Introduction to Wastewater Treatment Processes

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Preface

This book is an introductory presentation meant for both students and practicing engineers interested in the field of wastewater treatment. Most of the earlier books discuss the subject industry by industry, providing solutions to specific treatment problems. More recently, a scientific approach to the basic principles of unit operations and processes has been utilized. I have used this approach to evaluate all types of wastewater problems and to properly select the mode of treatment and the design of the equipment required.

In most cases, the design of specific wastewater treatment processes, e.g., the activated sludge process, is discussed following (1) a summary of the theory involved in the specific process, e.g., chemical kinetics, pertinent material and energy balances, discussion of physical and chemical principles; (2) definition of the important design parameters involved in the process and the determination of such parameters using laboratory-scale or pilot-plant equipment; and (3) development of a systematic design procedure for the treatment plant. Numerical applications are presented which illustrate the treatment of laboratory data, and subsequent design calculations are given for the wastewater processing plant. The approach followed, particularly in the mathematical modeling of biological treatment processes, is based largely on the work of Eckenfelder and associates.

Clarity of presentation has been of fundamental concern. The text should be easily understood by undergraduate students and practicing engineers. The book stems from a revision of lecture notes which I used for an introductory course on wastewater treatment. Not only engineering students of diverse backgrounds but also practicing engineers from various fields have utilized these notes at the different times this course was offered at Laval University and COPPE/UFRJ (Rio de Janeiro, Brazil). Favorable acceptance of the notes and the encouragement of many of their users led me to edit them for inclusion in this work.

I wish to express my appreciation to the secretarial staff of the Chemical Engineering Department of Laval University, Mrs. Michel, Mrs. Gagné, and Mrs. McLean, and to Miss Enidete Souza (COPPE/UFRJ) for typing the manuscript. I owe sincere thanks to Mr. Alex Légaré for the artwork, to Dr. and Mrs. Adrien Favre for proofreading the manuscript, and to Mr. Roger Thériault for his assistance in the correction of the galleys. The valuable suggestions made by Dr. M. Pelletier (Laval University) and Dr. C. Russo (COPPE/UFRJ) are gratefully acknowledged.

R. S. Ramalho

1

Introduction

1. Introduction

It was only during the decade of the 1960's that terms such as "water and air pollution," "protection of the environment," and "ecology" became household words. Prior to that time, these terms would either pass unrecognized by the average citizen, or at most, would convey hazy ideas to his mind. Since then mankind has been bombarded by the media (newspapers, radio, television), with the dreadful idea that humanity is effectively working for its self-destruction through the systematic process of pollution of the environment, for the sake of achieving material progress. In some cases, people have been aroused nearly to a state of mass hysteria. Although pollution is a serious problem, and it is, of course, desirable that the citizenry be concerned about it, it is questionable that "mass hysteria" is in any way justifiable. The instinct of preservation of the species is a very basic driving force of humanity, and man is equipped to correct the deterioration of his environment before it is too late. In fact, pollution control is not an exceedingly difficult technical problem as compared to more complex ones which have been successfully solved in this decade, such as the manned exploration of the moon. Essentially, the basic technical knowledge required to cope with pollution is already available to man, and as long as he is willing to pay a relatively reasonable price tag, the nightmare of self-destruction via pollution will never become a reality. Indeed, much higher price tags are being paid by humanity for development and maintenance of the war-making machinery.

This book is primarily concerned with the engineering design of process plants for treatment of wastewaters of either domestic or industrial origin. It is only in the last few years that the design approach for these plants has changed from empiricism to a sound engineering basis. Also, fundamental research in new wastewater treatment processes, such as reverse osmosis and electrodialysis, has only recently been greatly emphasized.

2. The Role of the Engineer in Water Pollution Abatement

2.1. THE NECESSITY OF A MULTIDISCIPLINARY APPROACH TO THE WATER POLLUTION ABATEMENT PROBLEM

Although it has been stated previously that water pollution control is not an exceedingly difficult technical problem, the field is a broad one, and of sufficient complexity to justify several different disciplines being brought together for achieving optimal results at a minimum cost. A systems approach to water pollution abatement involves the participation of many disciplines: (1) engineering and exact sciences [sanitary engineering (civil engineering), chemical engineering, other fields of engineering such as mechanical and electrical, chemistry, physics]; (2) life sciences [biology (aquatic biology), microbiology, bacteriology]; (3) earth sciences (geology, hydrology, oceanography); and (4) social and economic sciences (sociology, law, political sciences, public relations, economics, administration).

2.2. A SURVEY OF THE CONTRIBUTION OF ENGINEERS TO WATER POLLUTION ABATEMENT

The sanitary engineer, with mainly a civil engineering background, has historically carried the brunt of responsibility for engineering activities in water pollution control. This situation goes back to the days when the bulk of wastewaters were of domestic origin. Composition of domestic wastewaters does not vary greatly. Therefore, prescribed methods of treatment are relatively standard, with a limited number of unit processes and operations involved in the treatment sequence. Traditional methods of treatment involved large concrete basins, where either sedimentation or aeration were performed, operation of trickling filters, chlorination, screening, and occasionally a few other operations. The fundamental concern of the engineer was centered around problems of structure and hydraulics, and quite naturally, the civil engineering background was an indispensable prerequisite for the sanitary engineer.

This situation has changed, at first gradually, and more recently at an accelerated rate with the advent of industrialization. As a result of a new large variety of industrial processes, highly diversified wastewaters requiring more complex treatment processes have appeared on the scene. Wastewater treatment today involves so many different pieces of equipment, so many unit processes and unit operations, that it became evident that the chemical engineer had to be called to play a major role in water pollution abatement. The concept of unit operations, developed largely by chemical engineers in the past fifty years, constitutes the key to the scientific approach to design problems encountered in the field of wastewater treatment.

In fact, even the municipal wastewaters of today are no longer the "domestic wastewaters" of yesterday. Practically all municipalities in industrialized areas must handle a combination of domestic and industrial wastewaters. Economic and technical problems involved in such treatment make it very often desirable to perform separate treatment (*segregation*) of industrial wastewaters, prior to their discharge into municipal sewers.

Even the nature of truly domestic wastewaters has changed with the advent of a whole series of new products now available to the average household, such as synthetic detergents and others. Thus, to treat domestic wastewaters in an optimum way requires modifications of the traditional approach.

In summary, for treatment of both domestic and industrial wastewaters, new technology, new processes, and new approaches, as well as modifications of old approaches, are the order of the day. The image today is no longer that of the "large concrete basins," but one of a series of closely integrated unit operations. These operations, both physical and chemical in nature, must be tailored for each individual wastewater. The chemical engineer's skill in integrating these unit operations into effective processes makes him admirably qualified to design wastewater treatment facilities.

2.3. A CASE HISTORY OF INDUSTRIAL WASTEWATER TREATMENT

An interesting case history, emphasizing the role of the chemical engineer in the design of a wastewater treatment plant for a sulfite pulp and paper mill, is discussed by Byrd [2]. This pulp and paper plant was to discharge its wastewaters into a river of prime recreational value, with a well-balanced fish population. For this reason, considerable care was taken in the planning and detailed design of the wastewater treatment facilities. A study of assimilative capacity of the river was undertaken and mathematical models were developed.

Design of the treatment plant involved a study to determine which wastewater effluents should be segregated for treatment, and which ones should be combined. For the treatment processes a selection of alternatives is discussed [2]. Some of the unit operations and processes involved in the treatment plant, or considered at first but after further study replaced by other alternatives, were the following: sedimentation, dissolved air flotation, equalization, neutralization, filtration (rotary filters), centrifugation, reverse osmosis, flash drying, fluidized bed oxidation, multiple hearth incineration, wet oxidation, adsorption in activated carbon, activated sludge process, aerated lagoons, flocculation with polyelectrolytes, chlorination, landfill, and spray irrigation.

Integration of all these unit operations and processes into an optimally designed treatment facility constituted a very challenging problem. The treatment plant involved a capital cost of over $10 million and an operating cost in excess of $1 million per year.

2.4. THE CHEMICAL ENGINEERING CURRICULUM AS A PREPARATION FOR THE FIELD OF WASTEWATER TREATMENT [5]

Chemical engineers have considerable background that is applicable to water pollution problems. Their knowledge of mass transfer, chemical kinetics, and systems analysis is specially valuable in wastewater treatment and control. Thus, training in chemical engineering represents good preparation for entering this type of activity. In the past, the majority of engineers working in this field have been sanitary engineers with a civil engineering background.

The multidisciplinary nature of the field should be recognized. Chemical engineering graduates envisioning major activity in the field of wastewater treatment are advised to complement their background by studying microbiology, owing to the great importance of biological wastewater treatment processes, and also hydraulics [since topics such as open channel and stratified flow, mathematical modeling of bodies of water (rivers, estuaries, lakes, inlets, etc.) are not emphasized in fluid mechanics courses normally offered to chemical engineering students].

2.5. "INPLANT" AND "END-OF-PIPE" WASTEWATER TREATMENT [6]

2.5.1. Introduction

Frequently one may be tempted to think of industrial wastewater treatment in terms of an "end-of-pipe" approach. This would involve designing a plant without much regard to water pollution abatement, and then considering separately the design of wastewater treatment facilities. Such an approach should not be pursued since it is, in general, highly uneconomical.

The right approach for an industrial wastewater pollution abatement program is one which uncovers all opportunities for inplant wastewater treatment. This may seem a more complicated approach than handling wastewaters at the final outfall. However, such an approach can be very profitable.

2.5.2. What Is Involved in Inplant Wastewater Control

Essentially, inplant wastewater control involves the three following steps:

Step 1. Perform a detailed survey of all effluents in the plant. All pollution sources must be accounted for and cataloged. This involves, for each polluting stream, the determination of (a) flow rate and (b) strength of the polluting streams.

(a) *Flow rate.* For continuous streams, determine flow rates (e.g., gal/min). For intermittent discharges, estimate total daily (or hourly) outflow.

(b) *Strength of the polluting streams.* The "strength" of the polluting streams (concentration of polluting substances present in the streams) is expressed in a variety of ways, which are discussed in later chapters. For organic compounds which are subject to biochemical oxidation, the biochemical oxygen demand, BOD (which is defined in Chapter 2, Section 2.3) is commonly employed. In the case history summarized in Section 2.5.3 of this chapter, BOD is used to measure concentration of organics.

