.5 cm of water. The relatively low pressure drop for air flowing through a trickling filter also affects the design of the air distribution system.

Adequately sized underdrains are required to allow air to flow into or out of the bottom of the trickling filter. To ensure that air can move to and from the center of the trickling filter, underdrains are sized so that they are submerged no more than 50% at the peak hydraulic flow.[61,62] In addition, natural draft ventilation systems need a sufficiently large vent area to allow the air to flow to or from the underdrains. Typical criteria for trickling filters using plastic sheet media are the vent area per unit of media volume or the vent area per unit of peripheral length. Values such as 1 to 2 m^2/1,000 m^3 of media or 0.022 to 0.033 m^2/m of tower periphery are often used.[61,62] Another criterion for rock media trickling filters is a vent area equal to 15% of the trickling filter cross-sectional area.[61,62]

The considerations for sizing a forced draft ventilation system are somewhat different. First, air must be distributed equally across the entire cross section of the trickling filter. This differs from a natural draft ventilation system where the air-flow will naturally distribute itself, as discussed in Section 19.2.5. Natural drafting will tend to occur in a trickling filter using forced draft ventilation, resulting in short-circuiting, if the distribution of air across the trickling filter cross section is not uniform. In fact, the natural draft tendency in a trickling filter with forced draft ventilation can result in air flowing up in one portion of the trickling filter and down in another. The air-flow distribution system for a forced draft ventilation system consists of a piping network with a series of openings located across the cross section of the trickling filter. These openings are sized so that the air-flow rate through each is the same, thereby resulting in uniform air-flow distribution. A velocity of 0.3 to 0.6 m/sec is also typically maintained as another measure to provide good air-flow distribution. Air-flow requirements are generally calculated based on process oxygen requirements and the oxygen transfer efficiency, which are typically on the order of 2 to 10%. Forced draft ventilation systems can be designed as either up-flow or down-flow systems. Down-flow systems can be designed without covers, while covers are generally required for up-flow systems.

19.3.6 Coupled Trickling Filter/Activated Sludge Processes

As discussed in Section 19.1, coupled TF/AS processes consist of three components: a trickling filter, a suspended growth bioreactor, and a secondary clarifier. The trickling filter is sized using the procedures for sizing a stand-alone trickling filter. The secondary clarifier is sized in the same manner as one used with the activated sludge process. Since coupled TF/AS processes typically provide good control over the growth of filamentous microorganisms, the secondary clarifier will produce a thick RAS for recycle to the suspended growth bioreactor. A further consideration arises with TF/SC systems, however, because the primary function of the suspended growth bioreactor in them is to flocculate the poorly settleable suspended solids leaving the trickling filter. Experience indicates that the biological floc produced in some of these systems is relatively weak, possibly because few filamentous bacteria are present to serve as a backbone, as discussed in Section 10.2.1. Consequently, flocculating inlet wells are often provided in the secondary clarifiers used with TF/SC processes.[42] Such inlet wells can also be provided with other coupled TF/AS processes, but they are generally considered to be less necessary.

The design of the suspended growth bioreactor requires consideration of both the removal of biodegradable organic matter and the establishment of conditions necessary for good flocculation. As with activated sludge systems, the latter often controls the design.[35] Nevertheless, the trickling filter and the suspended growth bioreactor must also be viewed as a system and designed accordingly. A portion of the organic matter contained in the influent wastewater will be metabolized in the trickling filter, and the remainder must be metabolized in the suspended growth bioreactor. Likewise, a certain degree of flocculation will occur in the trickling filter, and the suspended growth bioreactor must be sized to achieve the remaining flocculation required. In general, the lower the TOL on the trickling filter, the greater the metabolism of organic matter and the higher the degree of flocculation that occurs there. This results in reduced requirements for organic matter metabolism and flocculation in the suspended growth bioreactor. Likewise, application of a higher TOL to the trickling filter results in a greater need for organic matter stabilization and flocculation in the suspended growth bioreactor. Table 19.6 provides general guidance concerning the relationship between the trickling filter TOL and the suspended growth bioreactor SRT required to achieve good flocculation in coupled TF/AS processes.

In coupled TF/AS processes, the SRT is calculated in the same fashion as for the activated sludge process. It is the mass of suspended solids in the suspended growth bioreactor divided by the rate at which suspended solids are wasted from the system, either intentionally or unintentionally. At steady-state, of course, the rate at which suspended solids are wasted from the process must equal the waste solids production rate, which is determined by the biological reactions occurring in both the trickling filter and the suspended growth bioreactor. Consequently, waste solids production calculations must consider the pollutant loadings placed on the entire system, not just the organic matter contained in the trickling filter effluent.[10,22] Biomass grown in the trickling filter will slough off and pass into the suspended growth bioreactor. Such growth must be accounted for in the total process waste solids production calculation. It is common to estimate solids production using a net process yield, Y_n, as discussed in Section 9.4.1. Furthermore, even though soluble substrate may remain in the trickling filter effluent, it is all typically removed in the suspended growth bioreactor. Consequently, the mass of MLSS in the suspended growth bioreactor can be calculated with Eq. 9.2 and the solids wastage rate can be calculated with Eq. 9.3. In both cases, the term (S_{SO} + X_{SO}) should reflect the concentrations of organic matter entering the

Table 19.6 Combinations of TOL and SRT Used to Achieve Good Bioflocculation in Coupled TF/AS Systems

System	TOL kg $BOD_5/(m^3 \cdot day)$	SRT days
RF/AS and BF/AS	3–4	3.0
RF/AS and BF/AS	2–2.5	2.0
TF/AS	0.6–1.0	1.0
ABF	<0.6	<1.0

system. Less information is available about net process yields in coupled TF/AS systems than is available for activated sludge systems. However, because the retention of biomass in a trickling filter increases as the TOL is decreased, the value of Y_n typically is influenced more by the TOL on the trickling filter than by the SRT of the suspended growth bioreactor and decreases as the TOL is decreased. Furthermore, values are typically on the order of 0.7 to 0.9 mg TSS/mg BOD_5. The following example illustrates this approach for the design of a TF/SC system.

Example 19.3.6.1

A high-quality effluent in terms of effluent BOD_5 and suspended solids is desired for the trickling filter sized in Example 19.3.2.2. Size a solids contact unit to achieve this goal, thereby converting the system into a coupled TF/SC system. Assume that the net process yield, Y_n, has a value of 0.70 mg TSS/mg BOD_5 and that the MLSS concentration in the suspended growth bioreactor is 2500 mg/L.

a. What SRT value would be appropriate for this application?
The TOL used for the trickling filter in Example 9.3.2.2 is 0.75 kg BOD_5/(m^3·day). Thus, from Table 19.6, a suspended growth bioreactor with an SRT of 1.0 day would be appropriate.

b. What should the volume of the suspended growth bioreactor be?
From Example 19.3.2.2, the flow rate is 5,000 m^3/day and the total BOD_5 concentration is 150 mg/L (150 g/m^3). The required reactor volume can be calculated with Eq. 9.2 by making use of the fact that the desired MLSS concentration is 2,500 mg/L (2,500 g/m^3):

$$V = \frac{(1.0)(0.70)(5{,}000)(150)}{2{,}500} = 210\ m^3$$

This gives an HRT of 1.0 hr. The acceptability of this size from a mixing energy and oxygen transfer perspective would have to be verified using the procedures discussed in Section 10.2.5.

c. What is the excess solids production rate?
This can be calculated with Eq. 9.3:

$$W_M = (5{,}000)(0.70)(150) = 525{,}000\ g/day = 525\ kg/day$$

A similar approach can be used to design a coupled RF/AS system, as illustrated in the following example.

Example 19.3.6.2

The roughing filter sized in Example 19.3.2.1 is to be used in a coupled RF/AS process. What size suspended growth bioreactor is required. Assume that the net process yield, Y_n, has a value of 0.80 mg TSS/mg BOD_5 and that the MLSS concentration in the suspended growth bioreactor is 2500 mg/L. The net process yield is higher than in the preceding example because the TOL on the trickling filter is higher. Also assume that because the subject wastewater is readily biodegradable, flocculation will control the design.

a. What SRT value would be appropriate for this application?
From Example 19.3.2.1, the TOL is 2.5 kg $BOD_5/(m^3 \cdot day)$. Consequently, from Table 19.6, an SRT of 2 days is necessary to obtain good flocculation.

b. What should the volume of the suspended growth bioreactor be?
From Example 19.3.2.1, the flow rate is 5,000 m^3/day and the total BOD_5 concentration is 150 mg/L (150 g/m^3). The required reactor volume can be calculated with Eq. 9.2 by making use of the fact that the desired MLSS concentration is 2,500 mg/L (2,500 g/m^3):

$$V = \frac{(2.0)(0.80)(5{,}000)(150)}{2{,}500} = 480 \text{ m}^3$$

This gives an HRT of 2.3 hr. The acceptability of this size from a mixing energy and oxygen transfer perspective would have to be verified using the procedures discussed in Section 10.2.5.

c. What is the excess solids production rate?
This can be calculated with Eq. 9.3:

$$W_M = (5{,}000)(0.80)(150) = 600{,}000 \text{ g/day} = 600 \text{ kg/day}$$

More solids are produced than in the TF/SC system because the TOL on the roughing filter is much higher, thereby making Y_n higher and the SRT of the suspended growth bioreactor is not large enough to reduce it.

While both process kinetics and flocculation must be considered when sizing a coupled TF/AS process, the criteria presented in Table 19.6 will typically control the process size since flocculation is generally the governing event. The concentration of soluble, biodegradable organic matter in the coupled TF/AS process effluent can be calculated by using the specific growth rate of the biomass in the suspended growth bioreactor, just as with other suspended growth bioreactors. However, for coupled TF/AS processes a significant fraction of the biodegradable organic matter contained in the process influent wastewater will be removed in the trickling filter, and this will result in a significant input of microorganisms into the suspended growth bioreactor. This must be considered in the calculation of the specific growth rate if accurate predictions of the effluent soluble substrate concentration are to be made. The effect of influent biomass on the specific growth rate in a suspended growth bioreactor is discussed in Section 5.2.3, and the results are presented as Eq 5.50. It is repeated here, with the only modification being that the source of the heterotrophic biomass in the suspended growth bioreactor influent is identified:

$$\mu_H = \frac{1}{\Theta_c} + b_H - \frac{X_{B,H,TF}}{\tau \cdot X_{B,H}} \tag{19.11}$$

where $X_{B,H,TF}$ is the heterotrophic biomass concentration in the trickling filter effluent resulting from a single pass of the wastewater over the trickling filter. In other words, it is the concentration that would result if the trickling filter were the only biochemical operation being used. All other symbols refer to the suspended growth bioreactor. As illustrated, the feed of microorganisms into the suspended growth bioreactor reduces the specific growth rate, thereby reducing the effluent substrate concentration below the value that would be obtained in a system with the same SRT, but with no biomass input.

If the characteristics of the trickling filter effluent, i.e., S_{Sc} and $X_{BH,TF}$, could be defined, the relationships presented in Section 5.2.3 could be used to size the suspended growth bioreactor of a coupled TF/AS system. Unfortunately this cannot be done easily. First, we saw in Section 19.3.1 that trickling filter design procedures focus on the removal of substrate rather than on the growth of biomass. Thus, while S_{Sc} may be well defined, $X_{BH,TF}$ is not. If an attempt were made to estimate biomass growth by using the true growth yield alone and neglecting cell decay, $X_{BH,TF}$ would be overestimated, thereby causing the substrate removal capability of the suspended growth bioreactor to be overestimated. Second, the suspended solids in the effluent from the trickling filter consist of heterotrophic biomass, cell debris, inert influent suspended solids, and unmetabolized substrate, with their relative quantities depending on the characteristics of the influent wastewater, as well as on the TOL and THL of the trickling filter. Precise prediction of the concentrations of these constituents, which are required for use of the equations in Section 5.2.3, is not currently possible. Third, both aerobic and anaerobic metabolism can occur within the biofilm of a trickling filter. Although the outer portion of the biofilm is aerobic, the inner portion may be anaerobic. Moreover, the relative importance of aerobic and anaerobic metabolism will vary depending on the nature and concentration of the biodegradable organic matter in the influent wastewater, the availability of oxygen, and hydraulic conditions affecting biofilm thickness. Since yields are quite different under aerobic and anaerobic conditions, biomass production can vary significantly. This further complicates prediction of the concentrations of various types of suspended solids in the trickling filter effluent. Finally, biomass can accumulate within a trickling filter and be sloughed periodically, as discussed in Section 19.4. This results in time-variant concentrations of biomass and other particulate constituents in the trickling filter effluent. In many instances, sloughing cycles occur over the course of several days, or even several weeks, a time interval that can significantly exceed the suspended growth bioreactor SRT. In fact, significant variations in suspended growth bioreactor MLSS concentrations have been observed as a result of trickling filter sloughing cycles.[10,22] These variations can affect the performance of the suspended growth bioreactor in a significant manner. Consequently, the suspended growth bioreactor is typically sized by using the net process yield approach as illustrated in Examples 19.3.6.1 and 19.3.6.2.

Some of the difficulties discussed above can be avoided when the focus is on nitrification because then only the autotrophic biomass concentration need be known. As a result, the relationships presented in Section 5.2.3 have been used successfully to characterize the removal of ammonia-N in a coupled TF/AS process accomplishing combined carbon oxidation and nitrification.[12] The concentration of nitrifiers in the trickling filter effluent was estimated as the concentration of ammonia-N nitrified in the trickling filter multiplied by the nitrifier true growth yield. This was permissible because there is little difference between the true growth yield and the observed yield for autotrophic bacteria. This concentration was used, along with Eq. 19.11, to predict the nitrifier specific growth rate in the suspended growth bioreactor allowing estimation of the effluent ammonia-N concentration. The performance relationship developed is presented in Figure 19.10 where the effluent ammonia-N concentration is plotted as a function of the suspended growth bioreactor SRT divided by the nitrifier minimum SRT. Several curves are presented corresponding to different trickling filter nitrification efficiencies. As seen, the sloughing of nitrifiers from the trick-

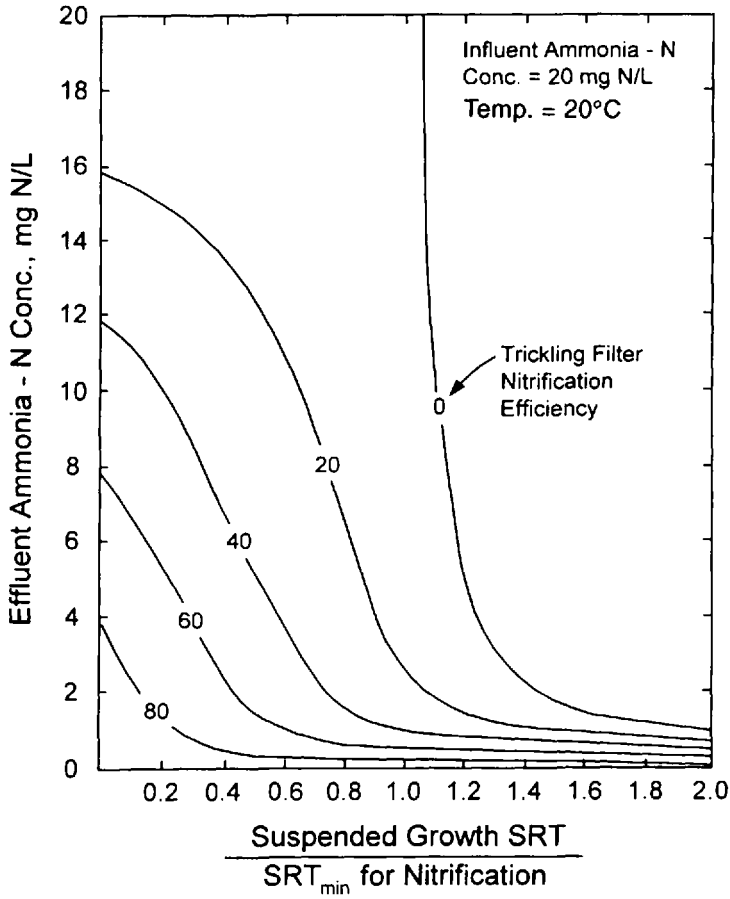


Figure 19.10 Effect of the SRT in a suspended growth bioreactor and the nitrification efficiency in an upstream trickling filter on the effluent ammonia-N concentration from a coupled TF/AS system. (From G. T. Daigger, L. E. Norton, R. S. Watson, D. Crawford, and R. B. Sieger, Process and kinetic analysis of nitrification in coupled trickling filter/activated sludge processes. *Water Environment Research* **65**:750–758, 1993. Copyright © Water Environment Federation; reprinted with permission.)

ling filter allows the suspended growth bioreactor to maintain nitrification even when operating at a nitrification design factor that would otherwise cause washout of the nitrifiers. Furthermore, the greater the seeding affect, i.e., the greater the ammonia-N conversion in the trickling filter, the lower the effluent ammonia-N concentration. The potential adverse affects of sloughing from the upstream trickling filter on the validity of this approach have been discussed in the literature.[9,40] Nevertheless, the results suggest that more fundamental procedures may be developed in the future for the design of suspended growth bioreactors in coupled TF/AS processes.

If the characteristics of the trickling filter effluent are described sufficiently, the procedures of Section 5.2.3 can also be used to calculate the oxygen requirement in the suspended growth bioreactor of a coupled TF/AS process. Just as with any sus-

pended growth process, the oxygen requirements must be compared with the energy input required for mixing and the larger of the two selected. The relationship between oxygen and mixing requirements differs among the various coupled TF/AS processes. The primary function of the suspended growth bioreactor in a TF/SC process is flocculation. Consequently, the TF/SC suspended growth bioreactor will generally be mixing limited. In contrast, substantial stabilization of biodegradable organic matter occurs in the suspended growth bioreactor of BF/AS or RF/AS systems, so that the energy input may be determined either by oxygen or mixing requirements. The difficulty, in any case, is in determining the degree of stabilization of biodegradable organic matter in the upstream trickling filter.

If the trickling filter effluent cannot be characterized sufficiently well to allow the relationships of Section 5.2.3 to be used to calculate the suspended growth bioreactor oxygen requirements, empirical correlations can be used. It will be recalled from Section 9.4.1 that the process oxygen stoichiometric coefficient, Y_{O2}, is often used with Eq. 9.4 to estimate the oxygen requirement for the activated sludge process in the absence of other data. Figure 9.8 shows how the value of that parameter varies with the SRT. A similar approach can be used to estimate the oxygen requirement in coupled TF/AS systems, except that an equation expresses the effect of SRT (or F/M ratio, U) on the process oxygen stoichiometric coefficient, rather than a figure. Based on studies of several full-scale coupled TF/AS processes, Harrison[20] used the following equation to estimate the overall oxygen stoichiometric coefficient that would occur in the suspended growth bioreactor in the absence of the trickling filter:

$$Y_{O2} = Y_{O2,S} + \frac{Y_{O2,d} \cdot b_H}{U} \tag{19.12}$$

where $Y_{O2,S}$ is the oxygen stoichiometric coefficient for synthesis, taken equal to 0.6 mg O_2/mg BOD_5; $Y_{O2,d}$ is the oxygen stoichiometric coefficient for decay, taken equal to 1.2 mg O_2/mg VSS; b_H is the decay coefficient, taken equal to $(0.115)(1.025)^{T-20}$ day^{-1}, where T is the temperature of the mixed liquor in the suspended growth bioreactor in °C; and U is the F/M ratio, based on the process influent BOD_5 loading and the MLVSS inventory in the suspended growth bioreactor. In addition, Harrison[20] developed the following equation relating the oxygen stoichiometric coefficient for the trickling filter, $Y_{O2,TF}$, to its TOL, Λ_S, expressed in units of kg $BOD_5/(m^3 \cdot day)$:

$$Y_{O2,TF} = \frac{0.25}{\Lambda_S} + 0.254 \tag{19.13}$$

The oxygen stoichiometric coefficient for the suspended growth bioreactor, $Y_{O2,SG}$, in a coupled TF/AS system can be estimated as the difference between Y_{O2}, as calculated with Eq. 19.12, and $Y_{O2,TF}$, as calculated with Eq. 19.13:

$$Y_{O2,SG} = Y_{O2} - Y_{O2,TF} \tag{19.14}$$

The oxygen requirement in the suspended growth bioreactor can then be estimated by multiplying $Y_{O2,SG}$ by the mass of BOD_5 entering the coupled TF/AS system per unit time, as indicated by Eq. 9.4. The following example illustrates the technique.

Example 19.3.6.3

What is the oxygen requirement for the suspended growth bioreactor in the RF/AS system sized in Example 19.3.6.2, as estimated using the procedure of Harrison? Assume that the temperature is 25°C and that the MLSS is 75% volatile

a. What is the F/M ratio for the process based on the organic matter entering the system and the mass of MLSS in the suspended growth bioreactor?
The F/M ratio can be calculated with Eq. 5.37 by extending it to account for the particulate contribution to the total BOD_5. From Example 19.3.6.2, the MLSS concentration 2,500 mg/L and the bioreactor volume is 480 m^3. Because the MLSS is 75% volatile, the MLVSS concentration is (0.75)(2,500) = 1,875 mg/L = 1,875 g/m^3. The wastewater flow rate is 5,000 m^3/day and the influent BOD_5 concentration is 150 mg/L = 150 g/m^3. Therefore,

$$U = \frac{(5{,}000)(150)}{(1{,}875)(480)} = 0.83 \text{ kg } BOD_5/\text{kg VSS}$$

b. What would the process oxygen stoichiometric coefficient be if no trickling filter was present?
This can be calculated with Eq. 19.12 after substituting the appropriate values for the oxygen stoichiometric coefficients for synthesis and decay:

$$Y_{O2} = 0.6 + \frac{(1.2)[(0.115)(1.025)^{T-20}]}{0.83} = 0.788$$

c. What is the oxygen stoichiometric coefficient for the trickling filter?
This can be calculated with Eq. 19.13. The roughing filter TOL is 2.5 kg $BOD_5/(m^3 \cdot \text{day})$. Therefore:

$$Y_{O2,TF} = \frac{0.25}{2.5} + 0.254 = 0.354$$

c. What is the oxygen stoichiometric coefficient for the suspended growth bioreactor?
The oxygen stoichiometric coefficient for the suspended growth bioreactor is just the difference between the overall and trickling filter oxygen stoichiometric coefficients, as reflected by Eq. 19.14:

$$Y_{O2,SG} = 0.788 - 0.354 = 0.434$$

d. What is the oxygen requirement for the suspended growth bioreactor?
The oxygen requirement for the suspended growth bioreactor can be calculated with Eq. 9.4 using the value of $Y_{O2,SG}$ calculated in part c above and the influent flow rate and BOD_5 concentration given in part a:

$$RO_{SG} = (5{,}000)(0.434)(150) = 326{,}000 \text{ g } O_2/\text{day} = 326 \text{ kg } O_2/\text{day}$$

One final point. Since flocculation rather than the removal of organic matter often controls the SRT required for the suspended growth bioreactor, a contact and sludge reaeration configuration similar to the contact/stabilization process can often be used.[42] This approach reduces the total volume of the suspended growth bioreactor.

19.4 PROCESS OPERATION

One of the often stated advantages of the trickling filter process is that it provides stable, reliable performance with relatively little operator attention. Operation with

little attention is possible because treatment capacity is determined primarily by the volume of media provided and its configuration. Consequently, daily operation typically consists of maintaining pumps and equipment. As a result, the perception exists that little can be done from an operational perspective if the desired degree of treatment is not being achieved; rather, the only solution is to add additional media volume. To a certain extent, this is true. However, in recent years it has been discovered that a greater degree of operational control is afforded by the process than originally thought and that proper operation is required to achieve optimum performance. This is particularly true for the TF/AS processes. This section provides an overview of trickling filter operation. Additional, detailed information is provided for the interested reader in various operational manuals of practice.[60,63]

19.4.1 Typical Operation

Experience indicates that the performance of a trickling filter can be improved significantly by proper control of the hydraulic regime applied to it. Section 19.2.7 discussed the recent results obtained with the use of mechanically driven distributors. For routine operation, a relatively rapid speed gives optimum treatment. The speed can be reduced periodically, however, causing increased hydraulic shear, with the associated sloughing of excess biomass from the media. This concept has been used historically to control trickling filter biofilm thickness.

Several approaches are available for temporarily increasing the hydraulic loading on a trickling filter. One that has worked well with a variety of high-rate media is to turn on all of the influent pumps while directing the flow to only one of the trickling filters. Increasing the hydraulic loading by a factor of 2.5 to 3 times for a period of about an hour is generally sufficient to slough off excess biomass. The frequency with which this should be done can be determined by periodically increasing the hydraulic loading and monitoring the trickling filter effluent. Samples can be collected every few minutes and examined visually. If dark colored (generally black) suspended solids are discharged from the trickling filter during the flushing event, the flushing frequency should be increased. On the other hand, if dark solids are not discharged, the flushing frequency can be decreased. A flushing frequency on the order of once per week is often found to be appropriate.

Another approach has been used with rock media trickling filters. A rope is tied to the end of the rotary distributor and used to restrain it. This reduces the rotational speed, causing the same effect as reducing the speed of a mechanically driven distributor. This practice has been referred to as "walking" the trickling filter. The flushing of dark, anaerobic solids from the trickling filter indicates the need for an increased flushing frequency, just as described above.

Proper control of the ventilation system is important to the maintenance of aerobic conditions at all times and to the minimization of heat loss during cold weather operation. As discussed in Section 19.2.4, the openings on natural draft ventilation systems must be reduced during cold weather operation to reduce airflow and the resulting heat loss. In contrast, openings must be increased as much as possible during more temperate conditions to encourage adequate air-flow.

Odors have frequently been detected from trickling filters. The reasons for this are many and varied, and may be complex for any particular installation. Good ventilation is one factor in odor control, along with control of biofilm thickness to

minimize anaerobic activity in it. Odors will also occur when septic wastewater is treated in a trickling filter. Volatile materials will be stripped from the wastewater when it is applied to the trickling filter, and air-flow through the trickling filter will disperse the stripped compounds. In such cases, pretreatment of the wastewater to remove odorous materials may be necessary.[10,22]

19.4.2 Coupled Processes

The operation of coupled TF/AS processes incorporates features of the operation of both components. As discussed above, preceding a suspended growth bioreactor by a trickling filter typically results in good control over the growth of filamentous microorganisms in the suspended growth bioreactor. This is generally a benefit and results in a process that is stable and reliable. However, it can result in a turbid effluent when insufficient filaments grow. The trickling filter/solids contact (TF/SC) process is particularly susceptible to this because most substrate is removed in the trickling filter, leaving little to be removed in the suspended growth bioreactor where filamentous microbial growth occurs. As a consequence, the mixed liquor from a TF/SC process is fragile and must be handled gently to avoid floc shear and elevated effluent suspended solids concentrations. Growth of insufficient filaments is less of a problem for processes such as roughing filter/activated sludge (RF/AS) and biofilter/activated sludge (BF/AS) because only a portion of the organic matter is removed in the trickling filter, leaving some to be metabolized in the suspended growth bioreactor. Consequently, a moderate population of filamentous microorganisms can be maintained in such processes, strengthening the floc.

A well designed coupled TF/AS process will provide the flexibility for operation in several modes. This is desirable because wastewater treatment facilities are not typically loaded to their design values on a consistent basis. During periods of low loading, the TOL on the trickling filter will be less than the design value, giving a higher degree of treatment there. Since less substrate removal will be required in the suspended growth bioreactor, the flexibility should be provided to reduce its SRT to lower aeration power requirements. This may also be necessary to avoid unwanted nitrification. Plant operations personnel must effectively use the operational flexibility provided in the design to optimize plant performance and minimize treatment costs. For example, during periods of reduced loading, many BF/AS facilities are operated either with less suspended growth bioreactor volume in service or in the ABF mode. Likewise, facilities designed as RF/AS processes can be operated as TF/SC processes during periods of reduced loading. Finally, a TF/SC system can be operated simply as a trickling filter during periods of reduced loading.

Periodic sloughing of biomass occurs in many trickling filter installations. In coupled TF/AS systems this periodic sloughing can lead to a sudden increase in the mass of MLSS in the suspended growth bioreactor.[10,22,42] Because of the good sludge settling properties of the suspended biomass, such sudden increases can often be tolerated with little or no adverse impact on effluent quality. However, if the process is loaded to its maximum treatment capacity, such sloughing events can lead to significant loss of suspended solids from the final clarifier, with the associated deterioration in effluent quality. This can be avoided by maintaining sufficient hydraulic loading on the trickling filter to periodically slough excess biomass, demonstrating the importance of this practice. In systems where controlled sloughing is not prac-

ticed, the mass of MLSS in the suspended growth bioreactor must be maintained at reduced levels in anticipation of periodic sloughing events.

19.4.3 Nuisance Organisms

An extremely diverse biota can develop in a trickling filter. It contains a variety of organisms, ranging from bacteria to protozoa, worms, adult and larval filter flies (often of the genus Psychoda), and snails. This diverse biota contributes to stabilization of the organic matter applied to the trickling filter by increasing the length of the food chain. For example, protozoa feed on bacteria, worms feed on protozoa, etc., with a yield associated with each trophic level, thereby reducing net biomass production. This is particularly important in rock media trickling filters where an extremely diverse biota can develop. However, in some instances some of these organisms can cause operational problems.

Filter flies can be a nuisance, particularly if a large hatch results in sufficient numbers to impact plant personnel and neighbors. Snails cause problems because their shells act like grit, accumulating in downstream unit operations, such as the suspended growth bioreactor of a coupled TF/AS process. They can also accumulate in the solids handling system, such as in anaerobic digesters, resulting in significant volume losses. Filter fly larvae, worms, and snails can cause performance problems in nitrifying trickling filters if they consume biomass faster than it is being produced, as has been observed in a number of pilot-scale and full-scale trickling filters. For these reasons, it is important to control nuisance organisms.

In spite of the need, few techniques have been developed to control nuisance organisms in trickling filters. In some cases, chlorination (for all media), flooding (especially rock media trickling filters), and hydraulic flushing have been successful. Media selection also affects the growth of nuisance organisms to some extent. Media with poor wetting characteristics are prone to the growth of nuisance organisms because they utilize the wet and dry areas of the media for different portions of their life cycle.[10,11,22,61,62]

19.5 KEY POINTS

1. A trickling filter is an attached growth, aerobic biochemical operation consisting of five major components: (1) the media bed, (2) the containment structure, (3) the wastewater application (or dosing) system, (4) the underdrain, and (5) the ventilation system.
2. Oxygen is provided either by natural draft or forced draft ventilation. Natural draft ventilation occurs because of differences in the density of air inside and outside the trickling filter. Forced draft ventilation requires a motive force for air, such as a fan.
3. Trickling filter systems often include a clarifier to separate produced biomass, although it may not be needed in nitrification applications because of the low growth yield.
4. The liquid-flow pattern through a trickling can be characterized as plug–flow with dispersion. Consequently, a spatial distribution of microorganisms will develop with heterotrophic bacteria in the upper portion of the

trickling filter and, if the organic loading is sufficiently low, autotrophic nitrifying bacteria in the lower portion.

5. Trickling filters are classified by their treatment objective: partial removal of organic matter (referred to as a roughing trickling filter), relatively complete removal of organic matter (carbon oxidation), combined carbon oxidation and nitrification, and separate stage nitrification. These applications are defined by the total organic loading (TOL) applied to the trickling filter and the characteristics of the wastewater.
6. Due to significant differences in performance and operational characteristics, trickling filter processes are also classified by media type.
7. Adequate pretreatment must be provided before wastewater is applied to a trickling filter. Most installations include primary clarifiers for this purpose.
8. Coupled trickling filter/activated sludge (TF/AS) systems consist of a trickling filter, a suspended growth bioreactor, and a clarifier. The trickling filter effluent passes directly into the suspended growth bioreactor without clarification. As a consequence, the biologies of the two biochemical operations interact directly.
9. Trickling filters can be used to treat a wide range of wastewaters. They are often used to treat high-strength, readily biodegradable wastewaters where they provide preliminary or roughing treatment prior to a suspended growth system. They can also be used for municipal wastewaters, although in some cases they must be coupled with a suspended growth bioreactor to produce the high-quality effluent required. Combined carbon oxidation and nitrification and separate stage nitrification can also be readily accomplished.
10. Trickling filter process loadings can be expressed as either the total organic loading (TOL) or the surface organic loading (SOL). The TOL is expressed per unit of media volume (e.g., kg COD or $BOD_5/(m^3 \cdot day)$), while the SOL is expressed per unit of media surface area (e.g., kg COD or $BOD_5/(m^2 \cdot day)$). Similar loadings can be defined for ammonia-N and total Kjeldahl nitrogen (TKN) for nitrification applications.
11. The total hydraulic loading (THL) is the applied flow rate per unit of cross-sectional area, and typically has units of m/hr. Application of a minimum THL is required to achieve effective use of all of the media. In some cases this requires recirculation of treated effluent. However, the recirculation rate is not a fundamental design parameter. Increased recirculation flow beyond the amount required to achieve the minimum THL will not improve performance.
12. For a constant TOL, trickling filter performance is affected by media depth only over a relatively narrow range. Performance improves as the depth is increased up to about 3 to 4 meters, but little improvement occurs as the depth is increased further.
13. Trickling filter performance is affected by the temperature of the wastewater flowing over the media. Because trickling filters are effective heat transfer devices, steps must be taken during winter operation to mitigate the cooling effects of the ambient air temperatures.

14. Trickling filters can use either rotary or fixed nozzle distributors. Rotary distributors possess significant performance and operational advantages, and they are used more frequently.
15. Natural draft ventilation systems are designed to provide sufficient vent and underdrain areas to allow adequate convective air movement for oxygen transfer. Forced draft ventilation systems utilize fans and duct systems to evenly distribute the needed air.
16. Trickling filter effluents may contain fine, colloidal suspended solids that settle poorly in conventional clarifiers. Coupled TF/AS systems provide greater control over these suspended solids by flocculating them for removal in the clarifier.
17. Process loading factors, such as the TOL and $J_{S,A}$, can be used to size trickling filters. Process performance data from either comparable full-scale applications or a pilot plant is used to select the appropriate loading factor.
18. The Velz/Germain[19,58] equation is an empirical model that has been used frequently to size trickling filters. Because of its frequent use, a significant database exists to allow selection of appropriate model coefficients.
19. The model of Logan et al.[30,31] is a more fundamental trickling filter model. Experience with it is currently limited, and this must be considered when using it to size full-scale applications.
20. A trade-off is inherent in the design of a coupled TF/AS system. Use of a large trickling filter allows use of a small suspended growth bioreactor, and vice versa. Consequently, the trickling filter TOL and suspended growth bioreactor SRT must be selected together.
21. Oxygen requirements in coupled TF/AS systems can be determined by either fundamental process calculations or empirical correlations. In either case, the trickling filter reduces the oxygen requirement in the suspended growth bioreactor.
22. Improved trickling filter performance can be obtained by proper control of the biofilm thickness through periodic flushing. Flushing can be accomplished in numerous ways.
23. Significant operating economies can be achieved when a trickling filter system is designed with the flexibility to be operated in more than one mode.
24. Nuisance organisms that can proliferate in trickling filters include flies, snails, and worms. Problems caused by the growth of these organisms range from simple nuisance to reduced performance. Techniques available to control their growth include control of the THL, flushing, chlorination, and flooding.

19.6 STUDY QUESTIONS

1. Describe the functions of the five major components of a trickling filter.
2. Describe the mechanisms responsible for natural draft ventilation in a trickling filter and define the conditions under which stagnation is likely to occur. How does the configuration of the air distribution system in a

trickling filter with natural draft ventilation differ from that in a trickling filter with forced draft ventilation? Why?

3. Describe the types of microorganisms present in the upper, middle, and lower portions of a trickling filter accomplishing combined carbon oxidation and nitrification and explain why the distribution develops?
4. Prepare a table that describes the various trickling filter process options, concentrating on differences associated with treatment objectives and media type. The table should define the essential features of each option.
5. Prepare a table that describes the various coupled TF/AS process options. What are the principal differences between the processes? Provide a quantitative definition of each process option.
6. Develop a table listing the benefits, drawbacks, and typical applications of the various trickling filter process options.
7. A wastewater with a flow rate of 3,500 m^3/day and a BOD_5 concentration of 120 mg/L is to be treated using a trickling filter sized at a TOL of 0.6 kg $BOD_5/(m^3 \cdot day)$. Develop a curve showing the recirculation ratio required to maintain a minimum THL of 1.8 m/hr as a function of trickling filter media depth. Define the practical range of media depths. Select a design media depth for this application and present the rationale for your selection.
8. Discuss the techniques that can be used to minimize the heat loss from a trickling filter during cold weather operation.
9. Prepare a table contrasting the characteristics of the various trickling filter media. What are the benefits and drawbacks associated with each media type, and when would each typically be used?
10. Prepare a table contrasting the relative benefits and drawbacks of rotary and fixed nozzle distributors.
11. Consider the trickling filter application discussed in Study Question 7. Using the recommended SK values presented in Table 19.5 calculate the necessary rotary distributor rotational speed for typical operation and for flushing. Assume that a four arm distributor is used.
12. Use the process loading factor approach to size a plastic sheet media trickling filter to accomplish tertiary nitrification. The secondary effluent to be nitrified has a low BOD_5 concentration, an ammonia-N concentration of 25 mg-N/L, and a flow rate of 15,000 m^3/day. The desired effluent ammonia-N concentration is 2 mg-N/L. Document and justify any assumptions necessary to complete these calculations.
13. Using the coefficients in Example 19.3.3.1 and Eq. 19.9, develop a curve demonstrating the impact of the TOL on the effluent soluble BOD_5 concentration from a trickling filter treating a wastewater with a soluble BOD_5 concentration of 125 mg/L and a flow rate of 4,000 m^3/day. The subject trickling filter has a media depth of 6.7 m. Maintain a THL of 1.8 m/hr.
14. Consider a wastewater with a flow of 15,000 m^3/day and a BOD_5 concentration of 175 mg/L. Size both the trickling filter and the suspended growth bioreactor for each of the four coupled TF/AS processes. Define the rationale for selection of the specific process loadings. For the BF/

AS and RF/AS processes, use the empirical relationships of Harrison to calculate the suspended growth bioreactor oxygen requirement.

15. Make a list of the critical considerations in the operation a trickling filter. What additional considerations are involved in the operation of a coupled TF/AS process?

REFERENCES

1. Albertson, O. E., Slow down that trickling filter. *Operations Forum* **6**(1):15–20, 1989.
2. Albertson, O. E., Slow-motion trickling filters gain momentum. *Operations Forum* **6**(8): 28–29, 1989.
3. Anderson, B., H. Aspegren, D. S. Parker, and M. P. Lutz, High rate nitrifying trickling filters. *Water Science and Technology* **29**(10/11):47–60, 1994.
4. Barker, L. S., G. T. Daigger, and R. C. Naef, A comparative evaluation of trickling filter media performance: a preliminary comparison of rock and plastic media. *Proceedings of the Utah Water Pollution Control Association Annual Meeting*, pp. 66–70, 1984.
5. Benzie, W., H. O. Larkin, and A. F. Moore, Effects of climatic and loading factors on trickling filter performance. *Journal, Water Pollution Control Federation* **35**:445–455, 1963.
6. Boller, M. and W. Gujer, Nitrification in tertiary trickling filters followed by deep bed filters. *Water Research* **20**:1363–1373, 1986.
7. Bruce, A. M. and J. C. Merkins, Recent studies of high rate biological filtration *Water Pollution Control* **69**:113–139, 1970.
8. Bruce, A. M. and J. C. Merkins, Further studies of partial treatment of sewage by high-rate biological filtration. *Water Pollution Control* **72**:499–523, 1973.
9. Daigger, G. T., Closure to discussion of 'Process and kinetic analysis of nitrification in coupled trickling filter/activated sludge processes'. *Water Environment Research* **67**: 380–382, 1995.
10. Daigger, G. T. and J. R. Harrison, Recent developments in trickling filter/activated sludge technology. *Proceedings of the Australian Water and Wastewater Association Annual Meeting*, Melbourne, Australia, pp. 297–304, 1985.
11. Daigger, G. T. and J. R. Harrison, A comparison of trickling filter media performance. *Journal, Water Pollution Control Federation* **59**:679–685, 1987.
12. Daigger, G. T., L. E. Norton, R. S. Watson, D. Crawford, R. B. Sieger, Process and kinetic analysis of nitrification in coupled trickling filter/activated sludge processes. *Water Environment Research* **65**:750–758, 1993.
13. Daigger, G. T., T. A. Heinemann, G. Land, and R. S. Watson, Practical experience with combined carbon oxidation and nitrification in plastic media trickling filters. *Water Science and Technology* **29**(10/11):189–196, 1994.
14. Daigger, G. T., L. E. Norton, R. S. Watson, D. Crawford, and R. B. Sieger, Process and kinetic analysis of nitrification in coupled trickling filter/activated sludge processes. *Water Environment Research* **65**:750–758, 1993.
15. Duddles, G. A., S. E. Richardson, and E. F. Barth, Plastic medium trickling filters for biological nitrogen control. *Journal, Water Pollution Control Federation* **46**:937–946, 1984.
16. Eckenfelder, W. W., Trickling filtration design and performance. *Transactions of the American Society of Civil Engineers* **128**, Part III:371–384, 1963.
17. Eckenfelder, W. W. and W. Barnhart, Performance of a high-rate trickling filter using selected media. *Journal, Water Pollution Control Federation* **35**:1535–1551, 1963.

18. Galler, W. S. and H. G. Gotaas, Analysis of biological filter variables. *Journal of the Sanitary Engineering Division, ASCE* **90**(SA6):59–79, 1964.
19. Germain, J. E., Economic treatment of domestic waste by plastic-medium trickling filters. *Journal, Water Pollution Control Federation* **38**:192–203, 1966.
20. Harrison, J. R., Survey of plants operating activated biofilter/activated sludge. Paper presented at the Northern Regional Conference and Training School of the California Water Pollution Control Association, 1980.
21. Harrison, J. R. and P. L. Timpany, Design considerations with the trickling filter solids contact process. *Proceedings of the Joint Canadian Society of Civil Engineers, American Society of Civil Engineers National Conference on Environmental Engineering*, Vancouver, B. C., Canada, pp. 753–762, 1988.
22. Harrison, J. R., Daigger, G. T., and J. W. Filbert, A survey of combined trickling filter and activated sludge processes. *Journal, Water Pollution Control Federation* **56**:1073–1079, 1984.
23. Hawkes, H. A., Film accumulation and grazing activity in the sewage filters at Birmingham. *Journal of the Institute of Sewage Purification* 88–102, 1957.
24. Hawkes, H. A., The effects of methods of sewage application on the ecology of bacteria beds. *Annals of Applied Biology* **47**:339–349, 1959.
25. Hawkes, H. A., *The Ecology of Waste Water Treatment*, Macmillan, New York, 1963.
26. Hinton, S. W. and H. D. Stensel, Discussion of: 'A fundamental model for trickling filter process design and engineering implications of a new trickling filter model'. *Journal, Water Pollution Control Federation* **61**:363–366, 1989.
27. Hinton, S. W. and H. D. Stensel, Experimental observations of trickling filter hydraulics. *Water Research* **25**:1389–1398, 1991.
28. Kincannon, D. F. and E. L. Stover, Design methodology for fixed film reactors, RBC's, and trickling filters. *Civil Engineering for Practicing and Design Engineers* **2**:107–124, 1983.
29. Kornegay, B. H., Modelling and simulation of fixed film biological reactors for carbonaceous waste treatment. In *Mathematical Modelling for Water Pollution Control Processes*, T. M. Keinath and M. Wanielista, eds. Ann Arbor Science, Ann Arbor, Michigan, pp. 271–315, 1975.
30. Logan, B. E., S. W. Hermanowicz, and D. S. Parker, Engineering implications of a new trickling filter model. *Journal, Water Pollution Control Federation* **59**:1017–1028, 1987.
31. Logan, B. E., S. W. Hermanowicz, and D. S. Parker, A fundamental model for trickling filter process design. *Journal, Water Pollution Control Federation* **59**:1029–1042, 1987.
32. Lumb, C and J. P. Barnes, Periodicity of dosing percolating filters. *Journal of the Institute of Sewage Purification* 1, 83–91, 1948.
33. Matasci, R. N., D. L. Clark, J. A. Heidman, D. S. Parker, B. Petrik, and D. Richards, Trickling filter/solids contact performance with rock filters at high organic loadings. *Journal, Water Pollution Control Federation* **60**:68–76, 1988
34. Metcalf & Eddy, Inc., *Wastewater Engineering: Treatment, Disposal, and Reuse*, Third Edition, McGraw-Hill, New York, 1991.
35. Newbry, B. W., G. T. Daigger, and D. Taniguchi-Dennis, Unit process tradeoffs for combined trickling filter and activated sludge processes. *Journal, Water Pollution Control Federation* **60**:1813–1821, 1988.
36. Norris, D. P., D. S. Parker, M. L. Daniels, and E. L. Owens, Production of high quality trickling filter effluent without tertiary treatment. *Journal, Water Pollution Control Federation* **54**:1087–1098, 1982.
37. NRC Subcommittee Report, Sewage treatment at military installations, Chapter V, Trickling Filters. *Sewage Works Journal* **18**:897–982, 1946.
38. Okey, R. W. and O. E. Albertson, Diffusion's role in regulating rate and masking temperature effects in fixed film nitrification. *Journal, Water Pollution Control Federation* **61**:500–509, 1989.

39. Parker, D. S. and D. T. Merrill, Effect of plastic media on trickling filter performance. *Journal, Water Pollution Control Federation* **56**:955–961, 1984.
40. Parker, D. S. and J. T. Richards, Discussion of 'Process and kinetic analysis of nitrification in coupled trickling filter activated sludge systems'. *Water Environment Research* **66**:934–935, 1994.
41. Parker, D. S. and T. Richards, Nitrification in trickling filters. *Journal, Water Pollution Control Federation* **58**:896–902, 1986.
42. Parker D. S., K. V. Brischke, and R. N. Matasci, Upgrading biological filter effluents using the TF/SC process. *Journal of the Institution of Water and Environmental Management* **7**(1):90–100, 1993.
43. Parker, D., M. Lutz, B. Andersson, and H. Aspegren, Effect of operating variables on nitrification rates in trickling filters. *Water Environment Research* **67**:1111–1118, 1995.
44. Parker, D. S., M. Lutz, R. Dahl, and S. Bernkopf, Enhancing reaction rates in nitrifying trickling filters through biofilm control. *Journal, Water Pollution Control Federation* **61**: 618–631, 1989.
45. Republic of West Germany, Arbeitsblatt A 135, Adwassertechnische Vereinigung *E. V. (ATV)*, **6**, 1980.
46. Richards, T. and D. Reinhardt, Evaluation of plastic media in trickling filters. *Journal, Water Pollution Control Federation* **58**:774–783, 1986.
47. Sampayo, F. F., Performance of nitrification towers at Sidney, Ohio and Lima, Ohio. *Proceedings of the Second International Conference on Fixed-Film Biological Processes*, Arlington, Virginia, pp. 1468–1490, 1984.
48. Sarner, E., *Plastic Packed Trickling Filters*, Ann Arbor Science Publishers, Ann Arbor, Michigan, 1980.
49. Sarner, E., Removal of particulate and dissolved organics in aerobic fixed-film biological processes. *Journal, Water Pollution Control Federation* **58**:165–172, 1986.
50. Schroeder, E. D. and G. Tchobanoglous, Mass transfer limitations on trickling filter design. *Journal, Water Pollution Control Federation* **48**:771–775, 1976.
51. Sedlak, R. I., ed., *Phosphorus and Nitrogen Removal from Municipal Wastewater: Principles and Practice*, Second Edition, Lewis Publishers, Ann Arbor, Michigan, 1991.
52. Stenquist, R. J., D. S. Parker, and J. J. Dosh, Carbon oxidation—nitrification in synthetic media trickling filters. *Journal, Water Pollution Control Federation* **46**:2327–2339, 1974.
53. Suschka, J., Hydraulic performance of percolating biological filters and consideration of oxygen transfer. *Water Research* **21**:865–873, 1987.
54. Tomlinson, T. G. and H. Hall, The effect of periodicity of dosing on the efficiency of percolating filters. *Journal of the Institute of Sewage Purification* 40–47, 1955.
55. U. S. Environmental Protection Agency, *Process Design Manual for Nitrogen Control*, U. S. Environmental Protection Agency, Washington, D.C., 1975.
56. U. S. Environmental Protection Agency, *Assessment of Single-Stage Trickling Filter Nitrification*, 430/9–91–005, U. S. Environmental Protection Agency, Cincinnati, Ohio, 1991.
57. U. S. Environmental Protection Agency, *Nitrogen Control Manual*, EPA/625/R-93/010, U. S. Environmental Protection Agency, Washington, D.C., 1993.
58. Velz, C. J., A basic law for the performance of biological filters. *Sewage Works Journal* **20**:607–617, 1948.
59. Wanner, O. and W. Gujer, Competition in biofilms. *Water Science and Technology* **17**(2/3):27–44, 1984.
60. Water Environment Federation, *Operation of Municipal Wastewater Treatment Plants*, Manual of Practice No. 11, Water Environment Federation, Alexandria, Virginia, 1990.
61. Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992.

62. Water Environment Federation, *Fixed Film Reactors*, Draft Facilities Design Manual, Water Environment Federation, Alexandria, Virginia, 1997.
63. Water Pollution Control Federation, *O & M of Trickling Filters, RBC's, and Related Processes*, Manual of Practice No. OM-10, Water Pollution Control Federation, Alexandria, Virginia, 1988.
64. Williamson, K. L. and P. L. McCarty, A model of substrate utilization by bacterial films. *Journal, Water Pollution Control Federation* **48**:9–24, 1976.