Step 2. Review data obtained in Step 1 to find all possible inplant abatement targets. Some of these are (1) increased recycling in cooling water systems; (2) elimination of contact cooling for off vapors, e.g., replacement of barometric condensers by shell-and-tube exchangers or air-cooling systems; (3) recovery of polluting chemicals: Profit may often be realized by recovering such chemicals, which are otherwise discharged into the plant sewers. A byproducts plant may be designed to recover these chemicals; (4) reuse of water from overhead accumulator drums, vacuum condensers, and pump glands. Devise more consecutive or multiple water uses; (5) design a heat recovery unit to eliminate quenching streams; and (6) eliminate leaks and improve housekeeping practices. Automatic monitoring and additional personnel training might be profitable.

Step 3. Evaluate potential savings in terms of capital and operating costs for a proposed "end-of-pipe" treatment, if each of the streams considered in Steps 1 and 2 are either eliminated or reduced (reduction in flow rates or in terms of strength of polluting streams). Then design the "end-of-pipe" treatment facilities to handle this reduced load. Compare capital and operating costs of such treatment facilities with that of an "end-of-pipe" facility designed to handle the original full load, i.e., the pollutant streams from a plant where inplant wastewater control is not practiced. The two case histories described in Ref. [6] are quite revealing in this respect.

For practicing inplant wastewater control, a deep knowledge of the process and ability to modify it, if necessary, are required. The chemical engineer is admirably well suited to handle this job.

2.5.3. Case Histories of Inplant Wastewater Control

Two interesting case histories are discussed by McGovern [6]. One of these, pertaining to a petrochemical plant, is summarized next.

A petrochemical plant already in operation conducted an effluent and inplant survey while evaluating a treatment plant to be designed and built, which would handle 20 million gal/day of wastewater with a BOD load of 52,000 lb/day. The plan called for an activated sludge unit to remove over 90% of the BOD load. This included vacuum filtration and incineration of the sludge, and chlorination of the total effluent.

Capital cost of the treatment facility was estimated at $10 million. Operating and maintenance costs were also estimated. All cost data were converted to an annual basis, using a 20-year project life and 15% interest rate.

Then a study of the possibility of reducing both the flow and the strength of the wastewaters was undertaken. This study followed the steps outlined under Section 2.5.2, with a number of changes being proposed for the process flowsheet. The reduction accomplished in flow rate and strength resulted in substantial savings in the total cost of the proposed treatment plant. Figure 1.1 shows a graph, prepared for this case history, illustrating the effect of reduction

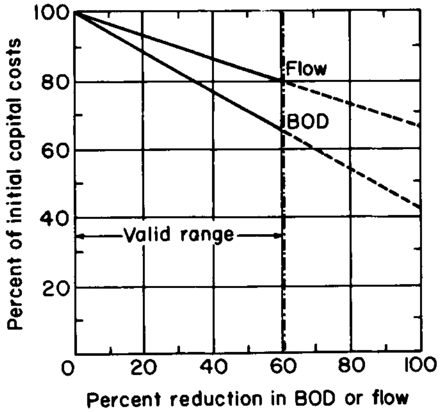

Fig. 1.1. Effect of waste load reductions on capital cost of treatment plant **[6]**. *(Excerpted by special permission from Chemical Engineering, May 14, 1973. Copyright by McGraw-Hill, Inc., New York, 10020.)*

TABLE 1.1
Savings from Inplant Wastewater Reductions[a]

Inplant savings	$/year
Flow reduction (1424 gal/min)	$410,000
BOD reduction (2000 lb/day)	302,000
Water use reduction	
Treated water (0.24 MGD)	34,000
River water (1.37 MGD)	14,000
Product recovery	14,000
Total inplant saving	$774,000
Cost of inplant control	$/year
Engineering	$ 15,000
Capital investment	150,000
Operating and maintenance	33,000
Total cost of inplant control	$198,000
Net savings: $774,000 − $198,000 = $576,000/year	

[a] Excerpted by special permission from *Chemical Engineering*, May 14, 1973; Copyright by McGraw-Hill, Inc., New York, 10020.

in BOD or flow rate upon the capital cost of the treatment facilities. This graph is valid to approximately 60% reduction in flow or BOD. Any further reduction probably requires a significantly different treatment system.

Savings from inplant wastewater control are tabulated in Table 1.1. Wastewater flow was cut to 85% of its value prior to inplant control and BOD load was cut to 50%. Moreover, the cost of these inplant controls was more than offset by economies in the treatment plant. As shown in Table 1.1 the program realized a net saving of $576,000/year.

2.6. A NEW CONCEPT IN PROCESS DESIGN: THE FLOWSHEET OF THE FUTURE

The considerations in Section 2.5 are leading engineers to a new concept in process design. The flowsheet of the future will no longer show a line with an arrowhead stating "to waste." Essentially everything will be recycled, by-products will be recovered, and water will be reused. Fundamentally the only streams in and out of the plant will be raw materials and products. The only permissible wastages will be clean ones: nitrogen, oxygen, carbon dioxide, water, and some (but not too much!) heat. In this connection, it is appropriate to recall the guidelines of the United States Federal Water Pollution Control Act of 1972: (1) best *practical* control technology, by July 1, 1977; (2) best *available* technology, by July 1, 1983; and (3) zero discharge by July 1, 1985.

3. Degrees of Wastewater Treatment and Water Quality Standards

The degree of treatment required for a wastewater depends mainly on discharge requirements for the effluent. Table 1.2 presents a conventional classification for wastewater treatment processes. Primary treatment is employed for removal of suspended solids and floating materials, and also conditioning the wastewater for either discharge to a receiving body of water or to a secondary treatment facility through neutralization and/or equalization. Secondary treatment comprises conventional biological treatment processes. Tertiary treatment is intended primarily for elimination of pollutants not removed by conventional biological treatment.

TABLE 1.2
Types of Wastewater Treatment

- Primary treatment
  - Screening
  - Sedimentation
  - Flotation
  - Oil separation
  - Equalization
  - Neutralization
- Secondary treatment
  - Activated sludge process
  - Extended aeration (or total oxidation) process
  - Contact stabilization
  - Other modifications of the conventional activated sludge process: tapered aeration, step aeration, and complete mix activated sludge processes
  - Aerated lagoons
  - Wastewater stabilization ponds
  - Trickling filters
  - Anaerobic treatment
- Tertiary treatment (or "advanced treatment")
  - Microscreening
  - Precipitation and coagulation
  - Adsorption (activated carbon)
  - Ion exchange
  - Reverse osmosis
  - Electrodialysis
  - Nutrient removal processes
  - Chlorination and ozonation
  - Sonozone process

These treatment processes are studied in following chapters. The approach utilized is based on the concepts of unit processes and operations. The final objective is development of design principles of general applicability to *any* wastewater treatment problem, leading to a proper selection of process and the design of required equipment. Consequently, description of wastewater treatment sequences for specific industries, e.g., petroleum refineries, steel mills, metal-plating plants, pulp and paper industries, breweries, and tanneries, is not included in this book. For information on specific wastewater treatment processes, the reader should consult Eckenfelder [3] and Nemerow [7].

Water quality standards are usually based on one of two criteria: stream standards or effluent standards. *Stream standards* refer to quality of receiving water downstream from the origin of sewage discharge, whereas *effluent standards* pertain to quality of the discharged wastewater streams themselves.

A disadvantage of effluent standards is that it provides no control over total amount of contaminants discharged in the receiving water. A large industry, for example, although providing the same degree of wastewater treatment as a small one, might cause considerably greater pollution of the receiving water. Effluent standards are easier to monitor than stream standards, which require detailed stream analysis. Advocates of effluent standards argue that a large industry, due to its economic value to the community, should be allowed a larger share of the assimilative capacity of the receiving water.

Quality standards selected depend on intended use of the water. Some of these standards include: concentration of dissolved oxygen (DO, mg/liter), pH, color, turbidity, hardness (mg/liter), total dissolved solids (TDS, mg/liter), suspended solids (SS, mg/liter), concentration of toxic (or otherwise objectionable) materials (mg/liter), odor, and temperature. Extensive tabulation of water quality standards for various uses and for several states in the United States is presented by Nemerow [7].

4. Sources of Wastewaters

Four main sources of wastewaters are (1) domestic sewage, (2) industrial wastewaters, (3) agricultural runoff, and (4) storm water and urban runoff. Although the primary consideration in this book is the study of treatment of domestic and industrial wastewaters, contamination due to agricultural and urban runoffs is becoming increasingly important. Agricultural runoffs carrying fertilizers (e.g., phosphates) and pesticides constitute a major cause of eutrophication of lakes, a phenomena which is discussed in Section 7 of this chapter. Storm runoffs in highly urbanized areas may cause significant pollution effects. Usually wastewaters, treated or untreated, are discharged into a natural body of water (ocean, river, lake, etc.) which is referred to as the receiving water.

5. Economics of Wastewater Treatment and Economic Balance for Water Reuse

In the United States average cost per thousand gallons of water is approximately \$0.20, which corresponds to \$0.05/ton. It is a relatively cheap commodity, and as a result the economics of wastewater treatment is very critical. In principle, by utilizing sophisticated treatment processes, one can obtain potable water from sewage. Economic considerations, however, prevent the practical application of many available treatment methods. In countries where water is at a premium (e.g., Israel, Saudi Arabia) some sophisticated water treatment facilities, which are not economically justified in North America, are now in operation. In evaluating a specific wastewater treatment process, it is important to estimate a *cost–benefit ratio* between the benefit derived from the treatment to obtain water of a specified quality, and the cost for accomplishing this upgrading of quality.

Reuse of water by recycling has been mentioned in connection with inplant wastewater control (Section 2.5). Selection of an optimum recycle ratio for a specific application involves an economic balance in which three factors must be considered [3]: (1) cost of raw water utilized in the plant; (2) cost of wastewater treatment to suitable process quality requirements (in Example 1.1, this is the cost of wastewater treatment preceding recycling to the plant for reuse); and (3) cost of wastewater treatment prior to discharge into a receiving water, e.g., in a river.

This economic balance is illustrated by Example 1.1.