This page intentionally left blank

20
Rotating Biological Contactor

The term rotating biological contactor (RBC) refers to a class of aerobic attached growth bioreactors containing circular shaped corrugated plastic media that are mounted on a horizontal shaft, partially submerged (typically 40%) in the wastewater, and rotated at a speed of one to two revolutions per minute to alternately expose them to the wastewater and to the atmosphere. Figure 20.1 provides a schematic diagram. A number of manufacturers produce RBC equipment, but they are all similar and produce similar results.

Microorganisms grow on the media and metabolize biodegradable organic material and nitrogen-containing compounds in the wastewater. Just as with a trickling filter, the produced biomass will slough off of the media and be transported by the wastewater to a clarifier where it is separated from the treated effluent. When the loading rate of biodegradable organic matter is sufficiently low, nitrifying bacteria will grow on the media and convert ammonia-N to nitrate-N. Wastewater that has been partially treated in another biochemical operation can be applied to an RBC system, where it will be nitrified.

20.1 PROCESS DESCRIPTION

The RBC process is a relatively recent development, although some credit the concept to work conducted in 1901.[32] Development of the current process began in the 1950s, with the first systems being installed in the early 1960s in Germany. Significant refinements in media type and equipment configuration occurred during the 1960s and early 1970s, resulting in our current system. Many facilities were installed in the United States in the 1970s, and over 600 currently exist. It is estimated that approximately 3,000 installations exist worldwide. Many of these facilities have not performed as expected, and the process has fallen into disfavor with many plant owners and operators. However, experience with existing installations has established the appropriate range of applications for the RBC process, as well as the range of operating conditions resulting in acceptable performance. Consequently, practitioners can upgrade and expand existing RBC facilities with confidence and properly evaluate RBCs as an alternative for new applications. A number of publications detail the development of RBC technology and document its appropriate operating range.[3,26,28] In addition, recent design manuals have summarized the current state-of-the-art.[27,31]

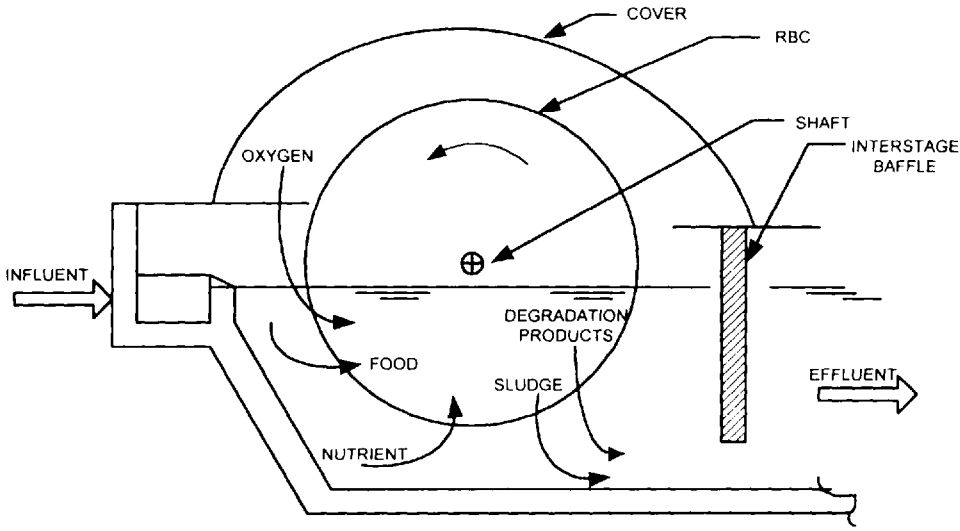


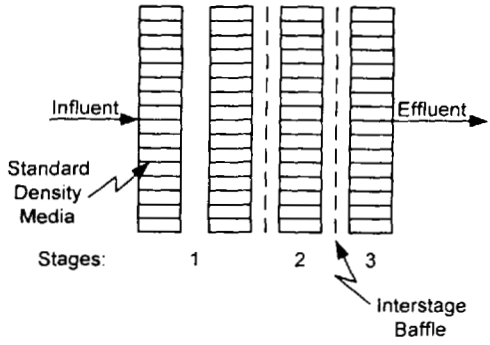
Figure 20.1 Schematic diagram of an RBC.

20.1.1 General Description

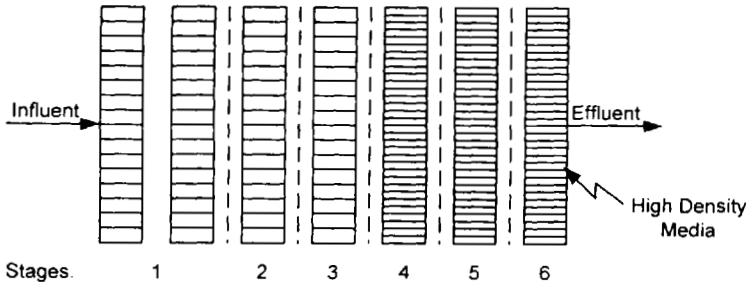
A description of the RBC process is provided in Chapter 17 and immediately above. As indicated in Figure 20.1, a cover (typically fiberglass) is provided over each individual RBC unit for physical protection and process enhancement. Alternatively, entire installations can be placed in buildings, but this can result in a humid, corrosive atmosphere, leading to accelerated corrosion. Covers provide protection against inclement weather, freezing, and sunlight, which accelerates media deterioration. Covers also reduce heat loss, allow the offgas to be collected for odor control, and minimize algae growth.

Nearly all manufacturers produce individual RBC units to standard dimensions. A typical media bundle is 3.66 m in diameter and 7.62 m long, on a shaft that is 8.23 m long. Consequently, at a typical rotational velocity of 1.6 rpm the peripheral velocity of the disc is 18.3 m/min. The media is manufactured from high-density polyethylene containing UV inhibitors such as carbon black. The individual sheets are corrugated in much the same fashion as plastic sheet trickling filter media. Corrugations increase the stiffness of each disc, increase the available surface area, improve mass transfer, and serve to define the spacing between individual disks. Just as with plastic sheet trickling filter media, the size of the corrugations defines how closely together the individual sheets of media can be placed, thereby determining the media density. Standard density media has a specific surface area of about 115 m^2/m^3, so each standard shaft provides 9,300 m^2 of media surface area. High-density media has a specific surface area of about 175 m^2/m^3, providing 13,900 m^2 of media surface area per shaft. Minor density differences occur from one manufacturer to another, so it is possible to purchase media with slightly larger or smaller surface areas per shaft. Like trickling filter installations, the media density used for a particular application is determined by the characteristics of the wastewater being treated and by the treatment objectives.

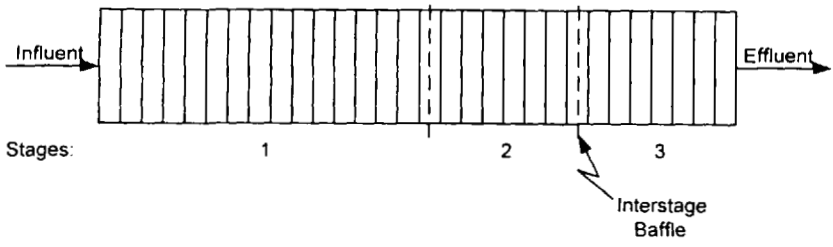
Individual RBC units (called shafts) generally are arranged in series to maximize capacity and treatment efficiency. Baffles are used to separate the RBC shafts into a series of completely mixed bioreactors, each referred to as a stage. The effects of staging on system performance are discussed in Sections 17.4 and 20.2.3. A single stage may contain more than one shaft, but because each stage is completely mixed, all shafts in a stage behave in the same manner. A volume of 45 m^3 per shaft is typically used to size the bioreactor. As illustrated in Figure 20.2a, a typical system for removal of biodegradable organic matter, i.e., secondary treatment, might use



(a)



(b)



(c)

Figure 20.2 Examples of RBC trains.

three stages in series, with the first stage containing two shafts. Such a series of stages is referred to as a treatment train. A mixture of standard and high-density RBC shafts can be used in a single train, although the initial shaft will generally contain standard density media. A larger number of RBC shafts in series will typically be used for advanced treatment applications, i.e., both carbon oxidation and nitrification, as illustrated in Figure 20.2b.

The axis of each individual RBC shaft is typically placed perpendicular to the direction of flow through the train. As indicated in Figure 20.1, the RBC shafts generally rotate in the direction that causes the top of the media to move opposite to the direction of flow. This minimizes short circuiting.

The baffles used to define the individual stages in a treatment train are typically not load bearing and are not capable of isolating an individual RBC shaft. Rather, they are often moveable to allow the number of stages and their sizes to be adjusted in response to long-term variations in process loadings. A typical inter-stage baffle is illustrated in Figure 20.1.

Staging can also be accomplished using a single RBC shaft, as illustrated in Figure 20.2c. The single shaft is placed in a bioreactor with a volume of 45 m^3, with the shaft parallel to the long dimension of the bioreactor. Flow is parallel to the shaft, and inter-stage baffles are placed at various points along the shaft to provide the necessary staging. This arrangement is used in small wastewater treatment plants where only a small number of RBC units is needed.

Individual RBC trains are arranged in parallel with flow split equally to each train, as illustrated in Figure 20.3. Larger wastewater treatment plants will use several trains of parallel shafts. In smaller facilities, the "end flow" configuration illustrated in Figure 20.2c is used for each train. For systems removing organic matter, the effluent from the RBC trains will typically be combined and conveyed to secondary

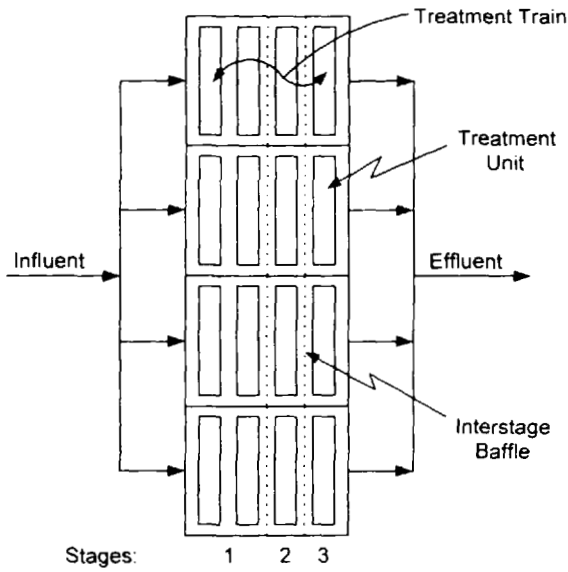


Figure 20.3 Typical configuration of an RBC treatment facility.

clarifiers for the removal of sloughed biomass. Clarification of the RBC effluent may not be necessary for tertiary nitrification applications.

20.1.2 Process Options

Treatment Objectives. RBCs are used to remove biodegradable organic matter and convert ammonia-N and organic-N to nitrate-N. As discussed in Section 20.2.1, operational problems caused by high unit organic loading rates restrict the use of RBCs for partial removal of organic matter, i.e., for "roughing" treatment. However they can be used quite effectively for substantial removal of organic matter. Process effluent, i.e., clarified, five day biochemical oxygen demand (BOD_5) and total suspended solids (TSS) concentrations can easily be reduced to less than 30 mg/L each, and even lower concentrations can be obtained in some instances. This degree of treatment can be accomplished by applying proper organic and hydraulic loading rates, as discussed below.

Combined carbon oxidation and nitrification can also be accomplished in an RBC system. As discussed in Section 15.4, heterotrophic and autotrophic bacteria compete for space within the aerobic portion of a biofilm, causing heterotrophs to predominate when both organic substrate and ammonia-N concentrations are high. Consequently, the oxidation of organic matter will generally occur in the initial stages of the RBC train, just as it occurs in the top of a trickling filter. However, if the organic loading on the train is sufficiently low, the organic substrate concentration will be reduced sufficiently so that autotrophs will be able to compete in the latter stages. As with other aerobic fixed film processes, this occurs when the soluble substrate concentration is reduced to about 20 mg/L as chemical oxygen demand (COD) (15 mg/L as BOD_5).[26,27,31] Thus, the primary distinction between a secondary treatment application (the removal of organic substrate alone) and a combined carbon oxidation and nitrification application (the removal of organic substrate and the oxidation of ammonia-N to nitrate-N) is the organic loading. A larger number of stages may be used for combined carbon oxidation and nitrification to increase the degree of staging and separate the heterotrophic and autotrophic bacteria.

Rotating biological contactors can also be used for separate stage nitrification; that is, to nitrify streams containing relatively high concentrations of ammonia-N and low concentrations of organic matter. Such applications may not require downstream clarification because of the low biomass production rates associated with nitrification. Separate stage nitrification applications are distinguished from combined carbon oxidation and nitrification applications by the characteristics of the wastewater being treated. If the concentration of organic substrate in the influent wastewater is low, relative to the concentration of ammonia-N, the biofilm that develops will be enriched in nitrifiers and the impact of the organic matter on process sizing will be negligible. Benchmarks for distinguishing a separate stage nitrification application for domestic wastewater treatment are an influent BOD_5/TKN ratio less than about 1.0 and/or an influent soluble BOD_5 concentration less than about 15 mg/L.

Rotating biological contactors have also been used to accomplish denitrification. In these applications, the RBC unit is entirely submerged and covers are provided to exclude air. The influent is generally a nitrified secondary effluent, so an

electron donor must be provided. These applications are quite limited and will not be discussed further.

Equipment Type. A motive force is necessary to rotate the RBC shaft. Two general approaches are used: mechanical drives and air drives. Mechanical drive systems consist of an electric motor, a speed reducer, and a belt or chain drive for each shaft. The electric motors are typically 3.7 or 5.6 kW, and the RBC rotational speed is typically 1.2 to 1.6 rpm. The capability to adjust speed and rotational direction can be provided by using speed reducers with multiple pulley or sprocket ratios or through a variable speed drive. These features can be used to control the buildup of excess biomass.

Air drive units increase the oxygen transfer capacity of an individual RBC unit and reduce the number of electric motors required. Cups are added to the periphery of the media and oriented to collect air injected under the RBC shaft. The cups are either 10 or 15 cm long, depending on the organic loading to the unit. The air flow per shaft ranges from 4.2 to 11.3 m^3/min under standard conditions, which is typically sufficient to provide rotational speeds of 1.0 to 1.4 rpm. Air is generally provided to all shafts by a centralized blower system. The quantity of air required varies depending on the specific configuration and operating conditions.

Mechanically driven systems provide reliable, consistent rotation of the RBC shaft and media. However, they are susceptible to biomass buildup when they are organically overloaded or subjected to high sulfide loading, as discussed in Section 20.2.6. Air drive systems provide enhanced oxygen transfer, and the injected air can assist with the removal of excess biomass. Both of these effects can be beneficial in a heavily loaded unit. The primary disadvantage of air drive systems is that they are more susceptible to loping, which is uneven rotation caused by the development of nonuniform biomass growth around the circumference of the RBC media. Uneven rotational speed results as the heavier portion of the disc is lifted out of the liquid, rotated to the top, and allowed to descend by gravity back into the liquid.

A recent innovation is the submerged biological contactor, within which between 70 and 90% of the media is submerged. They are generally aerated. Claimed advantages include reduced structural loadings on the shaft and bearings, improved biomass control, the ability to use larger media bundles, and increased treatment capacity. To date these systems have received limited use.

20.1.3 Comparison of Process Options

Table 20.1 summarizes the primary benefits and drawbacks of the RBC process. It is mechanically simple, which simplifies normal equipment maintenance. It is also an uncomplicated process, thereby lessening the need for intensive daily process control actions. The energy requirements are relatively low, being only those required to rotate the media. Finally, it is modular in nature, which simplifies design and construction.

Its principal drawbacks are that process performance is sensitive to wastewater characteristics and loadings, and that it possesses little operational flexibility to adjust to varying loading and operating conditions. As discussed in Section 20.2.1, high organic loadings can result in excessive biomass growth, which structurally overloads the media and shaft. This problem is exacerbated by elevated levels of hydrogen sulfide in the influent wastewater. Although significant deterioration in treatment

Table 20.1 RBC Process Benefits and Drawbacks

Benefits	Drawbacks
Mechanically simple	Performance susceptible to wastewater characteristics
Simple process, easy to operate	Limited process flexibility
Low energy requirements	Limited ability to scale-up
Modular configuration allows easy construction and expansion	Adequate pretreatment required

capacity and performance result from excessive biomass, the steps that can be taken to minimize its impact are relatively limited. Fortunately, the conditions leading to such cataclysmic declines in performance are now relatively well defined and can generally be avoided if the operating conditions for the facility are well defined.

An early claimed benefit of the RBC process was its ability to respond successfully to shock loads.[2] However, subsequent experience has demonstrated that its capability to respond to shock loads is much like that of the trickling filter process. Shock loads will not cause complete process failure, but the process does not generally possess sufficient reserve capacity to successfully treat shock loads.[9,11] Consequently, equalization should be provided upstream of an RBC process if the ratio of peak to average loading exceeds 2.5.[27,31]

The size of an individual RBC shaft limits the maximum plant size for which the RBC process is practical. As discussed above, each shaft can contain a media surface area of 9,300 or 13,900 m^2. Consequently, only for small to medium wastewater flow rates can sufficient media be provided by a reasonable number of RBC units. For example, a total media surface area of about 280,000 m^2 might be required to treat a typical municipal wastewater with a design flow of 15,000 m^3/day. This could be provided by 24 RBC shafts configured in six trains consisting of four shafts each. An equivalent media surface area would be provided by two trickling filters, each 16.4 m in diameter and 6.7 m deep. To treat a flow rate ten times as large, the number of RBC units required would increase ten-fold; in this case to 60 trains, each with 4 individual RBC shafts. In contrast, the equivalent trickling filter installation would still require only two trickling filters, although each would have to be 51.9 m in diameter and 6.7 m deep. Alternatively, four trickling filters, each 36.7 m in diameter and 6.7 m deep, could be used. From a cost, construction, and operational perspective, the smaller number of trickling filters would be more desirable. This factor tends to limit the use of RBCs to smaller wastewater treatment plants.

A final drawback of RBCs is the need for adequate preliminary treatment. Debris such as rags, plastics, and fibrous material can clog the RBC media if present in sufficient quantities, and grit will settle in the RBC bioreactor due to the relatively low level of turbulence provided. In general, the minimum degree of preliminary treatment required is fine screening (less than 1 mm opening) and excellent grit

removal. Primary clarification is provided in many instances. The cost of the necessary preliminary treatment facilities must be included in any cost evaluation of the RBC process.

The biomass produced in the RBC process generally settles and thickens readily. Consequently, the waste solids stream leaving the final clarifier can either be recycled to the primary clarifier to be settled and thickened with the primary solids, or it can be thickened separately. Other solids thickening options can also be applied successfully.

During the early years of their application, RBC systems experienced a number of mechanical and structural problems, such as detachment of the media from the shaft and/or fatigue and failure of the shafts. The causes for these problems are now well understood by RBC manufacturers, and the expertise exists to manufacture equipment that is devoid of these defects. Thus, while numerous references can be found to these structural problems in the literature, they are no longer a major factor in the evaluation and selection of RBC systems.

20.1.4 Typical Applications

Rotating biological contactors have typically been used to provide secondary treatment to municipal wastewater. They have also been used to nitrify municipal wastewaters, either in combined carbon oxidation and nitrification applications or in separate stage nitrification applications. Current estimates are that approximately 70% of the applications in the United States have been for removal of organic matter, 25% for combined carbon oxidation and nitrification, and 5% for separate stage nitrification.[13,32] Due to the poor economy of scale for this technology, it has been used most frequently for wastewater treatment plants with flows below about 40,000 m^3/day. It has also been used successfully in a number of industrial applications, particularly those involving wastewaters of moderate- to low-strength and with low concentrations of hydrogen sulfide.

The reliability of RBCs has improved considerably in recent years. While many RBC installations have provided acceptable performance, many others have not met performance expectations. For example, a survey indicated that over 80% of the RBCs designed before 1980 have experienced operational problems.[32] Many of these problems have been solved by improved construction techniques and the use of appropriate organic and hydraulic loading rates. Consequently, RBC technology is now sufficiently well defined so that it is possible to clearly evaluate existing facilities for upgrades and new applications. In fact, new approaches are being investigated for utilizing existing RBC facilities to accomplish higher levels of treatment.[15] Those interested in application of attached growth biochemical operations should monitor these developments for potential application elsewhere.

20.2 FACTORS AFFECTING PERFORMANCE

Many factors affect the performance of RBC wastewater treatment systems. This section emphasizes those factors that are particularly significant and/or are relatively unique to the RBC process.

20.2.1 Organic Loading

The performance of an RBC facility is significantly affected by the organic loading.[10,19,21,22,34] The organic loading on an RBC is typically expressed on the basis of the total media surface area, A_{st}, not just the wetted or submerged areas. Consequently, it is equivalent to the surface organic loading (SOL) as used in trickling filter design. The definition of the SOL is given by Eq. 19.2. Noting that in this case the area is the total media surface area, the SOL on an RBC is defined as:

$$\lambda_S = \frac{F(S_{SO} + X_{SO})}{A_{st}} \tag{20.1}$$

All terms in the equation are as previously defined. Typical units for SOL are g COD/(m^2·day) or g BOD_5/(m^2·day). Figure 20.4 illustrates the typical relationship between the SOL and the removal rate for readily biodegradable organic matter for a variety of full-scale RBC installations. The removal rate generally increases as the SOL increases, but at a decreasing rate. As a result, the substrate removal efficiency decreases as the SOL increases. Surface organic loading values can be calculated for an individual RBC stage or for the entire RBC treatment system. In the former case, the organic matter concentration is that entering the particular stage and A_{st} is the area of media in the stage. In the latter case, the organic matter concentration is that in the process influent and A_{st} is the media surface area for the entire system.

The organic loading cannot be increased indefinitely, as might be suggested by Figure 20.4. Rather, it is limited by the maximum oxygen transfer capacity of an individual RBC shaft. Analysis of data from full-scale RBC facilities indicates that oxygen transfer limitations occur at SOL values to individual RBC shafts of about 32 g BOD_5/(m^2·day). This value is commonly taken to correspond to a soluble BOD_5 SOL of 12 to 20 g/(m^2·day), which corresponds to a soluble biodegradable COD SOL of 20 to 35 g COD/(m^2·day).[20,26-28,31] Because of oxygen limitations, excessive growth of the nuisance organism *Beggiatoa* occurs at SOLs in excess of these values. The mechanisms for *Beggiatoa* growth and its impact on the performance of RBC

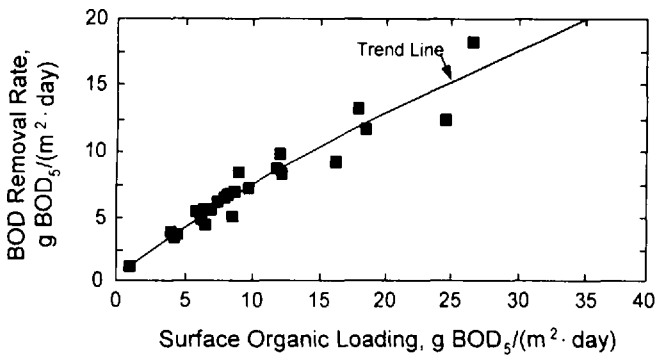


Figure 20.4 Effect of surface organic loading (SOL) on the surficial BOD_5 removal rate (flux) for full-scale RBC facilities treating domestic wastewater. (From Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992. Copyright © Water Environment Federation; reprinted with permission.)

systems are discussed in Section 20.2.6. Standard density RBC media (9,300 m^2/shaft) should be used in RBC stages that are highly loaded or where *Beggiatoa* growth is possible. Excess biological growths are more difficult to remove from high-density media and, consequently, use of high-density media can further exacerbate operational problems in highly loaded stages.

The organic loading affects nitrification in an RBC system. As discussed above, nitrifying bacteria can effectively compete for space in a biofilm once the concentration of soluble organic substrate is reduced below 20 mg COD/L (15 mg BOD_5/L). The SOL, among other factors, determines whether that occurs. An empirical relationship has been developed to show the effect of SOL (expressed as g total biodegradable COD/(m^2 · day)) on the nitrification rate that can develop in an RBC unit performing combined carbon oxidation and nitrification:[18]

$$f_{NH} = 1.43 - 0.1\lambda_S; \quad 4.3 < \lambda_S < 14.3 \tag{20.2}$$

where f_{NH} is fraction of the rate that would occur in the absence of simultaneous carbon oxidation. The limits on f_{NH} are one and zero. This equation indicates that no nitrification will occur when the SOL $\geq$ 14.3 g total biodegradable COD/(m^2 · day) and that unrestricted nitrification will occur at SOLs of 4.3 g total biodegradable COD/(m^2 · day) or less. At all SOL values between those extremes, significant competition occurs between heterotrophs and autotrophs, resulting in reduced nitrification rates.

For separate stage nitrification systems the classical relationship between the growth limiting ammonia-N concentration and the flux into the biofilm, J_{NH}, as described by Eq. 19.5, is observed. According to Eq. 20.2 this will occur when the loading of biodegradable COD is below 4.3 g COD/(m^2 · day). Figure 20.5 presents the relationship. As indicated there, at ammonia-N concentrations above about 5 mg/L nitrification proceeds in RBC units at a rate of about 1.5 g N/(m^2 · day). This represents a zero order biofilm in which the nitrification rate is not limited by the

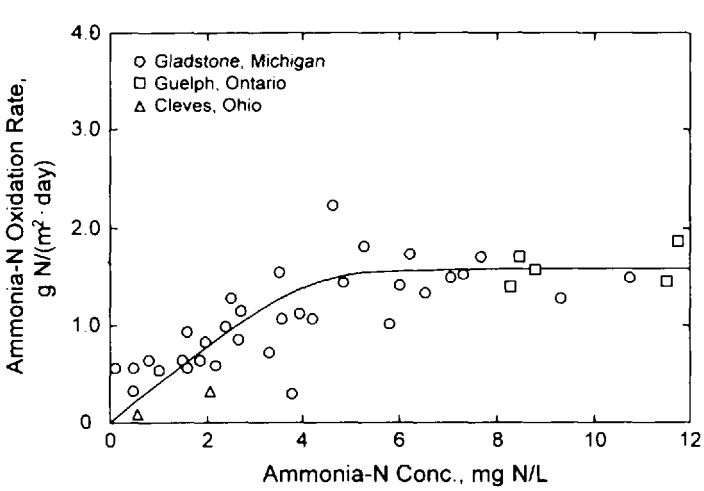


Figure 20.5 Effect of the ammonia-N concentration in an RBC stage on the ammonia-N oxidation rate in the absence of significant carbon oxidation. (Adapted from *Design Information on Rotating Biological Contactors.*[20])

bulk ammonia-N concentration. Rather, it is limited by the rate of oxygen transfer to the biofilm. As the bulk ammonia-N concentration decreases below 5 mg/L, it begins to become rate limiting. Examination of Figure 20.5 suggests that $K_{s,NH}$ for RBC biofilms is approximately 2 mg N/L.

During combined carbon oxidation and nitrification, the rate of nitrification must be obtained by multiplying the rate associated with the ammonia-N concentration, as given by Figure 20.5 or Eq. 19.5, by the value of f_{NH} obtained with Eq. 20.2. The use of these relationships is illustrated in the example that follows.

Example 20.2.1.1

One stage of an RBC system treating domestic wastewater contains a single shaft of high-density media (A_s = 13,900 m^2) and receives 2,000 m^3/day of flow containing 50 mg/L of biodegradable COD and 10 mg/L of ammonia-N. Determine the concentration of ammonia-N in the effluent from the stage. Assume that the curve in Figure 20.5 represents the observed effect of the ammonia-N concentration on the nitrification rate.

a. Will nitrification occur in the stage?
Two factors determine whether nitrification will occur in an RBC, the soluble COD in the stage and the SOL applied to it. The SOL must be less than 14.3 g COD/(m^2 · day). The SOL can be calculated with Eq. 20.1:

$$\lambda_S = \frac{(2{,}000)(50)}{13{,}900} = 7.19 \text{ g COD/}(m^2 \cdot \text{day})$$

Since the SOL is less than 14.3 g COD/(m^2 · day), nitrification will occur, provided the soluble COD in the stage is below 20 mg/L. Examination of Figure 20.4 shows the relationship between the SOL and the organic matter removal rate per unit area in an RBC. In that figure, the SOL is expressed in terms of BOD_5. Equation 8.33 shows that the biodegradable COD in domestic wastewater is 1.71 times the BOD_5. Thus an SOL of 14.3 g COD/(m^2 · day) is equivalent to an SOL of 8.36 g BOD_5/(m^2 · day). From Figure 20.4 it can be seen that an RBC receiving such a loading will have a organic removal rate around 6 g BOD_5/(m^2 · day), for a removal efficiency of approximately 73%. This suggests that the total biodegradable COD leaving the stage will be less than 15 mg/L, making the soluble COD well below that value. Thus, we can expect nitrification to occur in the stage.

b. At what fraction of the unrestricted nitrification rate will nitrification occur in the RBC?
The fractional nitrification rate can be estimated with Eq. 20.2 using the value of the SOL calculated in Part a:

$$f_{NH} = 1.43 - (0.1)(7.19) = 0.71$$

Thus, nitrification will occur in the RBC at 71% of the rate at which it would occur in an RBC performing only nitrification.

c. What is the concentration of ammonia-N leaving the RBC stage?
The effect of the ammonia-N concentration on the nitrification rate is shown in Figure 20.5. However, since the ammonia-N concentration in the RBC stage depends on the removal rate in the stage, and no analytical expression

is available relating the rate and the concentration, an iterative procedure must be used. Start by assuming that the concentration is above 5 mg/L so that the rate is independent of the concentration. In that case, the rate given by the figure is 1.5 g NH_3-N/(m^2 · day). The rate in the RBC will be only 71% of that value, however, because of the carbon oxidation occurring. Thus, the rate is 1.07 g NH_3-N/(m^2 · day). The effluent concentration resulting from that rate must be calculated from a mass balance on the stage and compared against the assumed value. If they do not agree, a new value must be assumed and the process repeated. Performing the mass balance:

Input rate = (2,000)(10) = 20,000 g/day

Removal rate = (1.07)(13,900) = 14,800 g/day

Output rate = 20,000 − 14,800 = 5,200 g/day

Effluent concentration = 5,200 ÷ 2,000 = 2.6 mg N/L

This concentration is lower than the assumed value. Thus, the removal rate will be lower than was assumed, which suggests that the concentration will be higher than 2.6 mg/L. For the second iteration, assume a concentration between the two previous values. Use the average of the two, which is 3.8 mg/L. Entering Figure 2.5 with that concentration, the unrestricted rate is 1.24 g NH_3-N/(m^2 · day). The rate in the presence of carbon oxidation is 71% of that value, or 0.88 g NH_3-N/(m^2 · day). Repeating the mass balance results in an effluent ammonia-N concentration of 3.9 mg/L. This is sufficiently close to the assumed value, so the effluent ammonia-N concentration can be considered to be around 3.8 to 3.9 mg/L.

20.2.2 Hydraulic Loading

The total hydraulic loading (THL) for an RBC system is defined slightly differently than for a trickling filter. It is defined with respect to the media surface area and is calculated as follows:

$$A_{H,RBC} = \frac{F}{A_s} \tag{20.3}$$

Comparison of Eqs. 20.1 and 20.3 indicates that the SOL and the THL are related to one another by the pollutant concentration in the influent wastewater.

The performance of RBC systems has historically been correlated with the THL. Manufacturers of RBC equipment have developed and published performance curves for domestic wastewater in which effluent quality is plotted as a function of the THL. The general shape of the effluent quality versus THL relationship is independent of the influent wastewater strength, but the quantitative relationship varies with it. Figure 20.6 presents a typical relationship; this particular one was developed by the Autotrol Corporation, which popularized the RBC process in the United States. Experience has indicated that the relationships developed by some of the RBC manufacturers accurately predict the performance trends observed by full-scale RBC facilities.[29,31] Thus, this general approach is a useful one for characterizing full-scale RBC facility performance. However, this same experience also indicates that the performance relationships published by many of the RBC equipment manufacturers provide optimistic estimates of facility performance.[29,31] Consequently, care should be exercised in the selection and application of such empirical relationships.

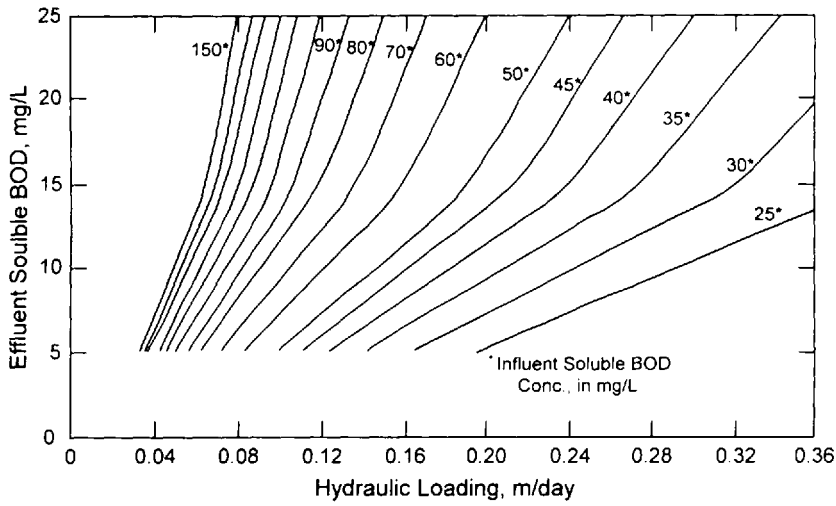


Figure 20.6 Typical design correlation for soluble BOD_5 removal in municipal wastewater treatment.

20.2.3 Staging

As discussed above and illustrated in Section 17.4, RBC facilities are typically staged to improve overall performance. In a staged system, the SOL for the initial stage is higher than the value for the entire system. As indicated in Figure 20.4, increased SOL values result in an increased removal rate. Consequently, by staging the bioreactor, the overall average removal rate can be increased, even though the SOL on the last stage is relatively low. Since the process effluent quality is determined by the SOL on the last stage, a good-quality effluent can be produced even with a relatively high overall average SOL. The principal constraint is that the SOL in any stage must not exceed the oxygen transfer capacity of a shaft, 32 g $BOD_5/(m^2 \cdot day)$, as discussed in Section 20.2.1. The performance benefits of staging RBCs were demonstrated early in the development of this technology.[4]

Staging is particularly effective in systems that are required to achieve combined carbon oxidation and nitrification. Staging in the upstream portion of the train produces a high SOL, resulting in a high rate of organic matter removal. This results in reduced SOLs in subsequent stages and concentrations of soluble biodegradable organic matter less than 20 mg/L as COD. This, in turn, allows nitrifying bacteria to become established in the later stages of the RBC train. As indicated in Figure 20.5, nitrification is first-order for low ammonia-N concentrations, but it is zero-order for ammonia-N concentrations greater than 5 mg/L. Staging does not increase the ammonia-N flux into the biofilm when the ammonia-N concentration exceeds 5 mg/L, but it does increase the overall flux when the ammonia-N concentration is less than 5 mg/L. Table 20.2 summarizes recommendations from several sources concerning the number of stages for particular applications.[27,31,32]

Staging also affects the nature of the biomass that develops on individual RBC shafts and their resulting pollutant removal capability. This is illustrated most graphically in an RBC system that is used for combined carbon oxidation and nitrification.

Table 20.2 Summary of Staging Recommendations

Carbon oxidation		Nitrification	
Effluent soluble BOD_5 conc. mg/L	Number of stages	Effluent NH_3-N conc. mg/L	Number of stages
> 25	1	> 5	1
15–25	1 or 2	< 5	Based on kinetics
10–15	2 or 3		
< 10	3 or 4		

As indicated in the preceding paragraph, in the initial stages the biofilm will consist primarily of heterotrophs, making it very active and capable of high organic matter fluxes. The later stages, however, will contain biofilms with increasing proportions of nitrifying bacteria. As a result, the organic removal capability of these later stages will be reduced, although the nitrification capability will be enhanced.

20.2.4 Temperature

The effect of temperature on the performance of RBC systems is similar to that of other attached growth processes. Since the reaction rate is influenced strongly by diffusion, wastewater temperature has little effect on process performance over a wide range of temperatures. The effect of temperature is generally neglected for temperatures over about 15°C, although removal rates generally decline for lower temperatures.[3,27,31] Temperature relationships have been presented in two ways. Figure 20.7 provides one relationship. To use it, first determine the media surface area required for the particular application at a temperature above 55°F (13°C). Then enter

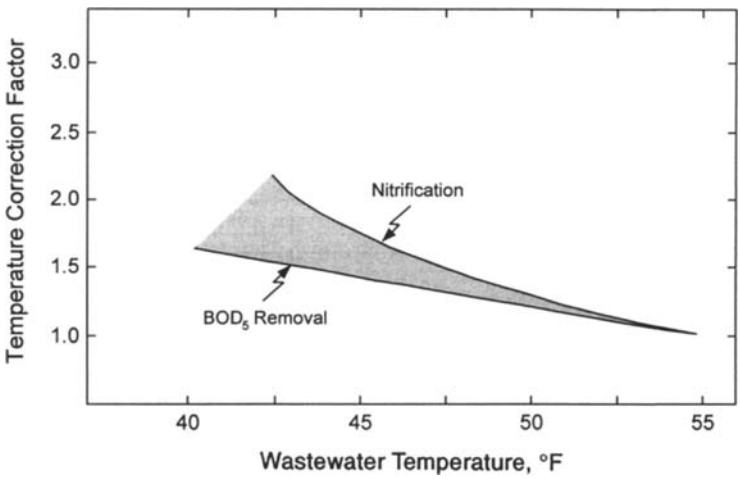


Figure 20.7 Effect of temperature on the area requirements for RBCs. (Adapted from *Nitrogen Control*.[27])

Table 20.3 Summary of Manufacturer's Temperature Correction Recommendations

Temp °C	Manufacturer		
	Envirex	LYCO	Walker process
	Temperature correction factor: BOD_5 removal		
17.8	1.00	1.00	1.00
12.8	1.00	1.00	1.00
10.0	0.87	0.83	0.87
7.2	0.76	0.71	0.73
5.5	0.67	0.66	0.65
4.4	0.65	0.62	—
	Temperature correction factor: nitrification		
17.8	1.00	1.00	1.39
12.8	1.00	1.00	1.00
10.0	0.78	0.78	0.78
7.2	0.48	0.57	0.56
5.5	0.42	—	0.43
4.4	—	—	—

Figure 20.7 and select the multiplier for the actual operating temperature. Finally, multiply the area by the multiplier to determine the appropriate area at the actual operating temperature. Table 20.3 provides three other relationships from the literature.[29] To use Table 20.3, determine expected biofilm flux, neglecting the impacts of temperature. Then enter the table to determine the correction coefficient for the expected temperature. Multiply the expected flux by that coefficient to obtain the design value, which can then be used to determine the required media area.

20.2.5 Wastewater Characteristics

The characteristics of the wastewater being treated will affect the performance of an RBC system in several ways, just as it will any other biochemical operation. Several of these effects were discussed in Chapter 15. For example, the flux into the biofilm may be smaller for large, slowly biodegradable compounds than for small, readily biodegradable compounds. The presence of particulate organic matter may reduce the flux of soluble substrate since the particulate matter occupies space within the biofilm, reducing the biomass concentration within the biofilm, which decreases the rate of biodegradation. Hydrolysis of entrapped particulate organic matter releases soluble organic matter, which reduces the diffusion of soluble organic matter from the bulk liquid into the biofilm.

Another wastewater constituent that has a particularly important impact on the performance of RBC systems is hydrogen sulfide. As discussed in the next section, growth of the sulfide oxidizing bacterium *Beggiatoa* can cause operating problems in an RBC system. The growth of these microorganisms is encouraged by elevated levels of hydrogen sulfide in the influent wastewater.

20.2.6 Biofilm Characteristics

The character of the biofilm that develops on an RBC can significantly affect its performance. This, in turn, is affected by the operating conditions imposed on each RBC shaft. In the initial stages of an RBC train, where the removal of organic matter occurs, a normal biofilm is grey-brown in color and of a stable and controllable thickness.[3] It is composed primarily of aerobic heterotrophic bacteria, entrapped particulate organic matter, and Eucarya. If the overall process organic loading is sufficiently low, a nitrifying biofilm will develop in the later stages. A nitrifying biofilm is generally thinner than a heterotrophic biofilm and reddish-brown in color.[23] In separate stage nitrification systems, the biofilm will be highly enriched in nitrifiers, and may be tan or bronze in color.

In some instances a tenacious whitish biofilm develops that will not slough off of the RBC media.[5,26,27,31] Since it does not slough off, it continues to build up, which can lead to structural overload and physical failure of the RBC unit. Failure occurs either in the media or the shaft. Media can fail directly, or it can detach from the rotating shaft. The result, in any case, is an inoperative unit. The tenacious biofilm develops from growth of the filamentous bacterium *Beggiatoa*. The filamentous nature of this microorganism reinforces the RBC biofilm, thereby increasing its resistance to hydraulic shear and decreasing sloughing rates. The whitish color of the biofilm is a result of the color of the *Beggiatoa*.

Beggiatoa obtains energy by oxidizing hydrogen sulfide and other compounds containing reduced sulfur. When sufficient quantities of hydrogen sulfide are present in the system, either from the influent wastewater or produced internally, *Beggiatoa* can compete effectively with heterotrophic bacteria for space within the aerobic biofilm. Hydrogen sulfide will be produced within the biofilm when the organic loading on individual RBC shafts exceeds 32 g $BOD_5/(m^2 \cdot day)$, which corresponds to the maximum oxygen transfer capacity, as described previously. Anaerobic conditions in the lower portion of the biofilm allow sulfate reducing bacteria to use sulfate as a terminal electron acceptor, resulting in hydrogen sulfide production. The hydrogen sulfide diffuses outward into the overlying aerobic layer where it is used as an energy source by *Beggiatoa*.

The nature of the wastewater collection system can aggravate *Beggiatoa* problems. For example, if the collection system consists of a series of force mains, septic conditions can result in both hydrogen sulfide production and solubilization of biodegradable organic matter. The resulting increased concentration of readily biodegradable organic matter can elevate the SOL in the initial stage of the RBC train, resulting in anaerobiosis and even more hydrogen sulfide production. The location of such a system in a warm climate will make the problem worse.

Several approaches are available to deal with excessive *Beggiatoa* growth. One is to eliminate the conditions that cause it. The influent hydrogen sulfide concentration can be reduced by chemical oxidation with chlorine, hydrogen peroxide, or dissolved oxygen; precipitation with iron; or alteration of the operation of the collection system. The reader is referred to a recent Manual of Practice[30] for more information. Another is to reduce the SOL on an overloaded stage, either by reducing the entire process SOL or by redistributing the influent load. Methods for reducing the overall SOL include placing more RBC trains in service and increasing the removal of organic matter in upstream unit operations, such as, by chemically en-

hancing the performance of a primary clarifier. Methods for redistributing the influent loading include changing the process staging, using effluent recirculation, or step-feeding load to downstream stages. Finally, a number of approaches are available for physically or chemically removing excess biomass from individual RBC shafts. Examples include oxidizing the biomass with chlorine, elevating the pH with caustic, scouring with air, changing the shear characteristics by reversing the rotational direction, recirculating effluent, and turning off and drying out individual shafts.[24,26,27,31]

Several approaches are available for monitoring the buildup of excessive biomass on individual RBC units. Visual inspection can be effective because of the distinctive appearance of *Beggiatoa* growth. Individual RBC shafts can also be equipped with load cells to periodically weigh them to detect increases in biofilm mass. This latter approach is particularly useful when excessive *Beggiatoa* growth occurs on a persistent basis. Regular measurements can be made, allowing periodic operational adjustments. Several references provide further details.[24,26,27,31]

20.3 PROCESS DESIGN

The complex nature of the events occurring in RBC systems is discussed in Chapter 17. Because of that complexity, fundamental design models for the RBC process are still in the developmental stage. Consequently, empirical approaches are currently used for design. Most of those approaches express the SOL in terms of BOD_5 so we will do the same herein. This section describes the design procedures for RBCs that remove organic matter and nitrify, both separately and together. The use of pilot plants to develop site-specific design data is also discussed. Some general comments on RBC design procedures are also provided.

20.3.1 Removal of Biodegradable Organic Matter

General Approach. The general approach to the design of an RBC system to remove biodegradable organic matter consists of the following steps:

- Select a design expression.
- Select an effluent quality goal. As discussed in Section 9.4.4, the selection of that goal should consider uncertainty and variability in process performance.
- Use the design expression and the effluent quality goal to calculate the total media surface area required.
- Determine the media surface area required in the first stage to keep the stage SOL below 32 g $BOD_5/(m^2 \cdot day)$ to prevent excessive growth of *Beggiatoa*.
- Select the number of trains to be used, the number stages in each train, and the number of shafts in each stage.

Several empirical design approaches have been presented in recent design manuals.[27,31] This section presents two equations which have been found to most accurately characterize the performance of full-scale systems treating domestic wastewaters, the first-order and the second-order models. Both can be used to estimate the total media area required for domestic wastewaters. The design of RBC systems

to treat industrial wastewaters generally requires full-scale experience with the same or a similar wastewater, or a pilot study.

After the total media surface area and the area in the first stage have been determined, engineering experience and judgement must be used to configure the system. Generally, a minimum of four trains is desirable from an operational perspective because when one train is out of service for maintenance, three-quarters of the total media volume will remain in service. This will generally be sufficient to produce an acceptable quality effluent in the short term. Guidance concerning the number of stages is provided in Table 20.2. Even though the total media volume may have been selected to give the desired effluent substrate concentration, staging that media as recommended in the table will provide a factor of safety in the design. Finally, the number of shafts in the first stage is determined by the minimum area required to prevent oxygen limitations. The number of shafts in the remaining stages can be selected based on the experience of the designer and other considerations. However, at no time should the loading on any stage exceed 32 g $BOD_5/(m^2 \cdot day)$.

First-Order Model. The first-order model is analogous to the Velz/Germain equation used to design trickling filters. It was first presented by Benjes,[5,25] as follows:

$$S_{Se}/S_{SO} = \exp\left[-k_1\left(\frac{V_M}{F}\right)^{0.5}\right] \tag{20.4}$$

where S_{Se} and S_{SO} are the concentrations of total biodegradable organic matter in the clarified process effluent and influent, respectively, V_M is the media volume, F is the influent flow rate, and k_1 is a first-order reaction rate coefficient. It should be noted that even though the first-order model is similar to the Velz/Germain equation, it is based on concentration of total organic matter entering and leaving the process rather than on the soluble organic matter as was done in Eq. 19.6. Based on a review of operating data from 27 full-scale municipal wastewater treatment plants, a value for k_1 of 0.3 was selected when S_{Se} and S_{SO} are measured as BOD_5, V_M is expressed in ft^3, and F is expressed in gallons per minute.[5,25] Figure 20.8[31] compares the predictions of this equation with the results from the plants and indicates a generally good fit. Note that the plot was prepared with $100[(S_{SO} - S_{Se})/S_{SO}]$ as the ordinate and the SOL as the abscissa. Because of the nature of the abscissa, a separate curve results from Eq. 20.4 for each influent substrate concentration and the plant data have been grouped into three sets to show that effect. Nevertheless, the data scatter indicates that some facilities may perform less efficiently than indicated by the equation. As a consequence, some design manuals suggest the use of a more conservative k_1 value in the range of 0.2 to 0.25.[29,31]

Because Eq. 20.4 was developed using standard density RBC shafts that contain media with a specific surface area of 35 ft^2/ft^3, i.e., 115 m^2/m^3, it can be converted into an expression based on media surface area. Using metric units and a value for k_1 of 0.3, the expression becomes:

$$S_{Se}/S_{SO} = \exp\left[-0.4\left(\frac{A_M}{F}\right)^{0.5}\right] \tag{20.5}$$

where A_M is the total media surface area expressed in m^2 and F is the wastewater flow rate in m^3/day. Even though Eq. 20.5 was developed for standard density media,

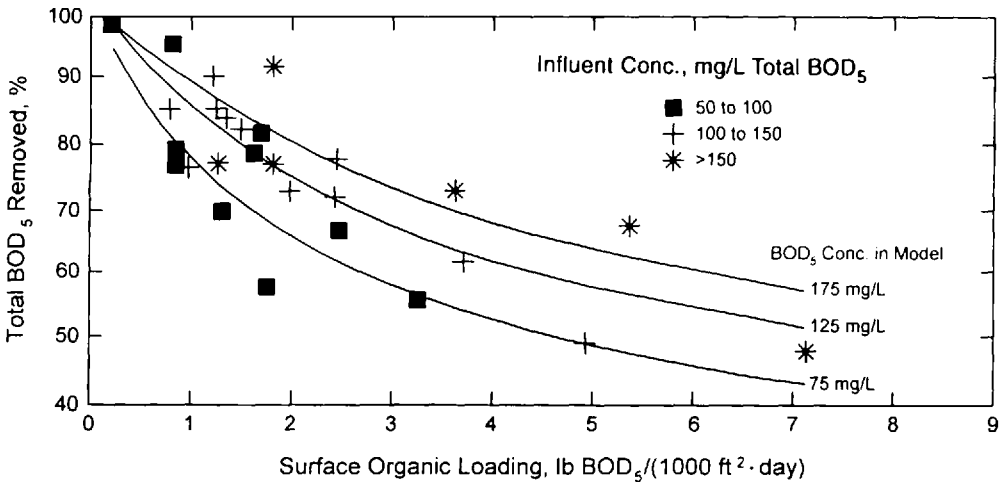


Figure 20.8 Comparison of predictions from first order model (Eq. 20.4, k_1 = 0.3) with data from 27 full-scale RBC installations. The curves are the model predictions and the points represent the data. (From Water Environment Federation, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992. Copyright © Water Environment Federation; reprinted with permission.)

the important outcome from its use is the total media surface area required. Consequently, once the required surface area has been determined, it can be achieved with a combination of standard and high-density media, as long as standard-density media is used in the most heavily loaded stages. Use of the first-order model, as expressed in Eq 20.5, to design an RBC for the removal of biodegradable organic matter is illustrated in the following example.

Example 20.3.1.1

Use the first-order equation of Benjes[5,25] to design an RBC to treat a wastewater with an average flow of 8,000 m^3/day and a BOD_5 concentration (after primary clarification) of 120 mg/L. The wastewater is domestic in origin. Secondary treatment is required, so the BOD_5 and TSS concentrations should not exceed 30 mg/L on a monthly average basis. A robust system capable of reliably meeting this requirement is desired.

a. What design effluent quality should be used?
Using the procedure of Roper et al., as discussed in Section 9.4.4, a design BOD_5 concentration of 20 mg/L is appropriate to allow a facility to reliably comply with a monthly average effluent BOD_5 limit of 30 mg/L.

b. What media surface area is required to produce the selected effluent quality?
The required media surface area can be calculated with a rearranged form of Eq 20.5.

$$A_M = F\left[\frac{-\ln(S_{SE}/S_{SO})}{0.4}\right]^2$$

$$A_{s,1} = 8{,}000\left[\frac{-\ln(20/120)}{0.4}\right]^2 = 160{,}500\ m^2$$

c. What is the minimum media area required in the first stage to avoid oxygen transfer limitations?
The limiting SOL is 32 g $BOD_5/(m^2 \cdot day)$. Using this value, the minimum media area required in the first stage can be calculated with Eq. 20.1:

$$A_{s,1} = \frac{(8{,}000)(120)}{32} = 30{,}000\ m^2$$

Standard density media must be used in the first stage because of the amount of biomass growth that will occur there. Since that media has a surface area of 9,300 m^2/shaft, 3.22 shafts would be required to obtain the needed minimum area. However, since whole shafts must be used, at least 4 shafts are needed, thereby providing 37,200 m^2.

d. How many trains should the facility have?
A minimum of four trains is desirable from an operational perspective. Since at least four shafts are required in the first stage, it is logical to have four trains, with at least one shaft in the first stage of each train.

e. How many stages should be used in each train?
From Table 20.2, two or three stages are needed to produce an effluent soluble BOD_5 of 10 to 15 mg/L (required to meet a total BOD_5 of 30 mg/L) and three or four stages are needed to produce an effluent soluble BOD_5 less than 10 mg/L (required to meet a total BOD_5 of 20 mg/L). Since the required effluent BOD_5 concentration is 30 mg/L and the design concentration is 20 mg/L, the use of three stages is reasonable.

f. How should the trains be configured?
From Part b above, the total media area for the system must be at least 160,500 m^2. Since four trains will be used, each train should have a total media area of at least 40,125 m^2. The simplest system would contain three shafts per train, one in each stage. As mentioned above, for carbon oxidation, standard density media (9,300 m^2/shaft) must be used in the first stage. However, high-density media (13,900 m^2/shaft) can be used in later stages if sufficient organic matter has been removed to prevent the growth of thick biofilms. With a total media area of 40,125 m^2 per train, if three stages were used with one shaft per stage, the average media area per shaft would have to be 13,375 m^2. There is no combination of standard- and high-density media that can provide this average area with three shafts. Only all high-density media can do so, which is unacceptable. Consequently, more than three shafts must be used in each train.

Four shafts per train requires an average area per shaft of about 10,000 m^2. This requirement cannot be met with standard-density media alone, but can easily be met with a combination of standard- and high-density media. Trying different combinations reveals that three standard-density shafts and one high-density shaft per train provides a total area of 41,800 m^2 per train, which is adequate. The total media area in the system would be 167,200 m^2. Two standard-density media and two high-density media shafts will provide 46,400 m^2 per train, or 185,600 m^2 for the system, which is approximately 15% greater than the needed area. There is an additional cost associated with

the use of high-density media, so the choice between the two options is an economic one in which the additional cost must be justified on the basis of improved performance or reliability.

The desired three stages might typically be achieved by using two standard density shafts as the first stage, with one other shaft in each subsequent stage. The use of two shafts in the first stage ensures that the SOL on that stage is well below the limit based on oxygen transfer. If only one high-density shaft is used in the system, it should be in the last stage. Regardless of which choice is made, each train should be configured with moveable baffles so that staging can be adjusted if necessary.

Second-Order Model. The second-order model, first proposed by Opatken,[16,17] assumes that the flux of soluble, biodegradable organic matter into a biofilm is second-order with respect to its bulk liquid phase concentration. The second-order flux equation was incorporated into the mass balance equation for soluble substrate in a single completely mixed bioreactor containing a biofilm. That equation was then solved and generalized to estimate the concentration of soluble, biodegradable organic matter in stage N of a multi-stage train, $S_{S,n}$:

$$S_{S,n} = \frac{-1 + \sqrt{1 + 4k_2 \cdot \tau_n \cdot S_{S,n-1}}}{2k_2 \cdot \tau_n} \tag{20.6}$$

where $S_{S,n-1}$ is the concentration of soluble substrate entering stage N from the preceding stage, k_2 is the second-order reaction rate coefficient, and τ_n is the HRT of stage N. When $S_{S,n-1}$ and $S_{S,n}$ are expressed in units of g/m^3 (or mg/L) of soluble BOD_5 and τ_n is in units of hrs, k_2 has been found to have a value of 0.083 m^3/(g · hr) for domestic wastewater when the tank volume to media surface area ratio is 4.89×10^{-3} m^3/m^2. This ratio is equivalent to 45 m^3/shaft for standard-density media.

As with the first-order model, Eq. 20.6 can be converted into one based on media surface area and wastewater flow rate. Using appropriate units conversions, and a value for k_2 of 0.083 m^3/(g · hr), the term $k_2 \cdot \tau_n$ becomes 0.00974 $A_{st,n}$/F, where $A_{st,n}$ is the media area (in m^2) in stage N and F is the flow to that stage, expressed in m^3/day. The units for this term are m^3/g of BOD_5. The complete expression is:

$$S_{S,n} = \frac{-1 + \sqrt{1 + (4)(0.00974)(A_{st,n}/F)(S_{S,n-1})}}{(2)(0.00974)(A_{st,n}/F)} \tag{20.7}$$

$S_{S,n}$ and $S_{S,n-1}$ must be expressed as soluble BOD_5 when using this expression.

The second-order model is used repetitively to estimate the effluent quality from one stage to another in an RBC train. Consequently, unlike the first-order model, it directly considers the benefits to be gained by staging. Thus, it is not uncommon for the second-order model to result in smaller systems than the first-order model for the same design situation. Because it minimizes the amount of media required, the media area in the first stage is typically set to give a SOL on that stage of 12 to 20 g soluble BOD_5/(m^2 · day) (equivalent to 32 g BOD_5/(m^2 · day) based on total BOD_5), the maximum that can be applied without oxygen transfer limitations. If no data are available about the soluble BOD_5 of the influent wastewater, it is often assumed that one-half of the total BOD_5 in the untreated wastewater is soluble. The influent soluble BOD_5 concentration is then used in conjunction with Eq 20.7 to estimate the effluent soluble BOD_5 concentration from the first stage. Various con-

figurations for the remaining stages are then evaluated to select the one that most economically meets the effluent quality goal, bearing in mind that the SOL on any stage must not exceed 12 to 20 g $BOD_5/(m^2 \cdot day)$ based on the soluble BOD_5. When the last stage is reached, since both the influent and the desired effluent concentrations are known, the required media area can be calculated directly. Rearrangement of Eq. 20.7 gives an equation for $A_{s,N}$, where N represents the last stage and $S_{S,n-1}$ is the soluble BOD_5 concentration entering the last stage:

$$A_{s,n} = (102.8)(F)\left(\frac{S_{S,n-1} - S_{Se}}{S_{Se}^2}\right) \tag{20.8}$$

If the calculated area is less than the area provided by a single shaft, the area of a single shaft must be used. This will produce a better effluent quality, which can be calculated with Eq. 20.7. Of course, the loading on the stage must be sufficiently low to prevent an oxygen limitation, but this will generally not be a problem for low effluent concentrations.

Equation 20.7 estimates the effluent soluble BOD_5 concentration. The total BOD_5 concentration in the process effluent must be estimated by adding the BOD_5 associated with the effluent suspended solids to the soluble BOD_5 concentration calculated with Eq. 20.7 for the last stage. One would expect the BOD_5 of the suspended solids to depend on the loading on the RBC system, just as the BOD_5 of suspended solids from trickling filters depends on the loading, as illustrated in Figure 19.9. Unfortunately, comparable data are not available for RBCs. Consequently, many designers assume that the soluble BOD_5 of a clarified RBC effluent is about one-half of the total BOD_5.

Example 20.3.1.2

Use the second-order model to size the RBC system considered in Example 20.3.1.1. Assume that four trains will be provided, just as in that example. Also assume that the primary clarifier removes 30% of the BOD_5 applied to it.

a. What must the design effluent soluble BOD_5 be?
In Part a of Example 20.3.1.1, it was determined that the design total BOD_5 must be 20 mg/L to meet the effluent criteria reliably. If we assume that the effluent soluble BOD_5 will be one-half of the total BOD_5, then the design effluent soluble BOD_5 must be 10 mg/L.

b. What is the soluble BOD_5 of the wastewater to be treated?
As discussed above, when no other information is available, designers typically assume that 50% of the BOD_5 in unsettled domestic wastewater is soluble. Because the primary clarifier removes 30% of the BOD_5 and the clarified wastewater has a total BOD_5 of 120 mg/L, the total BOD_5 of the untreated wastewater is:

$$\frac{120}{(1 - 0.3)} = 171 \text{ mg/L}$$

Assuming that 50% of that BOD_5 is soluble, the soluble BOD_5 is:

$$S_{SO} = (171)(0.5) = 86 \text{ mg/L}$$

c. What is the soluble BOD_5 concentration in the effluent from the first stage if it is made as small as possible without risk of oxygen limitation?
In Part c of Example 20.3.1.1 the minimum size of the first stage to avoid oxygen limitations was determined to be four standard density media shafts, with one in each of four trains. Use Eq 20.7 to determine the effluent soluble BOD_5 from this stage. The most straightforward way to approach the problem is to work with one train, since they will all be equal. Since the total flow to the system is 8,000 m^3/day, the flow per train is 2,000 m^3/day. Furthermore, since the first stage of each train contains one standard density shaft, the media area per stage is 9,300 m^2. Therefore, application of Eq. 20.7 gives:

$$S_{S,1} = \frac{-1 + \sqrt{1 + (4)(0.00974)(9{,}300/2{,}000)(86)}}{(2)(0.00974)(9{,}300/2{,}000)} = 33.9 \text{ mg/L}$$

d. What is the soluble BOD_5 concentration in the effluent from the second stage if it is also made as small as possible without risk of oxygen limitation?
To avoid oxygen limitations, no stage should have an SOL that exceeds 12 to 20 g soluble $BOD_5/(m^2 \cdot day)$. Using a conservative value of 12 g soluble $BOD_5/(m^2 \cdot day)$ for the second stage, and recognizing that the flow rate per train is 2,000 m^3/day, the minimum media surface area required in the second stage of each train can be calculated with Eq. 20.1:

$$A_{S,2} = \frac{(2{,}000)(33.9)}{12} = 5{,}600 \text{ m}^2$$

This is less than the media surface area provided by one standard density shaft, so the minimum acceptable media area is that associated with one standard density shaft, or 9,300 m^2. Using this area, the effluent soluble BOD_5 concentration from stage 2 can be calculated using Eq 20.7:

$$S_{S,2} = \frac{-1 + \sqrt{1 + (4)(0.00974)(9{,}300/2{,}000)(33.9)}}{(2)(0.00974)(9{,}300/2{,}000)} = 18.5 \text{ mg/L}$$

e. What size must the third stage be to produce an effluent with a soluble BOD_5 concentration of 10 mg/L?
Since the influent and the desired effluent BOD_5 concentrations are known, the required media area can be calculated with Eq. 20.8:

$$A_{S,3} = (102.8)(2{,}000)\left(\frac{18.5 - 10}{10^2}\right) = 17{,}500 \text{ m}^2$$

This is larger than the area provided by either a standard- or a high-density shaft. Thus, the desired effluent quality cannot be met with a single shaft stage. While two standard media shafts could be used in the third stage to provide the needed area, this would not be an economic design. It would be better to add more media area to the second stage by changing it from standard-density to high-density media, thereby reducing the BOD_5 concentration entering the third stage. This is possible because of the low BOD_5 concentration entering the second stage and the low SOL. Using Eq. 20.7 for a second stage with an area of 13,900 m^2 reveals that its effluent substrate concentration will be 16.2 mg/L. Using Eq. 20.8 with this value as the influent soluble BOD_5 into stage 3 reveals that the required media area 12,750 m^2, which can be met with one high-density shaft.

f. Summarize the design.
The design consists of four trains, each with three stages consisting of one shaft each. The first stage uses a standard-density shaft, while the following two stages use high-density shafts. The total media surface area for each train is therefore 37,100 m^2, giving 148,400 m^2 for the entire system.

Comparison of Examples 20.3.1.1 and 20.3.1.2 indicates that application of the first-order model results in a more conservative design than use of the second-order model. This is because the second-order model directly calculates the media area needed in each stage, thereby directly considering the benefits of staging. The first-order model, on the other hand, calculates the total media area needed in the entire system, and then divides it into stages with the empirical recommendations of Table 20.2. Nevertheless, both approaches resulted in a four train system containing three stages per train. The total media area calculated with the first stage model was 13% greater, however. Considering the range of assumptions required to complete the two examples, a 13% difference is not large and the similarity in the two resulting designs is encouraging. These two equations, and a variety of other process design approaches, including the relationships used by various manufacturers of RBC equipment, have been compared in design manuals.[29,31] Those comparisons indicate that the first- and second-order models provide relatively conservative and consistent estimates of RBC process performance that agree reasonably well with the performance of full-scale plants treating domestic wastewater. Consequently, use of those models was recommended for the design of municipal systems in the absence of relevant pilot- and/or full-scale data.[29,31]

20.3.2 Separate Stage Nitrification

The approach used to design RBC systems for separate-stage nitrification is the same as that outlined in Section 20.3.1, except that Figure 20.5 provides the performance relationship generally used. In order to accomplish separate stage nitrification, the concentration of soluble, biodegradable COD in the wastewater must be less than 20 mg/L (soluble BOD_5 less than 15 mg/L). Because of the low concentration of biodegradable organic matter, the first stage SOL will generally be less than 32 g BOD_5/(m^2·day) and thus will not need to be checked. The low first stage SOL will allow high-density media to be used throughout the process train.

Inspection of Figure 20.5 indicates that the flux of ammonia-N into the biofilm will be sustained at a value of 1.5 g NH_3-N/(m^2·day) until the ammonia-N concentration is lowered to about 5 mg/L. Thus, ammonia-N removal behaves in a zero-order manner at concentrations above that value, allowing the RBC media area required to lower the ammonia-N concentration from the influent concentration to 5 mg/L to be calculated directly using that flux. Then, a trial and error procedure based on the curve in Figure 20.5 must be used to determine the additional media area required to lower the ammonia-N concentration from 5 mg/L to the desired effluent concentration. Alternatively, the graphical procedure mentioned in Section 20.2.2 and presented in Section 20.3.4 could also be used. The computational procedure is illustrated in the next example.

Some researchers have advocated that the flow direction through separate stage nitrification trains be reversed periodically (say weekly) to maintain complete and

effective nitrifying biofilm development along the entire treatment train.[6] Such an approach will affect facility layout and configuration, but it may result in better system performance due to the maintenance of a more active nitrifying population.

Example 20.3.2.1

A wastewater treatment plant treating a flow of 8,000 m^3/day is achieving reliable secondary treatment (soluble BOD_5 concentration < 15 mg/L) but must be upgraded to provide nitrification. The ammonia-N concentration in the current effluent is 25 mg/L, and the effluent ammonia-N goal is 2.5 mg/L. Size a tertiary RBC system to accomplish this goal.

a. What surface area of media is required to lower the ammonia-N concentration from 25 mg/L to 5 mg/L?

From Figure 20.5, as long as the ammonia-N concentration is above 5 mg/L and the SOL is less than 4.3 g COD/(m^2 · day), the flux of ammonia-N into the biofilm is 1.5 g NH_3-N/(m^2 · day). The mass of ammonia-N to be removed is just the flow times the required concentration change of 20 mg/L, or 160,000 g N/day. From the definition of flux, the required RBC surface area is:

$$A_s = \frac{160{,}000}{1.5} = 106{,}700 \text{ m}^2$$

At 13,900 m^2/shaft for high-density media, this would require 7.7 shafts. Since only whole shafts can be used, use eight, for an area of 111,200 m^2.

Check the SOL to ensure that the assumption of a nitrification rate that is unrestricted by carbon oxidation is correct. As discussed in Section 20.2.1, nitrification will proceed at its maximum rate as long as the SOL is less than 4.3 g COD/(m^2 · day). The SOL can be calculated with Eq. 20.1, but only the soluble BOD_5 is specified in the problem statement. Thus, an estimate must be made of the COD concentration. As indicated earlier, the soluble BOD_5 is often assumed to be one-half of the total BOD_5. Furthermore, Eq. 8.33 states that the biodegradable COD can be estimated as 1.71 times the BOD_5. Using these conversion factors, a soluble BOD_5 of 15 mg/L is equivalent to a COD of 51 mg/L. Consequently, if all of the media is placed into a single stage, the SOL is:

$$\lambda_S = \frac{(8{,}000)(51)}{111{,}200} = 3.67 \text{ g COD/(m}^2 \cdot \text{day)}$$

Since this value is less than 4.3 g COD/(m^2 · day), nitrification would proceed at the unrestricted rate.

b. What media area is required to lower the ammonia-N concentration to 2.5 mg/L in a single stage?

Since a single stage is to be used, the ammonia-N concentration in the bioreactor will be 2.5 mg/L. From Figure 20.5, the ammonia-N flux at a concentration of 2.5 mg/L is approximately 0.75 g NH_3-N/(m^2 · day). Again, the mass of ammonia-N to be removed is just the flow times the required concentration change. Since the concentration change is 2.5 mg/L, the mass of ammonia-N to be removed is 20,000 g/day. From the definition of flux, the required RBC surface area is:

$$A_s = \frac{20{,}000}{0.75} = 26{,}700 \text{ m}^2$$

This area can be provided by two high-density shafts.

c. How should the system be configured?
Since a total of ten shafts are required and an integer number of shafts must be used in each train, configure the system as two trains containing five shafts each. Since no benefits are gained by staging as long as the ammonia-N removal rate is zero-order, i.e., as long as the ammonia-N concentration is 5 mg/L or above, the first four shafts in each train can be placed into a single stage. This will ensure that the SOL on that stage stays low enough so that carbon oxidation does not interfere with nitrification. The remaining fifth shaft per train should be placed in a separate stage to achieve the desired effluent ammonia-N concentration of 2.5 mg/L. Consequently, each train should have two stages with four shafts in the first stage and one in the second. Movable baffles should be used to achieve the staging. Furthermore, provisions could be made to allow the direction of flow to be reversed to maintain a more active biofilm in the last stage. If this were done, the baffle between the first and second stage would be moved to maintain four shafts in the first stage and one in the last.

20.3.3 Combined Carbon Oxidation and Nitrification

The design of combined carbon oxidation and nitrification systems incorporates the principles presented in the two previous sections. The major difference is that the effects of carbon oxidation on the nitrification rate must be considered through the use of Eq. 20.2, as was done in Example 20.2.1.1. The procedure is illustrated in the following example.

Example 20.3.3.1
Reconsider the wastewater for which the carbon oxidation designs were performed in Examples 20.3.1.1 and 20.3.1.2, and for which the separate-stage nitrification system was designed in Example 20.3.2.1. In this case, however, design a single RBC system to remove the biodegradable organic matter and to lower the effluent ammonia-N concentration to 2.5 mg/L. The concentration of ammonia-N available to be nitrified (after hydrolysis of organic nitrogen and uptake of nitrogen for cell synthesis) is 25 mg/L. Since four trains are used in Examples 20.3.1.1 and 20.3.1.2, configure the system as four trains in parallel.

a. How much media should the system contain?
Because of the competition for space between heterotrophs and autotrophs in the biofilm, nitrification will follow carbon oxidation to a considerable degree, even in systems in which both events are occurring. Consequently, we can obtain a conservative estimate of the media area required by summing the values determined in the preceding examples. From Example 20.3.1.1, a conservative design for the removal of biodegradable organic matter would require 160,500 m^2. From Example 20.3.2.1, a media surface area of 106,700 m^2 is required to reduce the ammonia-N concentration from 25 to 5 mg/L and 26,700 m^2 to reduce it from 5 to 2.5 mg/L in the absence of carbon oxidation. Use of the sum of these areas as an initial estimate of the required

media area, as follows:

$$A_s = 160{,}500 + 106{,}700 + 26{,}700 = 293{,}900 \text{ m}^2$$

b. How should the media be configured in each of the four trains?
The surface area required for each train is one-fourth of 293,900 m^2, or 73,500 m^2. Consider a layout consisting of two standard-density and four high-density shafts. The standard-density shafts will be placed in a single stage to ensure that the first stage SOL will be well below the value that will cause *Beggiatoa* growth. The remainder of the shafts will be high-density to minimize the total number of shafts. Using this configuration provides a total surface area per train of:

$$A_s = (2)(9{,}300) + (4)(13{,}900) = 74{,}200 \text{ m}^2$$

Consider a five-stage system using the two standard-density shafts as the first stage and each of the high-density shafts as an individual stage. Staging of the high-density shafts will maximize treatment capacity. As in the other designs, moveable baffles should be provided to maximize flexibility.

c. What is the BOD_5 removal profile through each train?
This information is needed to estimate where nitrification will occur. Use the second-order model, Eq. 20.7, to estimate the profile by calculating sequentially from stage to stage. The media surface area, A_s, and the value of A_s/F in each stage, calculated with a flow rate of 2,000 m^3/day for each train, are summarized in Table E20.1, along with the soluble BOD_5 concentration in each stage. The concentration of 4.6 mg/L in stage 5 indicates that the effluent total BOD_5 concentration will be low, provided that good removal of suspended solids occurs in the final clarifier.

d. What is the zero-order ammonia-N removal rate in each stage?
Two factors influence nitrification. One is the SOL, and its effects on the nitrification rate can be estimated by using Eq. 20.2. The second is the ammonia-N concentration, and its effect can be quantified using Figure 20.5. Furthermore, nitrification will be initiated only when the soluble BOD_5 concentration drops below about 15 mg/L.

Consider the impact of the SOL through use of Eq. 20.2. The SOL in that equation is expressed in units of g COD/(m^2 · day), but the substrate concentration leaving each stage in Table E20.1 is expressed as soluble BOD_5. Consequently, the COD concentration must be approximated from the BOD_5 information. This can be done as it was in Part a of Example 20.3.2.1, making the COD approximately 3.4 times the soluble BOD_5. The concentrations of soluble BOD_5 and the equivalent total substrate COD entering each stage are tabulated in Table E20.2. The SOL on each stage can be calculated with Eq. 20.1 using the appropriate media area in the stage and the flow rate of 2,000 m^3/day to each train. The SOL values are also tabulated in the table. Finally, once the SOL is known, the fractional nitrification rate, f_{NH}, can be calculated with Eq. 20.2. It should be recalled that no nitrification will occur when the SOL exceeds 14.3 g COD/(m^2 · day) and that nitrification will not be influenced by organic substrate removal when the SOL is less than 4.3 g COD/(m^2 · day). Furthermore, little if any nitrification will occur when the soluble BOD_5 concentration in a stage exceeds 15 mg/L. It is clear that no nitrification will occur in stage 1. For stage 2, the SOL is 12.7 g COD/(m^2 · day) and the soluble BOD_5 concentration in the stage is 13.5 mg/L. Thus,

Table E20.1 Soluble BOD_5 Concentrations Through the RBC Stages in Example 20.3.3.1

Stage	Surface area* m^2	A_s/F day/m	Soluble BOD_5 mg/L
0	—	—	86.0
1	18,600	9.30	25.8
2	13,900	6.95	13.5
3	13,900	6.95	8.6
4	13,900	6.95	6.1
5	13,900	6.95	4.6

*Per Train.

a small amount of nitrification may occur. However, it is not likely to be significant, so to be conservative f_{NH} was taken as zero. Examination of the SOL into stages 4 and 5 suggests that organic substrate removal will not restrict nitrification in them, making f_{NH} equal to 1.0. Only in stage 3 will significant competition occur between heterotrophs and autotrophs, giving an f_{NH} value of 0.77. The f_{NH} values are summarized in Table E20.2.

e. What is the ammonia-N profile through a train?
As long as the ammonia-N concentration is above 5 mg/L, calculation of the ammonia-N profile is straightforward because the nitrification rate is not influenced by the ammonia-N concentration. However, once the concentration drops below 5 mg/L, the effect of the concentration must be considered, which requires an iterative procedure, as illustrated in Example 20.2.1.1.

As indicated in Table E20.2, no nitrification will occur in stages 1 and 2, and thus the ammonia-N concentration in them is the same as the influent concentration, 25 mg/L. It is likely that the ammonia-N concentration will be high enough in stage 3 to allow nitrification to occur at a rate that is unaffected by the concentration. From Figure 20.5 the maximum rate in the absence of carbon oxidation is found to be 1.5 g NH_3-N/(m^2 · day). However, from Table E20.2 the value of f_{NH} is seen to be 0.77. Therefore, the ammonia-N removal rate is 1.16 g NH_3-N/(m^2 · day). Using this, the effluent ammonia-N concentration can be calculated from a mass balance on the stage.

Table E20.2 Ammonia-N Concentrations Through the RBC Stages in Example 20.3.3.1

Stage	Influent soluble BOD_5 mg/L	Influent COD mg/L	SOL g COD / m^2 · day	f_{NH}	NH_3-N mg/L
1	86.0	292	31.4	0.0	25.0
2	25.8	88	12.7	0.0	25.0
3	13.5	46	6.6	0.77	17.0
4	8.6	29	4.2	1.0	6.5
5	6.1	21	3.0	1.0	2.0

Input rate = (2,000)(25) = 50,000 g/day

Removal rate = (1.16)(13,900) = 16,100 g/day

Output rate = 50,000 − 16,100 = 33,900 g/day

Effluent concentration = 33,900 ÷ 2,000 = 17 mg N/L

The concentration is well above 5 mg/L, so the assumption of a zero-order rate is justified. Examination of the mass removal rate in stage 3 and comparison of it to the mass input rate into stage 4 suggests that the concentration in stage 4 will also be high enough to allow nitrification to occur at the zero-order rate. In this case, however, carbon oxidation has no effect, as indicated in Table E20.2. Thus, the rate will be 1.5 g NH_3-N/(m^2 · day). Repeating the procedure above for stage 4 reveals that the effluent ammonia-N concentration will be 6.5 mg/L. Since this is above 5 mg/L, the assumption that nitrification will occur at the zero-order rate is justified.

Since the ammonia-N concentration entering stage 5 is close to 5 mg/L, the effluent concentration from stage 5 will be less than 5 mg/L, causing the ammonia-N flux to be limited by the ammonia-N concentration. This requires use of an iterative procedure to estimate the effluent ammonia-N concentration, just as was done in Example 20.2.1.1. Application of that procedure reveals that the effluent ammonia-N concentration from stage 5 is 2.0 mg/L.

Although the effluent ammonia-N concentration is slightly lower than the desired effluent concentration, the necessity to use full shafts of fixed area makes it unlikely that the media area could be reduced while still meeting the effluent requirement. Consequently, this is considered to be an acceptable design.

20.3.4 Pilot Plants

Pilot studies and/or the analysis of full-scale data from similar systems can provide the basis for the design of systems treating a wide variety of wastewaters, such as industrial wastewater, contaminated groundwater, etc. Such data can also be used to optimize the expansion of an existing system. Pilot studies can be conducted with either a single-stage or a multi-stage system. The surface loading rates on the pilot unit are varied and the pollutant concentrations into and out of each stage are measured, thereby allowing the fluxes to be determined for the various loadings. For existing full-scale systems, similar information can be obtained by collecting interstage data at the operating loading rate and then increasing the loading to individual units to determine performance limits. Regardless of whether the studies are performed on pilot- or full-scale systems, the process loading rate should be maintained at a constant value until steady-state operating conditions are achieved before collecting data. This may require two to three weeks or more for each loading rate. Either a single-stage or a multi-stage pilot system may be rented or purchased. A multi-stage pilot unit requires fewer steady-state operating periods because of the unique loading to each stage, and thus need not be operated as long, but costs more. Care must be exercised when analyzing the data from a multi-stage system because different microbial communities may develop on the individual stages. If this occurs, separate performance relationships must be developed for each stage.

Pilot studies should be conducted using full-scale, i.e., 3.6 m diameter, units rotating at 1.4 to 1.6 rpm. This is because of the effects of disc diameter and rotational speed on process performance, as discussed in Section 17.2. In the past, small diameter pilot-scale units have been used without consideration for the differences in mass transfer characteristics between pilot-scale and full-scale units.[14] The result has been full-scale units that did not perform as expected. Furthermore, as illustrated in Figure 17.5, scale-up by maintaining the same THL and peripheral velocity in the pilot unit as in full-scale units does not work either.[12,34] Consequently, the only safe approach is to use pilot units with full diameter discs. The shaft length is simply shorter.

Data from full-scale or pilot-scale units can be analyzed using the models and empirical relationships presented above. For the removal of biodegradable organic matter this consists of the first-order and second-order models. For nitrification this involves comparison of the measured ammonia-N flux with the typical values presented in Figure 20.5. In some instances, a Monod-type model has been fit to data from pilot units,[28] although it should be recognized that in such cases the half-saturation coefficient incorporates the mass transfer effects, as indicated in Eq. 19.5 for trickling filters. Any result that indicates zero-order organic substrate or ammonia-N fluxes greater than those presented in this chapter should be viewed with skepticism. They indicate that, for some reason, the pilot unit is achieving a greater oxygen transfer rate than is typical in full-scale units.

Graphical procedures provide an alternative approach for using pilot-scale data to size an RBC system, and they have been used successfully in several instances.[1,7,9] They are based on a mass balance across an RBC stage. The basic procedure will be illustrated for organic matter removal, but applies equally well for nitrification. To begin the procedure, pilot-plant data are used to calculate the removal rate per unit area, i.e., the flux, of organic matter, J_S, for each stage and/or each loading. The rates are then plotted as a function of the residual concentration in the stage and a smooth curve is drawn through the data. Such a curve will generally have a shape like that illustrated in Figure 20.9. Since each stage in an RBC can be considered to be completely mixed, a steady-state mass balance on soluble substrate across stage N is:

$$F \cdot S_{S,n-1} - F \cdot S_{S,n} - J_{S,n} \cdot A_{st,n} = 0 \tag{20.9}$$

where $J_{S,n}$ is the flux in stage N, which is related to the substrate concentration $S_{S,n}$ in that stage as illustrated in Figure 20.9. All other terms have been defined previously. Rearrangement of this equation provides the rationale for the graphical approach:

$$\frac{J_{S,n}}{S_{S,n} - S_{S,n-1}} = \frac{-F}{A_{st,n}} = -\Lambda_{H,RBC,n} \tag{20.10}$$

As depicted by Eq. 20.10 and illustrated in Figure 20.9, a line drawn from $J_{S,n}$ to the input substrate concentration, $S_{S,n-1}$, on the abscissa will have a slope of $-F/A_{st,n}$, which is just the THL on the stage times -1.0. This line is called the operating line. Once the plot of flux versus substrate concentration has been drawn from the pilot-plant data, the THL required to decrease the substrate concentration from $S_{S,n-1}$ to $S_{S,n}$ in a single stage can be determined by drawing an operating line from $S_{S,n-1}$ on the abscissa to the flux associated with $S_{S,n}$ and measuring the slope. Since the

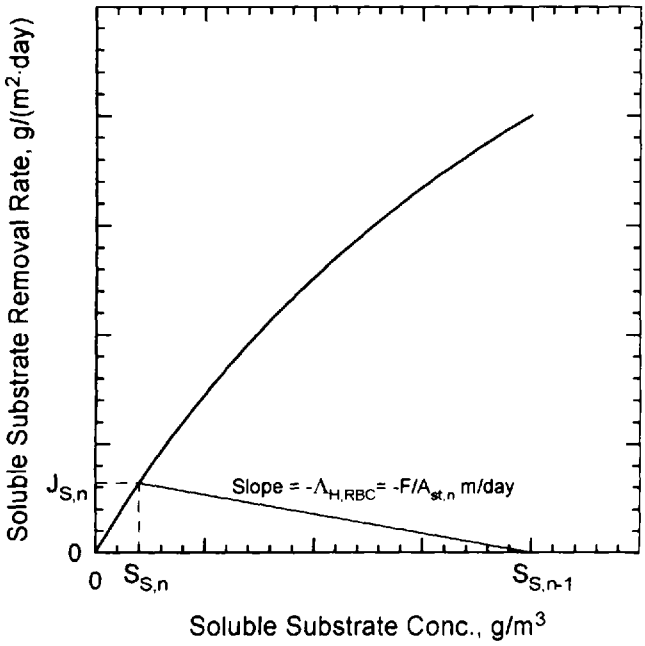


Figure 20.9 Illustration of the graphical procedure for determining the hydraulic loading required to reduce the soluble substrate concentration from $S_{S,n-1}$ to $S_{S,n}$ in a single-stage RBC.

flow rate is known, the required stage surface area can be determined. Furthermore, by recognizing the definition of the SOL, it can be seen that the intersection of the operating line with the ordinate is that loading. This allows the graphical procedure to also be used to determine the output concentration from any stage in an RBC train. The SOL on the stage is calculated with Eq. 20.1 and an operating line is drawn from the SOL on the ordinate to the influent substrate concentration on the abscissa. The intersection of that operating line with the curve gives the flux in that stage and the output substrate concentration, $S_{S,n}$.

The graphical procedure can also be used to determine the output concentration from a staged RBC system or to determine the size system required to achieve a desired effluent concentration. The former can be done directly whereas the latter requires an iterative approach. For an existing system, the THL is calculated for each stage. The procedure, illustrated in Figure 20.10, is initiated by plotting an operating line with slope $-F/A_{st,1}$ from the system influent substrate concentration, $S_{S,0}$ on the abscissa to the rate curve. The intersection of that operating line with the curve gives the output substrate concentration from stage 1, $S_{S,1}$. Since that concentration is also the influent concentration to stage 2 the procedure can be repeated to determine the output concentration from stage 2, and so on, until the final stage is reached, as shown in Figure 20.10. If all stages have the same hydraulic loading, then all of the operating lines will be parallel, as illustrated in the figure. For design of a new system, an iterative procedure must be used. Generally, the smallest system will result when the first stage is loaded with an SOL of 32 g $BOD_5/(m^2 \cdot day)$ because that will

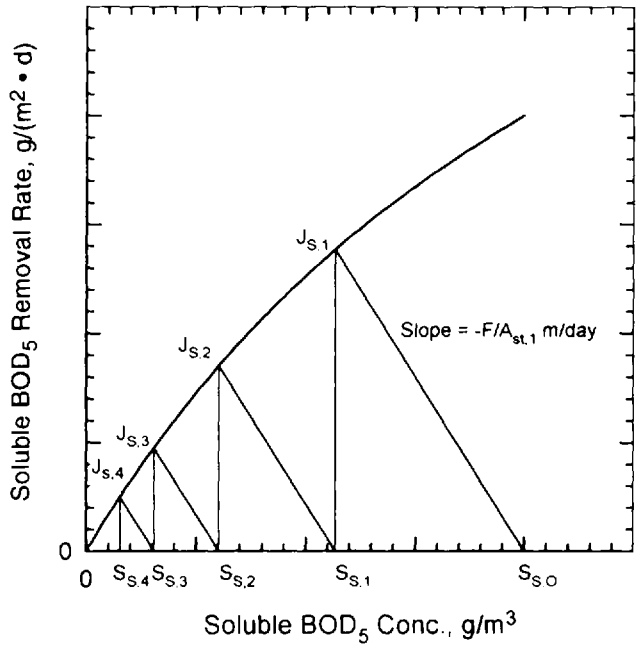


Figure 20.10 Illustration of the graphical procedure for determining the hydraulic loading required on each stage of a multi-stage RBC to reduce the soluble substrate concentration from $S_{S,O}$ to a desired effluent concentration, $S_{S,E}$.

minimize its size. Thus, to start the procedure, an operating line is drawn from the system influent substrate concentration on the abscissa to the limiting SOL on the ordinate. The intersection of that operating line with the rate curve gives the output substrate concentration from stage 1. A tentative decision is then made about the number of additional stages needed and their size. If they are all the same size they will all have the same THL, making the slopes of the operating lines the same for all stages. The graphical procedure is then employed to determine if the tentative design will achieve the effluent quality goal. If it does not, then a new size is selected and the procedure is repeated. The only constraint is that each stage must consist of at least one complete shaft of standard- (9,300 m^2) or high-density (13,900 m^2) media. The advantage of the graphical procedure is that it is rapid and allows the designer to visualize easily the impact of decisions. The procedure is illustrated in the following example.

Example 20.3.4.1

Data were collected with a single stage RBC pilot plant to determine the treatability of a soluble industrial wastewater. They were used to determine the relationship between the concentration of soluble BOD_5 and the related substrate flux into the biofilm, J_S, which is presented in Figure E20.1. The plotted points represent the data, whereas the curve represents the general relationship between the soluble BOD_5 flux and the soluble BOD_5 concentration. The wastewater to be treated has a design flow rate is 1,500 m^3/day and a soluble BOD_5 concentra-

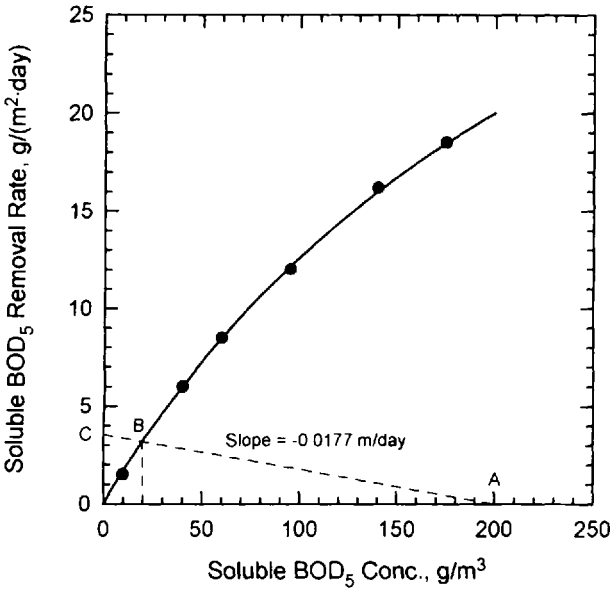


Figure E20.1 Application of the graphical procedure to determine the required hydraulic loading on the single-stage RBC in Example 20.3.4.1.

tion of 200 mg/L. Size both a single-stage and a multi-stage system to reduce the soluble BOD_5 concentration to 20 mg/L or less.

a. How large would a single-stage system have to be?
Two points must be located on the graph in Figure E20.1 to determine the size of a single-stage system. The first is point A, the location of the influent substrate concentration on the abscissa. The second is point B, corresponding to the substrate flux associated with the desired effluent substrate concentration of 20 mg/L. The value of the flux at point B is 3.19 g $BOD_5/(m^2 \cdot day)$. An operating line is then drawn between A and B. Its slope gives the THL, which can be estimated by applying Eq. 20.10:

$$-A_{TL,RBC,n} = \frac{3.19}{20 - 200} = -0.0177 \text{ m/day}$$

Since the flow rate is 1,500 m^3/day, the required surface area of media is 84,600 m^2. Because of the high influent substrate concentration, standard density media (9,300 m^2) should be used, requiring 9.10 shafts. Since whole shafts must be used, 10 shafts are required.

Given the low flux, it is unlikely that the SOL will be high enough to cause an oxygen limitation, but this can be easily checked by noting point C, the point at which the operating line intersects the ordinate. There it can be seen that the SOL is only 3.5 g $BOD_5/(m^2 \cdot day)$, which is well below the maximum allowable.

b. How large would a multi-stage system have to be?
This analysis is presented in Figure E20.2, which is drawn with an expanded ordinate. The SOL on the initial stage is limited to 32 g $BOD_5/(m^2 \cdot day)$ to

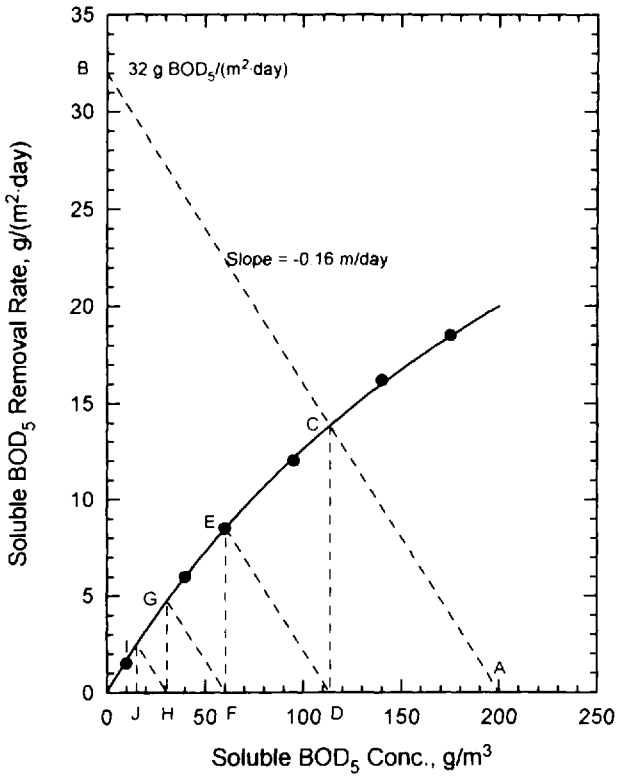


Figure E20.2 Application of the graphical procedure to determine the required hydraulic loading on each stage of the multi-stage RBC in Example 20.3.4.1.

avoid oxygen deficiencies and excessive *Beggiatoa* growth. The operating line for the first stage is constructed as line A-B, beginning at the influent BOD_5 concentration of 200 mg/L on the abscissa (point A) and terminating on the ordinate at the maximum allowable SOL of 32 g $BOD_5/(m^2 \cdot day)$ (point B). The slope of this line is −0.16 m/day, giving the THL on the first stage. The substrate flux into the biofilm on the first stage is 13.8 g $BOD_5/(m^2 \cdot day)$, giving an effluent soluble BOD_5 concentration of 113 mg/L, as indicated by points C and D in the figure. Since the flow rate is 1,500 m^3/day and the THL is 0.16 m/day, the required surface area in the first stage is 9,375 m^2. The difference between the required area and the area provided by a standard-density shaft (9,300 m^2) is within the error of the analysis. Consequently, a single standard-density shaft can be used.

As a first trial, assume that all stages will be equal sized, so that the THL on each stage will be the same. Consequently, the substrate flux into the biofilm and the output soluble BOD_5 concentration from each subsequent stage can be determined by using operating lines of equal slope, which will be parallel to the operating line for the first stage. From Figure E20.2 it can be seen that four stages are required to reduce the effluent soluble BOD_5 concentration to less than 20 mg/L. In fact, the effluent $SBOD_5$ will be about 15 mg/L. As in the first stage, each stage would contain a single, standard-density media shaft.

The multi-stage system requires four standard-density media shafts, whereas the single stage system requires ten. This illustrates the benefits of staging in RBC systems.

20.3.5 General Comments

The design of RBC systems must consider not only their normal operation, but also operation under unusual conditions. Consequently, operational flexibility must be provided. The following is a list of factors that should be considered and/or provided:[31]

- The need for supplemental aeration for mechanical drive systems. Supplemental aeration can help control biomass growth on the discs and allow the first stage to better accommodate SOLs approaching the oxygen transfer limit. Aeration is particularly needed when the influent H_2S concentration is elevated.
- A means for removing excess biomass, such as by air or water stripping, chemical addition, or rotational speed control and/or reversal.
- Multiple treatment trains. Multiple trains allow a portion of the system to be taken off-line for maintenance while minimizing the impact on effluent quality. They also allow the number of trains in service to be adjusted in proportion to the actual long-term plant load.
- Removable baffles between all stages. These allow the degree of staging to be altered.
- Positive influent flow control to each train.
- Alternate operating modes such as step feeding. These allow the loading to a particular stage to be changed.
- Positive air flow metering to each shaft when supplemental aeration or air drive units are used.
- Recirculation of treated effluent to the first stage. This reduces influent concentrations and uses the dissolved oxygen in the recirculation to oxidize H_2S in the influent.
- Methods for removing H_2S from the influent (e.g., by oxidation or precipitation) or for reducing the influent organic loading (e.g., by chemical addition to the upstream primary clarifier).
- The provision of hydraulic or electronic load cells on selected shafts if excess biomass buildup is a possibility.

20.4 PROCESS OPERATION

Efficient and effective operation of an RBC system is achieved by systematic application of the principles described above. Appropriate overall and individual stage organic and ammonia loadings must be maintained. This may require periodic removal of RBC trains from service or their return to service in response to long-term changes in process loadings. Inter-stage baffles may also be relocated as process loadings change. For carbon oxidation applications, it may be possible to save energy by removing trains from service when the long-term loading is less than the design

loading. Furthermore, this may be necessary when it is desirable to minimize nitrification, which will tend to occur in an under-loaded system. On the other hand, if nitrification is required, then a sufficient number of trains should always be in service both to remove the influent organic matter and to nitrify.

Achieving the objectives described above will require active management of the number of trains in service. If an RBC train is to be taken out of service for a short period of time, it can be idled by discontinuing its feed but continuing to rotate the media to keep the biofilm wet. The activity of the biomass will slowly drop as it undergoes decay, but the biofilm will remain attached to the media. In this way, when the unit is brought back on line, only a few days will be required to achieve full treatment capacity, with the number depending on the length of time the unit has been out of service. If the unit will be out of service so long that essentially all activity is lost, the unit should be drained and flushed. If this is not done and rotation is stopped, the biomass exposed to the air will dry out and partially detach from the media. This biomass will slough off of the media when rotation is resumed, and new growth will have to be established. Furthermore, odors will develop because the water in the bioreactor will become anaerobic due to lack of aeration. Once an RBC is devoid of biofilm, seven to ten days are required for establishment of a significant biofilm when it is placed back in service, and 20 to 30 days may be required before its full treatment capacity is achieved.

Like all biochemical operations, effective operation of an RBC system requires routine sampling and analysis, as well as the performance of preventative maintenance. These topics are covered well in a variety of sources.[32,33]

20.5 KEY POINTS

1. Rotating biological contactors (RBCs) contain vertical discs of corrugated plastic media mounted on a horizontal shaft, submerged approximately 40% in the process flow, and rotated at 1.6 rpm. Several manufactures produce RBC equipment, but they all use standard dimensions and similar configurations. A typical media bundle (usually referred to as a shaft) is 3.66 m in diameter and 7.62 m long. Standard-density media provides around 9,300 m^2 of surface area per shaft while high-density media provides 13,900 m^2.
2. Individual RBC units are arranged into treatment trains generally containing 4 to 8 shafts each. The treatment trains operate in parallel and independently of each other. Within a treatment train, wastewater flows in series from one RBC shaft to the next. Moveable baffles are used to establish physically and functionally distinct zones called stages. Each stage contains one or more RBC shafts.
3. RBC installations are classified using two criteria: the treatment objective and the equipment used. Treatment objectives include the removal of biodegradable organic matter, combined carbon oxidation and nitrification, and separate-stage nitrification. Equipment types include mechanically driven and air driven units.
4. Many RBC systems designed and installed prior to 1980 experienced significant equipment and/or performance problems. However, based on

today's understanding, RBC facilities can be properly designed and operated.

5. The surface organic loading (SOL), typically expressed as the g BOD_5/($m^2 \cdot$ day), is one of the principle factors determining the overall performance of an RBC. It determines the extent to which biodegradable organic matter is oxidized and whether nitrification will occur.
6. The SOL on any RBC stage must not exceed a limiting value, generally taken to be 32 g BOD_5/($m^2 \cdot$ day), which represents the maximum rate at which organic matter can be applied while still maintaining aerobic conditions within the RBC biofilm.
7. Nitrification occurs in RBC systems when the concentration of soluble, biodegradable organic matter surrounding the biofilm is reduced below about 20 mg COD/L (15 mg BOD_5/L). Equation 20.2 can be used to adjust nitrification rates for the presence of heterotrophs in the biofilm when both carbon oxidation and nitrification are occurring in the same RBC stage.
8. The zero-order ammonia-N flux in separate stage nitrification systems is approximately 1.5 g N/($m^2 \cdot$ day) and is limited primarily by the maximum rate of oxygen transfer into the biofilm. This flux is observed when ammonia-N concentrations exceed 5 mg/L. Lower fluxes are observed at lower ammonia-N concentrations.
9. As in most biochemical operations, staging allows advantage to be taken of the effect of bulk substrate concentration on the reaction rates in RBC systems, thereby reducing the required media volume. The higher SOLs in the initial stages of the RBC cause higher bulk substrate concentrations, which allow high reaction rates, while the lower SOLs on the downstream stages allow the desired effluent quality to be achieved.
10. The effect of temperature on RBC reaction rates is relatively small down to a temperature of about 15°C. This occurs because the flux to the biofilm is generally controlled by diffusion rather than by microbial reactions down to that temperature. Below 15°C, however, biological reaction rates slow sufficiently to make the flux dependent on temperature. Relationships are available to characterize the effects.
11. Excessive growths of the sulfur oxidizing bacteria *Beggiatoa* occur when hydrogen sulfide and biodegradable organic matter are present in sufficient concentrations. Hydrogen sulfide can be present in the influent wastewater or it can be produced in the biofilm by anaerobic conditions caused by organic overloading.
12. The conditions leading to excessive *Beggiatoa* growth are well defined and techniques are available for controlling it. One important one is to keep the SOL below 32 g BOD_5/($m^2 \cdot$ day).
13. Two models frequently used to characterize the flux of biodegradable organic matter into an RBC biofilm are the first-order model, which is similar to the Velz-Germain equation used to size trickling filters, and the second-order model. Typical values of the model parameters have been developed for domestic wastewaters. Site-specific values must be developed for industrial wastewaters.