Example 1.1 [3]

A plant uses 10,000 gal/hr of process water with a maximum contaminant concentration of 1 lb per 1000 gal. The raw water supply has a contaminant concentration of 0.5 lb/1000 gal. Optimize a water reuse system for this plant based on raw water cost of \$0.20/1000 gal. Utilize data in Fig. 1.2 to estimate costs for the two water treatment processes involved in the plant. The contaminant is nonvolatile.

The following conditions apply: (1) evaporation and product loss (stream E in Fig. 1.3): 1000 gal/hr of water; (2) contaminant addition (stream Y in Fig. 1.3): 100 lb/hr of contaminant; and (3) maximum discharge allowed to receiving water: 20 lb/hr of contaminant.

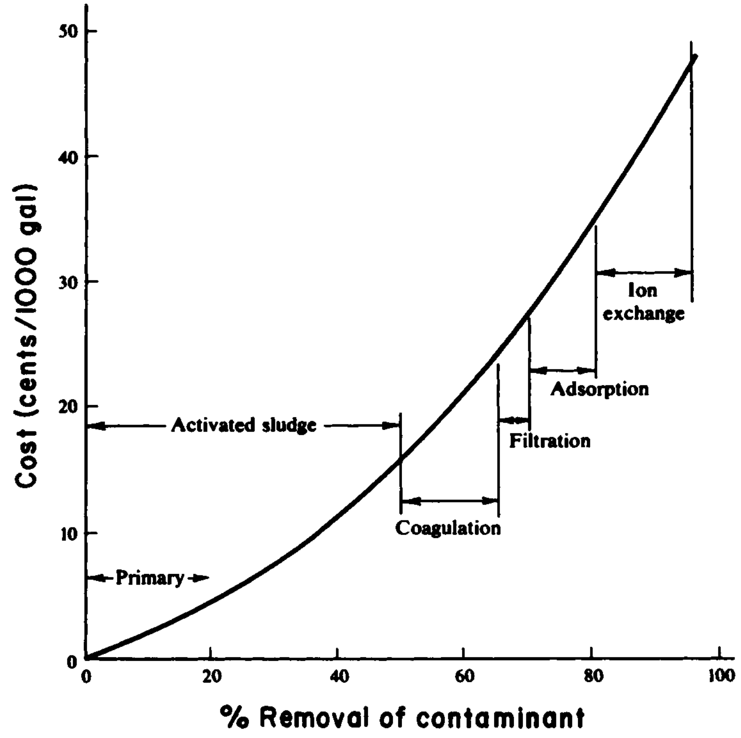

Fig. 1.2. Relationship between total cost and type of treatment [3].

SOLUTION A block flow diagram for the process is presented in Fig. 1.3. Values either assumed or calculated are underlined in Fig. 1.3. Values not underlined are basic data for the problem. Volumetric flow rates of streams 9, 10, and 11 are negligible.

The procedure for solution consists of assuming several values for the water recycle R (gal/hr). For each assumed value, the material balance is completed and the economic evaluation is made.

Step 1. Start assuming a 70% recycle, i.e., $R/A = 0.7$ (recycle ratio), where R is the recycle, i.e., stream 2 (gal/hr), and A is the combined feed, i.e., stream 3 (10,000 gal/hr). Then, calculate the recycle:

$$R = (0.7)(A) = (0.7)(10{,}000) = 7000 \text{ gal/hr} \quad [\text{stream } 2]$$

Thus, stream 5 in Fig. 1.3 also corresponds to a flow rate of 7000 gal/hr since the volumetric flow rate of contaminant removed [stream 11] is negligible.

Step 2. For this assumed recycle, the raw water feed [stream 1] is

$$F = A - R = 10{,}000 - 7000 = 3000 \text{ gal/hr}$$

Step 3. Effluent from the plant [stream 4] is

$$A - E = 10{,}000 - 1000 = 9000 \text{ gal/hr}$$

Step 4. From the material balance it follows that since stream 4 is split into streams 5 and 6,

Stream 6: 9000 − 7000 = 2000 gal/hr
Stream 7: 2000 gal/hr

Thus for 70% recycle, volumetric flow rates for all streams in Fig. 1.3 are now determined.

Step 5. Mass flow rate of contaminant in raw water [stream 1] is

$$F \times (0.5/1000) = 3000(0.5/1000) = 1.5 \text{ lb/hr}$$

Step 6. Mass flow rate of contaminant in stream 3 is

$$(1/1000) \times 10{,}000 = 10 \text{ lb/hr}$$

Step 7. Mass flow rate of contaminant in the recycle [stream 2] is

$$10 - 1.5 = 8.5 \text{ lb/hr}$$

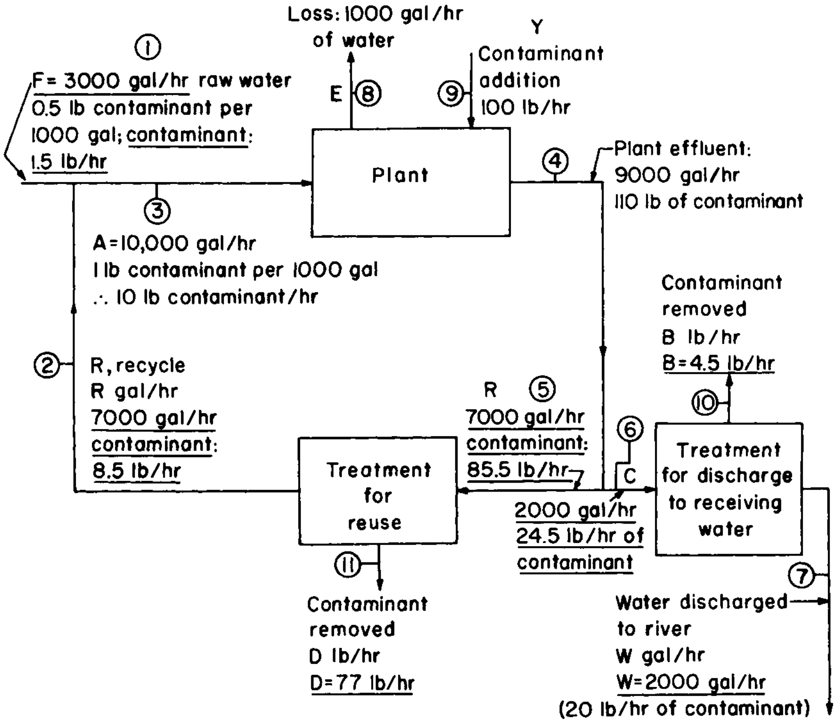

Fig. 1.3. ***Flow diagram for Example 1.1. Encircled numbers are streams.*** (*Adapted from Eckenfelder* [3].)

Step 8. Mass flow rate of contaminant in the plant effluent [stream 4] is

$$10 \text{ [from stream 3]} + 100 \text{ [from stream 9]} = 110 \text{ lb/hr}$$

Step 9. Mass flow rate of contaminant in streams 5 and 6 is

Stream 5: $(7000/9000) \times 110 = 85.5$ lb/hr
Stream 6: $110 - 85.5 = 24.5$ lb/hr

Step 10. Since the mass flow rate of contaminant in stream 7 is 20 lb/hr, that for contaminant in stream 10 is

$$24.5 - 20.0 = 4.5 \text{ lb/hr}$$

Step 11. Mass flow rate of contaminant removed in the treatment for reuse [stream 11] is

$$85.5 - 8.5 = 77.0 \text{ lb/hr}$$

Step 12. The % removal of contaminant in the two treatments is

Treatment for reuse: $(77/85.5) \times 100 = 90\%$
Treatment for discharge to receiving water: $(4.5/24.5) \times 100 = 18.4\%$

Step 13. The type of treatment required is essentially established from these % removals of contaminant (Fig. 1.2). In the treatment for reuse (90% removal), ion exchange is indicated. For discharge to receiving water (18.4% removal), Fig. 1.2 indicates that primary treatment is sufficient. Costs for these treatments are read from Fig. 1.2.

Treatment for reuse (90% removal): $0.42/1000 gal
Treatment for discharge to receiving water (18.4% removal): $0.05/1000 gal

Step 14. Daily cost for 70% recycle:

Raw water:
$$3000\,\frac{\text{gal}}{\text{hr}} \times \frac{\$0.20}{1000 \text{ gal}} \times 24\,\frac{\text{hr}}{\text{day}} = \$14.40/\text{day}$$

Cost

Effluent treatment for discharge to river:

$$2000\,\frac{\text{gal}}{\text{hr}} \times \frac{\$0.05}{1000 \text{ gal}} \times 24\,\frac{\text{hr}}{\text{day}} = \$\ 2.40/\text{day}$$

Treatment for reuse:
$$7000\,\frac{\text{gal}}{\text{hr}} \times \frac{\$0.42}{1000 \text{ gal}} \times 24\,\frac{\text{hr}}{\text{day}} = \$70.56/\text{day}$$

Total: $87.36/day

Step 15. This cost is plotted in Fig. 1.4 vs. 70% reflux. A similar series of calculations is made for freshwater inputs varying from 10,000 to 2000 gal/hr,

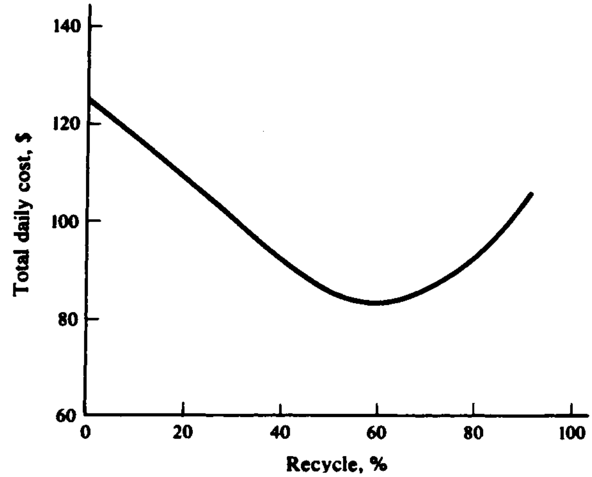

Fig. 1.4. Relationship between total daily water cost and treated waste recycle for reuse [3].

with recycles varying, respectively, from 0–80%. Figure 1.4 is obtained, which indicates that the optimum recycle is approximately 60% for a cost of about $83.00/day.

6. Effect of Water Pollution on Environment and Biota

Bartsch and Ingram [1] made an interesting study of the effect of water pollution on environment and biota. These effects are illustrated by Figs. 1.5–1.10, and a summary of their work is presented next. The source of pollution considered was raw domestic sewage for a community of 40,000 people, flowing to a stream with a volume flow of 100 ft^3/sec. Lowering of the concentration of dissolved oxygen (DO) and formation of sludge deposits are the most common environmental disturbances which may damage aquatic biota.

6.1. OXYGEN SAG CURVE

The curve in Fig. 1.5, referred to as dissolved oxygen curve, is a plot of dissolved oxygen concentration (mg/liter) for a stream. It is referred to hence as oxygen sag curve. Sewage is discharged at the point identified as zero (0) on the abscissa axis. The values to the right of point zero represent miles downstream of the point of sewage discharge. Complete mixing is assumed, and the water temperature is 25°C. An alternative scale for the abscissa, in terms of days of flow, is shown in Fig. 1.5.

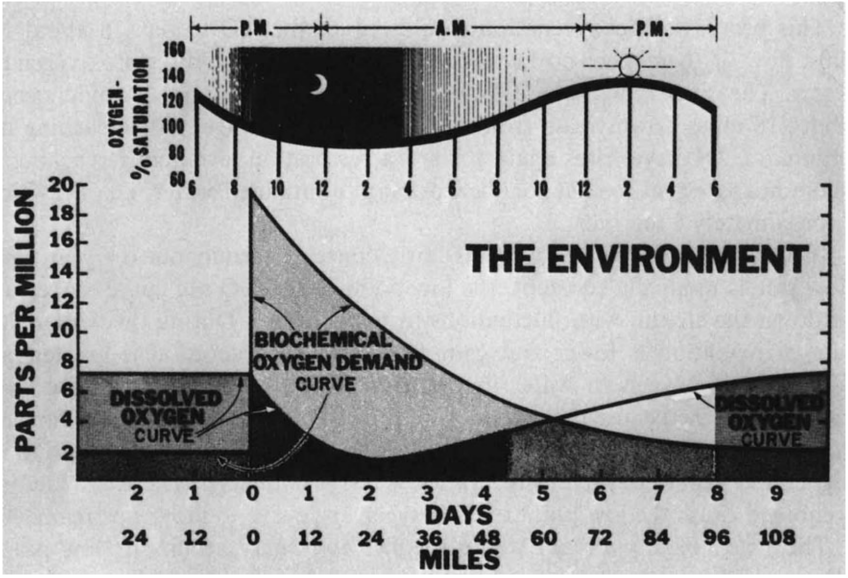

Fig. 1.5. DO and BOD curves for a stream [1].

Ordinate of the DO sag curve is in terms of mg/liter of dissolved oxygen. The shape of the DO sag curve, downstream of the point of sewage discharge, is understood from examination of Fig. 1.6. The DO sag curve is the net resultant of two curves: one corresponding to depletion of dissolved oxygen due to its utilization for oxidation of organic materials from the sewage discharge, and the other corresponding to oxygen gain by natural reaeration. Figure 1.5 shows that the DO sag curve reaches a low point about 27 miles downstream of the point of sewage discharge, corresponding to 2¼ days of flow and a DO of about 1.5 mg/liter.

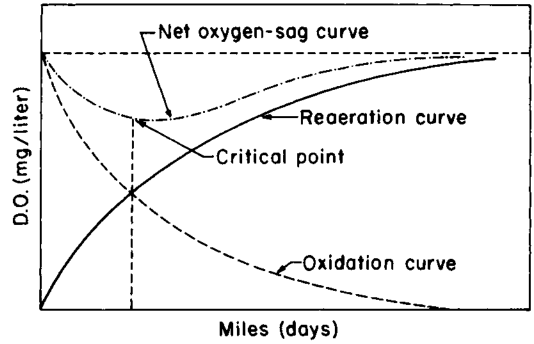

Fig. 1.6. Oxygen sag curve.

This process of deoxygenation would reduce the DO to zero in about 1½ days flow, if there were no factors in operation that could restore oxygen to water. The river reach where DO would be completely gone would occur about 18 miles downstream from the discharge of sewage. After reaching its minimum, DO level rises again toward a restoration, eventually reaching a value nearly equal to that for the upstream unpolluted water, i.e., a DO of approximately 7 mg/liter.

If population of the city remains fairly constant throughout the year, and flow rate is relatively constant, the low point of the DO sag curve moves up or down the stream with fluctuations in temperature. During the winter the rate of oxidation is lower and gain of oxygen by reaeration is greater, as solubility of oxygen in water increases at lower temperatures. These two factors combined cause the low point of the oxygen sag curve to move farther downstream. During the summer, on the other hand, the rate of oxidation is higher and gain of oxygen by reaeration is less pronounced. These two factors combined cause the low point of the oxygen sag curve to move upstream.

The reach of any stream where the DO sag curve attains its low point represents the stream environment poorest in DO resources. Living specimens that need a high DO, such as cold water fish, suffocate and move to other stream areas where the DO resources are greater.

The other curve shown in Fig. 1.5 corresponds to the *biochemical oxygen demand* (BOD). This important parameter is discussed in Chapter 2, Section 2.3. The biochemical oxygen demand is used as a measure of the quantity of oxygen required for oxidation by aerobic biochemical action of the degradable organic matter present in a sample of water. The BOD is low in the upstream unpolluted water (about 2 mg/liter), since there is not much organic matter present to consume oxygen. Then BOD increases abruptly at point zero (sewage discharge), and gradually decreases downstream from this point, as organic matter discharged is progressively oxidized, until reaching eventually a value of approximately 2 mg/liter, indicative of unpolluted water. At this point the raw sewage is stabilized. As indicated in Fig. 1.5, stabilization is achieved at approximately 100 miles downstream from the sewage discharge. BOD and DO are so interrelated that dissolved oxygen concentration is low where BOD is high, and the converse also is true.

Four distinct zones are shown in Fig. 1.5 underneath the DO curve: (1) clean water zone; (2) zone of degradation; (3) zone of active decomposition; and (4) zone of recovery.

6.2. EFFECT OF LIGHT

In Fig. 1.6 the effects of oxygen depletion by oxidation of organic materials and oxygen gain by reaeration are the only ones considered in explaining the shape of the oxygen sag curve. For a more complete analysis of the problem one needs, in addition, to consider the effect of light.

At any selected point in the stream, there is a variation in concentration of dissolved oxygen depending on the time of day. During daylight hours, algae and other plants give off oxygen into the water through the process of photosynthesis. This amount of oxygen may be so considerable that the water usually becomes supersaturated at some time during daylight hours. In addition to giving off oxygen, the process of photosynthesis results in the manufacture of sugar to serve as the basis of support for all stream life. This corresponds to the chemical reaction shown in Eq. (1.1).

$$6CO_2 + 6H_2O \rightarrow C_6H_{12}O_6 + 6O_2 \tag{1.1}$$

While photosynthesis occurs, so does respiration, which continues for 24 hr a day, irrespective of illumination. During respiration O_2 is taken in and CO_2 is given off. During daylight, algae may yield oxygen in excess of that needed for respiration, as well as in excess of that required for respiration by other aquatic life, and for satisfaction of any biochemical oxygen demand. This could be true in the recovery zone particularly. Under these conditions, supersaturation of oxygen may occur, and surplus oxygen may be lost to the atmosphere.

During the night, photosynthesis does not occur and the surplus DO is gradually used up by respiration of all forms of aquatic life, as well as for the satisfaction of biochemical oxygen demand. Therefore, concentration of dissolved oxygen is at its minimum during early morning hours. To take into account such DO variations, sampling of streams for sanitary surveys is conducted over a 24-hr period.

6.3. DECOMPOSITION OF CARBONACEOUS AND NITROGENOUS ORGANIC MATTER

Accelerated bacterial growth is a response to rich food supplies in the domestic sewage. During rapid utilization of food, bacterial reproduction is at an optimum, and utilization of DO becomes fairly proportional to the rate of food utilization. Figure 1.7 illustrates the progressive downstream changes of organic nitrogen to ammonia, nitrite, and finally nitrate. A high initial consumption of oxygen by bacterial feeding on proteinaceous compounds

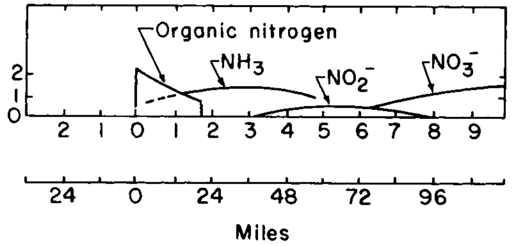

Fig. 1.7. Aerobic decomposition of nitrogenous organic matter [1].

available in upstream waters takes place due to the freshly discharged domestic sewage. With fewer and fewer of these compounds left in downstream waters, the DO concentration is progressively recovered, reaching eventually its initial value of approximately 7 mg/liter.

A similar process takes place with fat and carbohydrate foodstuffs. The final products of aerobic and anaerobic decomposition of nitrogenous and carbonaceous matter are

1. Decomposition of nitrogenous organic matter
   Aerobic (final products): NO_3^-, CO_2, H_2O, SO_4^{2-}
   Anaerobic (final products): mercaptans, indole, skatole, H_2S, plus miscellaneous products
2. Decomposition of carbonaceous matter
   Aerobic: CO_2, H_2O
   Anaerobic: acids, alcohols, CO_2, H_2, CH_4, plus miscellaneous products

Nitrogen and phosphorus in sewage proteins cause special problems in some receiving waters. High concentrations of these elements in water create conditions especially favorable for growing green plants. If the water is free flowing (rivers, brooks), green velvety coatings grow on the stones and possibly lengthy streamers, popularly known as mermaid's tresses, wave in the current. These growths are not unattractive and also constitute a miniature jungle in which animal life of many kinds prey on each other, with the survivors growing to become eventual fish food. If, however, the water is quiet (e.g., lakes), growth of very undesirable types of algae is stimulated. These algae make the water pea green, smelly, and unattractive. This phenomenon is discussed in Section 7 of this chapter. Sometimes, these blue-green algae develop poisons capable of killing livestock, wildlife, and fish.