14. Care must be exercised when interpreting data from RBC pilot plants. Since maximum fluxes in RBC systems are generally controlled by oxygen transfer rates, and the oxygen transfer characteristics of smaller pilot-scale units may exceed those of full-scale units, the pilot plant may overestimate full-scale RBC treatment capacity. The safest approach is to use pilot plants containing full diameter discs running at typical rotational speeds.
15. Design of an RBC system requires determination of the total media surface area required and distribution of that media into parallel treatment trains containing individual shafts in series. The designer has considerable latitude in selection of the system configuration.
16. Several features, listed in Section 20.3.5, should be considered for incorporation into any RBC facility. These features provide the flexibility required to allow the system to be adjusted to meet changing operational needs.
17. Operation of an RBC system requires maintenance of proper loadings on the treatment units. This will require units to periodically be taken out of service or returned to service.

20.6 STUDY QUESTIONS

1. Summarize the current developmental status and typical applications for RBC wastewater treatment technology.
2. Describe the relative advantages and disadvantages of mechanical versus air drives for RBC units.
3. Prepare a table summarizing the primary benefits and drawbacks of RBCs.
4. Describe the sequence of events that occurs when the SOL on the first stage of an RBC system exceeds 32 g $BOD_5/(m^2 \cdot day)$. Why do those events occur?
5. Explain why the concentration of soluble, biodegradable organic matter must be lowered to about 20 mg/L as COD before nitrification can occur in an RBC.
6. Describe why the substrate removal rate in RBCs is generally not affected by temperature until it is relatively low. At what temperature is an effect generally observed?
7. Describe the mechanism for growth of the nuisance microorganism *Beggiatoa* in RBC systems.
8. Using the graphical procedure illustrated in Section 20.3.4, explain why staging generally improves the performance of an RBC system. What constraints exist on the use of staging to improve RBC performance?
9. What wastewater characteristics are particularly important when an RBC is being considered for a specific application? How are these characteristics related to RBC performance?
10. Summarize the factors that should be considered in the physical design of an RBC system.

11. Domestic wastewater with a design flow rate of 15,000 m^3/day, a BOD_5 concentration of 90 mg/L, a soluble BOD_5 concentration of 60 mg/L, and an ammonia-N concentration of 20 mg/L must be treated.
    a. Use the first order model to size an RBC system to reduce the BOD_5 to 20 mg/L. In your design, select the total number of shafts required and the manner in which they should be arranged.
    b. Check the proposed design using the second order model and explain any differences that might exist between the two designs.
    c. Design a combined carbon oxidation and nitrification system to reduce the ammonia-nitrogen concentration to 2 mg/L. In your design, select the total number of shafts required and the manner in which they should be arranged.
12. A wastewater with a flow of 20,000 m^3/day, a BOD_5 of 15 mg/L, and an ammonia-N concentration of 30 mg/L must be treated to reduce the effluent ammonia-N concentration to less than 0.5 mg/L. Size an RBC system to accomplish this treatment goal. Assume that Figure 20.5 represents the effect of the ammonia-N concentration on the nitrification rate.
13. An industrial wastewater with a flow rate of 10,000 m^3/day and a soluble BOD_5 concentration of 150 mg/L must be treated to a soluble BOD_5 concentration of 10 mg/L. A pilot-plant study provided the data in Figure E20.2. Use the graphical procedure with that data to design a multi-stage RBC system to meet the treatment objective. Justify all decisions made for the design.

REFERENCES

1. Aliley, J. E., A pilot plant study of a rotating biological surface for secondary treatment of unbleached kraft mill waste. *Technical Association of the Pulp and Paper Industry* **57**(9):106–111, 1974.
2. Antonie, R. L., Response of the Bio-Disc process to fluctuating wastewater flow. *Proceedings of the 25th Industrial Waste Conference*, Purdue University Engineering Extension Series No. 137, pp. 427–435, 1970.
3. Antonie, R. L., *Fixed Biological Surfaces-Wastewater Treatment: The Rotating Biological Contactor*, CRC Press, Cleveland, Ohio, 1976.
4. Antonie, R. L. and F. M. Welch, Preliminary results of a novel biological process for treating dairy wastes. *Proceedings of the 24th Industrial Waste Conference*, Purdue University Engineering Extension Series No. 135, pp. 115–126, 1969.
5. Benjes, H. H., Sr., Small community wastewater treatment facilities-biological treatment systems. U.S. EPA Technology Transfer National Seminar, Chicago, Illinois, 1977.
6. Boller, M., W. Gujer and G. Nyhuis, Tertiary rotating biological contactors for nitrification. *Water Science and Technology* **22**(1/2):89–100, 1990.
7. Clark, J. H., E. M. Moseng, and T. Asano, Performance of a rotating biological contactor under varying wastewater flow. *Journal, Water Pollution Control Federation* **50**:896–911, 1978.
8. Daigger, G. T. and J. A. Buttz, *Upgrading Wastewater Treatment Plants*, Technomic Publishing, Lancaster, Pennsylvania, 1992.
9. Davies, B. T. and R. W. Vose, Custom designs cut effluent treating costs—case histories at Chevron USA Inc. *Proceedings of the 32nd Industrial Waste Conference, 1977, Pur-*

due University, Ann Arbor Science Publishers, Ann Arbor, Michigan, pp. 1035–1060, 1978.

10. Dupont, R. R. and R. E. McKinney, Data evaluation of a municipal RBC installation, Kirksville, Missouri. *Proceedings of the First National Symposium on Rotating Biological Contactor Technology*, Champion, Pennsylvania, pp. 205–234, 1980.
11. Filion, M. P., Murphy, K. L. and J. P. Stephenson, Performance of a rotating biological contactor under transient loading conditions. *Journal, Water Pollution Control Federation* **51**:1925–1933, 1979.
12. Freidman, A. A., L. E. Robbins and R. C. Woods, Effect of disk rotational speed on biological contactor efficiency. *Journal, Water Pollution Control Federation* **51**:2678–2690, 1979.
13. Hyned, R. J. and H. Iemura, Nitrogen and phosphorus removal with rotating biological contactors. *Proceedings of the First National Symposium on Rotating Biological Contactor Technology*, Champion, Pennsylvania, pp. 295–324, 1980.
14. Murphy, K. L. and R. W. Wilson, *Pilot Plant Studies of Rotating Biological Contactors Treating Municipal Wastewater*, Report SCAT-2, Environment Canada, Ottawa, Ontario, 1980.
15. Neu, K. E., Upgrading of rotating biological contactor (RBC) systems to achieve higher effluent quality, including biological nutrient enrichment and reduction techniques. *Proceedings of the 2nd International Specialized Conference on Upgrading of Wastewater Treatment Plant*, Berlin, Germany, pp. 231–241, 1993.
16. Opatken, E. J., An alternative RBC design: second order kinetics. *Environmental Progress* **5**:51–56, 1986.
17. Opatken, E. J., Rotating biological contactors—second order kinetics. *Proceedings of the First International Conference on Fixed-Film Biological Processes*, Vol I, EPA/600/9-82/023a, Kings Island, Ohio, pp. 210–232, 1992.
18. Pano, A. and J. E. Middlebrooks, Kinetics of carbon and ammonia nitrogen removal in RBCs. *Journal, Water Pollution Control Federation* **55**:956–965, 1983.
19. Poon, C. P. C., Y-L Chao and W. J. Mikucki, Factors controlling rotating biological contactor performance. *Journal, Water Pollution Control Federation* **51**:601–611, 1979.
20. Scheible, O. K. and J. J. Novak, Upgrading primary tanks with rotating biological contactors. *Proceedings of the First National Symposium on Rotating Biological Contactor Technology*, Champion, Pennsylvania, pp. 961–996, 1980.
21. Stover, E. L. and D. F. Kincannon, One step nitrification and carbon removal. *Water and Sewage Works* **122**(6):66–69, 1975.
22. Stover, E. L. and D. F. Kincannon, Rotating disc process treats slaughterhouse waste. *Industrial Wastes* **22**(3):33–39, 1976.
23. Stratta, J. M. and D. A. Long, *Nitrification Enhancement Through pH Control with Rotating Biological Contactors*, Final Report Prepared for USA Medical Research and Development Command, Contract No. DAMD17-79-C-9110 by Pennsylvania State University, 1981.
24. Surampalli, R. Y. and E. R. Baumann, Effectiveness of supplemental aeration and an enlarged first-stage in improving RBC performance. *Environmental Progress* **12**:24–29, 1993.
25. U.S. Environmental Protection Agency, *Capital and O&M Cost Estimates for Attached Growth Biological Wastewater Treatment Processes,* U. S. Environmental Protection Agency, EPA/600/52-89-003, Cincinnati, Ohio, 1989.
26. U.S. Environmental Protection Agency, *Design Information on Rotating Biological Contactors*, U. S. Environmental Protection Agency, EPA/600/2-84/106, Cincinnati, Ohio, 1984.
27. U.S. Environmental Protection Agency, *Nitrogen Control*, U. S. Environmental Protection Agency, EPA/625/R-93/010, Washington, D.C., 1993.

28. U.S. Environmental Protection Agency, *Review of Current RBC Performance and Design Procedures*, U. S. Environmental Protection Agency, EPA/600/2-85/033, Cincinnati, Ohio, 1985.
29. Water Environment Federation, *Aerobic Fixed Film Reactors* (Draft), Water Environment Federation, Alexandria, Virginia, 1994.
30. Water Environment Federation, *Hydrogen Sulfide Control Manual: Septicity, Corrosion, and Odor Control in Sewerage Systems*, Water Environment Federation, Alexandria, Virginia, 1989.
31. Water Environment Federation, *Wastewater Treatment Plant Design*, Manual of Practice No. 8, Water Environment Federation, Alexandria, Virginia, 1992.
32. Water Pollution Control Federation, *O & M of Trickling Filters, RBCs, and Related Processes*, Manual of Practice OM-10, Water Pollution Control Federation, Alexandria, Virginia, 1988.
33. Water Pollution Control Federation, *Operation of Municipal Wastewater Treatment Plants*, Manual of Practice No. 11, Water Pollution Control Federation, Alexandria, Virginia, 1990.
34. Wilson, R. W., K. K. Murphy and J. P. Stephenson, Scale-up in rotating biological contactor design. *Journal, Water Pollution Control Federation* **52**:610–621, 1980.

This page intentionally left blank

21

Submerged Attached Growth Bioreactors

Attached growth processes have been used for the aerobic treatment of wastewaters for over a century, primarily in the form of trickling filters and rotating biological contactors. Nevertheless, during the past two decades much has been learned about the mechanisms by which such processes operate, leading to the development of new bioreactor configurations, such as the fluidized bed biological reactor (FBBR) discussed in Chapter 18. Like the FBBR, this new generation of attached growth bioreactors tends to have media that is submerged in the process flow, hence the name submerged attached growth bioreactor (SAGB). This chapter introduces some of the more promising SAGBs. It is not meant to be an extensive review of all that have been proposed, but rather is intended to familiarize the reader with some of them.

21.1 PROCESS DESCRIPTION

Many SAGB configurations have been conceived of and evaluated over the past two decades. Some were specifically developed for and applied to anaerobic systems. They are discussed in Chapter 13. In general, interest in submerged attached growth systems stems from the high biomass concentrations that can be achieved, resulting in short hydraulic residence times (HRTs) in comparison to suspended growth systems with equivalent solids retention times (SRTs).[40] This results from the use of media with high specific surface areas, as discussed in Section 18.1.1. Short HRTs result in compact systems, which can be quite advantageous when the land area available to construct a wastewater treatment facility is limited. These highly compact systems generally use submerged media operated in either an upflow or a downflow mode. Combined suspended and attached growth systems, in which submerged media is added to a suspended growth system, are also receiving interest.

21.1.1 General Description

Figure 21.1 provides a schematic diagram of a typical SAGB. Its primary components are a reactor vessel, support media for biofilm growth, influent distribution system, and effluent withdrawal system. An oxygen transfer system may also be provided. Influent wastewater is added to the bioreactor and microorganisms grow attached to the submerged media, thereby removing soluble organic matter or oxidizing ammonia-N. The types of microorganisms that grow depend on the constit-

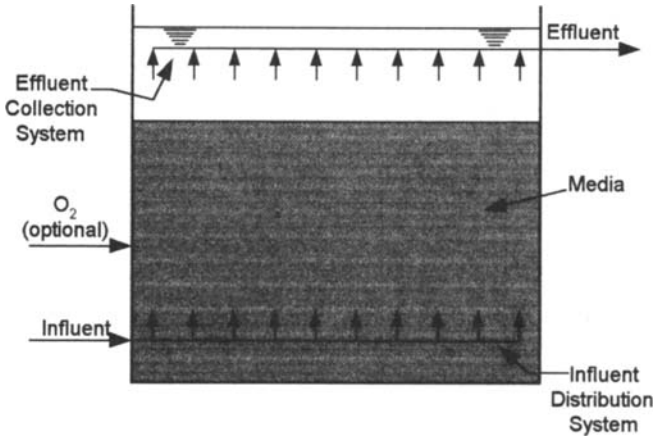


Figure 21.1 Submerged attached growth process. (Example shown is upflow. Reverse direction of arrows for downflow.)

uents in the wastewater and the environmental conditions provided in the bioreactor. Flow may be upward, downward, or horizontally across the media, and a wide variety of media types can be used. Table 21.1 summarizes the media characteristics and operating conditions for the SAGBs considered in this chapter. As indicated there, the media may be either packed or fluidized. Effluent is recirculated in some instances to maintain required fluid velocities through the system, as described in Chapter 18. A significant thing to note about the media for all of the SAGBs is the high specific surface area, which is about an order of magnitude higher than the specific surface areas of trickling filter or rotating biological contractor (RBC) media. This suggests that they can carry much more biomass, thereby allowing smaller HRTs to be used, and indeed, this has been their main appeal.

Many SAGBs are operated without a solids separation and recycle system. In such bioreactors, the HRT is less than the minimum SRT required for microbial growth on the substrates provided. By doing this, the growth of suspended microorganisms is minimized and the growth of attached microorganisms is maximized. For these systems, a means for controlled removal of the attached biomass must be provided. The specific means provided varies with system configuration and is discussed in conjunction with each system.

Other SAGBs use both suspended and attached growth biomass. These combined suspended and attached growth (CSAG) systems are illustrated schematically in Figure 21.2 and the characteristics of the media used in them are summarized in Table 21.2. The suspended component of the biomass is maintained by passing the bioreactor effluent through a liquid–solids separation device, such as a clarifier, and recycling the separated biomass to the bioreactor in the same manner as in the activated sludge process. Biomass that grows on the fixed media will occasionally slough off and be incorporated into the suspended biomass. The excess biomass produced is removed as waste solids, just like in the activated sludge process.

Heterotrophic microorganisms grow in these systems if biodegradable organic matter is present in the influent wastewater, but the type of metabolism they exhibit depends on the terminal electron acceptor provided. Aerobic metabolism predom-

Table 21.1 Packed and Fluidized Bed Reactors

Process	Media characteristics			Typical hydraulic loading rate, m/hr		Porosity %	Comments
	Type	Size mm	Specific surface area m^2/m^3	Average	Peak		
Packed bed downflow	Fired clay	2–6	1,000–1,500	2–3	5–6	40	Hydraulic loading rates vary with application. Specific surface area excludes internal micropores.
	Rounded sand	2–3	1,500	2–3	5–6	40	Hydraulic loading rates vary with application.
Upflow	Fired clay	2–6	1,000–1,500	2–3	5–6	40	Hydraulic loading rates vary with application. Specific surface area excludes internal micropores.
	Polystyrene beads	2–3	1,000–1,100	2–3	5–6	40	Hydraulic loading rates vary with application.
Fluidized bed	Sand	0.3–0.7	2,600–3,900	24 (min.)	36 (max.)	60	Hydraulic loading rate dictated by media size and density.
	GAC	0.6–1.4	2,300–3,300	27 (min.)	37 (max.)	60	Hydraulic loading rate dictated by media size and density. Specific surface area excludes internal micropores.

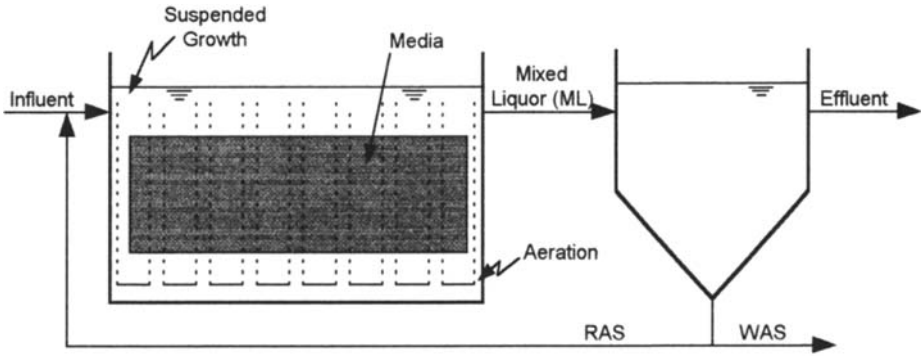


Figure 21.2 Combined suspended and attached growth (CSAG) process.

inates if oxygen is supplied to the system, either dissolved in the influent wastewater or transferred to the bioreactor contents by the addition of air or pure oxygen. Denitrification will occur if dissolved oxygen is not supplied but nitrate-N is provided. Anaerobic processes will occur if both oxygen and nitrate-N are excluded and suitable organic matter is available. Anaerobic systems are discussed in Chapter 13.

Nitrogenous compounds in the wastewater can also be transformed if appropriate environmental conditions are established in the SAGB. Influent ammonia-N can be nitrified if sufficient oxygen is supplied and the total organic loading (TOL) is low enough for nitrifying bacteria to grow. Alternatively, influent nitrate-N can be denitrified if oxygen is excluded and organic matter is available. The development of a microbial community capable of achieving biological phosphorus removal has also been demonstrated for at least one bioreactor configuration.[10] Thus, a wide range of biological conversions can be accomplished in these reactors.

21.1.2 Downflow Packed Bed Bioreactors

Figure 21.3 provides a schematic diagram of a downflow packed bed (DFPB) bioreactor. Wastewater is applied to the top and passes downward through the media to an effluent collection system. Relatively small diameter media and modest superficial velocities are used, as indicated in Table 21.1, which prevents biomass from being detached and transported through the media bed. These characteristics allow DFPB bioreactors to function as filters and to remove particulate matter contained in the wastewater. As a result, these bioreactors provide both biological treatment and filtration. Because suspended solids contained in the wastewater will accumulate in the bioreactor, they must periodically be removed by backwashing. Backwashing also removes microorganisms that grow within the media, thereby controlling the biomass inventory. As a result of the need for backwashing, the effluent collection system is designed much like the underdrain system of conventional granular media filters. This means that it must both collect effluent and distribute backwash water uniformly over the bottom of the media bed. Backwash collection facilities are also required.

Table 21.2 Example Media for Combined Suspended and Attached Growth (CSAG) Processes

Media	Media dimensions	Volume % of aeration basin	Biomass/activity	Comments
Reticulated polyurethane Foam pads		10–30	5–20 g/L of pad volume; 5 mg NH_3-N/(day · pad) for nitrification.	Higher biomass concentrations lead to anaerobic conditions.
Captor®	3 cm × 2.5 cm × 2.5 cm			
Linpor®	1.0 to 1.3 cm cubes			
Ringlace®	Looped PVC about 5 mm in diameter, placed 4 to 10 cm apart. Density of 120 to 500 m/m^3.	25–35	3–12 g/m of media length; up to 1.75 kg NH_3-N/(day · 1,000 m) for nitrification.	Low organic loadings lead to worm growth. Worms controlled by anoxia.
Trickling filter media	PVC vertical trickling filter media, 90 to 165 m^2/m^3	25–75 (with or without RAS recycle)	1,500–6,000 g/m^3 of media. See Figure 21.14 for activity.	Can use with or without suspended growth produced by RAS recycle from secondary clarifier.

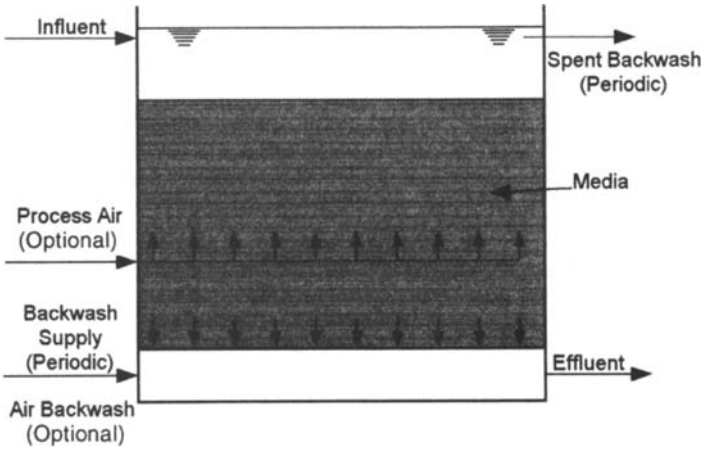


Figure 21.3 Downflow packed bed (DFPB) bioreactor.

Downflow packed bed bioreactors have been used for carbon oxidation, combined carbon oxidation and nitrification, and separate stage nitrification.[22,28,38,40] Oxygen must be supplied to accomplish nitrification, and is often supplied in carbon oxidation systems. This is done by adding air through a distribution system located about two-thirds of the way from the top of the bioreactor. Oxygen transfer can be quite efficient due to the tortuous paths that air bubbles must follow as they pass upward through the packed bed. The addition of air to the bioreactor results in its division into two zones. The upper two-thirds is the aerobic bioreactor section, while the lower one-third functions as a filter. One proprietary version of this bioreactor type uses media with an effective size of 2 to 6 mm that is manufactured from a fired clay.[20] Experience indicates that the porous media formed in the firing process provides internal surface area to retain bacteria in the packed bed during the backwashing process. Activated carbon has also been used as media. Aerated DFPB bioreactors are sometimes referred to as biological aerated filters (BAFs).

Downflow packed bed bioreactors can also be used for denitrification. One proprietary device containing 2 mm diameter rounded sand has proven effective for combined denitrification and suspended solids removal.[37,42] Because both biodegradable organic matter and nitrate-N must be present in the influent wastewater for denitrification to occur, external organic matter must be added when the wastewater has been nitrified in an upstream treatment system, as illustrated in Figure 21.4a. Materials such as methanol,[37,42] fermented wastewater, and fermented solids[1,14] can be used for this purpose. Alternately, effluent from a downstream nitrification process can be recirculated, mixed with influent wastewater, and applied to a DFPB denitrification bioreactor,[8,19,20,40] much in the manner of the modified Ludzack–Ettinger (MLE) suspended growth system, as illustrated in Figure 21.4b. In this case, however, a distinct microbial community will develop in each bioreactor. Nitrogen gas will accumulate in a DFPB bioreactor when it is used for denitrification. This gas must be purged periodically from the bioreactor or it will displace fluid from the interstitial spaces and restrict liquid flow. Purging is often accomplished by temporarily stopping the influent flow and using an air wash cycle.

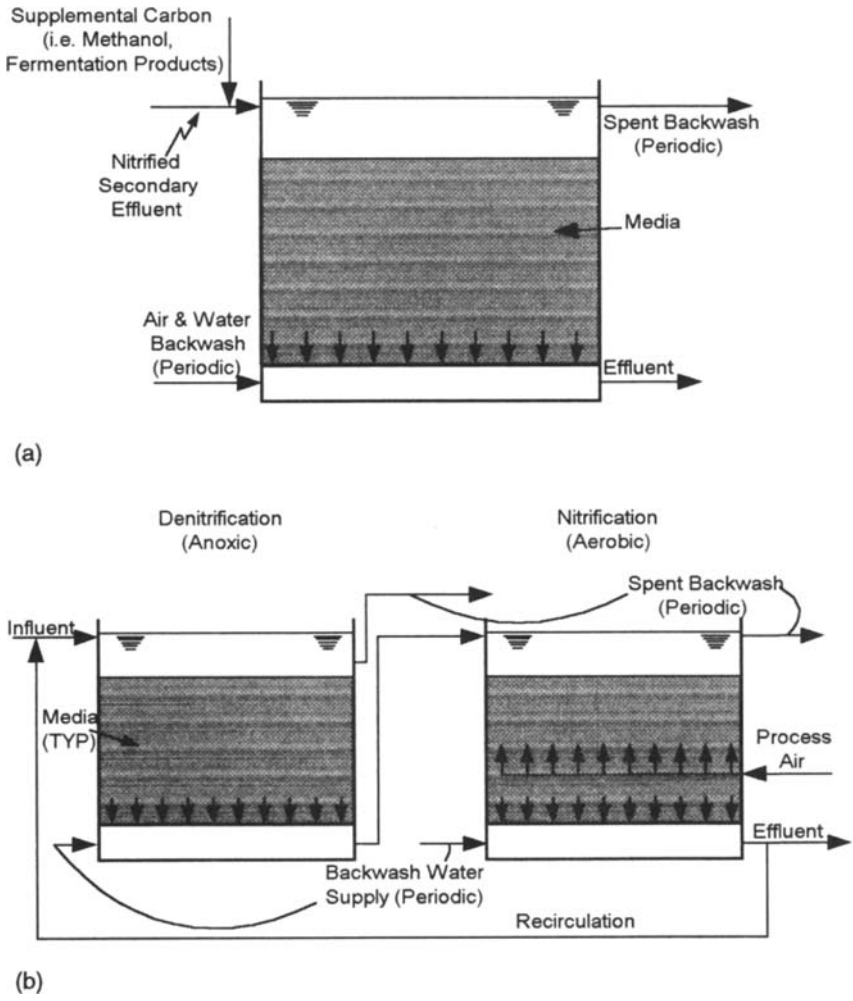


Figure 21.4 Downflow packed bed systems for denitrification.

21.1.3 Upflow Packed Bed Bioreactors

The upflow packed bed (UFPB) bioreactor, illustrated in Figure 21.5, is a more recent development. It is similar to the BAF units described above, except that flow is upward rather than downward. Similar media sizes and superficial velocities are used, as indicated in Table 21.1. Bed depths tend to be between 2 and 4 m. Aeration is provided in the same manner as in downflow BAF bioreactors, thereby creating an aerobic bioreaction zone in the top and a filtration zone in the bottom. Alternatively, aeration can be provided from the bottom, if the entire bed is to be aerobic and used for bioreaction. UFPB bioreactors are used for carbon oxidation, combined carbon oxidation and nitrification, separate-stage nitrification, and combined carbon oxidation, nitrification, and denitrification.[20,34,39,40] The last operational mode is achieved by sizing the aerated zone so that nitrification is achieved and recirculating nitrified

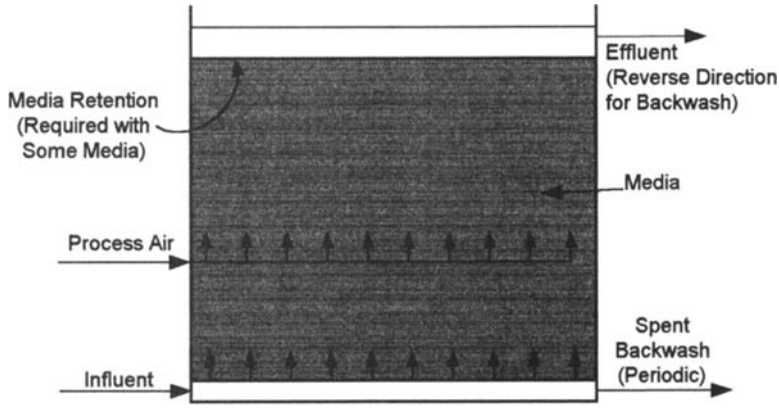


Figure 21.5 Upflow packed bed (UFPB) bioreactor.

effluent back to the bioreactor influent to supply nitrate-N to the lower portion of the bioreactor, thereby converting it into an anoxic zone.

One commercial version of the UFPB system uses the fired clay media described above.[20] This media is heavier than water and rests on an underdrain system. Another commercial system uses 2 to 5 mm plastic beads, which are lighter than water and tend to float.[19] As a result, a grid is required at the top of the bioreactor to retain the media, but an underdrain system is not required.

21.1.4 Fluidized Bed Biological Reactors

Chapter 18 provides a detailed description of fluidized bed biological reactors (FBBRs) and a theoretical analysis of their performance. This section summarizes information about FBBRs to allow their comparison with the other SAGBs. Anaerobic FBBRs are discussed in Chapter 13.

Figure 21.6 provides a schematic diagram of the FBBR process. Process influent flow, consisting of a mixture of wastewater and recirculated effluent, is added to the bottom of the bioreactor through a distribution system and flows upward to a collection system.[6] The process influent flow fluidizes and expands the media bed. A variety of media can be used but, as indicated in Table 21.1, silica sand with a diameter of 0.3 to 0.7 mm and granular activated carbon (GAC) with a diameter of 0.6 to 1.4 mm are used most often. The small carrier particle diameter provides a large specific surface area for biofilm growth. As discussed in Section 18.2.2, the diameter of a bioparticle increases, but its density decreases, as biofilm growth develops on it. The net result is a decrease in its settling velocity, resulting in the migration of particles with the greatest amount of biofilm to the top of the fluidized bed. Therefore, the mass of biomass in the bioreactor is controlled by periodically removing bioparticles from the top of the bed and processing them through a media cleaning system. The media cleaning system typically consists of a pump (where turbulence shears the attached biofilm) and a liquid–solids separation device (such as a cyclone) where the media is separated from the biomass. The removed biomass is directed to solids processing, while the cleaned media is returned to the bioreactor.

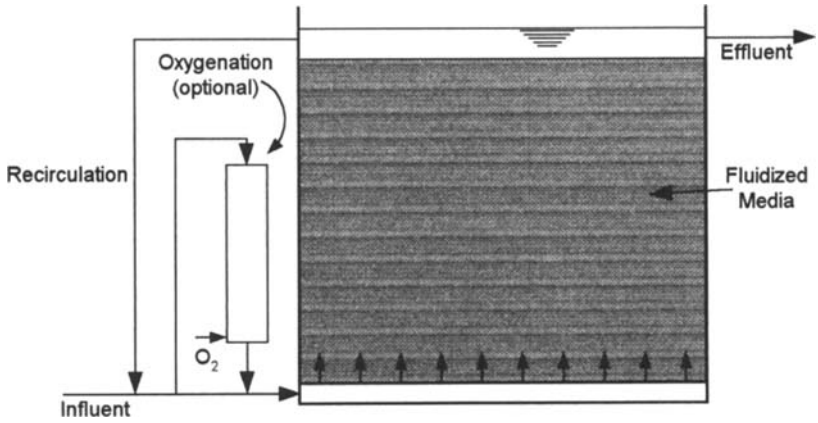


Figure 21.6 Fluidized bed biological reactor (FBBR).

Fluidized bed biological reactors can be applied to aerobic carbon oxidation,[6] combined carbon oxidation and nitrification,[15] separate-stage nitrification,[7,11] denitrification,[37] and anaerobic treatment.[29,31] For aerobic applications, a portion of the process influent flow is processed through a pressurized oxygen transfer vessel where it is saturated with pure oxygen, as illustrated in Figure 21.6. No oxygen is supplied for anoxic applications. However, electron donor must be supplied either by the influent wastewater or by addition of a supplement like methanol. For a wastewater containing both biodegradable organic matter and ammonia-N, combined nitrification and denitrification using the wastewater organic matter will require two FBBRs operating in series in the same manner as described for DFPB bioreactors and illustrated in Figure 21.4b. The first FBBR is anoxic and receives influent wastewater and recirculation from the downstream aerobic FBBR. Anaerobic operation of an FBBR requires the exclusion of both oxygen and nitrate-N.

21.1.5 Combined Suspended and Attached Growth Systems

Combined suspended and attached growth (CSAG) systems, illustrated in Figure 21.2, consist of an activated sludge bioreactor with attached growth media added to it.[18] Suspended biomass is removed in the secondary clarifier and recycled to the bioreactor to maintain a desired suspended biomass inventory, just as in the activated sludge process. Biomass also accumulates on the media and provides an additional biomass inventory. Excess attached biomass periodically sloughs from the media and is incorporated into the suspended biomass. Suspended biomass is wasted from the system to maintain either a desired mixed liquor suspended solids (MLSS) concentration or a desired SRT.

The attached biomass contributes both to the removal of biodegradable organic matter and to the production of a suspended biomass with improved settling characteristics, allowing sludge volume indexes (SVIs) on the order of 50 mL/g to be routinely observed.[17] Consequently, the addition of media increases the total biomass in the system in two ways: by providing sites for attached growth and by allowing

an increased MLSS concentration to be maintained through improved settling properties.

A variety of fixed media have been used in CSAG systems, as listed in Table 21.2. The principal media in use in 1997 include plastic sheet trickling filter media (see Figure 19.3), Ringlace,® and polyurethane porous foam pads in processes called Captor® and Linpor.®

Ringlace® is a flexible, looped, rope-like material, constructed of polyvinyl chloride woven into strands, that provides a high surface area for biological growth.[4,18,36,40] It is hung on racks suspended over the air diffusers in the bioreactor. This placement results in circulation of mixed liquor past the media, thereby transporting substrate and dissolved oxygen to the attached biogrowth. A similar media placement is used for plastic trickling filter media.[4,36,40] Figure 21.7 illustrates the placement and use of Ringlace® or trickling filter media in CSAG systems.

In the Captor® and Linpor® processes, polyurethane foam pads are placed in the bioreactor in a free-floating fashion and retained there by effluent screens, as illustrated in Figure 21.8.[4,9,36,40] Even when biomass accumulates inside them, the pads may float on the liquid surface due to their low density. The action of the diffused aeration system causes the pads to circulate within the bioreactor, although there may be a tendency for them to accumulate in the effluent end of the bioreactor. Air lift pumps are often used to recirculate the foam pads to counter this tendency and to maintain a uniform distribution throughout the bioreactor. Other mechanisms, such as a stream of air, are used to remove pads from effluent screens.

The oxygen requirements of both the suspended and attached biomass must be met by the oxygen transfer system installed in the bioreactor. As illustrated in Figure

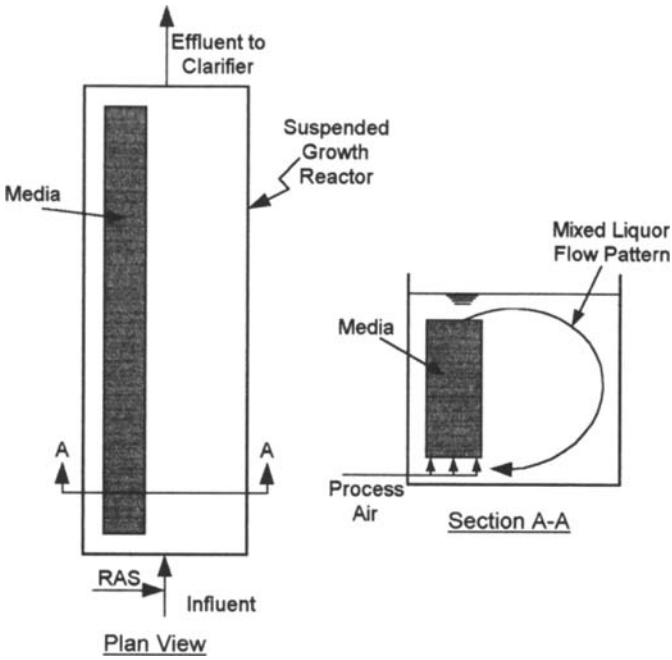


Figure 21.7 Placement of fixed media in combined suspended and attached growth systems.

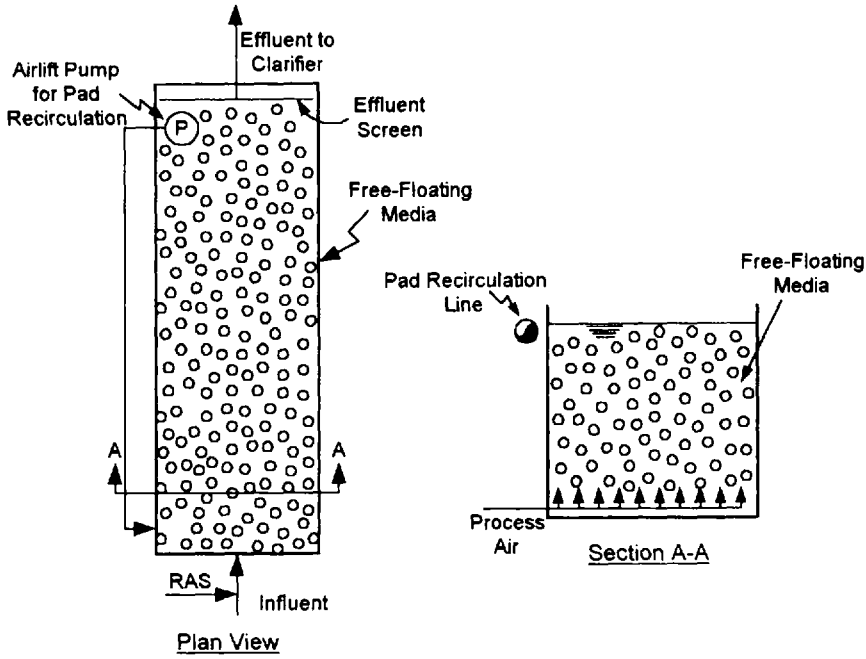


Figure 21.8 Configuration of combined suspended and attached growth system using free-floating media.

21.7, a spiral roll diffuser layout is required to ensure recirculation of mixed liquor through some media, but spiral roll systems are slightly less efficient at oxygen transfer than full floor coverage systems.[35] No evidence exists of enhanced oxygen transfer efficiency as a result of the presence of the attached growth media.

The location in which the media is placed can significantly impacts its effectiveness. Early experience with highly loaded suspended growth systems indicated that the media could be placed uniformly throughout the bioreactor.[17,18] More recent experience with more lowly loaded processes intended to accomplish nitrogen removal suggests that media placement is a critical factor.[26,27] For these systems, a mixed heterotrophic and autotrophic biofilm must develop because heterotrophic growth helps nitrifiers attach to the media. Consequently, the media should be placed in a location where the concentration of biodegradable organic matter has been lowered, but not completely removed, and where ammonia-N is still available.

21.1.6 Other Process Options

Many other attached growth bioreactor options have been evaluated, and some are in commercial development. Some consist of filtration systems that are operated as bioreactors, such as when methanol is fed for denitrification to an upflow moving bed filter.[13,16] Others involve media used in other attached growth systems, such as the submerged biofilter that uses trickling filter media (either rock or plastic) in a liquid-filled bioreactor.[12] Another example is the Surfpac® process in which RBCs are added to activated sludge bioreactors.[40] A wide variety of moving bed bioreactors

Table 21.3 Comparison of Submerged Attached Growth Processes

Process	Benefits	Drawbacks
Downflow packed bed (DFPB)	• Efficient biological oxidation. • Efficient oxygen transfer. • Filtration capability. • Denitrification possible. • Separate liquid–solids separation not required. • Process design basis well established. • Significant full-scale experience.	• Filtration less efficient than in UFPB. • More complex backwash than UFPB. • Less efficient mass transfer than FBBR. • Lower volumetric loadings than FBBR. • Uses conventional equipment (filtration), but in unconventional mode.
Upflow packed bed (UFPB)	• Efficient biological oxidation. • Efficient oxygen transfer. • Filtration capability. • Filtration more efficient than DFPB. • Backwash simpler than DFPB. • Denitrification possible. • Separate liquid–solids separation not required. • Process design basis well established. • Significant full-scale experience, although less than DFPB.	• Less efficient mass transfer than FBBR. • Lower volumetric loadings than FBBR. • Uses conventional equipment (filtration), but in unconventional mode.
Fluidized bed (FBBR)	• Efficient biological oxidation. • Efficient mass transfer. • Highest volumetric loading rates. • Denitrification possible. • Separate liquid/solids separation not required. • Process design basis well established. • Significant full-scale installations.	• Complex oxygen transfer system required for aerobic applications. • Essentially no filtration capability. • Poor removal of particulate substrates. • Long-term operating history somewhat limited.
Combined suspended and attached growth (CSAG)	• Efficient biological oxidation. • Denitrification possible. • Uses conventional wastewater treatment equipment. • Presence of suspended biomass provides excellent removal of colloidal and particulate substrate.	• Requires separate liquid/solids separation step. • No filtration capability. • Volumetric loadings higher than purely suspended growth systems but lower than other attached growth systems. • Process design basis not well established. • Limited full-scale application.

using a number of carrier media have also been considered. The air lift bioreactor is an FBBR that encourages excellent mass and oxygen transfer.[32] The KMT process uses 10 mm diameter porous plastic media in an aeration basin but without RAS recycle from the downstream clarifier.[24,25]

Much work is ongoing in this interesting and important area, and further significant developments are likely to occur. Due to space constraints we were unable to include citations of all relevant work. Consequently, the reader is referred to the proceedings of two International Association on Water Quality (IAWQ) conferences on biofilm processes for additional information.[2,21]

21.1.7 Comparison of Process Options

Table 21.3 summarizes the benefits and drawbacks of the various SAGBs. Both upflow and downflow packed bed bioreactors include biological treatment and filtration, which can be advantageous when a stream containing biodegradable organic matter and suspended solids is being treated. Efficient oxygen transfer occurs because low pressure air flows through tortuous paths within the packed bed. Furthermore, combined carbon oxidation, nitrification, and denitrification is possible, and biological phosphorus removal has also been achieved, demonstrating the versatility of these bioreactors. Some differences exist between the upflow and downflow options. Manufacturers who produce both are currently promoting the upflow option due to superior operating and loading characteristics.[20,33]

Fluidized bed biological reactors provide superior mass transfer characteristics and improved reaction rates for the removal of soluble materials. Oxygenation is more problematic than for packed bed bioreactors because all oxygen must be transferred to the process influent flow, which may be difficult to do for high-strength wastewaters. Combined carbon oxidation, nitrification, and denitrification can be obtained, but this requires the use of bioreactors in series with recirculation from the downstream aerobic FBBR to the upstream anoxic FBBR. Influent suspended solids are not efficiently removed in FBBRs.

Both packed and fluidized bed bioreactors provide the benefits of short HRTs, resulting in small bioreactor sizes. This can be quite important when space is limited and compact bioreactors are needed to achieve treatment goals.

Combined suspended and attached growth systems incorporate many of the benefits of the activated sludge process along with the shorter HRTs possible with attached growth bioreactors. These benefits include flexibility, high efficiency, and the capability to achieve a high-quality effluent. The shorter HRTs result in smaller bioreactor sizes than in the activated sludge process. In addition, the attached growth can significantly enhance the settling and thickening characteristics of the suspended biomass.[17] Thus, increased treatment capacity results from both the biomass that grows on the media and the better settling properties of the suspended biomass. A drawback of this technology is that separate clarifiers and solids recycle are required. Another drawback is that performance data and operating experience are currently limited, thereby restricting the range of applications in which it can be used.

21.1.8 Typical Applications

The bioreactors discussed in this chapter have been in use for more than two decades. Initial use of fluidized beds for wastewater treatment can be traced to the 1940s in

England. However, focused development work did not occur until the early 1970s when researchers at Manhattan College in New York City, the U.S. Environmental Protection Agency's Municipal Environmental Research Laboratory in Cincinnati, and the Water Research Centre in Medmanham, England began joint efforts.[30] A summary of work up to 1981 was published by Cooper and Atkinson.[6] Since then more than 80 full-scale FBBRs have been installed in North America and Europe, about two-thirds for industrial wastewater treatment and the remainder for municipal wastewater treatment.[31] Applications include aerobic carbon oxidation, nitrification, denitrification, and anaerobic treatment.

Industrial applications of FBBRs often focus on the removal of organic matter from industrial wastewaters or contaminated groundwater. They have also been used to pretreat wastewaters prior to treatment at the industrial site or in a municipal wastewater treatment plant. Operational problems with early municipal carbon removal applications have restricted the use of FBBRs for that application.[29] However, they are an accepted technology for the denitrification of nitrified municipal wastewaters.[37,42] The use of GAC as the media is well demonstrated at both bench- and pilot-scales, and several full-scale applications are now in place.[31] This modification incorporates the potential for adsorption of more slowly biodegradable, recalcitrant, or inhibitory materials with the benefits of FBBRs for treatment of biodegradable organic matter. In summary, after an extended development period, the use of FBBRs is now rapidly expanding.[30]

Submerged packed bed bioreactors were developed simultaneously with the FBBR, although developments in North America and Europe proceeded relatively independently. Submerged packed bed bioreactors have been used successfully in the United States for tertiary denitrification at municipal wastewater treatment plants.[37,42] They provide the added function of filtration, thereby producing an effluent that meets stringent water reuse requirements. European developments initially focused on use of the technology for carbon oxidation. Early work used GAC as the support media, but commercial applications now generally use other media, as discussed above. GAC proved to be expensive and relatively fragile, leading to a relatively high attrition rate, and adsorption proved to be an insignificant mechanism of organic matter removal. Aerobic packed bed bioreactor technology is now well proven and widely used in Europe, and new developments are focusing on the use of this technology for nutrient removal. Over 200 full-scale applications exist, and more are being added. Several commercial embodiments of the process are currently available.

The CSAG process is also a relatively old technology.[40] In the United States, Hayes first employed an aerated, submerged rock media biofilter which provided surface area for attached growth in an aeration basin. The process was later improved by Griffith, who used flat plates of wood, cement, plastic, or asbestos in a process called submerged contact aeration. During World War II more than 60 plants were constructed to serve military installations, and a number of municipal facilities were also placed in operation. By the late 1950s 37 facilities were in operation in New Jersey.[41] However, interest in the CSAG process subsequently decreased and no significant installations occurred until the develop of new media in the 1970s and 1980s. The Captor process was developed during the 1970s, and other embodiments of this process type were developed during the 1980s. Today only a small number of full-scale applications exist, but significant interest exists in the use of this technology

for upgrading existing suspended growth processes to increase capacity, improve performance, and achieve nutrient removal. Significant experience is expected to develop in the next few years.

21.2 FACTORS AFFECTING PERFORMANCE

Factors affecting the performance of submerged attached growth bioreactors are similar to those affecting the performance of other biological treatment processes. Those particularly important to these bioreactors include the total organic loading, the substrate flux and surface loading, the hydraulic loading, and the SRT.

21.2.1 Total Volumetric Loading

As with other attached growth bioreactors, the total volumetric loading is calculated as the mass of substrate applied to the bioreactor per unit time divided by the media volume. When the substrate is organic matter, the total volumetric loading is the TOL. Consequently, it is calculated with Eq. 19.1.

The performance of packed and fluidized bed bioreactors has traditionally been characterized by graphical relationships between the TOL and the removal efficiency. Figures 21.9 through 21.12 provide examples of such relationships for packed bed bioreactors treating municipal wastewater to achieve a variety of treatment goals. Figure 21.9 illustrates the relationship between the TOL (based on total chemical oxygen demand [COD]) and the effluent total COD concentration for twelve full-scale facilities and demonstrates the typical decrease in effluent quality (increased effluent total COD) observed with increasing TOL.[5] Performance could also be characterized by using the removal efficiency, where a decrease in efficiency would be

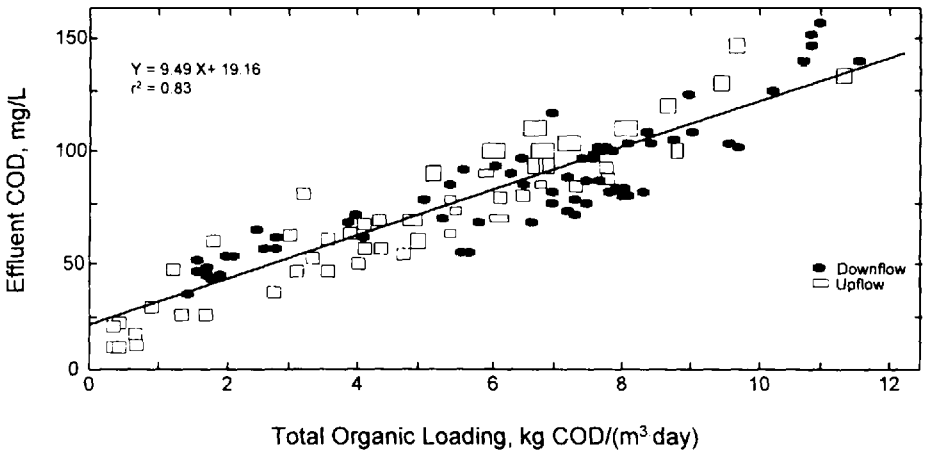


Figure 21.9 Effect of total organic loading (TOL) on effluent COD concentration. The data are from twelve full-scale DFPB and UFPB bioreactors containing fired clay media. (From J. P. Canler and J. M. Perret, Biological aerated filter: assessment of the process based on 12 sewage treatment plants. *Water Science and Technology* **29**(10/11):13–22, 1994. Copyright © Elsevier Science Ltd.; reprinted with permission.)

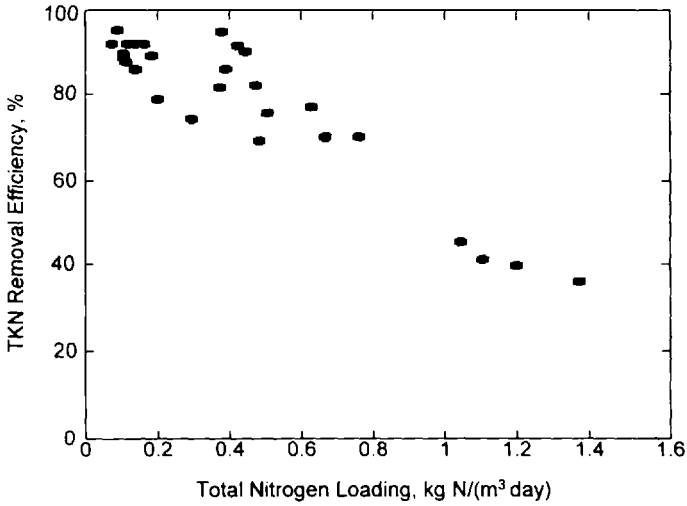


Figure 21.10 Effect of total nitrogen loading (TNL) on TKN removal efficiency during combined carbon oxidation and nitrification (COD/TKN ≈ 10). The data are from twelve full-scale DFPB and UFPB bioreactors containing fired clay media. (From J. P. Canler and J. M. Perret, Biological aerated filter: assessment of the process based on 12 sewage treatment plants. *Water Science and Technology* **29**(10/11):13–22, 1994. Copyright © Elsevier Science Ltd.; reprinted with permission.)