6.4. SLUDGE DEPOSITS AND AQUATIC PLANTS

A profile showing sludge depth vs. distance from the outfall of the sewage is shown in the bottom part of Fig. 1.8. Maximum depth occurs near the outfall, and then the sludge is gradually reduced by decomposition through the action of bacteria and other organisms, until it becomes insignificant about 30 miles below the municipality.

Also at the outfall there is great turbidity due to the presence of fine suspended solids. As these solids settle, the water becomes clear and approaches the transparency of upstream water, above the point of sewage discharge.

Distribution of aquatic plants is indicated in the upper part of Fig. 1.8. Shortly after the discharge, molds attain maximum growth. These molds and filamentous bacteria (*Sphaerotilus*) are associated with the sludge deposition

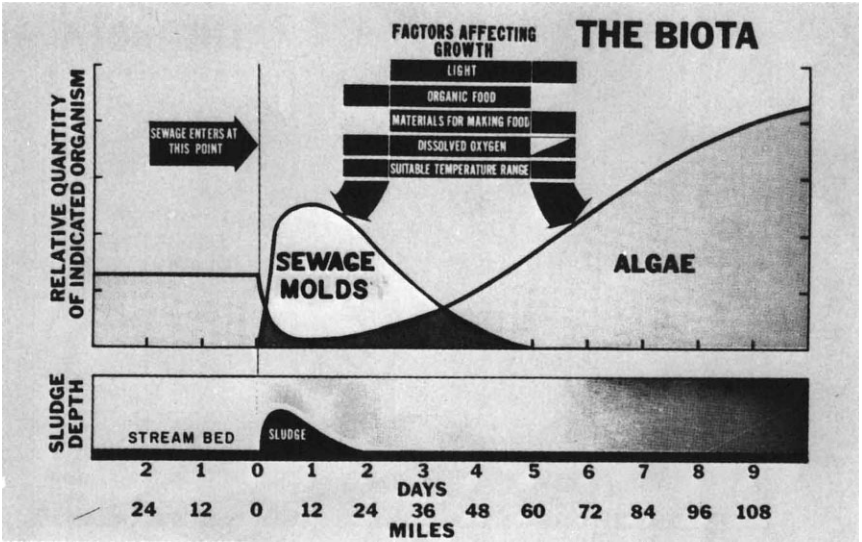

Fig. 1.8. Sludge deposits and aquatic plants [1].

shown in the lower curve. From mile 0 to mile 36, high turbidity is not conducive to production of algae, since they need sunlight in order to grow and light cannot penetrate the water effectively. The only type of algae that may grow are blue-green algae, characteristic of polluted waters. They may cover marginal rocks in slippery layers and give off foul odors upon seasonal decomposition.

Algae begin to increase in number at about mile 36. Plankton or free-floating forms become steadily more abundant. They constitute an excellent food supply for aquatic animals and also provide shelter for them. Thus, as plants respond downstream in developing a diversified population in the recovery and clean water zones, animals follow a parallel development, producing a great variety of species.

6.5. BACTERIA AND CILIATES

Figure 1.9 illustrates the interrelation between bacteria and other forms of animal plankton such as ciliated protozoans, rotifers, and crustaceans. Two die-off curves are shown, one for total sewage bacteria and the other for coliform bacteria only. The two bell-shaped curves pertain to ciliated protozoans and rotifers and crustaceans.

After entering the stream with the sewage, bacteria reproduce and become abundant, feeding on the organic matter of sewage. Ciliated protozoans, initially few in number, prey on the bacteria. Bacteria population decreases gradually, both by a natural process of "die-off," and from the predatory

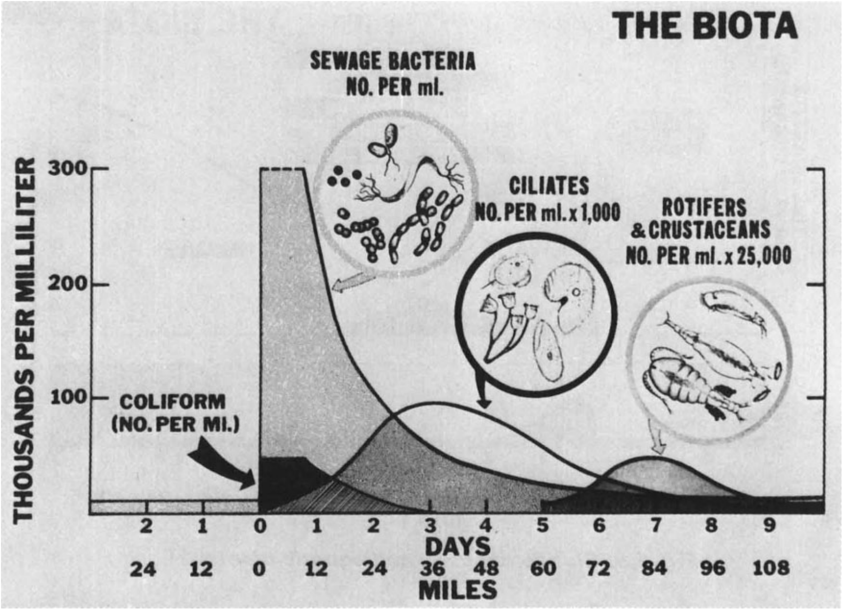

Fig. 1.9. Bacteria thrive and finally become prey of the ciliates, which, in turn, are food for the rotifers and crustaceans [1].

feeding by protozoans. After about 2 days flow, approximately 24 miles downstream of point zero, the environment becomes more suitable for ciliates, which form the dominant group of animal plankton. After about 7 days, 84 miles downstream of point zero, ciliates fall victim to rotifers and crustaceans, which become the dominant species. Thus, this sewage-consuming biological process depends on a closely interrelated succession of species of animal plankton, one kind of organism capturing and eating another.

This relationship between bacteria eaters and their prey is found in the operation of a modern sewage treatment plant. In fact, the stream can be thought of as a natural sewage treatment plant.

Stabilization of sewage in a plant is more rapid when ferocious bacteria-eating ciliates are present to keep the bacteria population at a low but rapidly growing state. In some sewage treatment plants, microscopic examination is made routinely to observe the battle lines between bacteria eaters and their prey.

6.6. HIGHER FORMS OF ANIMAL SPECIES

Figure 1.10 illustrates these types of organisms and their population along the course of the stream. Curve (a) represents the variety, i.e., the numbers of species of organisms found under varying degrees of pollution. Curve (b)

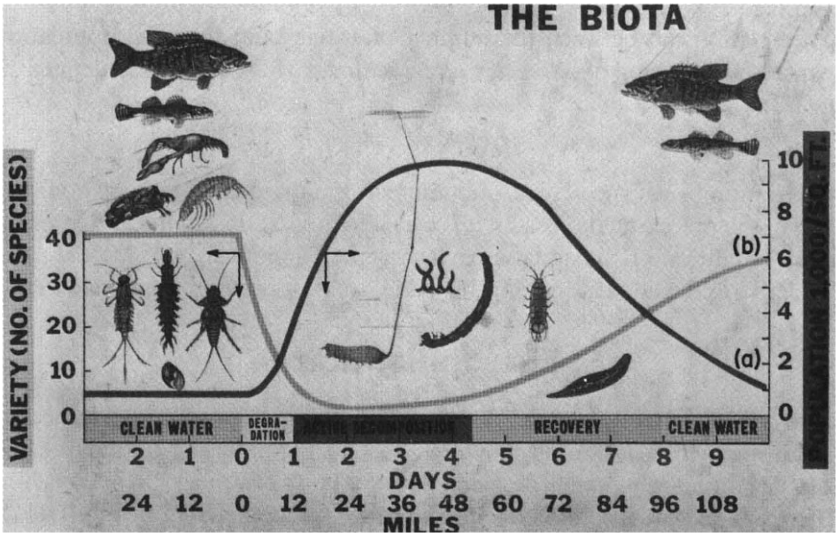

Fig. 1.10. Curve (a) shows the fluctuations in numbers of species; (b) the variations in numbers of each species [1].

represents the population in thousands of individuals of each species per square foot.

In the clean water, upstream of point zero, a great variety of organisms is found with very few of each kind present. At the point of sewage discharge, the number of different species is greatly reduced and there is a drastic change in the species makeup of the biota. This changed biota is represented by a few species, but there is a tremendous increase in the numbers of individuals of each kind as compared with the density of population upstream.

In clean water upstream there is an association of sports fish, various minnows, caddis worms, mayflies, stoneflies, hellgrammites and gill-breathing snails, each kind represented by a few individuals. In badly polluted zones this biota is replaced by an association of rattailed maggots, sludge worms, bloodworms, and a few other species, represented by a great number of individuals. When downstream conditions again resemble those of the upstream clean water zone, the clean water animal association tends to reappear and the pollution-tolerant group of animals become suppressed.

Pollution-tolerant animals are especially well adapted to life in thick sludge deposits and to conditions of low dissolved oxygen. The rattailed maggot, for example, possesses a "snorklelike" telescopic air tube which is pushed through the surface film to breathe atmospheric oxygen. Thus, even in total absence of dissolved oxygen it survives. These types of animals are found commonly around sewage treatment plants near the supernatant sludge beds.

The relationship between the number of species and the total population is expressed in terms of a *species diversity index* (SDI), which is defined in Eq. (1.2).

$$\mathrm{SDI} = (S-1)/\log I \tag{1.2}$$

where S, number of species; I, total number of individual organisms counted.

From the preceding discussion it is clear that the SDI is an indication of the overall condition of the aquatic environment. The higher its value the more productive is the aquatic system. Its value decreases as pollution increases.

7. Eutrophication [4]

Eutrophication is the natural process of lake aging. It progresses irrespective of man's activities. Pollution, however, hastens the natural rate of aging and shortens considerably the life expectancy of a body of water.

The general sequence of lake eutrophication is summarized in Fig. 1.11. It consists of the gradual progression ("ecological succession") of one life stage to another, based on changes in the degree of nourishment or productivity. The youngest stage of the life cycle is characterized by low concentration of plant nutrients and little biological productivity. Such lakes are called oligotropic lakes (from the Greek *oligo* meaning "few" and *trophein* meaning "to nourish," thus oligotropic means few nutrients). At a later stage in the succession, the lake becomes mesotrophic (*meso* = intermediate); and as the life cycle continues the lake becomes eutrophic (*eu* = well) or highly productive. The final life stage before extinction is a pond, marsh, or swamp.