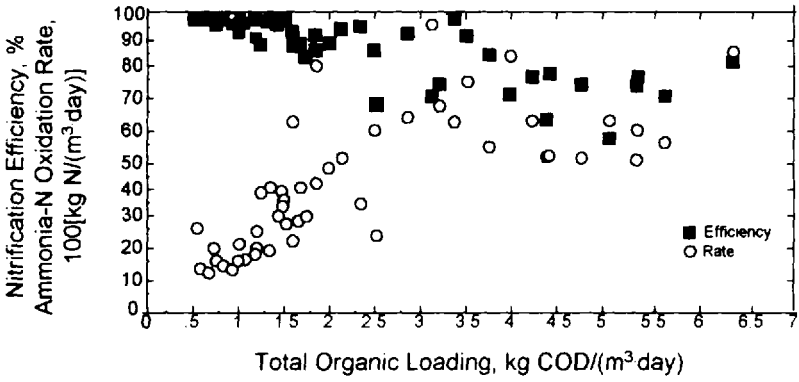


Figure 21.11 Effect of total organic loading (TOL) on nitrification efficiency and volumetric ammonia-N oxidation rate in packed bed bioreactors performing combined carbon oxidation and nitrification. (From R. Rogalla, M. Payraudeau, G. Bacquet, M. Bourbigot, J. Sibony, and P. Filles, Nitrification and phosphorus precipitation with biological aerated filters. *Research Journal, Water Pollution Control Federation* **62**:169–176, 1990. Copyright © Water Environment Federation; reprinted with permission.

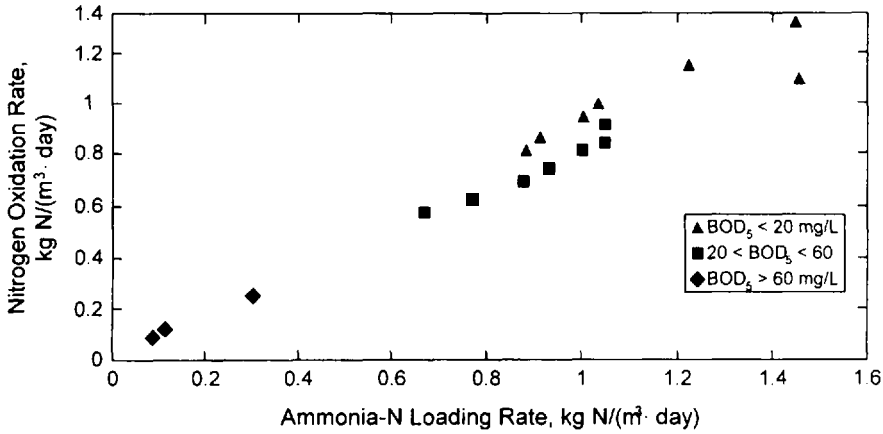


Figure 21.12 Effect of ammonia-N loading (ANL) on the volumetric ammonia-N oxidation rate in packed bed bioreactors. The data are from ten full-scale nitrification plants. (From R. Pujol, M. Hamon, X. Kandel, and H. Lemmel, Biofilters: flexible, reliable biological reactors. *Water Science and Technology* **29**(10/11):33–38, 1994. Copyright © Elsevier Science Ltd.; reprinted with permission.)

observed as the TOL was increased, or by the volumetric removal rate expressed in kg/(m^3·day). In the last case, a linear increase in the volumetric removal rate would initially be observed as the TOL was increased, much as in Figure 20.4. With further increases in TOL, however, the rate of increase in the volumetric removal rate would decrease until a relatively constant, plateau value was reached.

The total nitrogen loading (TNL) is calculated in the same manner as the TOL, except that the influent total Kjeldahl nitrogen (TKN) concentration is used instead of the influent COD or BOD_5 concentration:

$$\Lambda_N = \frac{F(S_{NHO} + S_{NSO} + X_{NSO})}{V_M} \tag{21.1}$$

where Λ_N is the TNL and the sum of S_{NHO}, S_{NSO}, and X_{NSO} is the TKN. Figure 21.10 illustrates the relationship between the TNL and the TKN removal efficiency for the same twelve full-scale facilities considered in Figure 21.9.[5] In this instance, data from periods when the facilities were accomplishing combined carbon oxidation and nitrification are presented. Notice that the TKN removal efficiency is high at low values of the TNL, but begins to deteriorate as the TNL increases. This is exactly as expected when one considers that the TOL was high when the TNL was high. Figure 21.11[23] presents nitrification performance data for another facility accomplishing combined carbon oxidation and nitrification, but in this case ammonia-N removal is presented as a function of the TOL expressed on a COD basis. Both the ammonia-N removal efficiency and the ammonia-N volumetric oxidation (removal) rate are presented. Initially, the volumetric removal rate increases with increasing TOL because as the TOL is increased, so is the TNL. Thus, as long as the TOL is low enough to allow nitrifiers to compete for space in the biofilms, more nitrification will occur as more nitrogen is added. Ultimately, however, as the TOL is increased further the ammonia-N oxidation rate plateaus and declines as the nitrifiers have more dif-

ficulty existing in the system. The ammonia-N removal efficiency parallels the trends in volumetric removal rate. Note that, for a combined carbon oxidation and nitrification application using a particular wastewater, the TOL and TNL are directly related by the COD/TKN ratio of the influent wastewater. Consequently, only one of the two loading rates needs to be specified to define the performance of the bioreactor.

For tertiary nitrification applications, in which the TOL is low enough to allow full growth of nitrifiers in the biofilm and most nitrogen is in the form of ammonia-N, nitrification performance can be correlated with the ammonia-N loading, TAL:

$$\Lambda_{NH} = \frac{F \cdot S_{NHO}}{V_M} \tag{21.2}$$

where Λ_{NH} is the TAL. Figure 21.12[20] illustrates the relationship between the TAL and the volumetric ammonia-N oxidation rate for ten full-scale packed bed nitrification applications. Most were tertiary nitrification applications. The range of TAL values applied in this instance was relatively limited. As a consequence, a linear increase in the volumetric oxidation rate with an increase in the TAL was observed. If a broader range of TALs had been used, the volumetric oxidation rate would have reached a plateau, just as the rate of nitrification did in RBCs (see Figure 20.5)

The relationship between loading rate and process performance will vary with both the application and the bioreactor type. For example, Pujol, et al.[20] indicate the following volumetric loading limitations for UFPB bioreactors using expanded clay media: organic matter removal, 10 kg COD/(m^3·day); nitrification, 1.5 kg NH_3-N/(m^3·day); and denitrification using methanol: >4 kg NO_3-N/(m^3·day). These are maximum loading rates, and lower values may be required depending on effluent requirements. By comparison, a loading rate of 6.4 kg NO_3-N/(m^3·day) or greater can be achieved for denitrification by FBBRs using sand media.[37,42]

The TOL concept can also be applied to CSAG bioreactors, but it is less useful than for bioreactors containing only attached biogrowth. This is because some means is required to equate the attached and suspended biomass within a CSAG bioreactor. Side-by-side comparisons between bioreactors with and without attached growth and/or with different types of media can be made based on the relative TOLs that can be achieved. However, the actual performance is a function of the relative amounts of attached and suspended biomass. When the total biomass concentration and its activity can be quantified, approaches typically applied to suspended growth bioreactors, such as the SRT, may be more useful process loading measures.

21.2.2 Substrate Flux and Surface Loading

Substrate flux and surface loading also influence the performance of a SAGB. Their relevance for characterizing the performance of attached growth bioreactors is discussed in Section 19.2.1 and the relationship between SOL and TOL is presented by Eq. 19.2. Similar arguments and expressions apply for the fluxes and surface loadings of total nitrogen and ammonia-N in nitrifying SAGBs. One of the principal values of fluxes is that they allow comparisons to be made between different types of attached growth bioreactors and media. An example of the use of the ammonia-N flux for this purpose is provided by Figure 21.13 where fluxes for three packed bed

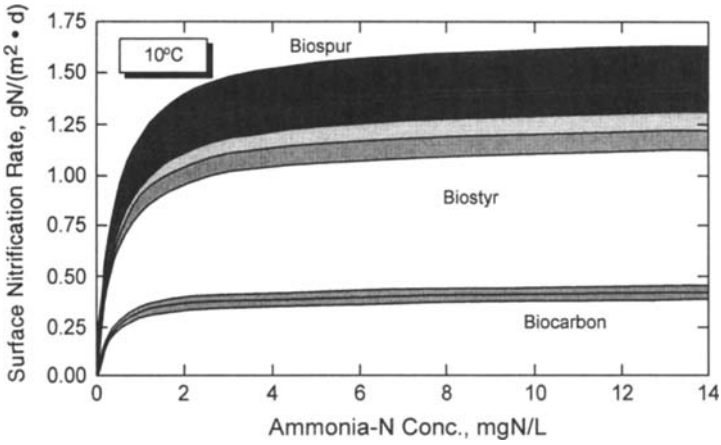


Figure 21.13 Effect of ammonia-N concentration on the surface nitrification rate (ammonia-N biofilm flux) in three types of SAGBs accomplishing tertiary nitrification. The range shown is for aeration rates of 15, 20, and 25 m/hr. (From M. Boller, W. Gujer, and M. Tschui, Parameters affecting nitrifying biofilm reactors. *Water Science and Technology* **29**(10/11):1–11, 1994. Copyright © Elsevier Science Ltd.; reprinted with permission.)

bioreactors using three types of media are compared as a function of the residual ammonia-N concentration.[3] Similar fluxes were obtained in two bioreactors but substantially lower fluxes were observed with the other. Figure 21.14 illustrates the use of a variation on the SOL to correlate the performance of attached biogrowth in CSAG systems.[36] The curves show the removal of BOD_5 by vertical plates in an activated sludge bioreactor as a function of the SOL divided by the HRT. In other words, the abscissa is the SOL per unit of residence time. Since no biomass recycle was used in the system from which these curves were obtained, they can be taken as being representative of the amount of the applied loading to a CSAG system that would be removed by the attached biomass. The empirical nature of this relationship is indicative of the fact that the development of these systems is in its infancy.

21.2.3 Total Hydraulic Loading

The THL for an SAGB, also referred to as the superficial velocity, is calculated in exactly the same manner as the THL for a trickling filter, as given by Eqs. 16.8 and 18.24. However, the THL affects the operation and performance of SAGBs in different ways, depending on the bioreactor type and application. For packed bed bioreactors it represents a constraint since an increase in the THL causes increased headloss, more frequent backwashing, and a larger total volume of backwash water. The specific upper limit depends on the media type, the influent suspended solids concentration, and the desired degree of treatment. For example, Pujol, et al.[20] have suggested the following maximum THLs for UFPB bioreactors using expanded clay media: organic matter removal, 6 m/hr; tertiary nitrification, 10 m/hr; and denitrification using methanol, 14 m/hr. In contrast, for FBBRs a minimum upflow superficial velocity must be maintained to achieve the necessary fluidization of the media, as

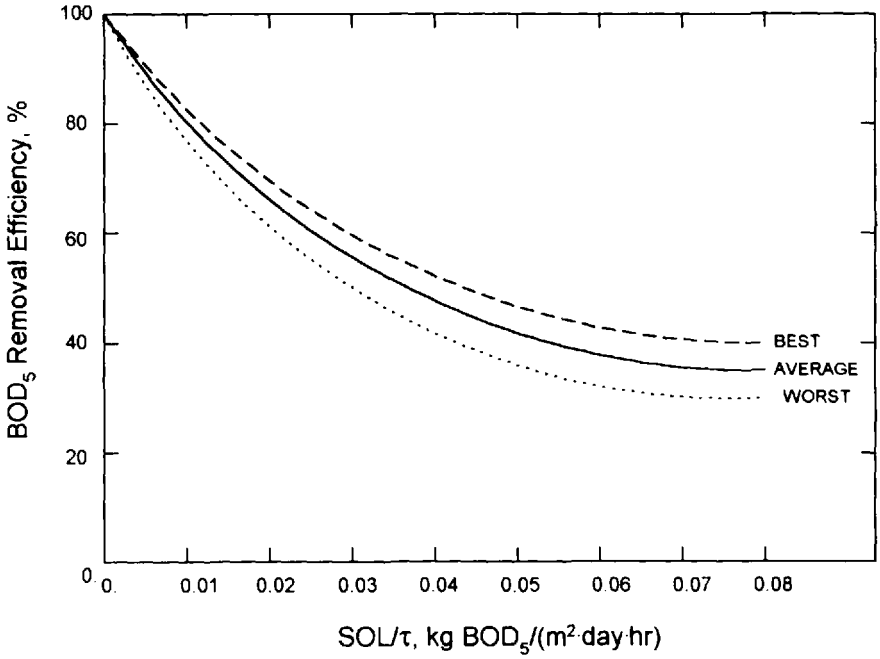


Figure 21.14 Effect of SOL per unit of residence time on the percent of BOD_5 applied that is removed by attached biomass growing on vertical plate media in CSAG systems.[40]

discussed in Section 18.2.1. This value is generally on the order of 35 m/hr for typical sand media. The THL is not a factor for CSAG processes.

21.2.4 Solids Retention Time

It is possible to measure the SRT for some attached growth bioreactors, such as FBBRs using sand media. In such instances, the SRT can be used to quantify and control process performance, just as with a suspended growth process, although it must be recognized that the thickness of the biofilm will determine the effectiveness of the biomass as discussed in Section 18.2.3. For an FBBR, samples of bioparticles can be collected and composited to produce a representative sample of the entire bioreactor. The volatile suspended solids (VSS) concentration in this sample can then be measured, and from it, the total biomass inventory determined. The biomass wasted from the system in both the effluent and by operation of the growth control system can also be measured, and the SRT calculated in the same fashion as for a suspended growth system. When this is done, bioreactor performance can be correlated with the SRT, as illustrated in Figure 21.15.[29] If data on the effect of SRT on effluent quality can be collected for a SAGB, kinetic parameters can be determined for the particular wastewater and bioreactor configuration. It must be borne in mind, however, that the parameters are not intrinsic, but implicitly incorporate the mass transfer characteristics for the particular system. Consequently, they cannot be extrapolated to other systems. Procedures for completing this analysis for FBBRs are

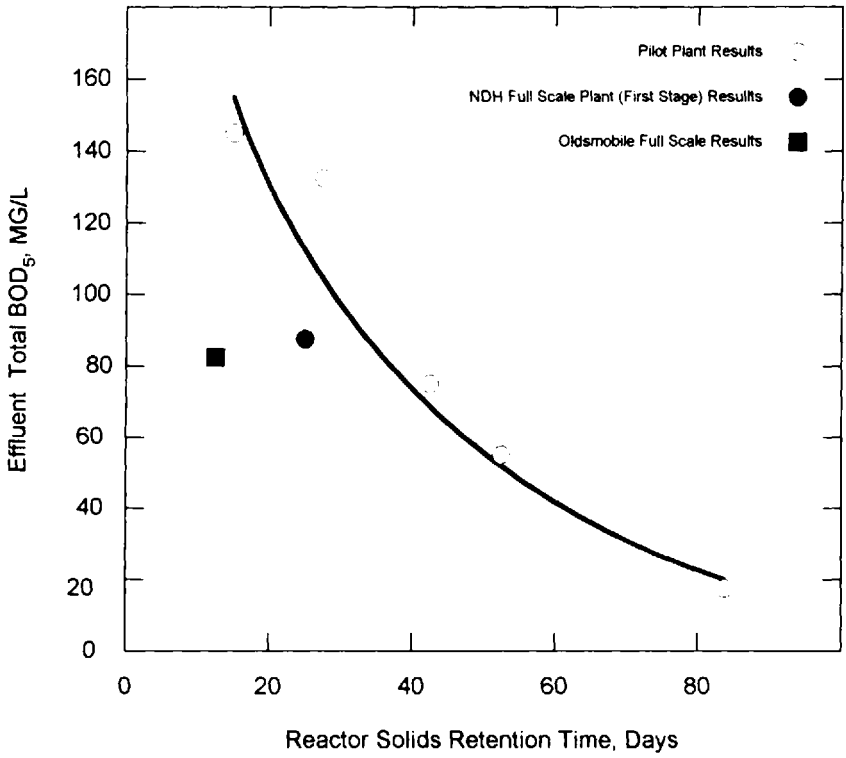


Figure 21.15 Effect of SRT on the total BOD_5 in the effluent from an FBBR treating automotive metal working wastewater. (From P. M. Sutton, *Biological Fluidized Beds for Water and Wastewater Treatment: A User's Forum*, Proceedings of the seminar held at the Sheraton University Inn, Ann Arbor, Michigan, P. M. Sutton & Associates, 1990. Reprinted with permission.)

conceptually the same as those used for suspended growth bioreactors, discussed in Chapter 8.

The SRT can also be used to characterize the performance of CSAG processes. Two values of the SRT can be calculated, one based only on the suspended biomass (suspended growth SRT) and the other based on the suspended plus the attached biomass (total SRT). It must be recognized, however, that the nature of the suspended and attached biomass may be different and, consequently, they may not be easily equated. Experience with CSAG systems indicates that the presence of the attached biomass allows them to achieve effective nitrification with suspended growth SRTs equal to or less than the minimum SRT for the growth of nitrifying bacteria.[9,17,27] The suspended biomass is predominantly heterotrophic and maintains a low bulk liquid phase organic substrate concentration. Since the suspended growth SRT is low, nitrifiers cannot grow in suspension, but they can form effective biofilms because the low organic substrate concentration allows them to compete effectively for space in the biofilm. Consequently, the biofilm provides the nitrifiers required to achieve process stability and to allow the system to successfully process peak ammonia-N

loadings. Experience also indicates that significant denitrification can occur in the biofilm of such systems.

21.2.5 Hydraulic Residence Time

In general, the performance of an attached growth process is not affected by the HRT, as long as the TOL and THL are satisfactory. This is because the amount of biomass in the system is determined by the surface area of media provided, the TOL, and other operating factors, not the length of time that the flow remains in the bioreactor. However, if the HRT exceeds the minimum SRT for biomass growth on the substrate provided, biomass will grow in suspension, which is undesirable in packed beds and FBBRs because the suspended growth will pass into the effluent, increasing its organic matter content and reducing effluent quality. The growth of suspended biomass will also interfere with biofilm development because it reduces the amount of substrate available to the attached biomass. This can be particularly troublesome during startup. This is because low organic loading rates may be needed initially to avoid process overloading as the biofilm develops. If the low TOL is achieved by applying only a portion of the wastewater flow, the resulting HRT may allow significant suspended growth to develop, hampering biofilm development and slowing startup. In such cases, the wastewater should be diluted to allow operation at a reduced TOL while maintaining a short HRT to avoid the growth of suspended biomass.

21.2.6 Other Factors

Like other biological processes, the performance of SAGBs is influenced by a variety of environmental conditions, such as temperature, pH, and dissolved oxygen (DO) concentration. As with trickling filters and RBCs, the importance of mass transport into the biofilms means that the impact of temperature is less significant than for suspended growth systems, until it is sufficiently low ($\sim$15°C) to make the biological reaction rates controlling. Relatively little information is available in the literature on the effects of temperature on substrate removal rates, but they can generally be assumed to be similar to those observed for RBCs (see Section 20.2.4).

When considering other environmental factors, it is important to recognize that the conditions inside the biofilm will be quite different from those in the bulk fluid. The concentrations of electron donors and electron acceptors will change with depth, as illustrated in Figure 15.19. This means that denitrification can occur in the biofilm even though DO may be present in the bulk liquid. The only way to fully analyze these effects is through the use of generalized models, such as those discussed in Part IV, but such models are not yet commonly used in design. Consequently, care must be exercised to ensure that the imposed operational conditions will achieve the desired result. This can be done most effectively with pilot studies.

21.3 PROCESS DESIGN

21.3.1 General Design Procedures

The design of SAGBs is largely empirical in nature. Loading versus performance correlations, such as those illustrated in Figures 21.9 through 21.14 are used to select

a volumetric loading (TOL, Eq. 19.1; TNL, Eq. 21.1; or TAL, Eq. 21.2) or a surface loading (SOL, SNL, or SAL), thereby providing the required bioreactor media volume. The basic relationship between volumetric loading and surface loading is given by Eq. 19.2. The bioreactor configuration is then selected based on minimum and/or maximum hydraulic loading rates and typical media bed depths. This procedure is similar to that used in the design of trickling filters (see Section 19.3). For packed bed bioreactors one constraint is the maximum THL since excessive headloss will occur if the bioreactor is operated at higher THLs. This sets the minimum cross-sectional area and the maximum media depth. For FBBRs a minimum THL must be maintained to fluidize the bed while a maximum THL must not be exceeded to prevent media loss. When coupled with recirculation requirements, these set the maximum and minimum cross-sectional areas, respectively. Procedures for determining the minimum and maximum THLs for FBBRs are presented in Section 18.2. For CSAG systems, the relative amounts of suspended and attached growth must be determined. This determination is generally based on experience with the specific system and application. Volume requirements are then determined in the same manner as for an activated sludge system, as discussed in Section 10.3.

Excess biomass production rates and heterotrophic oxygen requirements are calculated based on net process yields, Y_n, and process oxygen stoichiometric coefficients, Y_{O2}, just as is done in Eqs. 9.3 and 9.4 for the preliminary design of activated sludge systems. However, in some cases significant organic matter will be in the process effluent, requiring those equations to be corrected for its presence. In addition, if nitrification is occurring, the oxygen requirement for nitrification, RO_A, is calculated with a modified form of Eq. 10.16. Since the observed yield of nitrifiers is quite low, the right term within the brackets of that equation is typically ignored, allowing it to be simplified to:

$$RO_A = 4.57F(S_{N,a} - S_{NH}) \tag{21.3}$$

The concentration of nitrogen available to the nitrifiers, $S_{N,a}$, is calculated with Eq. 10.17. Use of that equation requires knowledge of the heterotrophic nitrogen requirement, NR, which for activated sludge systems is calculated with Eq. 5.36. However, in this case, because only the net process yield, Y_n, is known, NR can be estimated with:

$$NR = 0.087Y_n \tag{21.4}$$

Some denitrification applications require the addition of an electron donor, such as, methanol. For these systems, the concentration of methanol, S_{MeOH}, that must be added to remove the total amount of electron acceptor present is calculated from basic stoichiometry:

$$S_{MeOH} = 2.47S_{NOO} + 1.53S_{NO2O} + 0.87S_{OO} \tag{21.5}$$

where S_{NOO} is the influent nitrate-N concentration in mg/L, S_{NO2O} is the influent nitrite-N concentration in mg/L, and S_{OO} is the influent DO concentration in mg/L. The stoichiometric coefficients in Eq. 21.5 are based upon the net biomass yield for growth on methanol as the electron donor and carbon source. They can be calculated using the procedures presented in Chapters 3, 5, and 6. When exact information on the nature of the influent is not available, a methanol dose of 3 mg MeOH/mg NO_x-N is typically used, where NO_x-N represents the sum of the influent nitrate-N

and nitrite-N concentrations. The design of DFPB bioreactors must also accommodate frequent air purge cycles to release nitrogen gas that accumulates in the bioreactor. Procedures for accomplishing these computations are presented elsewhere.[37]

Bioreactor geometry is an important factor for several of the SAGBs and the resulting physical constraints must be considered during design. For example, packed and fluidized bed bioreactors require even flow distribution over the entire cross-sectional area, which is accomplished by nozzles located in either the influent or effluent regions. The physical constraints in accomplishing this distribution are very much like those involved in the design of a granular media filter. One result is a maximum allowable cross-sectional area corresponding to the area over which adequate flow distribution can be achieved. The depth of an aerobic packed bed bioreactor must be adequate to allow the aeration system to meet the oxygen requirements. Furthermore, the depths of both packed and fluidized bed bioreactors must also be sufficient to produce the required effluent quality, which they should do if the loading was appropriately chosen. Ongoing developments by system developers and suppliers are refining these restrictions.[39] Bioreactor geometry is generally less restrictive for the CSAG systems.

21.3.2 Packed Bed Bioreactors

The design of packed bed bioreactors is based on selection of an appropriate TOL and bioreactor configuration consistent with constraints on the maximum THL that can be applied to the bioreactor to avoid excessive headloss and backwash recycle volumes. These principles are illustrated in the following example where the technology is applied for a carbon removal application.

Example 21.3.2.1

Size an UFPB bioreactor using fired clay media to treat a wastewater with a flow rate of 10,000 m^3/day. The wastewater has been pretreated by chemically enhanced primary treatment, removing all suspended solids and leaving a COD concentration of 240 mg/L. The treatment objective is partial removal of organic matter without nitrification. For this application, Pujol et al.[20] recommend a maximum THL of 6 m/hr for the media to be used.

a. What TOL should be used?
Figures 21.9 and 21.11, and the recommendations of Pujol et al.[20] in Section 21.2.1, suggest that a TOL of 10 kg COD/(m^3·day) will allow significant COD removal with minimal nitrification. From Figure 21.9, the effluent COD will be approximately 115 mg/L.

b. What bioreactor volume (media volume) is required?
The media volume can be calculated from the definition of TOL, given by Eq. 19.1, after noting that a TOL of 10 kg COD/(m^3·day) is equivalent to 10,000 g COD/(m^3·day):

$$V_M = \frac{(10{,}000)(240)}{10{,}000} = 240 \text{ m}^3$$

c. Will the THL be acceptable if the bioreactor bed depth is 2 m?
For a 2 m depth, the bed cross-sectional area will be:

$$A_c = \frac{240}{2} = 120 \text{ m}^2$$

The THL is given by Eq. 16.8. Recirculation is not typically employed with a UFPB bioreactor, so:

$$A_H = \frac{10{,}000}{120} = 83.3 \text{ m/day} = 3.5 \text{ m/hr}$$

This is less than 6 m/hr, so it is acceptable.

d. How much oxygen must be provided?
The oxygen requirement can be estimated with Eq. 9.4. For a high-rate application such as this, synthesis oxygen requirements would be exerted but little decay will occur. Therefore, assume a value of 0.5 mg O_2/mg COD removed for Y_{O2}. Because the effluent COD is approximately 115 mg/L, which is a significant fraction of the influent COD, its presence must be accounted for. Therefore;

$$RO = (10{,}000)(0.5)(240 - 115) = 625{,}000 \text{ g/day} = 625 \text{ kg } O_2\text{/day}$$

e. What will the excess biomass production rate be?
From a COD balance, if Y_{O2} has a value of 0.5 g O_2/g COD removed, Y_n must have a value of 0.5 g biomass COD/g COD removed. Therefore, from Eq. 9.3 the excess biomass production rate will be:

$$W_M = (10{,}000)(0.5)(240 - 115) = 625{,}000 \text{ g/day}$$
$$= 625 \text{ kg biomass COD/day}$$

Note that a high TOL is required to avoid nitrification for this application. This high TOL limits the peak application rate to the bioreactor because substrate is already being removed rapidly. Consequently, if the TOL is increased transiently during peak loadings, the effluent substrate concentration will increase proportionally, as indicated in Figure 21.9. In addition, peak loadings to this type of bioreactor will also be limited by the maximum allowable THL of 6 m/hr. Since the design THL was 3.5 m/hr, the maximum flow that the system could handle from a hydraulic perspective would be 6.0/3.5 = 1.71 times the average flow. Thus, it can be seen that designing a system with a high TOL and a high THL limits its ability to handle peak loads.

A lower TOL is required to reliably accomplish nitrification. The following example illustrates the use of an UFPB bioreactor for such an application.

Example 21.3.2.2
Repeat Example 21.3.2.1, but size the bioreactor for nearly complete nitrification. The TKN concentration of the pretreated wastewater is 35 mg/L.

a. What TOL is required in this application?
Examination of Figure 21.11 reveals that the volumetric ammonia-N oxidation rate increases linearly with the TOL up to TOLs of about 3 to 3.5 kg COD/(m^3·day). The nitrification efficiency also remains above 80% up to a TOL of about 2.5 kg COD/(m^3·day). Consequently, select a TOL of 2.5 kg COD/(m^3·day), giving a nitrification rate of 0.6 kg NH_3-N/(m^3·day). From

Figure 21.9, this TOL will produce an effluent COD of approximately 40 mg/L.

b. What bioreactor volume (media volume) is required?
The media volume can be calculated from the definition of TOL, given by Eq. 19.1, after noting that a TOL of 2.5 kg COD/(m^3·day) is equivalent to 2,500 g COD/(m^3·day):

$$V_M = \frac{(10{,}000)(240)}{2{,}500} = 960 \text{ m}^3$$

c. What media depth would be required if a THL of 1.5 m/hr were chosen for this application?
The bed cross-sectional area can be calculated from the definition of THL, as given by Eq. 16.8, after noting that 1.5 m/hr is equivalent to 36 m/day:

$$A_c = \frac{10{,}000}{36} = 278 \text{ m}^2$$

The media depth is just the volume divided by the cross-sectional area:

$$L = \frac{960}{278} = 3.45 \text{ m}$$

This is an acceptable depth and will help ensure adequate oxygen transfer.

d. What is the process oxygen requirement?
For a low-rate process such as this, the process oxygen stoichiometric coefficient will be much greater than in Example 21.3.2.1 because more biomass decay will occur. Thus, to be conservative, use a value of 0.7 mg O_2/mg COD removed for Y_{O_2}. The heterotrophic oxygen requirement can be calculated with Eq. 9.4. Because the effluent COD is approximately 40 mg/L, which is a significant fraction of the influent COD, its presence must be accounted for. Therefore;

$$RO_H = (10{,}000)(0.7)(240 - 40) = 1{,}400{,}000 \text{ g/day}$$
$$= 1{,}400 \text{ kg } O_2\text{/day}$$

To this we must add the oxygen required for nitrification. This can be estimated with Eq. 21.3, which requires estimation of $S_{N,a}$, the nitrogen available to the nitrifiers, and S_{NH}, the effluent ammonia-N concentration. The nitrogen available to the nitrifiers can be calculated with Eq. 10.17, with the use of Eq. 21.4 for NR. To use Eq. 21.4 we must know Y_n. From a COD balance, since Y_{O_2} is 0.7 g O_2/g COD removed, Y_n must be 0.3 g biomass COD/g COD removed. The TKN in the influent is 35 mg/L. Therefore, from Eq 10.17:

$$S_{N,a} = 35 - (0.087)(0.3)(240 - 40) = 30 \text{ mg/L}$$

The effluent ammonia-N concentration can be estimated from a mass balance on nitrogen using the media volume and the volumetric nitrification rate of 0.6 kg NH_3-N/(m^3·day).

$$\text{Input rate} = (10{,}000)(30) = 300{,}000 \text{ g/day}$$
$$\text{Removal rate} = (0.6)(960) = 576 \text{ kg/day} = 576{,}000 \text{ g/day}$$

Since the removal rate exceeds the application rate, nitrification will occur until the ammonia-N concentration becomes rate limiting. Thus, the requirement for almost complete nitrification should be met. To be conservative in calculating the autotrophic oxygen requirement, assume that the effluent ammonia-N concentration is zero. Then the autotrophic oxygen requirement can be calculated with Eq. 21.3:

$$RO_A = (4.57)(10{,}000)(30) = 1{,}370{,}000 \text{ g/day} = 1{,}370 \text{ kg/day}$$

So, the total process oxygen requirement is:

$$RO = 1{,}400 + 1{,}370 = 2{,}770 \text{ kg } O_2\text{/day}.$$

e. What is the excess biomass production rate?
This can be calculated with Eq. 9.3 using the Y_H value of 0.3 g biomass COD/g COD removed that was used above:

$$W_{XH} = (10{,}000)(0.3)(240 - 40) = 600{,}000 \text{ g/day}$$
$$= 600 \text{ kg biomass COD/day}$$

Note the much larger bioreactor volume required and the greater media depth for this application in comparison to the carbon oxidation application. A substantial peak flow could be processed through this bioreactor without experiencing excessive headloss and backwash recycle volumes because the THL is also lower than in the previous case. The oxygen requirement is increased significantly, not only due to the oxygen requirement for nitrification, but because of the increased oxygen requirement for the oxidation of organic matter. Excess biomass production, on the other hand, is not significantly affected. This arises because of the higher degree of stabilization of the removed organic matter.

Designs that accomplish an intermediate level of organic matter removal are also possible. However, it must be recognized that they will also generally achieve at least partial nitrification.

Packed bed bioreactors can also be used for denitrification. The reader is referred to the U.S. EPA *Nitrogen Control Manual*[37] for a complete discussion of the design of such systems.

21.3.3 Fluidized Bed Biological Reactors

Although as presented in Section 18.5, procedures have been developed to optimize the size of an FBBR, many engineers approach their design in a manner similar to that used to design packed bed bioreactors. The first step is to select a design TOL or SOL and use it to calculate the bioreactor bed volume. The primary difference between the design of FBBRs and packed bed bioreactors is that both minimum and maximum values of the THL must be maintained for FBBRs whereas only the maximum THL was constrained in the design of packed bed bioreactors. For FBBRs a minimum THL must be maintained to fluidize the bed, while a maximum THL must not be exceeded to avoid bed washout. Table 21.1 lists typical minimum (in the column labeled "Average") and maximum THLs for FBBRs using sand and GAC media. The values differ only by a factor of 1.5 for sand media, suggesting that they represent the range of reasonable operating conditions, rather than extreme values. Indeed, examination of Figure 18.4 shows that these THLs will give reasonable

fluidization of typical particle sizes when they carry a biofilm of about 100 m thickness. Since this thickness often maximizes the effective biomass concentration, typical operating conditions for FBBRs are close to optimum, thereby justifying the simplified design approach commonly used.

The need to maintain specified THLs requires a trade-off between the FBBR bed depth and the recirculation ratio that is directly analogous to that for trickling filters, as discussed in Section 19.3. For a fixed bioreactor volume, the greater the bed depth the smaller the cross-sectional area and the less the need for recirculation to maintain a desired THL, and vice versa. Unlike trickling filters, FBBR bed depths are constrained primarily by economic considerations rather than by structural constraints. Bed depths in the range of 2 to 6 m are typical, although greater depths can be used. In some cases, maximum bed depths will be used to minimize recirculation. In other cases, reduced bed depths will be required to avoid excessive THLs. The following two examples illustrate these two different cases.

Example 21.3.3.1

An industrial wastewater with a flow rate of 7,500 m^3/day and a nitrate-N concentration of 100 mg/L is to be denitrified in a FBBR. The wastewater is devoid of organic matter so methanol will be added as an electron donor. Pilot studies have suggested that a nitrate loading rate of 6 kg NO_3-N/(m^3·day) will allow complete denitrification to be achieved with methanol in a bed of 0.6 mm sand particles. The THL is to be kept between 24 and 36 m/hr to achieve appropriate fluidization of the media with an optimum biofilm thickness.

a. What fluidized media volume is required?
The nitrate loading rate is equivalent to a TOL. Consequently, the fluidized media volume can be calculated with Eq. 19.1 after recognizing that 6 kg NO_3-N/(m^3·day) is equivalent to 6,000 g NO_3-N/(m^3·day):

$$V_M = \frac{(7{,}500)(100)}{6{,}000} = 125 \text{ m}^3$$

b. Can a bed depth of 6 m be used?
A bed depth of 6 m is permissible as long as it does not result in a THL greater than 36 m/hr without recirculation. Thus, the THL must be calculated with Eq. 16.8 to check it. This requires knowledge of the cross-sectional area. For the 6 m depth,

$$A_c = \frac{125}{6} = 20.8 \text{ m}^2$$

Therefore, the THL is:

$$A_H = \frac{(7{,}500)(1 + 0)}{20.8} = 360 \text{ m/day} = 15 \text{ m/hr}$$

This is less than the minimum THL, which suggests that a 6 m depth can be used, but that recirculation will be required to achieve the minimum THL.

c. What recirculation flow rate is needed to reach the minimum THL?
A THL of 24 m/hr is equivalent to 576 m/day. Using this in a rearranged form of Eq. 16.8 gives:

$$\alpha = \frac{(576)(20.8)}{7{,}500} - 1 = 0.60$$

Since the flow rate is 7,500 m^3/day, the recirculation flow rate required to maintain the minimum THL is 4,500 m^3/day.

d. How much methanol must be provided daily?
The required methanol concentration can be calculated with Eq. 21.5:

$$S_{MeOH} = (2.47)(100) = 247 \text{ mg/L}$$

The mass of methanol required daily is just the concentration times the flow rate or:

$$\text{Methanol requirement} = (7{,}500)(247) = 1{,}850{,}000 \text{ g/day}$$
$$= 1{,}850 \text{ kg/day}$$

In the previous example, recirculation was required even when the tallest feasible FBBR was used. This was due in part to the high concentration of nitrate-N that had to be removed. Treatment of a wastewater with a lower concentration will require a smaller total volume, which will result in a reduced cross-sectional area for a given height. The THL will increase as the cross-sectional area is reduced, and a point can be reached at which the THL exceeds the maximum allowable value. At this point, the FBBR cross-sectional area reaches a minimum value and further reductions in bioreactor volume can be achieved only by reducing the bed depth. The following example illustrates the design of a FBBR where the bed depth must be less than the maximum value to keep the THL below the maximum value.

Example 21.3.3.2
Rework Example 21.3.3.1, but with an influent nitrate-N concentration of 25 mg/L.

a. What fluidized media volume is required?
As before, the fluidized media volume can be calculated with Eq. 19.1 after recognizing that 6 kg NO_3-N/(m^3·day) is equivalent to 6,000 g NO_3-N/(m^3·day):

$$V_M = \frac{(7{,}500)(25)}{6{,}000} = 31.25 \text{ m}^3$$

b. Can a bed depth of 6 m be used?
A bed depth of 6 m is permissible as long as it does not result in a THL greater than 36 m/hr without recirculation. Thus, the THL must be calculated with Eq. 16.8 to check it. This requires knowledge of the cross-sectional area. For the 6 m depth,

$$A_c = \frac{31.25}{6} = 5.21 \text{ m}^2$$

Therefore, the THL is:

$$A_H = \frac{(7{,}500)(1 + 0)}{5.21} = 1{,}440 \text{ m/day} = 60 \text{ m/hr}$$

This exceeds the maximum allowable THL and thus an FBBR with a depth of 6 m cannot be used.

c. What bed depth is required if the THL is restricted to 36 m/hr without recirculation?
The FBBR cross-sectional area can be calculated with Eq. 16.8 by using $\alpha = 0$ and recognizing that 36 m/hr = 864 m/day:

$$A_c = \frac{(7{,}500)(1 + 0)}{864} = 8.68 \text{ m}^2$$

The height can be calculated as:

$$L = \frac{31.25}{8.68} = 3.6 \text{ m}$$

The height exceeds 2 m and is acceptable. If it had been too small it would have been necessary to reduce the TOL to achieve an acceptable size FBBR.

Further details on the design of fluidized bed reactors are presented elsewhere.[29,37,42]

21.3.4 Combined Suspended and Attached Growth Systems

Procedures for the design of CSAG systems are currently evolving. Consequently, the reader is cautioned to consult the most recent literature for updated information and procedures. The design example presented in this section illustrates the types of impacts that have been observed with these systems and the design approaches that have been suggested to deal with them.

Physical design considerations are important. For example, the design of the oxygen transfer system must be integrated with the placement of the attached growth media. The system must be configured to recirculate mixed liquor through fixed media so that dissolved oxygen, soluble organic matter, and ammonia-N are transported to the attached biomass. To date, diffused air systems in spiral roll configurations have been used to accomplish this, as illustrated in Figure 21.7. Baffles may be needed to ensure an appropriate quantity and distribution of mixed liquor flow through the media. The configuration of the oxygen transfer system can also affect the distribution of free floating media such as polyurethane pads, as discussed in Section 21.1.5. The oxygen transfer system must be designed to meet the entire oxygen requirement, i.e., for both the suspended and the attached biomass. Sufficient bioreactor volume must also be available to contain the added media.

The following example illustrates the use of fixed media to upgrade an overloaded activated sludge system. As discussed in Section 21.1.5, the attached biomass contributes both to the removal of organic matter and to the production of suspended biomass with improved settling characteristics. Consequently, addition of fixed media results in an increase in total biomass in the system by two mechanisms: directly by the attached biomass and indirectly by allowing an increased MLSS concentration to be handled by the final settler. The impact of improved solids settling characteristics on treatment capacity is determined first. Then, the media volume required to treat the remainder of the design organic matter loading is determined. Finally, the

required media volume is compared to the entire bioreactor volume to ensure that it can be placed in the existing bioreactor.

Example 21.3.4.1

The activated sludge system in a municipal wastewater treatment plant receives a flow of 10,000 m^3/day containing a BOD_5 concentration of 150 mg/L. The bioreactor volume is 1,000 m^3, giving an HRT of 2.4 hr. As a consequence, the system is overloaded and producing an effluent of poor-quality. Furthermore, the overloading has resulted in an activated sludge with very poor settling characteristics, thereby limiting the MLSS concentration that can be reliably maintained to 1,000 mg/L. This, in turn, has contributed to the problem because the low MLSS concentration results in a process loading factor (F/M ratio) of 1.5 kg BOD_5/(kg MLSS · day) and an SRT of about 1 day. Fixed media is to be installed in the activated sludge bioreactor to increase the biomass in the process and improve effluent quality. The plan is to use modular plastic trickling filter media with a specific surface area of 100 m^2/m^3. Experience indicates that this will allow a higher MLSS concentration to be maintained; a value of 3,000 mg/L is anticipated. The target F/M ratio for suspended biomass to achieve improved solids settleability and effluent quality is 0.3 kg BOD_5/(kg MLSS · day). How much attached growth media will be required to accomplish the goal?

a. How much BOD_5 could the suspended biomass remove at the revised operating condition?
At an MLSS concentration of 3,000 mg/L (3,000 g/m^3) the mass of suspended biomass in the system would be (1,000)(3,000) = 3,000,000 g = 3,000 kg. Consequently, if that suspended biomass were removing organic matter at a rate of 0.3 kg BOD_5/(kg MLSS · day), it could remove (0.3)(3,000) = 900 kg BOD_5/day.

b. How much BOD_5 must the attached biomass remove?
The attached biomass must remove everything not removed by the suspended biomass. The total BOD_5 loading on the process is:

$$(10{,}000)(150) = 1{,}500{,}000 \text{ g/day} = 1{,}500 \text{ kg } BOD_5\text{/day}$$

Therefore, the BOD_5 to be treated by the attached growth is:

$$1{,}500 - 900 = 600 \text{ kg } BOD_5\text{/day}$$

c. How much attached growth media is required to remove this mass of BOD_5?
Figure 21.14 presents performance curves for attached growth media in an activated sludge bioreactor. They represent the percent BOD_5 removal that can be attributed to the attached growth. Since the attached growth in the system under consideration must remove 600 kg/day of the 1,500 kg BOD_5/day applied to the system, the removal efficiency for the attached growth must be 40%. Entering Figure 21.14, with this figure reveals that the SOL per unit of residence time must be 0.06 kg BOD_5/(m^2 · day) per hour of aeration.
For this application, the BOD_5 loading is 1,500 kg BOD_5/day and the HRT is 2.4 hr. Thus, the required media surface area is:

$$A_5 = \frac{1{,}500}{(0.06)(2.4)} = 10{,}400 \text{ m}^2$$

If trickling filter media with a density of 100 m^2/m^3 is used, this corresponds to 104 m^3 of media. This is about 10% of the total bioreactor volume, which is acceptable.

Research is under way to develop procedures for using CSAG systems to upgrade suspended growth bioreactors to achieve nitrification and denitrification. Consequently, it is not yet possible to present a design procedure. Preliminary results suggest that the presence of the attached biomass allows the system to be operated at a suspended biomass SRT that would, by itself, provide only partial nitrification. The presence of the attached biomass allows sufficient nitrifiers to grow to achieve nearly complete nitrification. Thus, the evolving design procedure will likely involve the selection of a suspended growth SRT to achieve partial nitrification and the selection of a fixed media surface area to maintain an adequate additional mass of nitrifiers to achieve complete nitrification. As discussed in Section 21.1.5, preliminary experience also indicates that the fixed media must be placed in a location where soluble, biodegradable organic matter is present to allow heterotrophs to grow on the media as a flocculating agent. Ammonia-N must also be present to allow nitrifiers to grow. The reader is referred to Sen, et al.[26,27] for more details.

21.3.5 General Design Experience

Experience with the application and physical design of SAGBs is accumulating rapidly. This experience indicates that a significant degree of pretreatment is necessary to remove debris and fibrous material from the wastewater prior to applying it to the bioreactor. These materials can clog the inlet or outlet structures, or can become enmeshed in the media. The specific problems associated with each bioreactor configuration and media type differ slightly, but they all suggest the need for good preliminary treatment. For example, clogging of inlet or outlet structures is particularly an issue for packed and fluidized bed bioreactors. Media clogging, on the other hand, is a particular concern in CSAG bioreactors. In the latter case, fibrous material can form into long ropes that can become entangled with the media. Experience indicates that at least primary clarification should be provided for all SAGBs. Furthermore, some applications may require a higher level of pretreatment, such as chemically enhanced primary clarification or high-rate activated sludge. The use of screens in place of primary clarifiers should be regarded with caution due to their poor capture of fibrous materials. If they are used, at least two levels should be provided, with the second level containing relatively small openings.

Packed bed and fluidized bed bioreactors have different abilities for removing particulate organic matter. Packed bed bioreactors can remove these materials through filtration and subsequent biodegradation, whereas particulate organic material will generally pass through a FBBR.

Recent experience has revealed the potential for excessive growths of difficult to remove biomass in some packed bed bioreactor applications. This has occurred in applications with high loading rates, such as the one described in Example

21.3.2.1. Developments in this area should be followed carefully as this technology is applied.

21.4 PROCESS OPERATION

One of the frequently reported advantages of attached growth processes is that they are simpler to operate and require less attention than comparable suspended growth bioreactors. This suggests that operation of an attached growth process will be more focused on mechanical aspects than on process control. Experience with SAGBs confirms this. Process operation generally concentrates on the maintenance of appropriate organic and hydraulic loading rates, which is accomplished by keeping an adequate number of bioreactors in service. Other process requirements must also be met, such as an adequate supply of oxygen for aerobic applications and sufficient quantities of electron donor for denitrification applications. The bioreactor must also be monitored to ensure that it is maintained in proper operating condition. This can be accomplished by monitoring process performance and by visual observation of the bioreactor and its contents.

An issue of particular concern is the control of biomass within a SAGB. Each type of SAGB incorporates mechanisms for maintaining the quantity and activity of attached biomass at appropriate levels. For packed beds, this is accomplished by backwashing. In addition, for denitrification in packed beds, a purge cycle may be required to remove nitrogen gas. For FBBRs, biomass control is accomplished by operation of the growth control system. Biomass control in CSAG processes is accomplished by adjusting aeration rates, which circulate mixed liquor through stationary media, or by passing free-floating media through a cleaning device. Regular monitoring of the biomass concentration and activity is necessary to determine how the biomass control system should be operated. This requires continual evaluation of process efficiency. In addition, for packed bed bioreactors, it requires monitoring hydraulic capacity and headloss; for FBBRs, bed height and biomass development on the carrier particles; and for CSAG processes, observation of the media.

One problem observed with CSAG systems is the growth of worms on the media, which can consume the biofilm and decrease process efficiency.[17,26,27] Problems associated with worm growth can be minimized by maintaining a sufficiently high TOL to sustain active heterotrophic growth. Infestations can also be controlled by periods of anoxia since the worms are obligate aerobes. An anoxic period of twelve hours has proven effective in at least one application.[27]

It is expected that additional process control techniques will be developed as more experience accumulates with these processes.

21.5 KEY POINTS

1. Submerged attached growth bioreactors (SAGBs) provide high biomass densities, resulting in compact bioreactors that require significantly less land area than traditional suspended and attached growth systems.

2. Submerged attached growth bioreactors can be used for aerobic and anaerobic removal of biodegradable organic matter, for nitrification, and for denitrification. Research is underway on the development of SAGBs that will allow phosphorus accumulating microorganisms to grow.
3. Flow through the media in a SAGB can be upward, downward, or horizontal. Furthermore, granular media can be either packed or fluidized. Most SAGBs rely primarily on the attached biomass and are designed and operated with hydraulic residence times (HRTs) sufficiently short to minimize the growth of suspended biomass. However, combined suspended and attached growth (CSAG) processes use clarifiers to recycle suspended biomass, so that both suspended and attached biomass are utilized for treatment.
4. Packed bed bioreactors are usually used to treat wastewaters containing both soluble and particulate organic matter, such as domestic wastewaters. Fluidized bed biological reactors (FBBRs) are usually used to treat wastewaters containing mostly soluble organic matter, such as industrial wastewaters and contaminated groundwaters. Combined suspended and attached growth processes are usually used to upgrade existing suspended growth bioreactors.
5. The performance of SAGBs is often characterized by correlating it with the total organic loading, which is the mass flow rate of biodegradable organic matter divided by the bioreactor media volume.
6. The surface loading can also be used to characterize the performance of SAGBs. This parameter is particularly useful for characterizing the performance of attached media in CSAG processes.
7. The total hydraulic loading (THL) for many SAGBs must be controlled within specified limits. For packed beds, the THL must not exceed certain maximum values to avoid excessive headloss and unreasonable backwash volumes. For FBBRs, the THL must be sufficiently large to maintain the media in a fluidized condition, yet not so large as to wash out the bioparticles.
8. The solids retention time (SRT) can be used to characterize the performance of FBBRs using inert media such as sand, and the suspended growth component of a CSAG.
9. Many factors, such as pH, temperature, and dissolved oxygen concentration affect the performance of SAGBs. Their effects are similar to those observed with other attacheed growth processes.
10. Design procedures for SAGBs use the performance correlations described in items 5 through 8, above. Excess biomass production rates and oxygen requirements are estimated using net process yield factors and oxygen stoichiometric coefficients, similar to those used in the initial design of suspended growth systems.
11. Appropriate pretreatment is required to ensure long-term performance of SAGBs. Clogging of media and/or nozzles can result from debris or fibrous materials present in the wastewater. Pretreatment of municipal wastewater generally consists of primary clarification, either with or with-

out chemical enhancement. A high-rate biological process may also be used.

12. Experience with CSAG processes indicates that placement of the media can be important in determining the effectiveness of the biomass that develops on it.

21.6 STUDY QUESTIONS

1. Prepare a table summarizing the characteristics of the submerged attached growth bioreactors discussed in this chapter. The table should summarize the differences in biomass concentration that can be achieved with the various bioreactors, and the resulting differences in allowable organic loadings. Where possible, provide independent confirmation of the loading differences.
2. Prepare a table summarizing the benefits and drawbacks of SAGBs.
3. Describe how the TOL affects the removal of organic material and nitrification in an aerobic SAGB. Prepare a graph depicting the growth of heterotrophic and autotrophic bacteria in such a bioreactor, along with the removal of biodegradable organic matter and ammonia-N.
4. A wastewater with a flow of 25,000 m^3/day and a biodegradable COD concentration of 300 mg/L must be treated to reduce it to 60 mg/L. An UFPB bioreactor is to be used. Select the appropriate TOL and size the bioreactor. Determine the peak flow rate that can be processed. Determine the oxygen requirement and the excess biomass production rate. Will nitrification occur in this bioreactor?
5. The wastewater described in Study Question 4 must be treated to reduce the ammonia-N concentration to less than 5 mg/L. The initial TKN concentration is 40 mg/L. Evaluate two-stage versus single stage bioreactor configurations. Compare the total bioreactor volume, the oxygen required, and the excess biomass produced for the two options. Summarize the relative advantages and disadvantages of each option.
6. A submerged packed bed system is to be designed to accomplish combined nitrification and denitrification of the wastewater described in Study Questions 4 and 5. Describe the procedure that you would use to size this system.
7. An industrial wastewater with a flow rate of 7,500 m^3/day, a nitrate-N concentration of 150 mg/L, and no significant organic matter, must be treated to remove the nitrate-N prior to discharge to a municipal wastewater treatment plant. Size and configure a FBBR for this purpose. The maximum and minimum practical bed depths are 7 m and 5 m, respectively.
8. Repeat Study Question 7, but with an influent nitrate-N concentration of 50 mg/L.
9. A CSAG bioreactor is to be designed to treat a wastewater with a flow of 25,000 m^3/day and a BOD_5 concentration of 175 mg/L. Plastic trickling filter media with a specific surface area of 140 m^2/m^3 will be used as the

attached growth media, and will occupy a volume equal to 30% of the suspended growth bioreactor volume. The suspended growth process loading factor is to be maintained at less than 0.3 kg BOD_5/(kg MLSS · day), and the MLSS concentration is to be no more than 2,500 mg/L. The performance of the attached growth media can be characterized using Figure 21.14. Size the CSAG bioreactor.

10. Based on your knowledge of the factors that affect the behavior of biofilms, describe those factors that limit the performance of SAGBs. Based on this analysis, how could the performance of these systems be improved?

REFERENCES

1. Aesoy, A. and H. Odegaard, Denitrification in biofilms with biologically hydrolysed sludge as carbon source. *Water Science and Technology* **29**(10/11):93–100, 1994.
2. Bernard, J., ed., *Technical Advances in Biofilm Reactors, Water Science and Technology*, **22**(1/2), 1990.
3. Boller, M., W. Gujer, and M. Tschui, Parameters affecting nitrifying biofilm reactors. *Water Science and Technology* **29**(10/11):1–11, 1994.
4. Boyle, W. C. and A. T. Wallace, *Status of Porous Biomass Support Systems for Wastewater Treatment: An I/A Technology Assessment*, EPA/600/S2–86/019, U. S. Environmental Protection Agency, Cincinnati, Ohio, 1986.
5. Canler, J. P. and J. M. Perret, Biological aerated filter: assessment of the process based on 12 sewage treatment plants. *Water Science and Technology*, **29**(10/11):13–22, 1994.
6. Cooper, P. F. and B. Atkinson, eds., *Biological Fluidized Bed Treatment of Water and Wastewater*, Ellis Horwood Publishers, Chichester, England, 1981.
7. Cooper, P. F. and S. C. Williams, High-rate nitrification in a biological fluidized bed. *Water Science and Technology* **22**(1/2):431–442, 1990.
8. Dee, A., N. James, I. Jones, J. Strickland, J. Upton, and P. Cooper, Pre- or post-denitrification at biological filter works? A case study. *Water Science and Technology* **29**(10/11):145–155, 1994.
9. Golla, P. S., M. P. Reedy, M. K. Simms, and T. J. Laken, Three years of full-scale Captor® process operation at Moundsville WWTP. *Water Science and Technology* **29**(10/11):175–181, 1994.
10. Goncalves, R. F., L. Le Grand, and F. Rogalla, Biological phosphorus uptake in submerged biofilters with nitrogen removal. *Water Science and Technology* **29**(10/11): 135–143, 1994.
11. Green, M. K. and P. J. Hardy, The development of a high-rate nitrification fluidized bed process. *Water Pollution Control* **84**:44–53, 1985.
12. Haug, R. T. and P. L. McCarty, Nitrification with submerged filters. *Journal, Water Pollution Control Federation* **44**:2086–2102, 1972.
13. Hultman, B., K. Jonsson, and E. Plaza, Combined nitrogen and phosphorus removal in a full-scale continuous up-flow sand filter. *Water Science and Technology* **29**(10/11): 127–134, 1994.
14. Jansen, J. la C., S.-E. Jepsen, and K. D. Laursen, Carbon utilization in denitrifying biofilters. *Water Science and Technology* **29**(10/11):101–109, 1994.
15. Jeris, J. S., R. W. Owens, R. Hickey, and F. Flood, Biological fluidized bed treatment for BOD and nitrogen removal. *Journal, Water Pollution Control Federation* **49**:816-831, 1977.

16. Koopman, B., C. M. Stevens, and C. A. Wonderlick, Denitrification in a moving bed upflow sand filter. *Research Journal, Water Pollution Control Federation* **62**:239–245 1990.
17. Lessel, T. H., Upgrading and nitrification by submerged bio-film reactors-experiences from a large scale plant *Water Science and Technology* **29**(10/11):167–174, 1994.
18. Lessel, T. H., First practical experiences with submerged rope-type biofilm reactors for upgrading and nitrification. *Water Science and Technology* **23**(4/6):825–834, 1991.
19. Meaney, B. J. and J. E. T. Strickland, Operating experiences with submerged filters for nitrification and denitrification. *Water Science and Technology* **29**(10/11):119–125, 1994.
20. Pujol, R., M. Hamon, X. Kandel, and H. Lemmel, Biofilters: flexible, reliable biological reactors. *Water Science and Technology* **29**(10/11):33–38, 1994.
21. Rogalla, F. and P. Harremoës, eds., *Biofilm Reactors, Water Science and Technology* **29**(10/11), 1994.
22. Rogalla, F. and J. Sibony, Biocarbone aerated filters-ten years after: past, present, and plenty of potential. *Water Science and Technology* **26**(9/11):2043–2048, 1992.
23. Rogalla, R., M. Payraudeau, G. Bacquet, M. Bourbigot, J. Sibony, and P. Filles, Nitrification and phosphorus precipitation with biological aerated filters. *Research Journal, Water Pollution Control Federation* **62**:169–176, 1990.
24. Rusten, B., L. J. Hem, and H. Odegaard, Nitrogen removal from dilute wastewater in cold climate using moving-bed biofilm reactors. *Water Environment Research* **67**:65–74, 1995.
25. Rusten, B., L. J. Hem, and H. Odegaard, Nitrification of municipal wastewater in moving-bed biofilm reactors. *Water Environment Research* **67**:75–86, 1995.
26. Sen, D., G. D. Farren, R. R. Copithorn, and C. W. Randall, Full scale evaluation of nitrification and denitrification on fixed film media (Ringlace) for design of single-sludge nitrogen removal system. *Proceedings of the 66th Annual Water Environment Federation Conference & Exposition, Volume 3, Liquid Treatment Processes*, pp. 137–148, 1993.
27. Sen, D., C. W. Randall, K. Jensen, G. D. Farren, R. R. Copithorn, T. A. Young, and W. P. Brink, Design parameters for integrated fixed film activated sludge (IFAS) processes to enhance biological nitrogen removal. *Proceedings of the 67th Annual Water Environment Federation Conference & Exposition, Volume 1, Biological Treatment Systems and Biological Nutrient Removal*, pp. 713–724, 1994.
28. Stensel, H. D., R. C. Brenner, K. M. Lee, H. Melcer, and K. Rakness, Biological aerated filter evaluation. *Journal of the Environmental Engineering Division, ASCE* **114**:655–671, 1988.
29. Sutton, P. M., *Biological Fluidized Beds for Water and Wastewater Treatment: A User's Forum*, Proceedings of the seminar held at the Sheraton University Inn, Ann Arbor, Michigan, P.M. Sutton & Associates, 1990.
30. Sutton, P. M. and P. N. Mishra, Biological fluidized beds for water and wastewater treatment. *Water Environment and Technology* **3**(8):52–56, 1991.
31. Sutton, P. M. and P. N. Mishra, Activated carbon based biological fluidized beds for contaminated water and wastewater treatment: A state-of-the-art review. *Water Science and Technology* **29**(10/11):309–317, 1994.
32. Tijhuis, L., E. Rekswinkel, M. C. M. van Loosdrecht, and J. J. Heijnen, Dynamics of population and biofilm structure in the biofilm airlift suspension reactor for carbon and nitrogen removal. *Water Science and Technology* **29**(10/11):377–384, 1994.
33. Toettrup, H., R. Rogalla, A. Vidal, and P. Harremoës, The treatment trilogy of floating filters: From pilot to prototype to plant. *Water Science and Technology* **29**(10/11):23–32, 1994.
34. Tschui, M., M. Boller, W. Gujer, J. Eugster, C. Mader, and C. Stengel, Tertiary nitrification in aerated pilot filters. *Water Science and Technology* **29**(10/11):53–60, 1994.

35. U.S. Environmental Protection Agency, *Design Manual—Fine Pore Aeration Systems*, EPA/625/1-89/023, U.S. Environmental Protection Agency, Cincinnati, Ohio, 1989.
36. U.S. Environmental Protection Agency, *Inert Biomass Support Structures in Aerated Suspended Growth Systems: An Innovative/Alternative Technology Assessment*, EPA/600/X-87/078a&b, U.S. Environmental Protection Agency, Cincinnati, Ohio, 1986.
37. U.S. Environmental Protection Agency, *Nitrogen Control Manual*, EPA/625/R-93/010, U.S. Environmental Protection Agency, Cincinnati, Ohio, 1993.
38. U.S. Environmental Protection Agency, *Technology Assessment of the Biological Aerated Filter*, EPA/600/2-90/015, U.S. Environmental Protection Agency, Cincinnati, Ohio, 1990.
39. Vedry, B., C. Paffoni, M. Gousailles and C. Bernard, First months operation of two biofilter prototypes in the waste water plant at Acheres. *Water Science and Technology* **29**(10/11):39-46, 1994.
40. Water Environment Federation. *Aerobic Fixed Film Reactors (Draft)*, Water Environment Federation, Alexandria, Virginia, 1997.
41. Wilford, J. and T. P. Conlon, Contact aeration sewage treatment plants in New Jersey. *Sewage and Industrial Wastes* **29**:845-855, 1957.
42. Yoder, M. W., T. J. Simpkin, G. T. Daigger, and L. M. Morales, Denitrification trio. *Water Environment & Technology* 7(2):50-54, 1995.

Part VI
Future Challenges

The history of wastewater treatment has been one of responding to need. The first concern of environmental engineers was to protect human health, and early wastewater treatment systems focused primarily on that objective, with some consideration of aesthetics. Next, came the additional concern for the quality of the receiving water, with emphasis on the maintenance of adequate oxygen concentrations. This led to the development of biological treatment systems for the removal of oxygen demanding materials, e.g., organic matter and ammonia-N. Then, as eutrophication became a problem, it became necessary to remove nutrients. The first application of biological systems for this purpose was denitrification. However, as biological processes for the removal of phosphorus became available they began to displace physical/chemical systems because of their advantages. The most recent challenge has been the need to remove individual organic compounds to low levels, rather than just removing organic matter in general. This has been due primarily to the production and use of xenobiotic organic chemicals (XOCs). Although XOCs are foreign to the biosphere, they have many beneficial uses in society. Consequently, just as environmental engineers were given the responsibility of protecting human health and the aquatic environment from conventional pollutants, they have also been given the task of ensuring that XOCs do not enter the environment in excessive quantities through treated wastewaters. This has proven to be a complicated assignment. Part of the difficulty is a result of the very nature of XOCs; because they are foreign to the biosphere they are often hard to biodegrade. Additional difficulties arise because of the low concentrations in which XOCs are commonly present in wastewater and their amenability to removal by other mechanisms. As a way of introducing this topic, which is a major challenge both now and in the future, the last chapter of this book will address the factors influencing the fate and effects of XOCs in bioreactors.

This page intentionally left blank

22

Fate and Effects of Xenobiotic Organic Chemicals

It is clear from the material presented in the preceding chapters that our ability to design bioreactors for oxidation of biogenic organic matter, nitrification, and denitrification is well established. Furthermore, our understanding of biological phosphorus removal is advancing rapidly, leading to better models which will result in improved design procedures. Less well established, however, is our ability to design biological treatment systems for the biodegradation of xenobiotic organic chemicals (XOCs). The term xenobiotic means "foreign to the biosphere" and xenobiotic chemicals are those that have been produced by humankind through our chemical industry. While there is no question that the modern chemical industry has had a large beneficial effect on society, there have also been negative effects associated with the release of XOCs to the environment. One job of the environmental engineer is to minimize those negative effects and keep them at acceptable levels through effective destruction of chemical residues. Biological treatment systems are important tools in that effort. Consequently, in this chapter we will explore some of the factors that must be considered when applying biological treatment systems for the destruction of water-borne XOCs.

22.1 BIODEGRADATION

At first glance, it might seem strange to try and use biological treatment systems to destroy XOCs. After all, the destruction of organic matter in such systems is caused by microorganisms that use the organic matter as a carbon and energy source. The pathways employed for degrading biogenic organic matter have developed over long periods of time and are quite effective. Since XOCs have only been on the earth for a very short time period (most for 50 years or less), how is it possible for microorganisms to possess enzymatic pathways that can act on them? The answer to that question is two-fold. First, most XOCs have some structural similarity to biogenic materials, allowing them to fit into active sites on enzymes that did not evolve with them as substrates. Second, the specificity of enzymes is not exact.[1] While the type of reaction catalyzed by a given enzyme is very specific, the range of substrates upon which it can act is not. In other words, biodegradation of XOCs is largely fortuitous. With this in mind, we will first consider the requirements for biodegradation of an XOC.

22.1.1 Requirements for Biodegradation

Because initiation of biodegradation of an XOC requires the presence of an enzyme that is able to perform a transformation reaction, it follows that biodegradation of the XOC requires the presence of a microorganism with the genetic capability to synthesize that enzyme. Furthermore, if mineralization of the XOC is to occur, the transformation product from the first reaction must serve as the substrate for another transforming enzyme, etc., until ultimately a biogenic product is formed that will funnel into common metabolic pathways. The enzymes catalyzing reactions subsequent to the initial reactions may not reside in the same microorganism as the ones producing the compounds upon which they act, so that mineralization may require the concerted action of a community rather than a single type of organism.

Regardless of where the genes reside, maintenance of the appropriate genetic capability is an absolute requirement for biodegradation to occur. This means that the environmental conditions must be appropriate for growth of the required microorganisms. Not only must the required nutrients and electron acceptor be provided, but the system solids retention time (SRT) must be sufficiently long to maintain all of the microorganisms in the system. If the system SRT is less than that value, the required microorganisms will be lost and biodegradation will cease.[11] This is an important concept because many XOCs are biodegraded only slowly, making the minimum SRT relatively long, as illustrated in Figure 9.3. Furthermore, the biodegradation kinetics of some XOCs are particularly temperature sensitive, causing the minimum SRT to rise rapidly as the temperature drops.[51]

The requirement to maintain the appropriate microbial community can cause serious problems if the XOC is only intermittently discharged to a bioreactor. The capability to degrade any given organic chemical is not necessarily widely distributed in the microbial world. While the ability to use common sugars like glucose and fructose is widespread,[9] other biogenic organic compounds are utilized by only a fraction of all isolates tested.[9,17] Thus, one would anticipate that the ability to use an XOC would be even less widely distributed. When discharge of an XOC to a bioreactor stops, the niche by which the capable microorganisms are maintained may be eliminated, causing them to be lost from the system. The time constant for their loss is the SRT of the system, with the result that about 95% of them will be lost in three SRTs. Thus, it can be seen that most of the microorganisms capable of degrading an XOC can be lost if the XOC is absent from the bioreactor feed for only a few weeks, unless there is some other means of preserving their niche. This means that the capable population must be rebuilt upon reintroduction of the XOC in the influent, causing time lags during which release of the XOC in the effluent occurs.

Even if the capable microorganisms are retained, chances are that the required enzymes will not be synthesized in the absence of the XOC. This follows from the fact that with a few exceptions, most organic compounds are degraded by inducible enzymes, i.e., those that are synthesized only when they are needed. When the inducer is no longer present, their synthesis is stopped and the unused enzymes are degraded, freeing their amino acids for use in the synthesis of new enzymes. Because proteins have a relatively high turnover rate in bacteria, this loss can be quite rapid, causing a culture to lose the ability to degrade an XOC much faster than would be predicted from simple washout of the enzyme.[2] Then, when the inducer is reintro-

duced into the system, new enzyme synthesis must occur, leading to another type of lag. While this lag is considerably shorter than the lag associated with the regrowth of a population, it can still be sufficiently long to allow discharge of an XOC following its reintroduction in the feed to a bioreactor. All of this suggests that the very nature of biological systems dictates that they will be most effective in destroying XOCs when the XOCs are continuously discharged to treatment systems.

22.1.2 Factors Influencing Biodegradation

The primary factor determining the ability of microorganisms to degrade an XOC, as well as the kinetics of that biodegradation, is its molecular structure. The closer that structure is to the structure of a biogenic compound, the easier the XOC will be to biodegrade because the more readily it will fit into common metabolic pathways. Xenophores are substituents on organic molecules that are physiologically uncommon or entirely nonphysiological,[1] and their presence is one factor that can make a compound xenobiotic. Because they are alien to most organisms, they hinder the functioning of many enzymes and it is only the nonspecificity of those enzymes that allows them to function in the presence of xenophores. The nature, number, and position of xenophores all influence the biodegradability of an XOC. Halogens, nitro groups, and cyanide groups are typical xenophores and all reduce biodegradability in comparison to the unsubstituted compound. Furthermore, the greater the number of xenophores, the less susceptible the XOC is to biodegradation. However, it is difficult to generalize about the effect of the position of a xenophore. A given xenophore in one position may have little impact, whereas in another its effect may be large. Similarly, one xenophore in a given position may have a strong effect, whereas another xenophore in the same position may have none. Because of these widely diverse effects, there is strong interest in the development of structure-biodegradability relationships that can be used to deduce biodegradability from a compound's molecular structure.[1] It must be recognized, however, that the knowledge base for such relationships is still limited, thereby restricting their utility at this time. Prediction of biodegradability is a new science and much is still to be learned.

With a few exceptions, microorganisms are thought to degrade only organic compounds that are dissolved in the aqueous phase.[1] Furthermore, as seen in Chapter 3, the rate at which bacteria grow on a substrate is a function of its liquid phase concentration. This means that solubility has a profound effect on the biodegradability of any organic compound, whether xenobiotic or biogenic. Many xenophores reduce aqueous solubility, thereby reducing bioavailability and the rate of biodegradation. Furthermore, if a compound has very low solubility, it may be difficult to induce the enzymes required for its biodegradation.[1]

Finally, the environment has a strong effect on biodegradability. In addition to the usual effects of pH, temperature, and the availability of nutrients and electron acceptor, the presence or absence of molecular oxygen can have a strong effect. For example, some enzymatic steps, such as those carried out by oxygenase enzymes, require the presence of molecular oxygen, whereas others, such as reductive dehalogenation, require its absence. The engineer must be knowledgeable about the nature of the potential pathways for biodegradation of a given XOC so that the appropriate environment can be provided. Failure to do so will result in an inadequate system that cannot meet effluent goals.

22.1.3 Classes of Biodegradation and their Models

Engineers need to quantify biodegradation rates in order to design a biological process capable of achieving a desired effluent concentration of a given XOC. This requires the use of models and the evaluation of the parameters in them. For modeling purposes, biodegradation has been divided into two broad categories, growth-linked and cometabolic.

Growth-Linked Biodegradation. Most biodegradation occurs by growth-linked metabolism. By that we mean that the microorganisms performing the biodegradation receive their carbon and energy from degradation of the XOC and that the XOC can serve as the sole carbon and energy source for microbial growth. As a consequence, the transformation products from the initial biodegradative reactions with the XOC ultimately enter into the normal metabolic pathways wherein some of the carbon is incorporated into new cell material and the remainder is released as carbon dioxide. The result is mineralization of the XOC because it has been converted into those innocuous products. Engineered systems almost always employ natural microbial communities, rather than pure microbial cultures, and so it is not unusual for biodegradation of XOCs to occur by the coordinated action of several species within that community.[50] Such assemblages are called consortia. The models employed are similar to those used elsewhere in this text, which also reflect the activity of microbial communities.

When biodegradation occurs by growth-linked metabolism, the reactions involved may be depicted by the COD-based stoichiometric equation for biomass growth depicted by Eq. 3.8. Consequently, the rates of biomass growth, substrate removal, and oxygen utilization are all linked as expressed in Eq. 3.34. Furthermore, the rate of biomass growth is first-order with respect to the concentration of biomass actually involved in the biodegradation, as depicted in Eq. 3.35. In other words, everything is the same as growth on soluble chemical oxygen demand (COD), except that the heterotrophic biomass concentration must reflect only the concentration of the capable biomass, which will be only a fraction of the total heterotrophic biomass concentration as discussed in Section 22.1.1. The specific growth rate coefficient, μ, is related to the concentration of the XOC through the Monod equation (Eq. 3.36) if the XOC is not inhibitory to its own biodegradation, and by the Andrews equation (Eq. 3.39) if it is. The parameters in these models may be evaluated in batch reactors using the techniques discussed in Section 8.4.

As discussed in Section 8.4.1, two types of kinetic parameter estimates, intrinsic and extant, may be obtained, depending on the type of assay employed and its effect on the physiological state of the bacteria.[25] Intrinsic parameter values are those that reflect the ability of the bacteria to grow in unrestricted growth on an XOC as sole carbon and energy source. They are unaffected by the growth history of the biomass, except as it determines the composition of the microbial community, and are useful for comparing the relative biodegradability of organic compounds. Extant parameter values, on the other hand, reflect the physiological state of the biomass in the bioreactor from which it was obtained. They are most useful for predicting process performance, but their estimation requires knowledge of the concentration of capable biomass. Limited evidence suggests that extant kinetic parameter values measured with a test XOC while it is serving as the sole substrate are the same as those measured while the biomass is simultaneously using a complex mixture of biogenic

substrates at the same specific rate that it uses them in wastewater treatment bioreactors.[16] In other words, extant parameters determined in single substrate tests are adequate for describing biodegradation in the complex environment of a wastewater treatment system.

Extensive databases are not yet available giving intrinsic and extant kinetic parameter values for XOCs. However, available data reveal that, in general, extant parameter estimates are characterized by lower $\hat{\mu}$ and K_S values than intrinsic parameter estimates.[24] For example, intrinsic K_S values tend to be on the order of 1 to 10 mg/L whereas extant K_S values generally lie between 0.1 and 1.0 mg/L.[24] Extant estimates also generally predict lower effluent concentrations from completely mixed activated sludge (CMAS) systems at typical SRTs and are thought to be better predictors of bioreactor performance.[24] However, considerably more research is needed on this topic.

Cometabolic Biodegradation. Cometabolism is the transformation of an organic compound by microorganisms that are unable to use the compound or its transformation products as a source of carbon or energy.[31] Consequently, the microorganisms derive no nutritional benefit from a substrate that they cometabolize, and in fact, the transformation may be detrimental to the cells.[10,12] Cometabolism results from the lack of specificity of enzymes discussed previously.[13,31] In this case, however, the product of the fortuitous reaction is a dead-end product for the microorganism carrying out the reaction. As a consequence, the microorganism must have another substrate that it uses for energy and growth. Cometabolism is easy to demonstrate in pure culture because the transformation product accumulates. In mixed microbial communities, on the other hand, the transformation product may serve as a growth substrate for other types of microorganisms, leading ultimately to the mineralization of the original substrate. Because the microorganism carrying out the cometabolic transformation derives no benefit from it, there is no selective pressure to foster its growth or to retain it in the system. As a consequence, if it is necessary to depend on a cometabolic transformation, an environment must be created that will retain the desired microbe and foster its growth. This can be a major challenge for a design engineer.

Cometabolism of a large number of XOCs has been demonstrated in pure culture.[1,12] The major engineering interest, however, has been in the cometabolic transformation of trichloroethylene (TCE) and other halogenated aliphatic compounds by methanotrophs.[12] This transformation is particularly interesting because it requires reducing power in the form of NADH and also leads to inactivation of the microorganisms. As a consequence, an engineered system must provide an electron donor and/or a growth substrate to replenish the reducing power consumed in the cometabolic reaction and to continually provide new biomass to replace that lost to inactivation. Understanding the reactions involved and expression of them in forms that can be used in conceptual and mathematical models has been a major challenge involving a number of researchers in both engineering and microbiology. Although space does not permit further discussion here, a summary of this interesting topic may be found in the papers of Criddle[12] and Chang and Alvarez-Cohen.[10]

The modeling of cometabolic biodegradation is more complicated than the modeling of growth-linked biodegradation, but is built on the same concepts. Two separate reactions must be modeled, growth on the carbon and energy source and

the cometabolic transformation. Growth on the carbon and energy source is modeled in exactly the same way as any other growth-linked process, using the Monod equation as discussed throughout this text. The cometabolic transformation is usually modeled as an enzymatic reaction using either the Michaelis–Menten equation (analogous in form to the Monod equation) for a noninhibitory substrate or the Haldane equation (analogous in form to the Andrews equation) for an inhibitory one. Competition may occur between the growth substrate and the cometabolic substrate for the shared enzyme, and terms for this competition must be provided in the reaction rate expressions for both processes. In addition, provision must be made for the utilization of the growth substrate to provide reducing power if it is required. This has been handled as an additional decay term. Finally, inactivation of the cells must be modeled if it occurs, and this, too, has been handled as another decay term. Although our attempts at modeling this complex situation are relatively recent, they have been quite successful. Again, the reader is referred to the papers by Criddle[12] and Chang and Alvarez-Cohen[10] for detailed reviews. Procedures are given in them for the estimation of the required kinetic parameters, although the database from their application is small.

22.2 ABIOTIC REMOVAL MECHANISMS

Biodegradation may not be the only mechanism contributing to the loss of an XOC from a bioreactor. Because of the physicochemical properties of some XOCs, abiotic removal mechanisms such as volatilization to the atmosphere and sorption onto solids may contribute to their loss. Quantification of those losses is important for a variety of reasons. First, they represent ways in which the XOCs can enter the environment without alteration. Since XOCs may have negative environmental effects, such losses should be minimized. Second, depending on the configuration of the bioreactor, abiotic losses may influence the concentrations of the XOCs in the effluent and accurate prediction of those concentrations requires that the magnitude of the abiotic losses be known. In this section, we will review briefly techniques for quantifying such losses. The rate expressions presented can be used in mass balance equations for the XOCs along with the rate expression for biodegradation in order to obtain a complete picture of the fate of the XOCs.

22.2.1 Volatilization

Models for Volatilization. Volatilization is an interphase mass transfer process in which a constituent in the liquid phase is transferred to the gas phase and it may contribute to the loss of an XOC from an activated sludge bioreactor during the transfer of oxygen to the system. Models for interphase mass transfer are of the form:

$$r_{v,XOC} = -K_{l,XOC}a(S_{XOC} - S^*_{XOC})V \tag{22.1}$$

where $r_{v,XOC}$ is the rate of loss by volatilization of the XOC from the control volume V (mg/hr), $K_{l,XOC}a$ is the overall liquid phase mass transfer coefficient for the XOC (hr^{-1}), S_{XOC} is the liquid phase concentration of the XOC in the control volume (mg/L), and S^*_{XOC} is the liquid phase concentration that would exist if the liquid were

in equilibrium with the gas phase. Equilibrium between the gas and liquid phases is assumed to be governed by Henry's Law for dilute solutions:[38,48]

$$S^*_{XOC} = \frac{C_{G,XOC}}{H_{c,XOC}} \tag{22.2}$$

in which $C_{G,XOC}$ is the gas phase concentration of the XOC (mg/L) and $H_{c,XOC}$ is the dimensionless Henry's Law coefficient for the XOC. The Henry's Law coefficient in this context is the mass per volume based dimensionless coefficient:

$$[H_{c,XOC}] = \frac{\dfrac{\text{mg of XOC in gas phase}}{\text{L of gas phase}}}{\dfrac{\text{mg of XOC in liquid phase}}{\text{L of liquid phase}}} = \frac{\dfrac{\text{moles of XOC in gas phase}}{\text{L of gas phase}}}{\dfrac{\text{moles of XOC in liquid phase}}{\text{L of liquid phase}}} \tag{22.3}$$

It is related to the conventional Henry's Law coefficient, H_{XOC}, which has units of (atmospheres of XOC in gas phase $\cdot$ m_l^3)/(moles of XOC in liquid phase) by:[42]

$$H_{c,XOC} = \frac{H_{XOC}}{RT} \tag{22.4}$$

where R is the universal gas constant (= 82.06×10^{-6} atm $\cdot$ m^3/mole $\cdot$ K) and T is the absolute temperature (K).

The equation for the mass rate of loss of an XOC from an activated sludge bioreactor by volatilization depends on the type of oxygen transfer system employed. If mechanical surface aeration is used, the value of S^*_{XOC} can be considered to be zero because the atmosphere acts as an infinite sink for the XOC, making $C_{G,XOC}$ equal to zero.[33,42] Consequently, Eq. 22.1 can be simplified to:

$$r_{v,XOC} = -K_{L,XOC}a \cdot S_{XOC} \cdot V \tag{22.5}$$

and the mass loss rate can be considered to be first-order with respect to the liquid phase concentration of the XOC. When oxygen is transferred to an activated sludge bioreactor by diffused aeration, the situation is more complicated.[32,46] As the air bubble rises to the surface from the diffuser, the concentration of the XOC in it continually increases, thereby reducing the driving gradient for transfer. Consequently, the use of Eq. 22.5 would overestimate the loss of the XOC by volatilization. Under that circumstance, the mass removal rate by volatilization is given by:

$$r_{v,XOC} = C_{G,XOC} \cdot Q \tag{22.6}$$

where Q is the volumetric air flow rate through a bioreactor of volume V and:

$$C_{G,XOC} = S_{XOC} \cdot H_{c,XOC} \left[1 - \exp\left(-\frac{K_{L,XOC}a \cdot V}{Q \cdot H_{c,XOC}}\right)\right] \tag{22.7}$$

Estimation of Coefficients. Estimation of the volatile losses of an XOC from an activated sludge bioreactor by either type of oxygen transfer system requires knowledge of the mass transfer coefficient $K_{L,XOC}a$. Measurement of a mass transfer coefficient is time consuming and expensive. Since information is generally available about the overall mass transfer coefficient for oxygen to the bioreactor, $K_{L,O_2}a$, it would be advantageous to be able to estimate the mass transfer coefficient for the XOC from the oxygen mass transfer coefficient, and this is what is commonly done, using an expression of the form:[32,33,42,46]

$$K_{L,XOC}a = \Phi_{XOC} \cdot K_{L,O2}a \tag{22.8}$$

in which Φ_{XOC} is a proportionality factor for a given XOC.

According to the two-resistance theory of Lewis and Whitman,[37] the overall liquid phase mass transfer coefficient is the result of two resistances in series due to the liquid and gas boundary layers:

$$\frac{1}{K_{L,XOC}a} = \frac{1}{k_{L,XOC}a} + \frac{1}{k_{G,XOC}a \cdot H_{c,XOC}} \tag{22.9}$$

Many authors have assumed that the overall liquid phase mass transfer coefficient for any XOC can be approximated by $k_{L,XOC}a$, the liquid film coefficient, as can be done for oxygen, which is a slightly soluble gas. This cannot be done for all XOCs, however, because of their physicochemical characteristics,[32,33,42,46] and consideration must also be given to the gas phase film coefficient, $k_{G,XOC}$. The liquid film coefficients for two solutes may be related to each other by the ratio of their diffusivities in water raised to the n power. The relationship between the overall liquid phase coefficients, however, will also depend on the fraction of the mass transfer resistance for the XOC that can be attributed to the liquid phase.[33,42] As a consequence, Φ_{XOC} can be represented by:[33,42]

$$\Phi_{XOC} = \frac{\left[\frac{D_{w,XOC}}{D_{w,O2}}\right]^n}{\left[1 + \frac{1}{\left(\frac{k_{G,XOC}}{k_{L,XOC}}\right) H_{c,XOC}}\right]} \tag{22.10}$$

in which $D_{w,XOC}$ and $D_{w,O2}$ are the diffusivities of the XOC and oxygen in water, respectively. Caution must be used in the application of Eq. 22.10 because considerable variability exists in the literature regarding the values of both the diffusivity and the Henry's Law coefficient for XOCs. Furthermore, there is uncertainty about the value of n, although it appears to be approximately 0.5, which is in accordance with both the penetration and the surface renewal theories of mass transfer.[42]

The major difficulty in the application of Eq. 22.10 at this time is the lack of a broad base of information about the ratio of the gas to liquid phase film coefficients, $k_{G,XOC}/k_{L,XOC}$. Munz and Roberts[42] and Hsieh et al.[33] have both studied the effect of the power input per unit volume (P/V) for lab-scale mechanical surface aeration systems. Munz and Roberts[42] studied seven XOCs over a power range from 10 to 320 W/m^3 and found that:

$$\text{Log}_{10}\left(\frac{k_{G,XOC}}{k_{L,XOC}}\right) = -0.376\ \text{Log}_{10}\left(\frac{P}{V}\right) + 2.389 \tag{22.11}$$

Hsieh et al.[33] studied twenty XOCs over a power range from 30 to 500 W/m^3 and found that:

$$\text{Log}_{10}\left(\frac{k_{G,XOC}}{k_{L,XOC}}\right) = -1.85\ \text{Log}_{10}\left(\frac{P}{V}\right) + 2.43 \tag{22.12}$$

The difference in the correlations obtained by the two groups of researchers suggests

that the system configuration influences the effect of the power density of mechanical surface aeration on the value of $k_{G,XOC}/k_{L,XOC}$. Nevertheless, both correlations suggest that it is possible to obtain a rough estimate of the overall liquid phase mass transfer coefficient for an XOC from knowledge of the oxygen transfer coefficient, and that may be all that is needed to establish the relative importance of volatilization as a removal mechanism. Hsieh et al.[32] have also studied the effect of the air flow rate per unit volume (Q/V) in a diffused aeration system on the value of $k_{G,XOC}/k_{L,XOC}$. The range of $k_{G,XOC}/k_{L,XOC}$ values was relatively small (2.2–3.6) over a broad range of Q/V, suggesting that a fixed value of 2.6, which was the mean of their observations, could be used for all systems with little error.[32] The small values of $k_{G,XOC}/k_{L,XOC}$ for these systems, however, suggests that gas phase resistance is very important in them and should not be neglected in estimating overall liquid phase mass transfer coefficients for XOCs.

22.2.2 Sorption

Any XOC that is sorbed onto the biomass and other solids in a bioreactor will be removed from the system with the waste biomass. Since the sorbed XOC will not be chemically altered, the potential exists for it to desorb from the waste biomass during its handling, processing, and disposal, potentially leading to release of the XOC to the environment. Because of the potential of such releases, it is important to quantify them so that they may be controlled or eliminated if necessary.

Mechanisms and Models. Sorption of an XOC onto biomass is a complex process, involving both adsorption to the surface of the solids and absorption into cellular components, particularly the lipids. Because the exact mechanism is seldom known, the term "sorption" is typically used to describe the phenomenon and the determination of sorption coefficients is accomplished empirically. Unlike sorption to soil and sediments, which typically display two distinctly different rates with each contributing significantly to the removal,[34] sorption onto biomass is very rapid, with the vast majority of the sorption occurring in a matter of minutes, followed by slow sorption of an additional small amount over a period of hours.[14,55] For example, Wang[60] found that the liquid phase concentration of di-n-butyl phthalate was essentially the same after two minutes of contact with biomass as it was after 72 hours of contact, whereas others have found that equilibrium was approached within an hour.[15,59]

Desorption is the release of a sorbed chemical from the sorbent, and as such, is the opposite of sorption. In some cases, sorption is fully reversible, with the desorption relationship being the same as the sorption relationship. For example, the sorption of di-n-butyl phthalate,[60] lindane,[55] diazinon,[4,5] and 2-chlorobiphenyl[4,5] were all found to be fully reversible. In other cases, sorption may be irreversible, suggesting that chemical reactions are involved,[54] or the reversibility may change over time,[59] suggesting a shift in the relative importance of adsorption and absorption. All of this suggests that the reversibility of sorption will depend on the nature of the chemical and biosolids, and must be determined on a case by case basis. In spite of this, reversibility appears to be common and is often assumed.

A number of models are available to express the equilibrium relationship between the concentration of an XOC in the liquid phase and the quantity of the XOC on the solid phase.[38,48] All are referred to as isotherms because the equilibrium re-

lationship is influenced by temperature, requiring it to be quantified for a fixed temperature. For most purposes, the Freundlich isotherm is adequate:

$$C_{S,XOC} = k_{S,XOC} \cdot S_{XOC}^{n} \tag{22.13}$$

where $C_{S,XOC}$ is the concentration of the XOC on the solid phase, $k_{S,XOC}$ is the sorption coefficient, and n is an empirical coefficient. The units of $k_{S,XOC}$ depend on the value of n and the units of $C_{S,XOC}$ and S_{XOC}. Typically, $C_{S,XOC}$ has units of mg/g and S_{XOC} has units of mg/L. Thus, when n has a value of 1.0, $k_{S,XOC}$ has units of L/g. At the low concentrations at which XOCs are typically present in wastewaters, isotherms are often linear, allowing n to be taken as 1.0.[48] Such a value should not be assumed for all cases, however; particularly for higher XOC concentrations. The best policy is to determine the values of $k_{S,XOC}$ and n experimentally.

The major loss of an XOC due to sorption onto biomass in the activated sludge process comes from biomass wastage. Consequently, the rate of loss by sorption is given by:

$$r_{S,XOC} = -F_w \cdot X_{Mw} \cdot k_{S,XOC} \cdot S_{XOC}^{n} \tag{22.14}$$

where F_w and X_{Mw} are the flow rate and concentration of the wasted mixed liquor from the bioreactor.

Estimation of Coefficients. Although it is preferable to evaluate the coefficients $k_{S,XOC}$ and n experimentally for a given biomass, the procedure is tedious and expensive. Thus, it would be desirable to have a way to estimate them from the literature, particularly for preliminary engineering studies in which only an estimate of the relative importance of sorption as a removal mechanism is needed. Two methods exist. One allows transfer of information on a given XOC from one biomass to another, whereas the other allows extrapolation of information on one XOC to another XOC for the same biomass.

The vast majority of sorption to the biomass in an activated sludge system is to the organic fraction. Since the materials constituting the sorptive organic fractions of various sludges are similar, if a sorption coefficient is expressed per unit of organic carbon it can be used for any type of wastewater solids, even those from different plants.[15] Furthermore, since the mixed liquor volatile suspended solids (MLVSS) concentration is proportional to the organic carbon content of an activated sludge, the same should be true when the sorption coefficient is expressed per unit of MLVSS. Thus, a literature value for the sorption coefficient for a given XOC on activated sludge can be used to approximate the sorption coefficient on biomass from another plant, provided that the coefficients are expressed per unit of organic carbon or MLVSS.[15]

The octanol:water partition coefficient, k_{OW}, is a commonly reported characteristic of organic chemicals and is representative of their hydrophobicity, with larger k_{OW} values indicating more hydrophobic compounds.[48] Because sorption is related to the tendency of an XOC to leave the water phase, it is related to its hydrophobicity. Consequently, the sorption coefficient, $k_{S,XOC}$, is related to the octanol:water partition coefficient. A number of researchers have developed correlations of the type:

$$\text{Log}_{10}\, k_{S,XOC} = a\ \text{Log}_{10}\, k_{OW,XOC} + b \tag{22.15}$$

where $k_{S,XOC}$ is expressed on a per unit carbon basis. Schwarzenbach et al.[48] summarized values for the coefficients a and b for a number of types of XOCs likely to

be found in wastewater. The values of a were around 1.0 (0.81, 0.88, 1.01, and 1.12), suggesting that a value of 1.0 can be assumed with which to estimate the sorption coefficient for one XOC from a measured sorption coefficient for another on the same biomass. If a can be set equal to 1.0, then it follows from Eq. 22.15 that:

$$k_{s,XOC2} = k_{s,XOC1}\left(\frac{k_{OW,XOC2}}{k_{OW,XOC1}}\right) \tag{22.16}$$

Thus, the sorption coefficient for XOC #2 can be estimated from a measured sorption coefficient for XOC #1 using handbook values of the octanol:water partition coefficients for the two XOCs.

22.3 RELATIVE IMPORTANCE OF BIOTIC AND ABIOTIC REMOVAL

The relative contributions of biotic and abiotic removal mechanisms toward the removal of an XOC from a bioreactor can be obtained by substitution of the rate expressions for biodegradation, volatilization, and sorption into the mass balance equations for the XOC in the particular bioreactor type and solving those equations as discussed elsewhere in this text. The resulting output will depend on the configuration of the bioreactor and the method of oxygen transfer employed, and will be specific to each system. Often, however, during preliminary engineering studies it would be advantageous to have a rough estimate of the relative importance of the three removal mechanisms. Such an estimate can be obtained easily for a CMAS system operated with the Garrett flow scheme and being oxygenated by mechanical surface aeration, by combining the rate expressions for the two abiotic removal mechanisms with biodegradation as depicted by the simple, traditional model of Chapter 5, and making a few simplifying assumptions.[23]

In most situations, the concentration of a particular XOC in the influent to an activated sludge system will be only a small fraction of the total biodegradable COD entering the plant. Consequently, loss of some of the XOC due to the abiotic mechanisms will have little, if any, effect on the mixed liquor suspended solid (MLSS) concentration, X_M. This is important because sorption of the XOC will occur on the entire MLSS, not just the capable biomass. Furthermore, based on the discussion in Section 22.1, it is likely that only a small fraction of the biomass in the MLSS will be involved in the biodegradation of the XOC.[7] By assuming that this capable biomass arises only from degradation of the XOC, its concentration, $X_{B,XOC}$, can be calculated as a separate entity. A steady-state mass balance on the capable biomass leads to Eq. 5.7, the familiar expression linking the specific growth rate to the SRT. Substitution of this expression into the Monod equation (Eq. 3.36) leads to the familiar equation for the concentration of a soluble substrate in a CMAS bioreactor:

$$S_{XOC} = \frac{K_{S,XOC}(1/\Theta_c + b_{XOC})}{\hat{\mu}_{XOC} - (1/\Theta_c + b_{XOC})} \tag{5.13}$$

In this case, however, the kinetic coefficients are specific to the XOC. The important point about Eq. 5.13 in this context is that the effluent concentration of an XOC from a CMAS system is determined solely by the system SRT and is independent of the abiotic removal mechanisms. This is an important concept.

The impact of the abiotic mechanisms in a CMAS bioreactor is to reduce the concentration of the biomass capable of degrading the XOC. In this case, the mass balance equation on that XOC must contain three loss terms, rather than the single loss term used in Eq. 5.17. Loss of the XOC by biodegradation is still given by Eq. 3.43, while loss by volatilization and sorption are given by Eqs. 22.5 and 22.14, respectively. Furthermore, since the Garrett flow scheme is being used, the concentration of the waste biomass, X_{M_w}, is equal to the MLSS concentration, X_M. Substitution of these loss terms into the mass balance equation for the XOC and simplification lead to the equation for the capable biomass concentration in the presence of abiotic losses:

$$X_{B,XOC} = \frac{Y_{XOC}}{(1/\Theta_c + b_{XOC})}\left[\frac{1}{\tau}(S_{XOC,0} - S_{XOC}) - K_{L,XOC}a \cdot S_{XOC} - \frac{X_M \cdot k_{s,XOC} \cdot S_{XOC}}{\Theta_c}\right] \tag{22.17}$$

In the absence of abiotic removal mechanisms, both $K_{L,XOC}a$ and $k_{s,XOC}$ are equal to zero, causing Eq. 22.17 to simplify to Eq. 5.19, the familiar equation for the biomass concentration in a CMAS bioreactor:

$$X'_{B,XOC} = \frac{Y_{XOC}}{(1/\Theta_c + b_{XOC})}\left[\frac{1}{\tau}(S_{XOC,0} - S_{XOC})\right] \tag{5.19}$$

where the prime on $X_{B,XOC}$ denotes that it represents the concentration of capable biomass that would result from the biodegradation of the XOC in the absence of abiotic removal mechanisms. Comparison of Eqs. 22.17 and 5.19 makes it clear that the role of abiotic removal mechanisms in a CMAS bioreactor is to reduce the concentration of the capable biomass.

Equations 22.17 and 5.19 can be used to estimate the contribution of abiotic mechanisms to the overall removal of an XOC by a CMAS system. Examination of Eq. 5.19 reveals that in the absence of abiotic removal mechanisms, the capable biomass concentration, $X'_{B,XOC}$, is directly proportional to the removal of the XOC by the bioreactor ($S_{XOC,0} - S_{XOC}$). Furthermore, since the concentration of the XOC in the bioreactor, S_{XOC}, is not affected by abiotic mechanisms and is determined solely by the SRT as denoted in Eq. 5.13, the decrease in the capable biomass concentration associated with the action of the abiotic mechanisms is directly proportional to the amount of the XOC removed by those mechanisms. Thus, if we let $\Delta X_{B,XOC}$ represent the decrease in the capable biomass concentration as a result of the action of the abiotic removal mechanisms, then γ, the fraction of the XOC removal attributable to abiotic removal mechanisms can be calculated as:

$$\gamma = \frac{\Delta X_{B,XOC}}{X'_{B,XOC}} \tag{22.18}$$

The value of $\Delta X_{B,XOC}$ can be obtained by subtracting Eq. 22.17 from Eq. 5.19. Substitution of it into Eq. 22.18 and rearrangement leads to:

$$\gamma = \frac{\alpha_a \left(\frac{S_{XOC}}{S_{XOC,O}} \right)}{\left(1 - \frac{S_{XOC}}{S_{XOC,O}} \right)} \tag{22.19}$$

in which α_a is a dimensionless abiotic loss coefficient. It is made up of the dimensionless volatilization loss coefficient, α_v, and the dimensionless sorption coefficient, α_s:

$$\alpha_a = \alpha_v + \alpha_s \tag{22.20}$$

where:

$$\alpha_v = \tau \cdot K_{L,XOC} a \tag{22.21}$$

and

$$\alpha_s = \frac{\tau \cdot k_{s,XOC} \cdot X_M}{\Theta_c} \tag{22.22}$$

Equation 22.19 lends itself to graphical presentation, as shown in Figure 22.1 where the fraction of XOC removal due to abiotic mechanisms, γ, is plotted versus the fraction of the XOC remaining in the effluent, $S_{XOC}/S_{XOC,O}$, with the abiotic loss coefficient, α_a, as a parameter. Values of α_a from 0.001 to 1000 are provided to cover a broad range of conditions. Examination of Figure 22.1 makes it clear that the smaller the fraction of the XOC remaining in the CMAS effluent, the less important abiotic removal mechanisms are. This follows directly from the fact that the concentration of the XOC in the CMAS effluent is controlled by the SRT as expressed in Eq. 5.13, and from the fact that both volatilization and sorption are directly proportional to the XOC concentration in the CMAS bioreactor. However, Eq. 5.13 also tells us that the concentration of the XOC in the CMAS system is independent of the influent XOC concentration. As a consequence, at a fixed SRT, the lower the influent XOC concentration, $S_{XOC,O}$, the greater the contribution of abiotic removal mechanisms to the overall removal of the XOC, although the total mass of the XOC removed by those mechanisms will be constant.

Equation 22.20 shows that the effects of the two abiotic removal mechanisms are additive. Consequently, the relative importance of each is simply α_v/α_a or α_s/α_a. This can be very helpful in assessing their relative contributions in a given system. For example, *m*-xylene is a volatile XOC that is not very sorptive. In a lab-scale CMAS system, the value of α_v was 76.8 whereas the value of α_s was only 0.007.[23] Consequently, only 0.009% of the abiotic losses were due to sorption. Conversely, di-*n*-butyl phthalate is highly sorptive, but not very volatile. In the same lab-scale CMAS system, the value of α_v was 0.048, whereas the value of α_s was 0.229.[23] Thus, in that case, 82.7% of the abiotic losses were due to sorption. Thus, it can be seen that the relative importance of the abiotic mechanisms depends strongly on the physicochemical properties of the XOC as well as on the nature of the CMAS system.