Enrichment and sedimentation are the principal contributors to the aging process. Shore vegetation and higher aquatic plants utilize part of the inflowing nutrients, grow abundantly, and, in turn, trap the sediments. The lake gradually fills in, becoming shallower by accumulation of plants and sediments on the bottom, and smaller by the invasion of shore vegetation, and eventually becoming dry land. The extinction of a lake is, therefore, a result of enrichment, productivity, decay, and sedimentation. The effect of nitrogen- and phosphorus-rich wastewater discharges on accelerating eutrophication has been discussed in Section 6 of this chapter.

8. Types of Water Supply and Classification of Water Contaminants

According to their origin, water supplies are classified into three categories: (1) surface waters, (2) ground waters, and (3) meteorological waters. Surface waters comprise stream waters (e.g., rivers), oceans, lakes, and impoundment

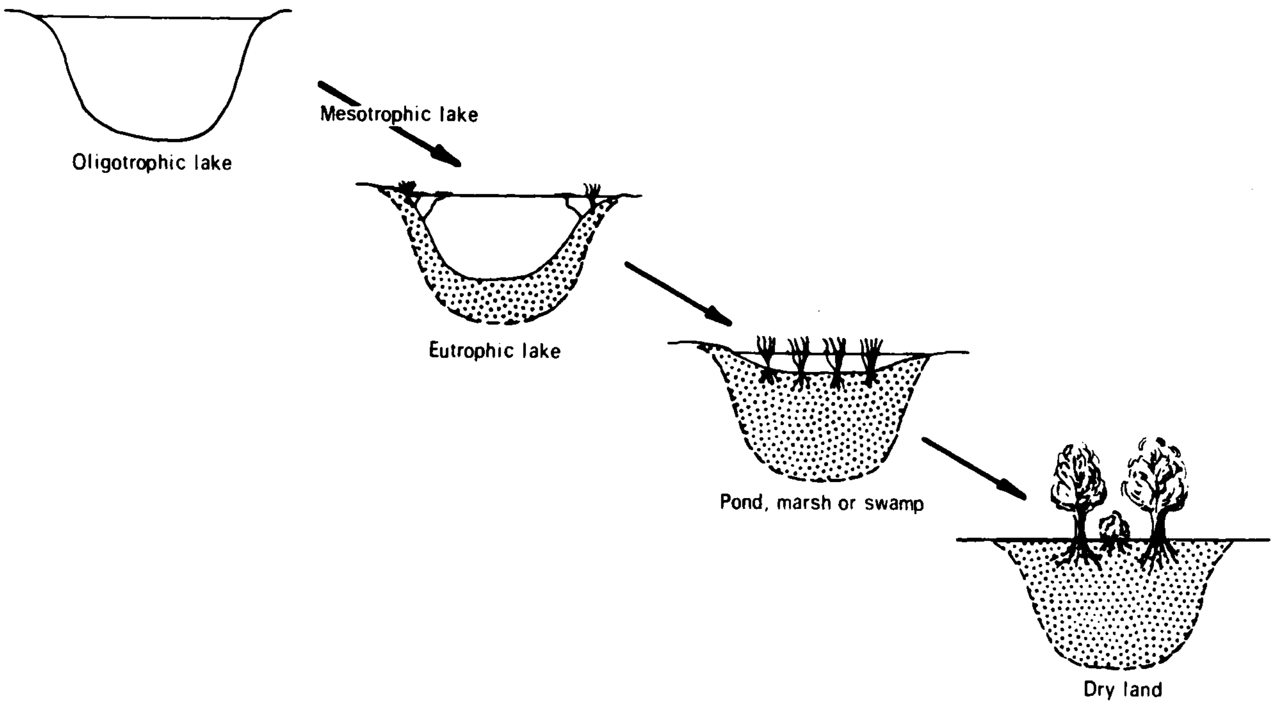

Fig. 1.11. Eutrophication—the process of aging by ecological succession [4]. (*Reprinted with permission. Copyright by American Water Resources Association.*)

waters. Stream waters subject to contamination exhibit a variable quality along the course of the stream, as discussed in Section 6. Waters in lakes and impoundments, on the other hand, are of a relatively uniform quality. Ground waters show, in general, less turbidity than surface waters. Meteorological waters (rain) are of greater chemical and physical purity than either surface or ground waters.

Water contaminants are classified into three categories: (1) chemical, (2) physical, and (3) biological contaminants. Chemical contaminants comprise both organic and inorganic chemicals. The main concern resulting from pollution by organic compounds is oxygen depletion resulting from utilization of DO in the process of biological degradation of these compounds. As discussed in Section 6, this depletion of DO leads to undesirable disturbances of the environment and the biota. In the case of pollution resulting from the presence of inorganic compounds the main concern is their possible toxic effect, rather than oxygen depletion. There are, however, cases in which inorganic compounds exert an oxygen demand, so contributing to oxygen depletion. Sulfites and nitrites, for example, take up oxygen, being oxidized to sulfates and nitrates, respectively [Eqs. (1.3) and (1.4)].

$$SO_3^{2-} + \tfrac{1}{2}O_2 \rightarrow SO_4^{2-} \tag{1.3}$$

$$NO_2^{-} + \tfrac{1}{2}O_2 \rightarrow NO_3^{-} \tag{1.4}$$

Heavy metal ions which are toxic to humans are important contaminants. They occur in industrial wastewaters from plating plants and paint and pigment industries. These include Hg^{2+}, As^{3+}, Cu^{2+}, Zn^{2+}, Ni^{2+}, Cr^{3+}, Pb^{2+}, and Cd^{2+}. Even their presence in trace quantities (i.e., minimum detectable concentrations) causes serious problems.

Considerable press coverage has been given to contamination of water by mercury. Microorganisms convert the mercury ion to methylmercury (CH_3Hg) or dimethylmercury [$(CH_3)_2Hg$]. The dimethyl compound, being volatile, is eventually lost to the atmosphere. Methylmercury, however, is absorbed by fish tissue and might render it unsuitable for human consumption. Mercury content in fish tissue is tolerable up to a maximum of 15–20 ppm. Methylmercury present in fish is absorbed by human tissues and eventually concentrates in certain vital organs such as the brain and the liver. In the case of pregnant women it concentrates in the fetus. Recently in Japan, there were several reported cases of deaths from mercury poisoning, due to human consumption of mercury-contaminated fish. Analysis of fish tissue revealed mercury concentrations of approximately 110–130 ppm. These high mercury concentrations, coupled with the large fish intake in the typical Japanese diet, caused this tragedy.

Contamination by nitrates is also dangerous. Fluorides, on the other hand, seem actually beneficial, their presence in potable waters being responsible for appreciable reduction in the extent of tooth decay. There is, however, considerable controversy concerning fluoridization of potable water.

Some physical contaminants include (1) temperature change (thermal pollution). This is the case of relatively warm water discharged by industrial plants after use in heat exchangers (coolers); (2) color (e.g., cooking liquors discharged by chemical pulping plants); (3) turbidity (caused by discharges containing suspended solids); (4) foams [detergents such as alkylbenzene sulfonate (ABS) constitute important cause of foaming]; and (5) radioactivity.

Biological contaminants are responsible for transmission of diseases by water supplies. Some of the diseases transmitted by biological contamination of water are cholera, typhoid, paratyphoid, and shistosomiasis.

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2

Characterization of Domestic and Industrial Wastewaters

1. Measurement of Concentration of Contaminants in Wastewaters

Contaminants in wastewaters are usually a complex mixture of organic and inorganic compounds. It is usually impractical, if not nearly impossible, to obtain complete chemical analysis of most wastewaters.

For this reason, a number of empirical methods for evaluation of concentration of contaminants in wastewaters have been devised, the application of which does not require knowledge of the chemical composition of the specific wastewater under consideration. The most important standard methods for analysis of organic contaminants are described in Sections 2 and 3. For discussion of analytical methods for specific inorganic contaminants in wastewaters, determination of physical parameters (total solids, color, odor), and bioassay tests (coliforms, toxicity tests), the reader is referred to Ref. [13].

Special attention is given in this chapter to the biochemical oxygen demand of wastewaters (BOD). A mathematical model for typical BOD curves is discussed, as well as the evaluation of feasibility of biological treatment for an industrial wastewater (Sections 4–9). Average characteristics of municipal sewage and the procedure followed in industrial wastewater surveys are described in Sections 10 and 11. Since both flow rate and sewage strength may follow an aleatory pattern of variation, it may be desirable to perform a statistical correlation of such data. This subject is discussed in Section 12.

Analytical methods for organic contaminants are classified into two groups:

Group 1. Oxygen parameter methods

1. Theoretical oxygen demand (ThOD)
2. Chemical oxygen demand (COD) [standard dichromate oxidation method; permanganate oxidation test; rapid COD tests; instrumental COD methods ("AquaRator")]
3. Biochemical oxygen demand (BOD) (dilution methods; manometric methods)
4. Total oxygen demand (TOD)

Group 2. Carbon parameter methods

1. Theoretical organic carbon (ThOC)
2. Total organic carbon (TOC) (wet oxidation method; carbon analyzer determinations)

2. Measurement of Organic Content: Group 1—Oxygen Parameter Methods

2.1. THEORETICAL OXYGEN DEMAND (ThOD)

Theoretical oxygen demand (ThOD) corresponds to the stoichiometric amount of oxygen required to oxidize completely a given compound. Usually expressed in milligrams of oxygen required per liter of solution, it is a calculated value and can only be evaluated if a complete chemical analysis of the wastewater is available, which is very rarely the case. Therefore, its utilization is very limited.