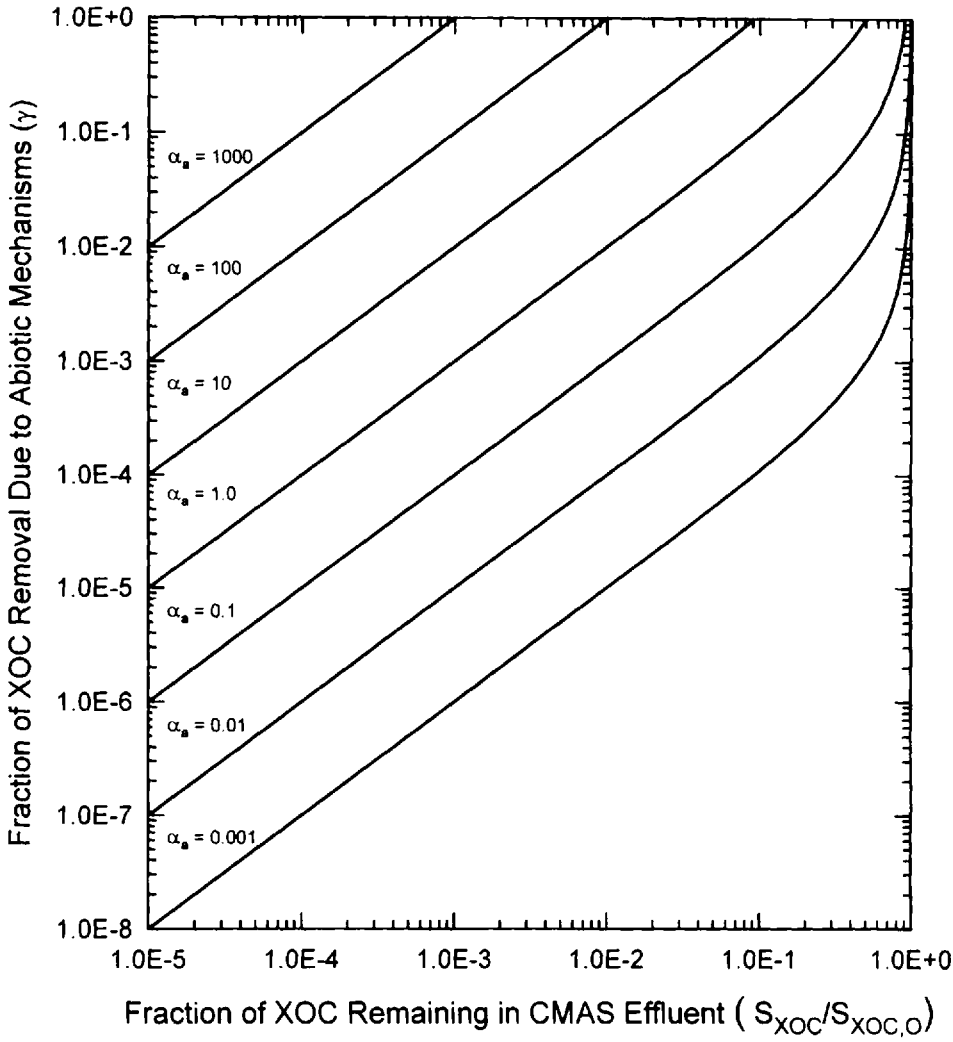


Figure 22.1 Graphical presentation of Eq. 22.19 showing the fraction of XOC removal due to abiotic mechanisms as a function of the fraction of the XOC remaining in the effluent from a CMAS bioreactor. The abiotic loss coefficient, α_a, is defined in Eqs. 22.20–22.22. (From C. P. L. Grady Jr., B. S. Magbanua, S. Brau, and R. W. Sanders II, A simple technique for estimating the contribution of abiotic mechanisms to the removal of SOCs by completely mixed activated sludge. *Water Environment Research*, **69**:1232–1237, 1997. Copyright © Water Environment Federation; reprinted with permission.)

22.4 EFFECTS OF XENOBIOTIC ORGANIC CHEMICALS

So far we have focused on the fate of XOCs in bioreactors. However, XOCs can also affect the performance of a bioreactor by retarding the removal of biogenic organic matter and nitrification. This can occur through inhibition and toxicity. In-

hibition is when the presence of a chemical reduces the rate of a microbial process, such as growth and substrate utilization. Toxicity is when the presence of a chemical causes all microbial activity to cease. Inhibition increases as the concentration of an inhibitor increases and toxicity occurs when the concentration of the inhibitor becomes sufficiently high to stop microbial activity. Currently, inhibition is treated by environmental engineers as if it were fully reversible, but whether this is generally true is unclear. Toxicity, on the other hand, is usually considered to be irreversible, although little evidence exists concerning this either.

22.4.1 Mechanisms and Models for Inhibition and Toxicity

The mechanisms by which XOCs inhibit microbial growth and substrate removal are not well-defined and have only recently become of interest. Consequently, our knowledge of the subject is rather limited. Nevertheless, one would think that the mechanisms are probably the same for inhibition and toxicity, with reaction rates becoming progressively slower as more damage occurs until ultimately the cumulative damage is sufficient to disrupt all activity and we say that toxicity has occurred.

Some XOCs have very specific effects on microbial cells, whereas others have a more general or nonspecific effect. Among the specific effects are radical forming reactions by transition metals, forming hydroxyl radicals that react with a broad range of macromolecules; reactions of organic acids with thiol groups in enzymes, changing their conformation, and thus their reaction rates; formation of covalent bonds with amino acid side-chains in enzymes, altering the conformation of the active site; and interference with protein synthesis. In addition, XOCs that are analogs of biogenic organic compounds can bind irreversibly with the active site on an enzyme, blocking its activity.

A good example of a nonspecific effect is provided by the action of hydrocarbons.[49] Many hydrocarbons are amphiphilic, i.e., they contain both hydrophobic and hydrophilic moieties. As such they behave like the phospholipids that form cellular membranes, making them soluble in those membranes. As hydrocarbons dissolve in the membrane, they disrupt its structural integrity, thereby interfering with its major functions, which are to serve as a barrier separating the cytoplasm from the environment, provide for energy transduction, and provide spatial organization for certain enzymes. At low concentrations, alteration of these functions merely reduces the activity of the cell. However, as the concentration of the hydrocarbon in the environment is increased its concentration in the membrane also increases, thereby causing greater disruption and a greater effect. Ultimately, when the concentration is sufficiently high, the cell stops functioning.

Models for inhibition quantify the effects by altering the values of $\hat{\mu}$ and K_S associated with carbon oxidation or nitrification. Although several have been proposed, the most general is that of Han and Levenspiel,[27] which was given as Eq. 3.42:

$$\mu = \hat{\mu}\left(1 - \frac{S_I}{S_I^*}\right)^n \left[\frac{S_S}{S_S + K_S\left(1 - \frac{S_I}{S_I^*}\right)^m}\right] \tag{3.42}$$

Table 22.1 Types of Inhibition

Inhibition type	Effect on $\hat{\mu}$	Value of n	Effect on K_S	Value of m
Competitive	None	0	Increase	<0
Noncompetitive	Decrease	>0	None	0
Uncompetitive	Decrease	>0	Decrease	>0
Mixed	Decrease	>0	Increase	<0

In this equation, S_I^* is the concentration of the inhibitor, i.e., the XOC, causing all microbial activity to cease. In other words, it is the toxic concentration. The magnitudes of the coefficients m and n can be altered to represent the type of inhibition occurring. Four types are generally modeled,[29,45,56] and they are defined in Table 22.1 in terms of their effects on $\hat{\mu}$ and K_S. Also shown are the characteristics m and n must assume to model the effects. Identification of the inhibitor type is important because it determines the manner in which the substrate and inhibitor concentrations interact in regulating substrate removal and biomass growth, thereby governing the response of a bioreactor to an inhibitory shock load.[47] For example, if an inhibitor acts in a competitive manner it will have no effect when the substrate concentration is high because it does not effect $\hat{\mu}$. On the other hand, a mixed inhibitor is the worst type because it will have a negative effect regardless of the substrate concentration.

22.4.2 Effects of Xenobiotic Organic Chemicals on Carbon Oxidation and Nitrification

Many studies have been conducted in which the concentration of an XOC causing a 50% reduction in microbial activity (IC50) was quantified and these have been tabulated for a variety of microbial groups, including aerobic heterotrophs, nitrifiers, and methanogenic bacteria.[8] Such tables of IC50 values are very helpful for comparing the relative inhibition caused by different XOCs in a semiquantitative manner and they have demonstrated that nitrifying bacteria are much more sensitive than either of the other groups to a broad range of XOCs, with IC50 values being as much as two orders of magnitude lower. Furthermore, they have also shown that methanogens are sometimes more sensitive than aerobic heterotrophs, but not always. The disadvantage of IC50 values is that they cannot be translated directly into a quantitative effect on the kinetic parameters in the Monod equation.[56] Thus, they cannot be used to predict the effect of an inhibitor on the performance of a bioreactor. Unfortunately, relatively few studies have been done to quantify the effects of XOCs on the kinetic parameters in the Monod equation. Procedures are now available by which this can be done,[36,57] so hopefully this situation will soon change.

Volskay et al.[58] studied the effects of 14 XOCs on the kinetics of carbon oxidation by aerobic heterotrophic bacteria. Three of them were found to be uncompetitive inhibitors, one to be a noncompetitive inhibitor, and ten to be mixed inhibitors. Mixed inhibitors are the worst type, because they decrease the value of $\hat{\mu}$ and increase the value of K_S, both of which cause the substrate concentration associated

with a given SRT in an activated sludge bioreactor to increase. The values of the toxic XOC concentration, S_i^*, were relatively high, however, with the smallest measured value being that for tetrachloroethylene at 126 mg/L. This suggests that XOCs must be present at relatively high concentration to have a deleterious effect on aerobic heterotrophs, although mixtures of XOCs will have an additive effect so that the total concentration is what is important.[26] Volskay et al.[58] also found that the values of m and n in Eq. 3.42 were not equal to 1.0, and thus the effects of the XOC on the kinetics of carbon oxidation were nonlinear. Furthermore, in some cases the absolute values of m and n were greater than 1.0, suggesting that the effect increases more and more as the XOC concentration is increased.

Kim et al.[35] studied the effects of three chlorinated phenols and three chlorinated anilines on aceticlastic methanogens and their results were similar to those in the preceding paragraph. All six compounds acted as mixed inhibitors, all had nonlinear effects, and all had relatively high toxic concentrations. The major difference was that the absolute values of m and n were larger (generally greater than 1.0), meaning that the effects increased drastically as the concentration of the XOC was increased.

Most studies on inhibition of nitrification have resulted in IC50 values rather than in specific effects on $\hat{\mu}$ and K_S. Nevertheless, there is a general perception in the environmental engineering literature that XOCs inhibit nitrification in a noncompetitive manner. This perception arises in part from studies that reached that conclusion even though the manner in which the experiments were run did not provide the type of data required to justify it.[43,53] Although others have found that a variety of XOCs do impact nitrification in a noncompetitive manner,[3,19] the amount of data available do not justify a general conclusion about the nature of the inhibition associated with XOCs. It is clear, however, that the oxidation of ammonia-N is much more sensitive to XOCs than is oxidation of nitrite-N.[30] Thus ammonia oxidation must be considered to be the weak link in nitrification.

22.5 EXPERIENCE WITH XENOBIOTIC ORGANIC CHEMICALS

At this time we have about eighty years experience in the design and operation of biological systems for the removal of organic matter from domestic wastewater. As a result, we have had the opportunity to try many things, some of which worked and some of which didn't. Nevertheless, through that experience and the research that it fostered, we have been able to establish the many fundamental principles that are the basis for this book. Consequently, a designer who applies them can be confident that a proposed design will succeed. Our experience with biological nutrient removal systems, on the other hand, dates back about thirty-five years, to the early 1960s. Since then we have learned much about them, allowing generalizations to be made in the form of International Association on Water Quality (IAWQ) activated sludge models Nos. 1 and 2, which are facilitating further research and development. While our experience base is not as old as that for systems focused on removal of organic matter, designers can still have reasonable confidence that a proposed system will function as planned, particularly if its design has been based on pilot studies. In contrast to the above, our experience with XOCs in biological wastewater treat-

ment systems is fairly recent, dating from the late 1970s and early 1980s. This is not to suggest that such chemicals were not present in wastewaters prior to that date. Rather, that was when we began to be concerned with the discharge of individual compounds in treated effluents rather than just with COD or biochemical oxygen demand (BOD). Thus, even though considerable effort has been expended by many researchers, it is not yet possible to reach broad generalizations like those presented in the preceding chapters.[21,22]

Publicly owned treatment works normally receive XOCs at very low concentrations, typically in the μg/L range. Industrial wastewater treatment facilities, on the other hand, often receive them at high concentrations, reaching to hundreds of mg/L. In spite of this broad range, biological treatment systems have been found to be very robust and to do a very good job reducing many types of XOCs to low levels.[21,22] Beyond that, it is difficult to generalize. This difficulty stems from the fact that the ability of a particular bioreactor system to remove a given XOC depends strongly on the configuration of the system, the physicochemical characteristics of the XOC, and the XOC's biodegradation kinetics. In spite of the complexity that suggests, our experience with XOCs is consistent with the principles set forth in the preceding sections of this chapter. For example, compounds like chloroform, trichloroethane, dichlorobenzene, toluene, xylene, and ethylbenzene have high Henry's Law coefficients and low octanol:water partition coefficients. Consequently, we would expect volatilization to be a more important removal mechanism in the activated sludge process than sorption, and that is exactly what happens.[44] Furthermore, chlorine is a more powerful xenophore than methyl and hydroxyl groups, and thus we would expect the halogenated volatile compounds to be removed more by volatilization than by biodegradation, whereas we would expect the reverse for the nonhalogenated ones, and again, this is what happens.[44] On the other hand, polycyclic aromatic hydrocarbons have low Henry's Law coefficients, high octanol:water partition coefficients, and low rates of biodegradation. Thus, we would expect most of their removal to be due to sorption onto solids, and this is what has been oberved.[39] Finally, pesticides like diazinone, 2,4-dichlorophenoxyacetic acid, and 2,4,5-trichlorophenoxyacetic acid tend to be very resistant to biodegradation, and have low Henry's Law coefficients and low octanol:water partition coefficients. Consequently, we would expect them to pass through activated sludge systems with little removal. This, too, has been observed.[41] These examples make it clear that the biodegradability and physicochemical characteristics of an XOC determine its fate. For simple CMAS systems, Eq. 22.19 and Figure 22.1 can assist an engineer in determining what that fate is likely to be. More complex systems require the use of models.[20,40]

The type of bioreactor employed will also influence the fate of XOCs. For example, trickling filters are not as capable of removing nonvolatile XOCs as activated sludge systems, although they have equal capability for the removal of volatile XOCs.[28] This suggests that volatilization may be an important removal mechanism in trickling filters, which one would expect, given their configuration. Although experimental studies on large-scale facilities have not confirmed it, modeling studies have suggested that attached growth bioreactors should be more resistant to inhibitory, but biodegradable, XOCs than suspended growth systems.[18,52] This is because the inner layers in a biofilm experience a lower substrate concentration than the outer layers, which gives a higher reaction rate for an inhibitory substrate. Thus, the overall reaction rate will be higher than the bulk substrate concentration would suggest.

Even within a given type of named biochemical operation, bioreactor configuration can have important effects. Tanks-in-series or plug–flow activated sludge systems have concentration gradients in them whereas CMAS systems are uniform throughout. As a consequence, volatilization is less important as a removal mechanism in the latter.[6]

In summary, although we have not been concerned about individual XOCs for very long, our knowledge base is growing rapidly. Furthermore, the things that we are learning are consistent with the principles of bioreactors presented throughout this text. By applying those principles in practice, progress in this important area should be rapid and we should soon be able to design biological wastewater treatment systems for XOC removal with much more confidence. After all, look how much easier it was to learn to design biological nutrient removal systems than it was to design the first biological treatment systems. Certainly, progress has been made!

22.6 KEY POINTS

1. The ability of microorganisms to biodegrade a xenobiotic organic chemical (XOC) is the result of two things. First, most XOCs have some structural similarity to biogenic materials, allowing them to fit into active sites on enzymes that did not evolve with them as substrates. Second, the specificity of enzymes is not exact.
2. The molecular structure of an XOC is the primary determinant of its biodegradability, but factors such as solubility, the presence of molecular oxygen, and the availability of appropriate nutrients and electron acceptors have strong effects on whether biodegradation can occur.
3. When biodegradation occurs by growth-linked metabolism, the rates of biomass growth, substrate removal, and oxygen utilization are all linked and are all first-order with respect to the concentration of biomass actually involved in the biodegradation. In other words, everything is the same as growth on soluble chemical oxygen demand (COD) as depicted throughout this text.
4. The modeling of cometabolic biodegradation requires two separate reactions, (1) growth on the carbon source and energy source and (2) the cometabolic transformation. Growth on the carbon source and energy source is done in exactly the same way as any other growth-linked process. The cometabolic transformation is usually modeled as an enzymatic reaction using either the Michaelis-Menten or Haldane equation. It may also be necessary to consider competition between the growth substrate and the cometabolic substrate for the shared enzyme, utilization of the growth substrate to provide reducing power, and inactivation of the cells.
5. Because of the physicochemical properties of some XOCs, abiotic removal mechanisms such as volatilization to the atmosphere and sorption onto solids may contribute to their loss from a bioreactor. They permit the XOCs to enter the environment without alteration and their contribution should be minimized.
6. The mass transfer coefficient for loss of an XOC from a bioreactor by volatilization can be estimated from knowledge of the mass transfer co-

efficient for oxygen, the diffusivity of the XOC, and the power per unit volume expended for oxygen transfer.

7. The sorption coefficient for loss of an XOC from a bioreactor through biomass wastage can be estimated from knowledge of the sorption coefficient for another XOC on the same biomass by using the octanol: water partition coefficients for the two XOCs.
8. The effluent concentration of an XOC from a CMAS system is determined solely by the system solids retention time (SRT) and is independent of the abiotic removal mechanisms. The impact of the abiotic mechanisms is to reduce the concentration of the biomass capable of degrading the XOC. Consequently, the fraction of XOC removal due to abiotic mechanisms is equal to the fractional reduction in capable biomass concentration due to them.
9. Abiotic losses of an XOC from a completely mixed activated sludge (CMAS) system can be characterized by a dimensionless abiotic loss coefficient, α_a, which is the sum of the dimensionless abiotic loss coefficients for sorption and volatilization. Consequently, the effects of the two abiotic removal mechanisms are additive.
10. Xenobiotic organic chemicals can affect the performance of a bioreactor by retarding the removal of biogenic organic matter and nitrification. This can occur through inhibition and toxicity. Inhibition is when the presence of a chemical reduces the rate of a microbial process, such as growth and substrate utilization. Toxicity is when the presence of a chemical causes all microbial activity to cease.
11. The kinetic effects of inhibitors may be classified as competitive, noncompetitive, uncompetitive, or mixed, depending on how the inhibitors affect the $\hat{\mu}$ and K_S values describing carbon oxidation or nitrification. Identification of the inhibitor type is important because it determines the response of a bioreactor to an inhibitory shock load. The model of Han and Levenspiel (Eq. 3.42) may be used as a general model for inhibition and toxicity.
12. Nitrifiers are much more sensitive to inhibition than are aerobic heterotrophic and methanogenic bacteria. Furthermore, ammonia oxidation is considerably more sensitive than nitrite oxidation.
13. Biological treatment systems have been found to be very robust and to do a very good job removing many types of XOCs to low levels. However, because of our relatively limited experience with XOCs and because the ability of a particular bioreactor system to remove a given XOC depends strongly on the configuration of the system, the physicochemical characteristics of the XOC, and the XOC's biodegradation kinetics, it is difficult to generalize about their fate and effects in bioreactors.

22.7 STUDY QUESTIONS

1. List and discuss the requirements for biodegradation of an XOC.
2. List and discuss the factors determining whether biodegradation of a given XOC will occur.

3. Explain the differences between growth-linked and cometabolic biodegradation and the implications of those differences to the modeling of each.
4. Explain why the rate expression for loss of an XOC by volatilization from a bioreactor using mechanical surface aeration for oxygen transfer is different from the rate expression for volatile losses from a bioreactor using diffused aeration.
5. Describe how to estimate the mass transfer coefficient for removal of an XOC by volatilization from knowledge of the mass transfer coefficient for oxygen.
6. Describe the two methods for estimating the coefficient describing sorption of an XOC onto the biomass in a bioreactor.
7. Explain why abiotic removal mechanisms become less important as the fraction of an XOC remaining in a CMAS effluent becomes smaller.
8. A CMAS bioreactor has a volume of 6667 m^3, receives wastewater at a flow rate of 40,000 m^3/day, and operates with a mixed liquor suspended solid (MLSS) concentration of 2750 mg/L when the SRT is 4 days. Oxygen is transferred to it by mechanical surface aeration with a power input of 350 kW, which achieves an oxygen transfer coefficient of 10 h^{-1}. Two of the constituents in the wastewater are dichloromethane and diethyl phthalate, both of which are present in the influent at a concentration of 1.0 mg/L. The concentration of dichloromethane in the effluent is 5.0 μg/L while the concentration of diethyl phthalate is 15.0 μg/L. Sorption studies on the same MLSS with di-n-butyl phthalate gave a linear isotherm with a sorption coefficient of 0.98 L/g. Using physicochemical characteristics available in the literature, estimate the contributions of volatilization and sorption to the removal of dichloromethane and diethyl phthalate.
9. Define the four types of inhibition in terms of their effects on $\hat{\mu}$ and K_S and then describe the characteristics of the coefficients m and n in the Han and Levenspiel model required to model each type. What is the significance of a large absolute value of m or n?

REFERENCES

1. Alexander, M. *Biodegradation and Bioremediation*, Academic Press, New York, 1994.
2. Arbuckle, W. B. and M. S. Kennedy, Activated sludge response to a parachlorophenol transient. *Journal, Water Pollution Control Federation* **61**:476–480, 1989.
3. Beg, S. A. and M. M. Hassan, Effects of inhibitors on nitrification in a packed bed biological flow reactor. *Water Research* **21**:191–198, 1987.
4. Bell, J. P. and M. Tsezos, Removal of hazardous organic pollutants by biomass adsorption. *Journal, Water Pollution Control Federation* **59**:191–198, 1987.
5. Bell, J. P. and M. Tsezos, Removal of hazardous organic pollutants by adsorption on microbial biomass. *Water Science and Technology* **19**(3/4):409–416, 1988.
6. Bell, J., H. Melcer, H. Monteith, I. Osinga, and P. Steel, Stripping of volatile organic compounds at full-scale municipal wastewater treatment plants. *Water Environment Research* **65**:708–716, 1993.

7. Blackburn, J. W., R. K. Jain, and G. S. Sayler, Molecular microbial ecology of a naphthalene-degrading genotype in activated sludge. *Environmental Science & Technology* **21**:884–890, 1987.
8. Blum, D. J. W. and R. E. Speece, A database of chemical toxicity to environmental bacteria and its use in interspecies comparisons and correlations. *Research Journal, Water Pollution Control Federation* **63**:198–207, 1991.
9. Bochner, B., Biolog, Inc., Hayward, CA, Personal communication, June 1995.
10. Chang, H.-L. and L. Alvarez-Cohen, Model for the cometabolic biodegradation of chlorinated organics. *Environmental Science and Technology* **29**:2357–2367, 1995.
11. Chudoba, J., J. Albokova, and J. S. Cech, Determination of kinetic constants of activated sludge microorganisms responsible for the degradation of xenobiotics. *Water Research* **23**:1431–1438, 1989.
12. Criddle, C. S., The kinetics of cometabolism. *Biotechnology and Bioengineering* **41**: 1048–1056, 1993.
13. Dalton, H. and D. I. Sterling, Co-metabolism. *Philosophical Transactions of the Royal Society, London* **297**:481–496, 1982.
14. Dobbs, R. A., M. Jelus, and K.-Y. Cheng, Partitioning of toxic organic compounds on municipal wastewater treatment plant solids. *Proceedings of the International Conference on Innovative Biological Treatment of Toxic Wastewaters*, Arlington, Virginia, June 1986, pp. 585–601.
15. Dobbs, R. A., L. Wang, and R. Govind, Sorption of toxic organic compounds on wastewater solids: correlation with fundamental properties. *Environmental Science and Technology* **23**:1092–1097, 1989.
16. Ellis, T. G., B. F. Smets, and C. P. L. Grady Jr., Influence of simultaneous multiple substrate biodegradation on the kinetic parameters for individual substrates. *Proceedings of the Water Environment Federation 68th Annual Conference and Exposition, I, Wastewater Treatment Research and Municipal Wastewater Treatment*, Water Environment Federation, Alexandria, Virginia, pp. 167–178, 1995.
17. Fredrickson, J. K., D. L. Balkwill, J. M. Zachara, S.-M. W. Li, F. J. Brockman, and M. A. Simmons, Physiological diversity and distributions of heterotrophic bacteria in deep Cretaceous sediments of the Atlantic coastal plain. *Applied and Environmental Microbiology* **57**:402–411, 1991.
18. Gantzer, C. J., Inhibitory substrate utilization by steady-state biofilms. *Journal of Environmental Engineering* **115**:302–319, 1989.
19. Gokare, M. A., *A Protocol for Testing the Effect of Organic Chemicals on Nitrification*, M.S. Thesis, Clemson University, Clemson, South Carolina, 1990.
20. Govind, R., L. Lai, and R. Dobbs, Integrated model for predicting the fate of organics in wastewater treatment plants. *Environmental Progress* **10**:13–23, 1991.
21. Grady, C. P. L. Jr., Biodegradation of hazardous wastes by conventional biological treatment. *Hazardous Waste & Hazardous Materials* **3**:333–365, 1986.
22. Grady, C. P. L. Jr., Biodegradation of toxic organics: status and potential. *Journal of Environmental Engineering* **116**:805–828, 1990.
23. Grady, C. P. L. Jr., B. S. Magbanua, S. Brau, and R. W. Sanders II, A simple technique for estimating the contribution of abiotic mechanisms to the removal of SOCs by completely mixed activated sludge. *Water Environment Research* **69**:1232–1237, 1997.
24. Grady, C. P. L. Jr., B. S. Magbanua, B. F. Smets, S. Brau, R. L. Buddin, A. G. Rodieck, R. W. Sanders, W. W. Sowers, and J. C. Stanfill, Relative efficacy of extant and intrinsic kinetic parameter estimates for predicting the removal of synthetic organic chemicals by activated sludge. *Proceedings of the Water Environment Federation 69th Annual Conference and Exposition, I, Wastewater Treatment Research and Municipal Wastewater Treatment*, Water Environment Federation, Alexandria, Virginia, pp. 103–114, 1996.

25. Grady, C. P. L. Jr., B. F. Smets, and D. S. Barbeau, Variability in kinetic parameter estimates: a review of possible causes and a proposed terminology. *Water Research* **30**: 742–748, 1996.
26. Hall, E., B. Sun, J. Prakash, and N. Nirmalakhandan, Toxicity of organic chemicals and their mixtures to activated sludge microorganisms. *Journal of Environmental Engineering* **122**:424–429, 1996.
27. Han, K. and O. Levenspiel, Extended Monod kinetics for substrate, products and cell inhibition. *Biotechnology and Bioengineering* **32**:430–437, 1988.
28. Hannah, S. A., B. A. Austern, A. E. Eralp, and R. A. Dobbs, Removal of organic toxic pollutants by trickling filter and activated sludge. *Journal, Water Pollution Control Federation* **60**:1281–1283, 1988.
29. Hartmann, L. and G. Laubenberger, Toxicity measurements in activated sludge. *Journal of the Sanitary Engineering Division, ASCE* **94**:247–256, 1968.
30. Hockenbury, M. R. and C. P. L. Grady Jr., Inhibition of nitrification—Effects of selected organic compounds. *Journal, Water Pollution Control Federation* **57**:768–777, 1977.
31. Horvath, R. S., Microbial co-metabolism and the degradation of organic compounds in nature. *Bacteriological Reviews* **36**:146–155, 1972.
32. Hsieh, C.-C., R. W. Babcock, Jr., and M. K. Stenstrom, Estimating emissions of 20 VOCs. II. Diffused aeration. *Journal of Environmental Engineering* **119**:1099–1118, 1993.
33. Hsieh, C.-C., K. S. Ro, and M. K. Stenstrom, Estimating emissions of 20 VOCs. I. Surface aeration. *Journal of Environmental Engineering* **119**:1077–1098, 1993.
34. Karickhoff, S. W., Sorption dynamics of hydrophobic pollutants in sediment suspensions. *Environmental Toxicology and Chemistry* **4**:469–479, 1985.
35. Kim, I. S., J. C. Young, and H. H. Tabak, Kinetics of acetogenesis and methanogenesis in anaerobic reactions under toxic conditions. *Water Environment Research* **66**:119–132, 1994.
36. Kong, Z., P. Vanrolleghem, P. Willems, and W. Verstraete, Simultaneous determination of inhibition kinetics of carbon oxidation and nitrification with a respirometer. *Water Research* **30**:825–836, 1996.
37. Lewis, W. K. and W. E. Whitman, Principles of gas absorption. *Industrial and Engineering Chemistry* **16**:1215–1220, 1924.
38. Lyman, W. J., W. F. Reehl, and D. H. Rosenblatt, *Handbook of Chemical Property Estimation Methods—Environmental Behavior of Organic Compounds*, McGraw-Hill, New York, 1982.
39. Melcer, H., P. Steel, and W. K. Bedford, Removal of polycyclic aromatic hydrocarbons and heterocyclic nitrogen compounds in a municipal treatment plant. *Water Environment Research* **67**:926–934, 1995.
40. Melcer, H., P. Steel, J. P. Bell, D. Thompson, C. M. Yendt, and J. Kemp, Modeling volatile organic contaminant's fate in wastewater treatment plants. *Journal of Environmental Engineering* **120**:588–609, 1994.
41. Monteith, H. D., W. J. Parker, J. P. Bell, and H. Melcer, Modeling the fate of pesticides in municipal wastewater treatment. *Water Environment Research* **67**:964–970, 1995.
42. Munz, C. and P. V. Roberts, Gas- and liquid-phase mass transfer resistances of organic compounds during mechanical surface aeration. *Water Research* **23**:589–601, 1989.
43. Oslislo A. and Z. Lewandowski, Inhibition of nitrification in the packed bed reactors by selected organic compounds. *Water Research* **19**:423–426, 1985.
44. Parker, W. J., D. J. Thompson, J. P. Bell, and H. Melcer, Fate of volatile organic compounds in municipal activated sludge plants. *Water Environment Research* **65**:58–65, 1993.
45. Patterson, J. W. and P. L. Brezonik, Discussion of 'Toxicity measurements in activated sludge'. *Journal of the Sanitary Engineering Division, ASCE* **95**:775–780, 1969.

46. Roberts, P. V., C. Munz, and P. Dändliker, Modeling volatile organic solute removal by surface and bubble aeration. *Journal, Water Pollution Control Federation* **56**:157–163, 1984.
47. Santiago, I. and C. P. L. Grady Jr., Simulation studies of the transient response of activated sludge systems to biodegradable inhibitory shock loads. *Proceedings of the 44th Industrial Waste Conference, 1989, Purdue University*, Lewis Publishers, Chelsea, Michigan, pp. 191–198, 1990.
48. Schwarzenbach, R. P., P. M. Gschwend, and D. M. Imboden, *Environmental Organic Chemistry*, Wiley, New York, 1993.
49. Sikkema, J., J. A. M. de Bont, and B. Poolman, Mechanisms of membrane toxicity of hydrocarbons. *Microbiological Reviews* **59**:201–222, 1995.
50. Slater, J. H., Mixed cultures and microbial communities. In *Mixed Culture Fermentations*, M. E. Bushell and J. H. Slater, eds. Academic Press, New York, pp. 1–24, 1981.
51. Sock, S. M., *A Comprehensive Evaluation of Biodegradation as a Treatment Alternative for the Removal of 1,4-Dioxane*, M.S. Thesis, Clemson University, Clemson, South Carolina, 1993.
52. Stevens, D. K., Interaction of mass transfer and inhibition in biofilms. *Journal of Environmental Engineering* **114**:1352–1358, 1988.
53. Tomlinson, T. G., A. G. Boon, and C. N. A. Trotman, Inhibition of nitrification in the activated sludge process of sewage disposal. *Journal of Applied Bacteriology* **29**:266–291, 1966.
54. Tsezos, M and J. P. Bell, A mechanistic study on the fate of malathion following interaction with microbial biomass. *Water Research* **25**:1039–1046, 1991.
55. Tsezos, M. and X. Wang, Study on the kinetics of hazardous pollutants adsorption and desorption by biomass: mechanistic considerations. *Journal of Chemical Technology and Biotechnology* **50**:507–521, 1991.
56. Volskay, V. T. Jr. and C. P. L Grady Jr., Toxicity of selected RCRA compounds to activated sludge microorganisms. *Journal, Water Pollution Control Federation* **60**: 1850–1856, 1988.
57. Volskay, V. T. Jr. and C. P. L. Grady Jr., Respiration inhibition kinetic assay. *Water Research* **24**:8630874, 1990.
58. Volskay, V. T. Jr., C. P. L. Grady Jr., and H. H. Tabak, Effect of selected RCRA compounds on activated sludge activity. *Research Journal, Water Pollution Control Federation* **62**:654–664, 1990.
59. Wang, K., B. Rott and F. Korte, Uptake and bioaccumulation of three PCBs by *Chlorella fusca*. *Chemosphere* **11**:525–530, 1982.
60. Wang, X., *The Effects of Biosorption on the Biodegradation of di-n-Butyl Phthalate*, Ph.D. Dissertation, Clemson University, Clemson, South Carolina, 1993.

Appendix A
Acronyms

Acronym	Definition	Section where first used
A/AD	Anoxic/Aerobic Digestion	1.3.1
ABF	Activated Biofilter	19.1.2
AC	Anaerobic Contact	1.3.1
AD	Anaerobic Digestion	1.3.1
ADP	Adenosine Diphosphate	2.4.1
AEL	Aerobic Lagoon	14.1.2
AER	Aerobic	10.1.1
AF	Anaerobic Filter	1.3.2
ANA	Anaerobic	11.1.2
ANL	Anaerobic Lagoon	1.3.1
ANX	Anoxic	11.1.2
AOL	Areal Organic Loading	14.2.3
A/O	Anaerobic/Oxic	7.7.1
A^2/O	Anaerobic/Anoxic/Oxic	11.1.3
AR	Anoxic mixed liquor Recirculation	11.1.3
ASIM	Activated Sludge SIMulation program	6.1.5
ASM	Activated Sludge Model	3.7
ATAD	Autothermal Thermophilic Aerobic Digestion	12.1.2
ATP	Adenosine Triphosphate	2.4.1
BAF	Biological Aerated Filter	21.1.2
BF/AS	Biofilter/Activated Sludge	19.1.2
BNR	Biological Nutrient Removal	1.3.1
BOD	Biochemical Oxygen Demand	3.2.10
BPR	Biological Phosphorus Removal	7.1.1
CAD	Conventional Aerobic Digestion	1.3.1
CAS	Conventional Activated Sludge	1.3.1
CFSTR	Continuous-Flow Stirred Tank Reactor	4.3.1
CMAL	Completely Mixed Aerated Lagoon	1.3.1
CMAS	Completely Mixed Activated Sludge	1.3.1
COD	Chemical Oxygen Demand	1.2.1
COD/ΔP	COD to Phosphorus removal ratio	11.2.2
CSAG	Combined Suspended and Attached Growth	21.1.1
CSAS	Contact Stabilization Activated Sludge	10.1.1
CSTR	Continuous Stirred Tank Reactor	1.2.3

Acronym	Definition	Section where first used
DFPB	Downflow Packed Bed	21.1.2
DO	Dissolved Oxygen	1.0
DSFF	Downflow Stationary Fixed Film	13.1.4
DSVI	Diluted Sludge Volume Index	10.2.1
EAAS	Extended Aeration Activated Sludge	1.3.1
ECP	Exocellular Polymer	10.2.1
ED	Entner–Doudoroff pathway	2.4.6
EMP	Embden–Meyerhof–Parnas pathway	2.4.6
F/AL	Facultative/Aerated Lagoon	1.3.1
FBBR	Fluidized Bed Bioreactor	1.2.3
FB/EB	Fluidized Bed/Expanded Bed	13.1.4
F/M	Food to Microorganism ratio	2.3.1
FSS	Fixed Suspended Solids	8.7
GAC	Granular Activated Carbon	21.1.4
HO	Horizontal trickling filter media	19.2.6
HPOAS	High-Purity Oxygen Activated Sludge	1.3.1
HRT	Hydraulic Residence Time	4.3.2
IAWPRC	International Association on Water Pollution Research and Control	6.0
IAWQ	International Association on Water Quality	6.0
IIM	Insoluble Inorganic Matter	1.1
IOM	Insoluble Organic Matter	1.1
MLE	Modified Ludzack–Ettinger	7.1.1
MLR	Mixed Liquor Recirculation	11.1.2
MLSS	Mixed Liquor Suspended Solids	2.1
MUCT	Modified University of Cape Town	11.1.3
MW	Molecular Weight	3.1.1
NAD	Nicotinamide Adenine Dinucleotide	2.4.1
NADH	Reduced Nicotinamide Adenine Dinucleotide	2.4.1
NADP	Nicotinamide Adenine Dinucleotide Phosphate	2.4.1
NADPH	Reduced Nicotinamide Adenine Dinucleotide Phosphate	2.4.1
NR	Nitrified mixed liquor Recirculation	11.1.3
NRC	National Research Council	19.3.1
NUR	Nitrate Utilization Rate	8.5.3
NUR_g	Nitrate Utilization Rate for Growth	8.5.3
NUR_h	Nitrate Utilization Rate for Hydrolysis	8.5.3
OUR	Oxygen Uptake Rate	8.3.1
OUR_g	Oxygen Uptake Rate for Growth	8.5.3
OUR_h	Oxygen Uptake Rate for Hydrolysis	8.5.3

Acronym	Definition	Section where first used
PAO	Phosphate Accumulating Organism	2.2.1
PFR	Plug–Flow Reactor	1.2.3
PHA	Polyhydroxyalkanoic Acid	6.1.4
PHB	Poly-β-hydroxybutyrate	2.4.2
Poly-P	Polyphosphate	2.4.6
RA	Random trickling filter media	19.2.6
RAS	Return Activated Sludge	10.1.1
RBC	Rotating Biological Contactor	1.3.2
RDR	Rotating Disk Reactor	1.2.3
RF/AS	Roughing Filter/Activated Sludge	19.1.2
RNA	Ribonucleic Acid	2.4.2
RTD	Residence Time Distribution	4.3.2
SAGB	Submerged Attached Growth Bioreactor	21.0
SAL	Surface Ammonia-N Loading	19.2.1
SAS	Selector Activated Sludge	1.3.1
SBR	Sequencing Batch Reactor	1.2.3
SBRAS	Sequencing Batch Reactor Activated Sludge	1.3.1
SBRSIM	Sequencing Batch Reactor SIMulation	7.1.1
SFAS	Step Feed Activated Sludge	1.3.1
SIM	Soluble Inorganic Matter	1.0
SNL	Surface Nitrogen Loading	19.2.1
SP	Soluble Phosphorus	11.2.2
SOL	Surface Organic Loading	19.2.1
SOM	Soluble Organic Matter	1.0
SOUR	Specific Oxygen Uptake Rate	8.4.3
SRT	Solids Retention Time	5.1.1
SS	Suspended Solids	2.4.1
SSSP	Simulation of Single Sludge Processes	6.1.5
$SSVI_{3.5}$	Stirred Sludge Volume Index at 3.5 g/L MLSS	10.2.1
SVI	Sludge Volume Index	10.2.1
TAL	Total Ammonia-N Loading	19.1.2
TCA	Tricarboxylic Acid Cycle	2.4.6
TF	Trickling Filter	1.3.2
TF/AS	Trickling Filter/Activated Sludge	19.1.2
TF/SC	Trickling Filter/Solids Contact	19.1.2
THL	Total Hydraulic Loading	13.2.3
TKN	Total Kjeldahl Nitrogen	8.6
TNL	Total Nitrogen Loading	19.1.2
TOL	Total Organic Loading	19.1.2
TP	Total Phosphorus	11.2.2
TSS	Total Suspended Solids	8.2.1
USAB	Upflow Anaerobic Sludge Blanket	1.3.1
UASB/AF	USAB and Anaerobic Filter	13.1.4

Acronym	Definition	Section where first used
UCT	University of Cape Town	11.1.3
UFPB	Upflow Packed Bed	21.1.3
VFA	Volatile Fatty Acid	11.2.1
VFC	Vertical Fully Corrugated	19.2.6
VIP	Virginia Initiative Plant	11.1.3
VOL	Volumetric Organic Loading	13.1.4
VSC	Vertical Semi-Corrugated	19.2.6
VS	Volatile Solids	13.1
VSS	Volatile Suspended Solids	2.4.1
WAS	Waste Activated Sludge	10.1.1
XF	Crossflow	19.2.6
XOC	Xenobiotic Organic Compound	22.0

Appendix B
Symbols

Symbol	Definition	Units	Equation where first used
a	Empirical coefficient	—	16.4
a_i	Molar stoichiometric coefficient for reactant A_i	mole/mole	3.1
a_v	Specific surface area	L^{-1}	15.42
A	Constant in Arrhenius equation	Varies	3.91
A_c	Cross-sectional area	L^2	4.7
A_i	Constituent i in a stoichiometric equation	mole or M	3.1
A_L	Surface area of a lagoon	L^2	14.1
A_n	Planar biofilm surface area normal to the direction of diffusion	L^2	15.7
A_s	Total planar surface area in the submerged sector of an RDR	L^2	17.8
$A_{s,t}$	Total planar surface area in an RBC	L^2	20.1
$A_{s,t,n}$	Total planar surface area in stage N of an RBC	L^2	20.7
b	Decay coefficient, traditional approach	T^{-1}	3.56
b	Empirical coefficient	—	22.15
b	Empirical exponent	—	16.4
b_A	Decay coefficient for autotrophs, traditional approach	T^{-1}	6.2
b_C	Competition coefficient for a biofilm	T^{-1}	15.59
b_D	Detachment coefficient	T^{-1}	15.21
b_H	Decay coefficient for heterotrophs, traditional approach	T^{-1}	5.5
b_L	Decay coefficient, lysis:regrowth approach	T^{-1}	3.63
$b_{L,A}$	Decay coefficient for autotrophs, lysis: regrowth approach	T^{-1}	Table 6.1
$b_{L,H}$	Decay coefficient for heterotrophs, lysis: regrowth approach	T^{-1}	Table 6.1
b_{XOC}	Decay coefficient for biomass grown on an XOC	T^{-1}	5.13 in Ch. 22
Bi	Biot number	—	15.11
BOD_5	5-Day BOD concentration	ML^{-3}	8.28
$BOD_5/\Delta P$	BOD_5 to phosphorus removal ratio	$M_{BOD5}M_P^{-1}$	11.4
BOD_u	Ultimate BOD concentration	ML^{-3}	8.29

Symbol	Definition	Units	Equation where first used
c	Empirical coefficient	—	16.7
C	Temperature coefficient	$°C^{-1}$	3.93
C	Empirical coefficient in the modified Velz/Germain equation for a trickling filter	Varies	19.7
C_A	Concentration of reactant A	ML^{-3}	4.3
C_{AO}	Concentration of reactant A in the influent	ML^{-3}	4.3
C'_A	Concentration of reactant A in the stream passing through the biomass separator	ML^{-3}	5.4
C_D	Drag coefficient	—	18.2
$C_{G,XOC}$	Concentration of an XOC in the gas phase	ML^{-3}	22.2
COD_{BO}	Influent biodegradable COD concentration	ML^{-3}	8.31
COD_{IO}	Influent inert COD concentration	ML^{-3}	8.34
COD_{TO}	Influent total COD concentration	ML^{-3}	8.23
$C_{S,XOC}$	Solid phase concentration of an XOC	MM^{-1}	22.13
d	Characteristic dimension in the Reynolds number	L	Sec. 15.2.1
d	Empirical exponent	—	16.7
d_b	Diameter of a bioparticle	L	18.7
d_p	Diameter of a carrier particle	L	18.1
D_e	Effective diffusivity in a biofilm	L^2T^{-1}	15.5
D_l	Axial dispersion coefficient	L^2T^{-1}	4.22
D_w	Diffusivity in water	L^2T^{-1}	15.1
$D_{w,O2}$	Diffusivity of oxygen in water	L^2T^{-1}	22.10
$D_{w,XOC}$	Diffusivity of an XOC in water	L^2T^{-1}	22.10
e	Empirical coefficient	Varies	17.2
E'_0	Standard oxidation-reduction potential	mV	Table 2.3
E(t)	Fraction of fluid elements leaving a reactor that have resistance times between t and t + dt	—	4.16
f	Empirical exponent	—	17.2
f_A	Active fraction of the biomass	—	5.26
f_{AL}	Fraction of packed tower cross-sectional area occupied by liquid film	—	16.6
f_D	Fraction of active biomass contributing to biomass debris, traditional approach	—	3.53
f'_D	Fraction of active biomass contributing to biomass debris, lysis:regrowth approach	—	3.60
f_e	Fraction of electron donor used for energy	—	3.14
f_{MP}	Fraction of active biomass contributing to biomass associated products, traditional approach	—	3.74
f'_{MP}	Fraction of active biomass contributing to biomass associated products, lysis:regrowth approach	—	3.74

Symbol	Definition	Units	Equation where first used
f_{NH}	Fraction of the nitrification rate that would occur in an RBC biofilm in the absence of carbon oxidation	—	20.2
$f_{NO,D}$	Fraction of nitrate-N dentrified in an initial anoxic zone	—	11.7
$f_{NO,R}$	Fraction of nitrate-N recirculated to an initial anoxic zone	—	11.8
f_s	Fraction of electron donor captured through synthesis	—	3.14
$f_{V,A}$	Fraction of system volume in which nitrification occurs	—	10.35
$f_{V,ANX}$	Fraction of system volume that is anoxic	—	11.17
$f_{V,SS}$	Fraction of system volume in which biomass synthesis occurs on readily biodegradable substrate	—	10.31
$f_{V,XS}$	Fraction of system volume in which biomass synthesis occurs on slowly biodegradable substrate	—	10.27
$f_{XA,C}$	Fraction of autotrophic biomass in the contact tank of a CSAS system	—	10.48
f_{XBi}	Fraction of biomass density in a biofilm due to species i	—	15.51
$f_{XM,AER}$	Fraction of MLSS maintained under aerobic conditions	—	11.1
$f_{XM,ANA}$	Fraction of MLSS maintained under anaerobic conditions	—	11.3
$f_{XM,ANX}$	Fraction of MLSS maintained under anoxic conditions	—	11.2
$f_{XM,C}$	Fraction of MLSS in the contact tank of a CSAS system	—	10.45
$f_{XM,N}$	Fraction of MLSS in the last tank of a chain of CSTRs	—	10.41
$f_{XS,H}$	Fraction of additional slowly biodegradable substrate oxidized during a transient load	—	10.12
F	Volumetric flow rate	L^3T^{-1}	4.3
F_c	Volume applied to an SBR each cycle	L^3	7.11
F_f	Effective flow rate of liquid film on the aerated sector of an RDR	L^3T^{-1}	17.8
F_O	Influent volumetric flow rate	L^3T^{-1}	4.3
F_r	Volumetric flow rate of biomass recycle stream	L^3T^{-1}	5.69
F_R	Volumetric flow rate of recirculation stream	L^3T^{-1}	13.5
F_w	Volumetric flow rate of biomass wastage stream	L^3T^{-1}	5.1
F_{wi}	Volumetric flow rate of biomass wastage stream from bioreactor i	L^3T^{-1}	7.1
$F(t)$	Fraction of fluid elements leaving a reactor that have residence times less than t	—	4.16

Symbol	Definition	Units	Equation where first used
g	Gravitational acceleration	LT^{-2}	18.1
G	Root mean-square velocity gradient	T^{-1}	10.6
Ga	Galileo number	—	18.13
h	Depth of liquid above diffuser	L	10.6
H	Depth of a suspended growth bioreactor	L	10.36
H′	Heat transfer coefficient	$MT^{-3}{}^{\circ}F^{-1}$	14.3
H_{Bb}	Height of a fluidized bed containing bioparticles	L	18.5
H_{Bbc}	Calculated height of a fluidized bed containing bioparticles	L	Fig. 18.9
H_{Bp}	Height of a fluidized bed containing clean carrier particles	L	18.3
H_{XOC}	Henry's law coefficient for an XOC	Atm L^3Mol^{-1}	22.3
$H_{c,XOC}$	Dimensionless Henry's law coefficient for an XOC	—	22.2
H_{Rp}	Reference bed height of a fluidized bed	L	18.3
$i_{N,XB}$	Mass of nitrogen per mass of COD in active biomass	$M_N M_{COD}^{-1}$	3.55
$i_{N,XD}$	Mass of nitrogen per mass of COD in biomass debris	$M_N M_{COD}^{-1}$	3.55
$i_{O,EA}$	Mass of COD per mass of electron acceptor	$M_{COD} M_{EA}^{-1}$	9.5
$i_{O,XBI}$	Mass of COD per mass of biomass measured as TSS	$M_{COD} M_{TSS}^{-1}$	Sec. 8.2.1
$i_{O,XBV}$	Mass of COD per mass of biomass measured as VSS	$M_{COD} M_{VSS}^{-1}$	Sec. 8.2.1
$i_{O,XMLI}$	Mass of COD per mass of MLSS measured as TSS	$M_{COD} M_{TSS}^{-1}$	Sec. 9.4.2
$i_{P,XB}$	Mass of phosphorus per mass of COD in active biomass	$M_P M_{COD}^{-1}$	3.89
J_{NH}	Flux of ammonia-N	$ML^{-2}T^{-1}$	19.5
$J_{NH,max}$	Maximum flux of ammonia-N	$ML^{-2}T^{-1}$	19.5
J_S	Flux of substrate	$ML^{-2}T^{-1}$	15.1
J_{SR}	Reference flux for normalized loading curves	$ML^{-2}T^{-1}$	Sec. 15.2.3
J_S^*	Dimensionless substrate flux	—	15.31
J_{SR}^*	Dimensionless reference flux for normalized loading curves	—	Sec. 15.2.3
$J_{S,deep}^*$	Dimensionless substrate flux into a deep biofilm	—	15.29
$J_{S,first}^*$	Dimensionless substrate flux into a first-order biofilm	—	15.48
$J_{S,fp}^*$	Dimensionless substrate flux into a fully penetrated biofilm	—	15.43
$J_{S,zero}^*$	Dimensionless substrate flux into a zero-order biofilm	—	15.50

Symbol	Definition	Units	Equation where first used
k	Rate coefficient	Varies	3.91
k	BOD first-order rate coefficient	T^{-1}	8.29
k	Empirical coefficient in the modified Velz/ Germain equation for a trickling filter	$M^{1-n}L^{3n}T^{-1}$	19.6
k_1	First-order reaction rate coefficient	T^{-1}	20.4
k_2	Second-order reaction rate coefficient	$M^{-1}L^{3}T^{-1}$	20.6
k_a	Ammonification rate coefficient	$M^{-1}L^{3}T^{-1}$	3.79
k_c	Mean reaction rate coefficient	$M^{-1}L^{3}T^{-1}$	3.45
$k_{G,XOC}$·a	Gas film mass transfer coefficient for an XOC	T^{-1}	22.9
k_h	Hydrolysis coefficient	T^{-1}	3.77
k_L	Liquid phase mass transfer coefficient	LT^{-1}	15.2
$k_{L,a}$	Liquid phase mass transfer coefficient in the aerated sector of an RDR	LT^{-1}	17.7
$k_{L,s}$	Liquid phase mass transfer coefficient in the submerged sector of an RDR	LT^{-1}	17.1
$k_{L,XOC}$·a	Liquid film mass transfer coefficient for an XOC	LT^{-1}	22.9
k_L^*	Dimensionless liquid phase mass transfer coefficient	—	15.40
k_{OW}	Octanol:water partition coefficient	—	22.15
$k_{s,XOC}$	Sorption coefficient for an XOC	L^3M^{-1}	22.13
K	Empirical coefficient in Velz model for a packed tower	—	16.9
K	Empirical coefficient in the modified Velz-Germain equation for a trickling filter	Varies	19.8
K_1	Empirical coefficient in Eckenfelder model for a packed tower	—	16.12
K_2	Empirical coefficient in Kornegay model for a packed tower	—	16.14
K_3	Empirical coefficient	—	16.16
K_A	Half-saturation coefficient for acetic acid	ML^{-3}	3.82
K_g	Pseudo half-saturation coefficient in Kornegay model	ML^{-3}	16.14
$K_{g,NH}$	Pseudo half-saturation coefficient for ammonia-N	ML^{-3}	19.5
K_I	Inhibition coefficient	ML^{-3}	3.39
K_{IO}	Inhibition coefficient for oxygen against denitrification	ML^{-3}	3.48
K_{IPP}	Inhibition coefficient for Poly-P storage	$M_pM_{COD}^{-1}$	3.87
K_{L,O_2}a	Overall liquid phase mass transfer coefficient for oxygen	T^{-1}	22.8
$K_{L,XOC}$·a	Overall liquid phase mass transfer coefficient for removal of an XOC by volatilization	T^{-1}	22.1
K_{NH}	Half-saturation coefficient for ammonia-N	ML^{-3}	3.51
K_{NO}	Half-saturation coefficient for nitrate-N	ML^{-3}	3.48
K_O	Half-saturation coefficient for oxygen	ML^{-3}	Sec. 3.2.9

Symbol	Definition	Units	Equation where first used
$K_{O,A}$	Half-saturation coefficient for oxygen for autotrophs	ML^{-3}	Sec. 3.2.9
$K_{O,H}$	Half-saturation coefficient for oxygen for heterotrophs	ML^{-3}	Sec. 3.2.9
K_P	Half-saturation coefficient for soluble phosphate	ML^{-3}	3.85
K_{PHB}	Half-saturation coefficient for PHB	MM^{-1}	3.85
K_{PMAX}	Maximum mass of Poly-P per unit mass of PAO	$M_P M_{COD}^{-1}$	3.87
K_{PP}	Half-saturation coefficient for Poly-P	ML^{-3}	3.82
K_S	Half-saturation coefficient for substrate	ML^{-3}	3.36
$K_{S,XOC}$	Half-saturation coefficient for growth of an XOC	ML^{-3}	5.13 in Ch. 22
K_X	Half-saturation coefficient for hydrolysis of slowly biodegradable substrate	MM^{-1}	3.77
L	Length of a suspended growth bioreactor; depth of a packed tower	L	4.12
L_f	Biofilm thickness	L	15.6
L_{fc}	Characteristic biofilm thickness	L	18.17
L_{fd}	Assumed biofilm thickness	L	Fig. 18.9
L_f^*	Dimensionless biofilm thickness	—	15.44
L_w	Stagnant liquid film thickness	L	15.1
m	Exponent depicting the effect of an inhibitor on K_S	—	3.42
m	Empirical exponent in Eckenfelder model	—	16.12
M	Mass velocity	$ML^{-2}T^{-1}$	16.2
M	Number of RDRs in series	—	Sec. 17.3
M_p	Mass of carrier particles in a fluidized bed	M	18.4
n	Exponent depicting the effect of an inhibitor on $\hat{\mu}$	—	3.42
n	Empirical exponent in Eckenfelder model	—	16.12
n	Empirical exponent in the Richardson-Zaki equation	—	18.6
n	Empirical exponent in the modified Velz/ Germain equation for a trickling filter	—	19.7
n	Empirical exponent in Eq. 22.10	—	22.10
n	Empirical exponent in the Freundlich equation	—	22.13
N	Number of CSTRs in series	—	4.21
N	Number of discs in an RDR	—	17.11
N_a	Number of arms on a trickling filter rotary distributor	—	19.4
N_c	Number of cycles per day applied to an SBR	T^{-1}	7.11
NR	Nitrogen requirement per unit of COD removed	$M_N M_{COD}^{-1}$	5.36

Symbol	Definition	Units	Equation where first used
ON_{TO}	Influent total organic nitrogen concentration	ML^{-3}	8.37
$\bar{p}_{CO2}$	Partial pressure of carbon dioxide	atm	13.7
P	Power input	ML^2T^{-3}	10.1
P′	Weight fraction moisture content of a biofilm	—	18.9
P_t	Inorganic phosphate	ML^{-3}	Fig. 2.5
q	Specific substrate removal rate	T^{-1}	3.43
$q_{NO,XB}$	Specific nitrate-N utilization rate associated with biomass decay	T^{-1}	11.12
$q_{NO,XS}$	Specific nitrate-N utilization rate associated with slowly biodegradable substrate utilization	T^{-1}	11.11
$\hat{q}$	Maximum specific substrate removal rate	T^{-1}	3.44
$\hat{q}_A$	Maximum specific acetic acid uptake rate	T^{-1}	3.82
$\hat{q}_{PP}$	Maximum specific Poly-P storage rate	$M_P M_{COD}^{-1} T^{-1}$	3.87
Q	Air flow rate	L^3T^{-1}	10.2
r	Generalized reaction rate	$ML^{-3}T^{-1}$	3.10
r	Radius of a rotating disc	L	17.1
r_A	Reaction rate for reactant A	$ML^{-3}T^{-1}$	4.3
r_i	Reaction rate for reactant A_i	$ML^{-3}T^{-1}$	3.10
r_i	Inner radius of the submerged sector of an RDR	L	17.4
r_j	Generalized reaction rate for reaction j	$ML^{-3}T^{-1}$	3.12
r_o	Outer radius of the submerged sector of an RDR	L	17.4
$r_{X,XOC}$	Rate of loss of an XOC by volatilization	$ML^{-3}T^{-1}$	22.1
r_{SA}	Reaction rate for acetic acid	$ML^{-3}T^{-1}$	3.82
r_{SMP}	Reaction rate for soluble microbial products	$ML^{-3}T^{-1}$	3.72
r_{SNH}	Reaction rate for ammonia-N	$ML^{-3}T^{-1}$	3.59
r_{SNS}	Reaction rate for soluble organic-N	$ML^{-3}T^{-1}$	3.78
r_{SO}	Reaction rate for dissolved oxygen	$ML^{-3}T^{-1}$	3.34
$r_{SO,i}$	Volumetric oxygen transfer rate in tank i	$ML^{-3}T^{-1}$	10.37
r_{SP}	Reaction rate for soluble phosphate	$ML^{-3}T^{-1}$	3.84
r_{SS}	Reaction rate for readily biodegradable substrate	$ML^{-3}T^{-1}$	3.34
r_{XB}	Reaction rate for active biomass	$ML^{-3}T^{-1}$	3.34
r_{XBP}	Reaction rate for PAO biomass	$ML^{-3}T^{-1}$	3.85
r_{XD}	Reaction rate for biomass debris	$ML^{-3}T^{-1}$	3.57
r_{XNS}	Reaction rate for particulate organic nitrogen	$ML^{-3}T^{-1}$	3.66
r_{XPP}	Reaction rate for stored polyphosphate	$ML^{-3}T^{-1}$	3.84
r_{XPHB}	Reaction rate for PHB	$ML^{-3}T^{-1}$	3.83
r_{XS}	Reaction rate for slowly biodegradable substrate	$ML^{-3}T^{-1}$	3.65
R	Gas constant	$kJMole^{-1}K^{-1}$	3.91
R	Overall stoichiometric equation	—	3.14
R_a	Half-reaction for the electron acceptor	—	3.14

Symbol	Definition	Units	Equation where first used
R_c	Half-reaction for cell material	—	3.14
R_d	Half-reaction for the electron donor	—	3.14
Re	Reynolds number	—	16.2
Re_t	Terminal Reynolds number	—	18.10
REA	Mass rate of electron acceptor utilization	MT $^{-1}$	9.5
Ri	Rittmann number	—	15.34
$RN_{A,max}$	Maximum mass nitrification rate	MT $^{-1}$	10.34
RO	Mass rate of oxygen utilization	MT $^{-1}$	5.32
RO_A	Steady-state oxygen requirement for autotrophs	MT $^{-1}$	6.2
$RO_{A,C}$	Autotrophic oxygen requirement in the contact tank of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
$RO_{A,D}$	Autotrophic oxygen requirement for biomass decay	MT $^{-1}$	10.33
$RO_{A,S}$	Autotrophic oxygen requirement in the stabilization basin of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
$RO_{A,D,i}$	Autotrophic oxygen requirement for biomass decay in tank i	MT $^{-1}$	10.38
$RD_{A,D,S}$	Autotrophic oxygen requirement for biomass decay in a selector	MT $^{-1}$	Ex. 10.3.4.3
$RO_{A,max}$	Maximum mass rate of oxygen utilization by autotrophs	MT $^{-1}$	10.15
$RO_{A,SS}$	Autotrophic oxygen requirement for biomass synthesis	MT $^{-1}$	10.32
$RO_{A,SS,i}$	Autotrophic oxygen requirement for biomass synthesis in tank i	MT $^{-1}$	10.38
$RO_{A,SS,C}$	Autotrophic oxygen requirement for biomass synthesis in the contact tank of a CSAS system	MT $^{-1}$	10.51
$RO_{A,SS,S}$	Autotrophic oxygen requirement for biomass synthesis in a selector	MT $^{-1}$	Ex. 10.3.4.3
$RO_{A,SS,S}$	Autotrophic oxygen requirement for biomass synthesis in the stabilization basin of a CSAS system	MT $^{-1}$	10.52
$RO_{A,TS}$	Transient-state oxygen requirement for autotrophs	MT $^{-1}$	10.13
$RO_{A,XNS}$	Autotrophic oxygen requirement for biomass synthesis from particulate organic nitrogen	MT $^{-1}$	10.53
$RO_{A,XNS,C}$	Autotrophic oxygen requirement for biomass synthesis from particulate organic nitrogen in the contact tank of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
$RO_{A,XNS,S}$	Autotrophic oxygen requirement for biomass synthesis from particulate organic nitrogen in the stabilization basin of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
RO_C	Oxygen requirement in the contact tank of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
RO_H	Steady-state oxygen requirement for heterotrophs	MT $^{-1}$	9.10

Symbol	Definition	Units	Equation where first used
$RO_{H,C}$	Heterotrophic oxygen requirement in the contact tank of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
$RO_{H,D}$	Heterotrophic oxygen requirement for biomass decay	MT $^{-1}$	10.25
$RO_{H,S}$	Heterotrophic oxygen requirement in the stabilization basin of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
$RO_{H,D,C}$	Heterotrophic oxygen requirement for biomass decay in the contact tank of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
$RO_{H,D,i}$	Heterotrophic oxygen requirement for biomass decay in tank i	MT $^{-1}$	10.26
$RO_{H,D,S}$	Heterotrophic oxygen requirement for biomass decay in a selector	MT $^{-1}$	Ex. 10.3.4.3
$RO_{H,D,S}$	Heterotrophic oxygen requirement for biomass decay in the stabilization basin of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
$RO_{H,SS}$	Heterotrophic oxygen requirement for biomass synthesis from readily biodegradable substrate	MT $^{-1}$	10.23
$RO_{H,SS,i}$	Heterotrophic oxygen requirement for biomass synthesis from readily biodegradable substrate in tank i	MT $^{-1}$	10.38
$RO_{H,SS,S}$	Heterotrophic oxygen requirement for biomass synthesis from readily biodegradable substrate in a selector	MT $^{-1}$	Ex. 10.3.4.3
$RO_{H,TS}$	Transient state oxygen requirement for heterotrophs	MT $^{-1}$	10.12
$RO_{H,XS}$	Heterotrophic oxygen requirement for biomass synthesis from slowly biodegradable substrate	MT $^{-1}$	10.24
$RO_{H,XS,C}$	Heterotrophic oxygen requirement for biomass synthesis from slowly biodegradable substrate in the contact tank of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
$RO_{H,XS,i}$	Heterotrophic oxygen requirement for biomass synthesis from slowly biodegradable substrate in tank i	MT $^{-1}$	10.38
$RO_{H,XS,S}$	Heterotrophic oxygen requirement for biomass synthesis from slowly biodegradable substrate in a selector	MT $^{-1}$	Ex. 10.3.4.3
$RO_{H,XS,S}$	Heterotrophic oxygen requirement for biomass synthesis from slowly biodegradable substrate in the stabilization basin of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4
RO_i	Oxygen requirement in tank i	MT $^{-1}$	10.37
RO_P	Peak oxygen requirement for a system	MT $^{-1}$	10.20
RO_S	Oxygen requirement in the stabilization basin of a CSAS system	MT $^{-1}$	Ex. 10.3.5.4

Symbol	Definition	Units	Equation where first used
$RX_{I,I}$	Input rate of inert organic solids	MT^{-1}	Eq. 14.13
S	Soluble constituent concentration	ML^{-3}	Sec. 5.1.1
S_A	Acetic acid concentration	ML^{-3}	3.82
S_{ALK}	Alkalinity	Mole	Sec. 6.1.1
S_{BALK}	Bicarbonate alkalinity concentration	ML^{-3}	13.7
Sc	Schmidt number	—	16.2
S_C	Soluble COD concentration	ML^{-3}	Sec. 8.2.1
S_{CO}	Influent soluble COD concentration	ML^{-3}	Sec. 8.2.1
S_D	Biomass associated product concentration	ML^{-3}	Sec. 2.4.7
S_I	Inhibitory chemical concentration	ML^{-3}	3.42
S_I^*	Inhibitor concentration that causes all microbial activity to cease	ML^{-3}	3.42
S_I	Soluble inert organic matter concentration	ML^{-3}	5.42
S_{IO}	Influent soluble inert organic matter concentration	ML^{-3}	5.42
S_{MeOH}	Methanol concentration	ML^{-3}	21.5
S_{MP}	Soluble microbial product concentration	ML^{-3}	3.70
$S_{N,a}$	Influent nitrogen concentration available to the nitrifying bacteria	ML^{-3}	10.16
$S'_{N,a}$	Influent soluble nitrogen concentration available to the nitrifying bacteria	ML^{-3}	10.50
S_{NH}	Ammonia-N concentration	ML^{-3}	3.51
S_{NHC}	Ammonia-N concentration in the contact tank of a CSAS system	ML^{-3}	10.48
S_{NHO}	Influent ammonia-N concentration	ML^{-3}	6.2
S_{NIO}	Influent soluble inert organic-N concentration	ML^{-3}	8.38
S_{NS}	Soluble organic-N concentration	ML^{-3}	3.78
S_{NSO}	Influent soluble organic-N concentration	ML^{-3}	8.25
S_{NO}	Nitrate-N concentration	ML^{-3}	3.48
S_{NOO}	Influent nitrate-N concentration	ML^{-3}	6.3
S_{NO2O}	Influent nitrite-N concentration	ML^{-3}	21.5
S_O	Dissolved oxygen concentration	ML^{-3}	3.31
S_{OO}	Influent dissolved oxygen concentration	ML^{-3}	21.5
S_S	Readily biodegradable (soluble) substrate concentration	ML^{-3}	3.31
$S_{S,a}$	Applied substrate concentration	ML^{-3}	16.5
S_{Sb}	Bulk liquid phase substrate concentration	ML^{-3}	15.1
S_{Sbmin}	Minimum bulk substrate concentration required to maintain a steady-state biofilm	ML^{-3}	15.22
$S_{Sbmin,2C}$	Minimum bulk liquid concentration of substrate 2 required to allow coexistence of two species in a biofilm	ML^{-3}	15.61
S_{SCMAL}	Soluble substrate concentration leaving a CMAL	ML^{-3}	14.15
S_{Se}	Effluent substrate concentration	ML^{-3}	16.5
S_{SF}	Readily biodegradable substrate concentration in a fictitious CSTR	ML^{-3}	10.28

Symbol	Definition	Units	Equation where first used
S_{Si}	Readily biodegradable (soluble) substrate concentration in bioreactor i	ML^{-3}	7.5
S_{SLR}	Soluble substrate concentration in the liquid film on the aerated sector of an RDR at its point of return to the tank	ML^{-3}	17.8
S_{Smin}	Minimum attainable soluble substrate concentration	ML^{-3}	5.14
$S_{S,n}$	Soluble substrate concentration in stage N of a multi-stage RBC	ML^{-3}	20.6
S_{SN}	Soluble substrate concentration in the Nth cell of a multicell benthal stabilization basin	ML^{-3}	14.15
S_{SO}	Influent readily biodegradable substrate concentration	ML^{-3}	5.15
S_{SOi}	Concentration of readily biodegradable substrate entering bioreactor i	ML^{-3}	7.5
S_{Ss}	Substrate concentration at the liquid–biofilm interface	ML^{-3}	15.1
S_S^*	Concentration of an inhibitory substrate giving μ^*	ML^{-3}	3.41
S_{Sb}^*	Dimensionless bulk substrate concentration	—	15.39
S_{Sbmin}^*	Dimensionless minimum bulk substrate concentration required to maintain a steady-state biofilm	—	15.33
S_{Ss}^*	Dimensionless substrate concentration at the liquid–biofilm interface	—	15.29
S_T	Tracer concentration	ML^{-3}	4.18
S_{TAlk}	Total alkalinity concentration	ML^{-3}	13.9
S_{TO}	Influent tracer concentration	ML^{-3}	4.18
S_{TN}	Tracer concentration leaving the Nth reactor	ML^{-3}	4.21
S_{VFA}	Volatile fatty acid concentration	ML^{-3}	13.9
S_{XOC}	Liquid phase concentration of an XOC	ML^{-3}	22.1
$S_{XOC,O}$	Influent liquid phase concentration of an XOC	ML^{-3}	22.17
S_{XOC}^*	Concentration of an XOC that would exist in the liquid phase if it were in equilibrium with the gas phase	ML^{-3}	22.1
SK	Spülkraft	L	19.4
t	Time	T	4.3
T	Temperature	°C, K	3.91
T_A	Weekly average air temperature	°F	14.3
T_i	Weekly average influent wastewater temperature	°F	14.3
T_L	Weekly average lagoon temperature	°F	14.3
TO	Rate of oxygen transfer	MT^{-1}	11.15
u	Temperature coefficient in Arrenhius equation	$kJMole^{-1}$	3.91

Symbol	Definition	Units	Equation where first used
U	Process loading factor	T^{-1}	5.37
U_{ANX}	Process loading factor for the anoxic zone	T^{-1}	11.11
U_i	Process loading factor in bioreactor i	T^{-1}	7.6
U_s	Process loading factor in a selector	T^{-1}	10.39
v	Velocity	LT^{-1}	4.23
v_{mf}	Minimum fluidization velocity	LT^{-1}	18.1
v_t	Terminal settling velocity	LT^{-1}	18.2
V	Volume	L^3	4.3
V_{AER}	Volume of aerobic zone	L^3	Ex. 11.3.1.2
V_{ANA}	Volume of anaerobic zone	L^3	Sec. 11.3.2
V_{ANX}	Volume of anoxic zone	L^3	Sec. 11.3.1.2
V_{br}	Volume of biomass retained per cycle in an SBR	L^3	7.12
$V_{br,min}$	Minimum possible volume of biomass retained per cycle in an SBR	L^3	10.58
V_C	Volume of contact tank in CSAS system	L^3	10.44
V_F	Volume of a fictitious CSTR	L^3	10.31
V_i	Volume of bioreactor i	L^3	7.1
V_{LFS}	Lower feasible bioreactor volume based on floc shear	L^3	10.4
V_{LOT}	Lower feasible bioreactor volume based on oxygen transfer	L^3	10.5
V_M	Volume of media in a trickling filter	L^3	19.1
V_{nr}	Volume of nitrate containing fluid retained per cycle in an SBR	L^3	7.13
V_N	Volume of the last tank in a chain of CSTRs	L^3	10.40
V_s	Volume of a selector	L^3	10.39
V_S	Volume of the stabilization basin in a CSAS system	L^3	10.47
V_T	Total volume of a chain of CSTRs	L^3	7.2
V_U	Upper feasible bioreactor volume	L^3	10.3
W	Width of a suspended growth bioreactor	L	10.36
W_M	Mass wastage rate of MLSS in COD units	MT^{-1}	5.48
$W_{M,T}$	Mass wastage rate of MLSS in TSS units	MT^{-1}	9.12
W_T	Mass wastage rate of total biomass	MT^{-1}	5.29
x	Distance from a reference point	L	4.7
X	Particulate constituent concentration	ML^{-3}	5.1
X_B	Active biomass concentration	ML^{-3}	3.35
X_{BA}	Active autotrophic biomass concentration in COD units	ML^{-3}	Sec. 6.1.1
$X_{BA,T}$	Active autotrophic biomass concentration in TSS units	ML^{-3}	10.15
$X_{BA,V}$	Active autotrophic biomass concentration in VSS units	ML^{-3}	12.3
X_{BH}	Active heterotrophic biomass concentration	ML^{-3}	3.31

Symbol	Definition	Units	Equation where first used
$X_{B,Hb}$	Active heterotrophic biomass concentration in the bulk liquid phase of a biofilm reactor	ML^{-3}	15.17
$X_{B,Hf}$	Active heterotrophic biomass concentration in a biofilm	ML^{-3}	15.6
$X_{B,H,TFL}$	Active heterotrophic biomass concentration in trickling filter effluent	ML^{-3}	19.11
$X_{B,H,0}$	Influent active heterotrophic biomass concentration	ML^{-3}	5.50
$X_{B,H,T}$	Active heterotrophic biomass concentration in TSS units	ML^{-3}	8.40
$X_{B,H,V}$	Active heterotrophic biomass concentration in VSS units	ML^{-3}	8.41
$X_{B,i}$	Density of species i at some point in a biofilm	ML^{-3}	15.51
$X_{B,P}$	PAO biomass concentration	ML^{-3}	3.82
$X_{B,XOC}$	Concentration of biomass capable of degrading an XOC	ML^{-3}	22.17
$X'_{B,XOC}$	Concentration of biomass capable of degrading an XOC that would be present in the absence of abiotic removal mechanisms	ML^{-3}	5.19 in Ch. 22
X_D	Biomass debris concentration in COD units	ML^{-3}	3.53
$X_{D,V}$	Biomass debris concentration in VSS units	ML^{-3}	12.3
X_{Df}	Density of biomass debris in a biofilm	ML^{-3}	Sec. 15.4
$X_{D,0}$	Influent biomass debris concentration	ML^{-3}	5.53
$X_{F,0}$	Influent fixed solids concentration	ML^{-3}	Sec. 8.7
X_I	Particulate, inert organic matter concentration in COD units	ML^{-3}	5.43
$X_{I,V}$	Particulate, inert organic matter concentration in VSS units	ML^{-3}	12.3
X_{IO}	Influent particulate inert organic matter concentration in COD units	ML^{-3}	5.43
$X_{IO,T}$	Influent particulate inert material concentration in TSS units	ML^{-3}	9.11
X_M	MLSS concentration i in COD units	ML^{-3}	5.46
X_{Me}	MLSS concentration in effluent stream in COD units	ML^{-3}	8.1
$X_{M,F}$	FSS concentration	ML^{-3}	12.15
$X_{M,F0}$	Influent concentration of FSS	ML^{-3}	12.16
X_{Mi}	MLSS concentration in bioreactor i in COD units	ML^{-3}	7.1
X_{Mr}	MLSS concentration in biomass recycle stream in COD units	ML^{-3}	5.69
X_{Mw}	MLSS concentration in biomass wastage stream in COD units	ML^{-3}	5.69
$X_{M,T}$	MLSS concentration in TSS units	ML^{-3}	9.11
$X_{M,T,ANA}$	MLSS concentration in TSS units in the anaerobic zone of a BNR system	ML^{-3}	11.19

Symbol	Definition	Units	Equation where first used
X_{MLANX}	MLSS concentration in TSS units in the anoxic zone of a BNR system	ML^{-3}	11.19
X_{MLC}	MLSS concentration in TSS units in the contact tank of a CSAS system	ML^{-3}	10.44
X_{MLI}	MLSS concentration in TSS units in a fictitious CSTR	ML^{-3}	10.30
$X_{ML,i}$	MLSS concentration in TSS units in tank i of a chain of CSTRs	ML^{-3}	10.42
X_{MLL}	Lower limit on feasible MLSS concentrations in TSS units	ML^{-3}	Ex. 10.3.3.3
X_{MLN}	MLSS concentration in TSS units in the last tank of a chain of CSTRs	ML^{-3}	10.40
X_{MLU}	Upper limit on feasible MLSS concentrations in TSS units	ML^{-3}	Ex. 10.3.3.3
X_{MR}	MLSS concentration in TSS units in the RAS stream	ML^{-3}	10.56
$X_{MR,max}$	Maximum attainable MLSS concentration in TSS units in the RAS stream	ML^{-3}	10.57
X_{MV}	MLSS concentration in VSS units	ML^{-3}	12.1
X_{MVO}	Influent MLSS concentration in VSS units	ML^{-3}	12.4
X_{MVP}	Biodegradable MLSS concentration in VSS units	ML^{-3}	12.1
X_{MVBO}	Influent biodegradable MLSS concentration in VSS units	ML^{-3}	12.5
X_{MVI}	Nonbiodegradable MLSS concentration in VSS units	ML^{-3}	12.1
X_{MVIO}	Influent nonbiodegradable MLSS concentration in VSS units	ML^{-3}	12.4
X_{NIO}	Influent inert organic-N concentration in COD units	ML^{-3}	8.38
X_{NS}	Particulate organic-N concentration	ML^{-3}	3.66
X_{NSO}	Influent particulate organic-N concentration	ML^{-3}	8.25
X_{PN}	Photosynthetic microorganism concentration in Nth tank	ML^{-3}	14.14
X_{PO}	Photosynthetic microorganism concentration in influent	ML^{-3}	14.14
X_{PP}	Polyphosphate concentration in biomass	ML^{-3}	3.82
X_{PHB}	PHB concentration in biomass	ML^{-3}	3.85
X_S	Slowly biodegradable (particulate) substrate concentration in COD units	ML^{-3}	3.77
X_{SV}	Slowly biodegradable (particulate) substrate concentration in VSS units	ML^{-3}	12.3
X_T	Total biomass concentration in COD units	ML^{-3}	5.24
X_{Te}	Total biomass concentration in effluent	ML^{-3}	Sec. 8.2.1
X_{Tw}	Total biomass concentration in wastage stream	ML^{-3}	Sec. 8.2.1
X_w	Particulate constituent concentration in the wastage stream	ML^{-3}	5.1

Symbol	Definition	Units	Equation where first used
Y	True growth yield	$M_{COD}M_{COD}^{-1}$	Sec. 2.4.1
Y_A	True growth yield for autotrophs on COD/N basis	$M_{COD}M_N^{-1}$	3.27
$Y_{A,T}$	True growth yield for autotrophs on TSS/N basis	$M_{TSS}M_N^{-1}$	8.48
$Y_{A,V}$	True growth yield for autotrophs on VSS/N basis	$M_{VSS}M_N^{-1}$	8.48
Y_H	True growth yield for heterotrophs on COD/COD basis	$M_{COD}M_{COD}^{-1}$	3.16
Y_{Hobs}	Observed yield for heterotrophs on COD/COD basis	$M_{COD}M_{COD}^{-1}$	5.27
$Y_{H,T}$	True growth yield for heterotrophs on TSS/COD basis	$M_{TSS}M_{COD}^{-1}$	8.42
$Y_{H,T,UB}$	True growth yield for heterotrophs on TSS/BOD_u basis	$M_{TSS}M_{BOD_u}^{-1}$	8.44
$Y_{H,T,SB}$	True growth yield for heterotrophs on TSS/BOD_s basis	$M_{TSS}M_{BOD_5}^{-1}$	8.46
$Y_{H,V}$	True growth yield for heterotrophs on VSS/COD basis	$M_{VSS}M_{COD}^{-1}$	8.43
$Y_{H,V,UB}$	True growth yield for heterotrophs on VSS/BOD_u basis	$M_{VSS}M_{BOD_u}^{-1}$	8.45
$Y_{H,V,SB}$	True growth yield for heterotrophs on VSS/BOD_5 basis	$M_{VSS}M_{BOD_5}^{-1}$	8.47
Y_{MP}	Microbial product yield	$M_{COD}M_{COD}^{-1}$	3.70
$Y_{Nitrobacter}$	True growth yield for *Nitrobacter*	$M_{COD}M_N^{-1}$	3.26
$Y_{Nitrosomonas}$	True growth yield for *Nitrosomonas*	$M_{COD}M_N^{-1}$	3.25
Y_n	Net process yield	MM^{-1}	9.2
Y_{obs}	Observed yield	$M_{COD}M_{COD}^{-1}$	5.28
Y_{O2}	Process oxygen stoichiometric coefficient	$M_{O_2}M_{BOD_5}^{-1}$	9.4
$Y_{O2,d}$	Oxygen stoichiometric coefficient for decay	$M_{O_2}M_{VSS}^{-1}$	19.12
$Y_{O2,s}$	Oxygen stoichiometric coefficient for synthesis	$M_{O_2}M_{BOD_5}^{-1}$	19.12
$Y_{O2,SG}$	Oxygen stoichiometric coefficient for suspended growth bioreactor in a TF/AS system	$M_{O_2}M_{BOD_5}^{-1}$	19.14
$Y_{O2,TF}$	Oxygen stoichiometric coefficient for a trickling filter	$M_{O_2}M_{BOD_5}^{-1}$	19.13
Y_P	Poly-P requirement (S_P release) for PHB storage	$M_PM_{COD}^{-1}$	3.84
Y_{PAO}	True growth yield for PAOs	$M_{COD}M_{COD}^{-1}$	3.86
Y_{XOC}	True growth yield for biomass degrading an XOC	$M_{COD}M_{COD}^{-1}$	22.17
z	Dimensionless length	—	4.23
α	Biomass recycle ratio	—	7.7
α	Empirical coefficient	$L^{0.333}T^{-0.667}$	17.3
α'	Empirical parameter in pseudoanalytical approach to biofilms	—	15.35

Symbol	Definition	Units	Equation where first used
α_1	Empirical coefficient	$LT^{-0.667}$	17.5
α_2	Empirical coefficient	L	17.6
α_3	Empirical coefficient	LT^{α_4}	17.6
α_4	Empirical exponent	—	17.6
α_5	Dimensionless abiotic loss coefficient for an XOC	—	22.19
α_6	Coefficient in COD mass balance to account for the type of nitrogen source	—	3.90
α_7	Dimensionless sorption coefficient for an XOC	—	22.20
α_8	Dimensionless violatilization coefficient for an XOC	—	22.20
β	Mixed liquor recirculation ratio	—	7.13
β'	Empirical parameter in pseudoanalytical approach to biofilms	—	15.35
γ	Liquid specific weight	$ML^{-2}T^{-2}$	10.6
γ	Fraction of XOC removal from a CMAS system attributable to abiotic removal mechanisms	—	22.19
$\Gamma_{A,S}$	Areal organic loading rate	$ML^{-2}T^{-1}$	14.1
$\Gamma_{A,XB}$	Areal loading rate for biodegradable suspended solids	$ML^{-2}T^{-1}$	14.12
$\Gamma_{V,S}$	Volumetric organic loading rate	$ML^{-3}T^{-1}$	13.1
δ	Anoxic mixed liquor recirculation ratio	—	11.19
$\Delta(F \cdot S_{NH})_{d,TS}$	Transient increase in available ammonia-N	MT^{-1}	10.14
$\Delta(F \cdot S_{NH0})$	Transient increase in ammonia-N loading	MT^{-1}	10.14
$\Delta(F \cdot S_{NSO})$	Transient increase in soluble organic-N loading	MT^{-1}	10.14
$\Delta(F \cdot S_{SO})$	Transient increase in the loading of readily biodegradable substrate	MT^{-1}	10.12
$\Delta(F \cdot X_{NSO})$	Transient increase in particulate organic-N loading	MT^{-1}	10.14
$\Delta(F \cdot X_{SO})$	Transient increase in the loading of slowly biodegradable substrate	MT^{-1}	10.12
$\Delta G^{0'}$	Gibbs free energy change	kJ	Sec. 2.4.1
ΔN	Mass rate of nitrate-N removal by denitrification	MT^{-1}	11.7
ΔN_{SS}	Mass rate of nitrate-N removal associated with biomass synthesis from readily biodegradable substrate	MT^{-1}	11.10
ΔN_{XB}	Mass rate of nitrate-N removal associated with biomass decay	MT^{-1}	Ex. 11.3.1.4
ΔN_{XS}	Mass rate of nitrate-N removal associated with slowly biodegradable substrate utilization and decay	MT^{-1}	Ex. 11.3.1.3

Symbol	Definition	Units	Equation where first used
$\Delta N/\Delta S$	Amount of nitrate-N required to serve as electron acceptor for a unit of substrate COD	$M_N M_{COD}^{-1}$	11.6
$\Delta S/\Delta N$	Amount of substrate COD required to remove a unit of nitrate-N by denitrification	$M_{COD} M_N^{-1}$	6.4
$\Delta X_{H,XOC}$	Decrease in the concentration of biomass degrading an XOC	ML^{-3}	22.18
ε	Porosity	—	16.2
ε_M	Porosity at incipient fluidization	—	18.1
ε_R	Porosity associated with the reference bed height	—	18.3
E_{XMV}	VSS destruction efficiency	percent	12.4
ζ	Fraction of a cycle in an SBR devoted to fill plus react	—	7.8
η_c	Effectiveness factor for biofilms	—	15.6
η_{cE}	External effectiveness factor	—	Sec. 15.2.2
η_{cI}	Internal effectiveness factor	—	Sec. 15.2.2
η_{cO}	Overall effectiveness factor	—	Sec. 15.2.2
η_{cOa}	Overall effectiveness factor in the aerated sector of an RDR	—	17.12
η_{cOs}	Overall effectiveness factor in the submerged sector of an RDR	—	17.8
η_{cZ}	Zero-order effectiveness factor	—	18.20
η_{c1}	First-order effectiveness factor	—	18.22
η_g	Anoxic growth factor	—	Table 6.1
η_h	Anoxic hydrolysis factor	—	Table 6.1
η_P	In-process energy efficiency for mechanical aeration systems	T^3L^{-2}	10.1
η_O	Field oxygen transfer efficiency for diffused aerations systems	—	10.2
θ	Temperature coefficient	—	3.95
θ	Dimensionless time	—	4.22
θ	Circumferential angle in RDR	°	Sec. 17.1
θ	Angle transcribed by the aerated sector of an RDR	°	17.11
Θ_c	Solids retention time	T	5.1
$\Theta_{c,AER}$	Aerobic solids retention time	T	11.1
$\Theta_{c,AER,eq}$	Equivalent aerobic solids retention time	T	11.18
$\Theta_{c,ANA}$	Anaerobic solids retention time	T	11.3
$\Theta_{c,ANX}$	Anoxic solids retention time	T	11.1
Θ_{ce}	Effective solids retention time in an SBR	T	7.9
Θ_{cmin}	Minimum solids retention time at which biomass can grow on a given influent substrate concentration	T	5.16

Symbol	Definition	Units	Equation where first used
Θ_{c}	Solids retention time required to obtain a desired effluent quality	T	10.10
$\Theta_{c,SS}$	Solids retention time required to degrade slowly biodegradable substrate	T	10.27
κ	Parameter depicting the deviation of the Thiele modulus from first order kinetics	—	15.13
λ_N	Surface total nitrogen loading to an attached growth bioreactor	$ML^{-2}T^{-1}$	Sec. 19.2.1
λ_{NH}	Surface ammonia-N loading to an attached growth bioreactor	$ML^{-2}T^{-1}$	Sec. 19.2.1
λ_S	Surface organic loading to an attached growth bioreactor	$ML^{-2}T^{-1}$	19.2
Λ_H	Total hydraulic loading to a bioreactor, superficial velocity	LT^{-1}	13.5
$\Lambda_{H,RBC}$	Total hydraulic loading on an RBC	LT^{-1}	20.3
Λ_N	Total nitrogen loading to an attached growth bioreactor	$ML^{-3}T^{-1}$	Sec. 19.1.2
Λ_{NH}	Total ammonia-N loading to an attached growth bioreactor	$ML^{-3}T^{-1}$	Sec. 19.1.2
Λ_{OR}	Oxidation rate in a trickling filter	$ML^{-3}T^{-1}$	19.3
Λ_S	Total organic loading to an attached growth bioreactor	$ML^{-3}T^{-1}$	19.1
μ	Specific growth rate coefficient	T^{-1}	3.35
$\mu_{A,C}$	Specific growth rate coefficient for autotrophs in the contact tank of a CSAS system	T^{-1}	10.48
μ_H	Specific growth rate coefficient for heterotrophs	T^{-1}	5.5
$\mu_{H,C}$	Specific growth rate coefficient for heterotrophs in the contact tank of a CSAS system	T^{-1}	10.44
$\mu_{H,F}$	Specific growth rate coefficient for heterotrophs in a fictitious CSTR	T^{-1}	10.28
μ_{Hi}	Specific growth rate coefficient for heterotrophs in bioreactor i	T^{-1}	7.5
$\mu_{H,N}$	Specific growth rate coefficient for heterotrophs in the last equivalent tank of a SFAS system	T^{-1}	10.40
$\mu_{H,max}$	Maximum specific heterotrophic growth rate associated with a given influent substrate concentration	T^{-1}	5.15
μ_P	Specific growth rate coefficient for photosynthetic microorganisms	T^{-1}	14.14
μ_w	Absolute viscosity of water	$MT^{-1}L^{-1}$	10.6
$\mu_{w,c}$	Absolute viscosity of water in centipoise	$MT^{-1}L^{-1}$	12.10
$\hat{\mu}_{A,m}$	Maximum specific growth rate coefficient for autotrophs at optimum pH	T^{-1}	3.50

Symbol	Definition	Units	Equation where first used
$\hat{\mu}_H$	Maximum specific growth rate coefficient for heterotrophs	T^{-1}	Sec. 3.2.10
$\hat{\mu}_P$	Maximum specific growth rate coefficient for PAOs	T^{-1}	3.85
$\hat{\mu}_{XOC}$	Maximum specific growth rate coefficient for degradation of an XOC	T^{-1}	5.13 in Ch. 22
μ^*	Maximum observed specific growth rate	T^{-1}	3.40
ξ	Parameter relating the dimensionless substrate flux to the dimensionless flux to a deep biofilm	—	15.31
Ξ	Fraction of substrate aerobically stabilized	—	14.10
Π	Volumetric power input	T^{-1} or $ML^{-1}T^{-3}$	Sec. 10.2.5
Π_L	Lower limit on volumetric power input	T^{-1} or $ML^{-1}T^{-3}$	Sec. 10.2.5
$\Pi_{L,P}$	Lower limit on volumetric power input for mechanical aeration systems	$ML^{-1}T^{-3}$	10.3
$\Pi_{L,O}$	Lower limit on volumetric power input for diffused aeration systems	T^{-1}	10.3
Π_U	Upper limit on volumetric power input	T^{-1} or $ML^{-1}T^{-3}$	Sec. 10.2.5
$\Pi_{U,P}$	Upper limit on volumetric power input for mechanical aeration systems	$ML^{-1}T^{-3}$	10.4
$\Pi_{U,O}$	Upper limit on volumetric power input for diffused aeration systems	T^{-1}	10.4
ρ_b	Density of a bioparticle	ML^{-3}	18.8
ρ_{bd}	Dry density of the biofilm on an FBBR bioparticle	ML^{-3}	18.9
ρ_{bw}	Wet density of the biofilm on an FBBR bioparticle	ML^{-3}	18.8
ρ_p	Density of a carrier particle	ML^{-3}	18.1
ρ_w	Density of water	ML^{-3}	Sec. 15.2.1
ς_{DO}	Dissolved oxygen safety factor	—	10.9
ς_{PL}	Peak load safety factor	—	10.8
ς_U	Safety factor for uncertainty	—	10.10
τ	Hydraulic residence time	T	4.15
τ_c	Effective hydraulic residence time in an SBR	T	7.8
τ_{min}	Minimum allowable hydraulic residence time	T	Ex. 5.1.3.1
τ_n	Hydraulic residence time in stage N	T	20.6
υ	Fraction of CSAS system volume in the contact tank	—	10.46

Symbol	Definition	Units	Equation where first used
Y_i	COD-based stoichiometric coefficient for reactant A_i	MM $^{-1}$	3.4
Y_{ij}	COD-based stoichiometric coefficient for reactant A_i in reaction j	MM $^{-1}$	Sec. 3.1.3
ϕ	Thiele modulus	—	15.12
ϕ_l	Modified Thiele modulus	—	15.14
ϕ_{Zm}	Modified zero-order Thiele modulus	—	18.19
ϕ_{1m}	Modified first-order Thiele modulus	—	18.21
Φ_{XOC}	Proportionality factor relating the mass transfer coefficient for an XOC to the mass transfer coefficient for oxygen	—	22.8
Ψ_i	Mass-based stoichiometric coefficient for reactant A_i	MM $^{-1}$	3.2
Ψ_{ij}	Mass-based stoichiometric coefficient for reactant A_i in reaction j	MM $^{-1}$	3.11
ω	Rotational speed of an RDR	RevT $^{-1}$	17.1
ω_d	Rotational speed of a rotary distributor on a trickling filter	RevT $^{-1}$	19.4
Ω	S_{Smin} in a CSTR receiving active biomass in the influent expressed as a fraction of S_{Smin} in absence of such biomass	—	5.56

Appendix C
Unit Conversions

U.S. Units to Metric Units

Multiply U.S. units	By	To obtain metric units
ac	4.047×10^{3}	m^2
ac	0.4047	ha
BTU	0.2520	kcal
BTU	1.055	kJ
BTU	2.931×10^{-4}	kW · hr
degrees F	0.5556 (°F − 32)	degrees C
ft	0.3048	m
ft/hp	0.4087	m/kW
ft^2	9.290×10^{-2}	m^2
ft^2/ft^3	3.281	m^2/m^3
ft^3	2.832×10^{-2}	m^3
$ft^3/(ft^2 \cdot min)$	0.3048	$m^3/(m^2 \cdot min)$
gal	3.785×10^{-3}	m^3
gal/(ac · day)	9.357×10^{-7}	$m^3/(m^2 \cdot day)$
$gal/(ft^2 \cdot day)$	4.074×10^{-2}	$m^3/(m^2 \cdot day)$
$gal/(ft^2 \cdot min)$	58.674	$m^3/(m^2 \cdot day)$
$gal/(ft^2 \cdot min)$	6.791×10^{-4}	$m^3/(m^2 \cdot sec)$
gal/min	5.451	m^3/day
gal/min	6.308×10^{-5}	m^3/sec
gal/(min · hp)	8.460×10^{-5}	$m^3/(sec \cdot kW)$
hp	0.7457	kW
hp/(1000 ft^3)	26.334	kW/(1000 m^3)
hp/(10^6 gal)	0.1973	kW/(1000 m^3)
lb (mass)	0.4536	kg
lb/(ac · day)	1.121	kg/(ha · day)
lb/(ac · day)	1.121×10^{-4}	$kg/(m^2 \cdot day)$
lb/(1000 ft^2 · day)	4.882	$g/(m^2 \cdot day)$
lb/(1000 ft^2 · day)	4.882×10^{-3}	$kg/(m^2 \cdot day)$
lb/(1000 ft^3 · day)	1.602×10^{-2}	$kg/(m^3 \cdot day)$
lb/(hp · hr)	0.6083	kg/(kW · hr)
Mgd	3.785×10^{3}	m^3/day
Mgd	4.381×10^{-2}	m^3/sec

Metric Units to U.S. Units

Multiply metric units	By	To obtain U.S. units
degrees C	1.800 (°C + 32)	degrees F
g/(m^2 · day)	0.2048	lb/(1000 ft^2 · day)
ha	2.471	ac
kcal	3.968	BTU
kg	2.205	lb (mass)
kg/(ha · day)	0.8922	lb/(ac · day)
kg/(kW · hr)	1.644	lb/(hp · hr)
kg/(m^2 · day)	8922	lb/(ac · day)
kg/(m^2 · day)	204.8	lb/(1000 ft^2 · day)
kg/(m^3 · day)	62.428	lb/(1000 ft^3 · day)
kJ	0.9478	BTU
kW	1.341	hp
kW · hr	3412	BTU
kW/(1000 m^3)	3.797×10^{-2}	hp/(1000 ft^3)
kW/(1000 m^3)	5.068	hp/(10^6 gal)
m	3.281	ft
m/kW	2.447	ft/hp
m^2	2.471×10^{-4}	ac
m^2	10.76	ft^2
m^2/m^3	0.3048	ft^2/ft^3
m^3	35.31	ft^3
m^3	264.2	gal
m^3/day	0.1835	gal/min
m^3/day	2.642×10^{-4}	Mgd
m^3/(m^2 · day)	1.069×10^{0}	gal/(ac · day)
m^3/(m^2 · day)	24.55	gal/(ft^2 · day)
m^3/(m^2 · day)	1.704×10^{-2}	gal/(ft^2 · min)
m^3/(m^2 · min)	3.281	ft^3/(ft^2 · min)
m^3/(m^2 · sec)	1472	gal/(ft^2 · min)
m^3/sec	1.585×10^{4}	gal/min
m^3/sec	22.83	Mgd
m^3/(sec · kW)	1.182×10^{4}	gal/(min · hp)

Index

This page intentionally left blank

about the first edition . . .

"...extremely well presented in clear and crisp language without compromising scholarly values."
—Food Technology

about the second edition . . .

Completely rewritten to reflect contemporary interest in bioreactors used in wastewater treatment while maintaining the features that made its predecessor the unrivalled choice of educators, students, and practitioners for nearly two decades, this ***Second Edition*** of a classic **textbook** integrates the principles of biochemical processes with their applications in the real world—communicating **modern, rational** approaches to the conception, design, operation, and optimization of **existing** and **novel** biochemical unit operations *in a comprehensive yet lucid manner.*

about the authors . . .

C. P. LESLIE GRADY, JR. is an R. A. Bowen Professor of Environmental Engineering and Science at Clemson University, South Carolina. His extensive research experience has focused on the biodegradation of toxic organic chemicals, the fate of pollutants in wastewater treatment systems, and the impact of alternative carbon sources on the kinetics of xenobiotic organic compound degradation. The author of over 200 technical publications, Dr. Grady is widely recognized for his research and teaching in the discipline. He is a Fellow of the American Academy of Microbiology, a Diplomate of the American Academy of Environmental Engineers, and a member of the Water Environment Federation, among others. He received the Ph.D. degree (1969) from Oklahoma State University, Stillwater.

GLEN T. DAIGGER is Senior Vice President at CH2M Hill, Englewood, Colorado. Previously Professor in, and Chair of, the Department of Environmental Systems Engineering at Clemson University, South Carolina, he has established major technical initiatives in such areas as selector activated sludge systems, coupled trickling filer/activated sludge process technology, nutrient control, and plant analysis technologies. The holder of two U.S. patents, his publications include three books and numerous journal papers. A member of the Water Environment Federation and the International Association on Water Quality, among others, Dr. Daigger received the Ph.D. degree (1979) in environmental engineering from Purdue University, West Lafayette, Indiana.

HENRY C. LIM is a Professor of Chemical, Biochemical, and Environmental Engineering at the University of California, Irvine, where as Founding Chair he established the Department of Chemical and Biochemical Engineering. His previous experiences include industrial research and development work in bioreactions and bioseparations at Pfizer, Inc., Groton, Connecticut, and 21 years at Purdue University, West Lafayette, Indiana, where he initiated research in biochemical engineering. The author of over 150 publications, his research has focused on the optimization and control of bioreactors and fermentation processes, particularly fedbatch cultures and engineering applications for recombinant DNA technology. He is also the recipient of numerous teaching and research awards. Professor Lim received the Ph.D. degree (1967) from Northwestern University, Evanston, Illinois.

Printed in the United States of America

MARCEL DEKKER
NEW YORK

MARCEL DEKKER

DK4768
ISBN 0-8247-8919-9
90000
9 780824 789190