To illustrate the calculation of ThOD, consider the simple case of an aqueous solution of a pure substance: a solution of 1000 mg/liter of lactose. Equation (2.1)* corresponds to the complete oxidation of lactose.

$$\begin{array}{lcc} & (CH_2O) + O_2 \rightarrow CO_2 + H_2O & (2.1) \\ \text{Molecular weight:} \quad 30 \qquad\qquad 32 & & \end{array}$$

ThOD value is readily obtained from a stoichiometric calculation, based on Eq. (2.1):

$$\frac{30\ (\text{wt. lactose})}{1000} = \frac{32\ (\text{wt. } O_2)}{\text{ThOD}}$$

$$\therefore \text{ThOD} = (32/30)1000 = 1067 \text{ mg/liter}$$

2.2. CHEMICAL OXYGEN DEMAND (COD)

Chemical oxygen demand (COD) corresponds to the amount of oxygen required to oxidize the organic fraction of a sample which is susceptible to permanganate or dichromate oxidation in an acid solution. Since oxidation performed in a COD laboratory test does not necessarily correspond to the stoichiometric Eq. (2.1), COD value is not expected to equal ThOD.

Standard COD tests (Sections 2.2.1 and 2.2.2) yield values which vary

TABLE 2.1
Average Values of Oxygen Parameters for Wastewaters as a Fraction of the Theoretical Oxygen Demand (Taken as 100)[a]

ThOD	100
TOD	92
COD (standard method)	83
COD (rapid tests)	70
BOD_{20}	
With nitrification	65
Nitrification suppressed	55
BOD_5	
With nitrification	58
Nitrification suppressed	52

[a] For carbon parameters the TOC represents an average of about 95% of the theoretical organic carbon (ThOC). Relationships between ThOD and ThOC are discussed in Section 3.

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* For simplicity in Eq. (2.1), lactose was represented by one sugar unit (CH_2O). Multiplying this unit by a factor of 12 one obtains $C_{12}H_{24}O_{12}$, which is the molecular formula for lactose.

from 80–85% of the ThOD, depending on the chemical composition of the wastewater being tested. Rapid COD tests, discussed in Section 2.2.3, yield values equal to approximately 70% of ThOD value.

Approximate relationships between the various oxygen and carbon parameters are presented in Table 2.1, as estimated from a graph in Eckenfelder and Ford [4]. Values indicated in Table 2.1 are typical average values; correct relationships should be determined for the wastewater in question, as they are dependent upon its chemical composition. Thus, values in Table 2.1 are only utilized for rough estimates in the absence of actual data. Four types of COD tests are described next.

2.2.1. Standard Dichromate Oxidation Method [5, 8, 13]

The standard dichromate COD test is widely used for estimating the concentration of organic matter in wastewaters. The test is performed by heating under total reflux conditions a measured sample with a known excess of potassium dichromate ($K_2Cr_2O_7$), in the presence of sulfuric acid (H_2SO_4), for a 2-hr period. Organic matter in the sample is oxidized and, as a result, yellow dichromate is consumed and replaced by green chromic [Eq. (2.2)]. Silver sulfate (Ag_2SO_4) is added as catalyst.

$$Cr_2O_7^{2-} + 14H^+ + 6e \rightleftharpoons 2Cr^{3+} + 7H_2O \qquad (2.2)$$

Measurement is performed by titrating the remaining dichromate or by determining colorimetrically the green chromic produced. The titration method is more accurate, but more tedious. The colorimetric method, when performed with a good photoelectric colorimeter or spectrophotometer, is more rapid, easier, and sufficiently accurate for all practical purposes.

If chlorides are present in the wastewater, they interfere with the COD test since chlorides are oxidized by dichromate according to Eq. (2.3).

$$6Cl^- + Cr_2O_7^{2-} + 14H^+ \rightarrow 3Cl_2 + 2Cr^{3+} + 7H_2O \qquad (2.3)$$

This interference is prevented by addition of mercuric sulfate ($HgSO_4$) to the mixture, as Hg^{2+} combines with Cl^- to form mercuric chloride ($HgCl_2$), which is essentially nonionized. A 10:1 ratio of $HgSO_4$:Cl^- is recommended. This corresponds to the following chemical reaction [Eq. (2.4)].

$$Hg^{2+} + 2Cl^- \rightarrow HgCl_2\downarrow \qquad (2.4)$$

The presence of the Ag_2SO_4 catalyst is required for oxidation of straight-chain alcohols and acids. If insufficient quantity of $HgSO_4$ is added, the excess Cl^- precipitates the Ag_2SO_4 catalyst, thus leading to erroneously low values for the COD test. This corresponds to the following chemical reaction [Eq. (2.5)].

$$Ag^+ + Cl^- \rightarrow AgCl\downarrow \qquad (2.5)$$

Standard ferrous ammonium sulfate [$Fe(NH_4)_2 \cdot (SO_4)_2 \cdot 6H_2O$] is used for the titration method. Ordinarily, standard ferrous sulfate loses strength with age, due to air oxidation. Daily standardization and mathematical correction in the calculation of COD to account for this deterioration are recommended [13]. Cadmium addition to the stock bottle of ferrous sulfate completely prevents deterioration. Ferrous sulfate available from Hach Chemical Company for the COD test is preserved in this manner, so that no further standardization checks are required.

The recommended procedure is to cool the sample after the 2-hr digestion with $K_2Cr_2O_7$, add five drops of ferroin indicator, and titrate with the standard ferrous ammonium sulfate solution until a red-brown color is obtained. The end point is very sharp. Ferroin indicator solution may be purchased already prepared (it is an aqueous solution of 1,10-phenanthroline monohydrate and $FeSO_4 \cdot 7H_2O$). The red-brown color corresponding to the end point is due to formation of a complex of ferrous ion with phenanthroline. Equation (2.6) corresponds to oxidation of ferrous ammonium sulfate by dichromate.

$$Cr_2O_7^{2-} + 14H^+ + 6Fe^{2+} \rightleftharpoons 2Cr^{3+} + 6Fe^{3+} + 7H_2O \tag{2.6}$$

Equation (2.7) corresponds to formation of the ferrous–phenanthroline complex, which takes place as soon as all dichromate is reduced to Cr^{3+}, and therefore further addition of ferrous ammonium sulfate results in an excess of Fe^{2+} (ferrous ion).

$$\underset{\substack{\text{phenanthroline–ferric}\\ \text{(pale blue)}}}{Fe(C_{12}H_8N_2)_3^{3+}} + e \rightleftharpoons \underset{\substack{\text{phenanthroline–ferrous}\\ \text{(red-brown)}}}{Fe(C_{12}H_8N_2)_3^{2+}} \tag{2.7}$$

Details concerning preparation and standardization of reagents and calculation procedure are given in Refs. [5], [8], and [13]. Reproducibility of the COD test is affected by the reflux time. COD value obtained increases with reflux time up to about 7 hr and then remains essentially constant [4]. Instead of refluxing for 7 hr or more, a practical reflux time of 2 hr is recommended in the standard procedure.

2.2.2. Permanganate Oxidation Test

Recommended as the standard method until 1965, this test has been replaced by the dichromate test just described. This test utilizes potassium permanganate ($KMnO_4$) instead of dichromate as the oxidizing agent.

The wastewater sample is boiled with a measured excess of permanganate in acid solution (H_2SO_4) for 30 min. The pink solution is cooled and a known excess of ammonium oxalate [$(NH_4)_2C_2O_4$] is added, the solution becoming colorless. Excess oxalate is then titrated with $KMnO_4$ solution until the pink color returns. Oxalate used is calculated by difference, and permanganate utilized is calculated from simple stoichiometry. Equation (2.8) corresponds to oxidation of the oxalate.

$$5C_2O_4^{2-} + 2MnO_4^- + 16H^+ \rightarrow 10CO_2 + 2Mn^{2+} + 8H_2O \qquad (2.8)$$

2.2.3. Rapid COD Tests

Several rapid COD tests have been proposed involving digestion with dichromate for periods of time shorter than the 2 hr prescribed in the standard test. In one of these techniques, the wastewater is digested with the $K_2Cr_2O_7$–H_2SO_4–$AgSO_4$ solution at 165°C for 15 min. The solution is diluted with distilled water and titrated with ferrous ammonium sulfate, as in the standard method.

In this test, COD yield for domestic sludge corresponds to approximately 65% of the value obtained by the standard method. For other wastewaters, COD yield ratio between the rapid and the standard test varies depending on the nature of the wastewater.

2.2.4. Instrumental COD Methods [11, 14, 15]

Instrumental COD methods are very fast and yield reproducible results. In this section, the Precision AquaRator developed by the Dow Chemical Company and licensed to the Precision Scientific Company is described. The COD measurement requires only about 2 min and data are reproducible to within ±3% or better. Results correlate well with those of the standard COD method and are much more consistent than BOD tests, which typically vary by ±15%.

The AquaRator is designed to measure oxygen demand in the range of 10–300 mg/liter. Samples of higher concentration are handled by preliminary dilution of the sample. A flow diagram of the Precision AquaRator is shown in Fig. 2.1.

A 20-μl sample (20×10^{-6} liter ≈ 0.02 cm^3), homogenized if necessary, is injected by a syringe into the Precision AquaRator. (See sample injection port, SIP.) The sample is swept through a platinum catalytic combustion furnace (SF) by a stream of dry CO_2, which oxidizes the contaminants to CO and H_2O. Water is stripped out in a drying tube (DT), and reaction products are then passed through a second platinum catalytic treatment. The CO concentration is measured by an integral nondispersive infrared analyzer (IA), sensitized for carbon monoxide. The resultant reading is directly converted to COD by use of a calibration chart.

Carbon dioxide flow is set at approximately 130 cm^3/min by the flow control system. Any trace of oxygen present in the feed gas is reduced by a "purifying" carbon furnace (PCF), yielding a background gas stream of CO

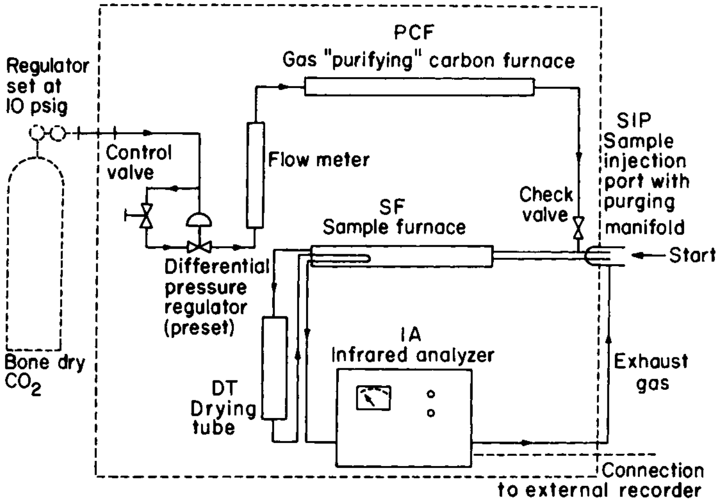

Fig. 2.1. ***Flow diagram of Precision AquaRator*** [11]. (*Courtesy of Precision Scientific Company.*)

and CO_2 which is indicated as a normal baseline of the recorder. The sample is injected into the sample furnace (SF), where contaminants and CO_2 react to form a typical mixture of CO, CO_2, and H_2O. The infrared analyzer (IA) determines the increase of CO content in the gas stream, which is directly related to COD of the sample. Exhaust gas is then discharged through a sample inlet purging manifold.

The AquaRator theory is discussed in Stenger and Van Hall [14, 15]. Equations (2.9) and (2.10) indicate the types of reactions that take place when organic material is combusted in atmospheres of oxygen and carbon dioxide, respectively.

$$C_aH_bN_cO_d + (n/2)O_2 \rightarrow aCO_2 + (b/2)H_2O + (c/2)N_2 \qquad (2.9)$$

$$C_aH_bN_cO_d + mCO_2 \rightarrow (m+a)CO + (b/2)H_2O + (c/2)N_2 \qquad (2.10)$$

If oxygen required in Eq. (2.9) could be determined exactly, it would represent the ThOD of the sample. Ideally, the dichromate COD determination approaches this value, but some compounds are difficult to oxidize by the dichromate treatment. Oxidation which takes place in the AquaRator is more vigorous than dichromate oxidation, and thus results represent a more realistic level of oxygen demand of the contaminants present.

The originators of the method used in the AquaRator [14, 15] demonstrated that $(m+a)$ in Eq. (2.10) is equal to n in Eq. (2.9); that is, the number of moles of carbon monoxide produced is the same as the number of oxygen atoms required. Therefore, instrument readings of carbon monoxide formed are directly related to chemical oxygen demand. Calibration is carried out by injecting standard solutions of sodium acetate trihydrate, for which oxygen demand in milligrams per liter can be calculated. A graph of oxygen demand vs. recorder output (chart divisions) is all that is required for determining the unknown contaminant demand.

2.3. BIOCHEMICAL OXYGEN DEMAND (BOD)

Biochemical oxygen demand is used as a measure of the quantity of oxygen required for oxidation of biodegradable organic matter present in the water sample by aerobic biochemical action. Oxygen demand of wastewaters is exerted by three classes of materials: (1) carbonaceous organic materials usable as a source of food by aerobic organisms; (2) oxidizable nitrogen derived from nitrite, ammonia, and organic nitrogen compounds which serve as food for specific bacteria (e.g., *Nitrosomonas* and *Nitrobacter*). This type of oxidation (nitrification) is discussed in Section 8; and (3) chemical reducing compounds, e.g., ferrous ion (Fe^{2+}), sulfites (SO_3^{2-}), and sulfide (S^{2-}), which are oxidized by dissolved oxygen.

For domestic sewage, nearly all oxygen demand is due to carbonaceous organic materials and is determined by BOD tests described in Sections 2.3.1 and 2.3.2. For effluents subjected to biological treatment, a considerable part of the oxygen demand may be due to nitrification (Section 8 of this chapter).

2.3.1. BOD Dilution Test

Detailed description of the dilution test as well as preparation of reagents is given in Ref. [13]. Procedure is given below.

1. Prepare several dilutions of the sample to be analyzed with distilled water of high purity. Recommended dilutions depend on estimated concentration of contaminants responsible for oxygen demand. For highly contaminated waters, dilution ratios (ml of diluted sample/ml of original sample) may be of 100:1. For river waters, the sample may be taken without dilution for low pollution streams, and in other cases dilution ratios of 4:1 may be utilized.

2. Incubation bottles (250- to 300-ml capacity), with ground-glass stoppers are utilized. In the BOD bottle one places (a) the diluted sample (i.e., the "substrate"), (b) a *seed* of microorganisms (usually the supernatant liquor from domestic sewage), and (c) nutrient solution for the microorganisms. This solution contains sodium and potassium phosphates and ammonium chloride (nitrogen and phosphorus are elements needed as nutrients for microorganisms).

The pH of the solution in the BOD bottle should be about 7.0 (neutral).

Phosphate solution utilized is a buffer. For samples containing caustic alkalinity or acidity, neutralization to about pH 7 is made with dilute H_2SO_4 or NaOH prior to the BOD test.

For each BOD bottle a control bottle, which does not contain the substrate, is also prepared.

3. Bottles are incubated at 20°C. Each succeeding 24-hr period, a sample bottle and a corresponding control bottle are taken from the incubator, and dissolved oxygen in both is determined as described at the end of this section. The difference between concentrations of dissolved oxygen (mg/liter) in control bottle and in sample bottle corresponds to the oxygen utilized in biochemical oxidation of contaminants [Eq. (2.11)].

$$y\,(\text{mg/liter}) = \text{DO (control bottle)} - \text{DO (sample bottle)} \qquad (2.11)$$

Values of y (BOD, mg/liter) are plotted vs. incubation time t (days). A typical BOD curve for oxidation of carbonaceous materials is shown in Fig. 2.2. Curves for cases where nitrification takes place are discussed in Section 8.

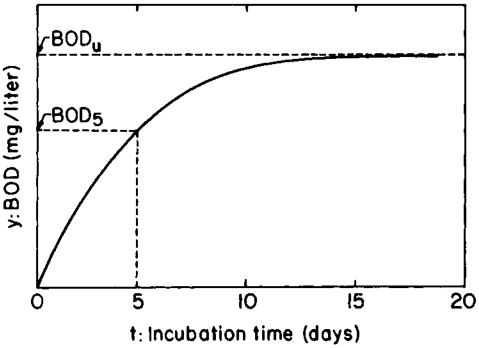

Fig. 2.2. Typical BOD curve for oxidation of carbonaceous materials.

Oxygen utilization in the BOD test is very slow. A typical curve (Fig. 2.2) only reaches the limiting BOD in about 20 days or more. This value is called *ultimate BOD*, denoted as BOD_u.

It is impractical to monitor continuously a process stream in terms of BOD because of the time factor involved in the test. In practice, BOD is reported in terms of 5-day BOD, denoted as BOD_5 (Fig. 2.2). Even 5 days is too long a period of time to wait for the result of a test.

It is important to notice that the value of BOD_u is not equal to ThOD, because in the BOD bottle not all substrate is oxidized. Ratios of values of BOD_u (or BOD_5) to ThOD depend on the chemical composition of the wastewater. Average values are given in Table 2.1.

The ratio of BOD_5 to BOD_u also varies according to the substrate. For domestic sewage, this ratio is approximately 0.77 [Eq. (2.12)].

$$BOD_5/BOD_u = 0.77 \tag{2.12}$$

Considerable experience is required to obtain reliable results in the BOD dilution test. In general, reproducibility of results is not better than ±15%. Some of the difficulties involved in the BOD dilution test are discussed in the next sections. Because of these fluctuations it is recommended that several BOD bottles be taken from the incubator every 24 hr and that statistical averaging of results be performed.

a. Ratio of COD and BOD_u

It has just been stated that values of BOD_u and ThOD are not equal. Similarly, the value of BOD_u is generally lower than that for COD obtained by the standard dichromate oxidation method, as indicated in Table 2.1. The reasons are that (1) many organic compounds which are oxidized by $K_2Cr_2O_7$ are *not* biochemically oxidizable and (2) certain inorganic ions such as sulfides (S^{2-}), thiosulfates ($S_2O_3^{2-}$), sulfites (SO_3^{2-}), nitrites (NO_2^-), and ferrous ion (Fe^{2+}) are oxidized by $K_2Cr_2O_7$, thus accounting for inorganic COD, which is not detected by the BOD test.

b. Effect of Seeding and Acclimation of Seed on the BOD Test

One of the most frequent reasons for unreliable BOD values is utilization of an insufficient amount of microorganism seed. Another serious problem for industrial wastes is acclimation of seed. For many industrial wastes, the presence of toxic materials interferes with growth of the microorganism population. BOD curves obtained exhibit a time lag period (Fig. 2.3).

Low BOD values are obtained if adequate corrective action is not taken. It becomes necessary to acclimate the microorganism seed to the specific

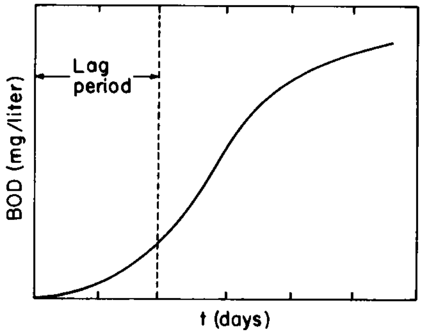

Fig. 2.3. Lag period in BOD test.

waste. This is achieved by starting with a sample of settled domestic sewage which contains a large variety of microorganisms, and adding a small amount of industrial effluent. Air is bubbled through this mixture. The operation is performed in bench scale reactors of either continuous or batch type. These reactors are described in Chapter 5, Section 6.1.

The process is repeated with gradual increase in the proportion of industrial waste to domestic sewage, until a microbial culture acclimated to the specific industrial waste is developed. This may be a long and difficult procedure for very toxic industrial wastewaters. When an acclimated culture has been developed, the BOD curve does not present a lag period, thus becoming a typical BOD curve of the general shape shown in Fig. 2.2.

c. Effect of Presence of Algae on the BOD Test

Presence of algae in the wastewater being tested affects the BOD test. If the sample is incubated in the presence of light, low BOD values are obtained owing to production of oxygen by photosynthesis, which satisfies part of the oxygen demand. On the other hand, if incubation is performed in darkness, algae survive for a while. Thus, short-term BOD determinations show the effect of oxygen on them. After a period in the dark, algae die and algal cells contribute to the increase of total organic content of the sample, thus leading to high BOD values. Therefore, the effect of algae on the BOD test is difficult to evaluate.

d. Glucose–Glutamic Acid Check

The quality of dilution water, which if contaminated leads to incorrect BOD values, the effectiveness of the seed, and the analytical technique are checked periodically by using pure organic compounds for which BOD is known or determinable. One of the most commonly used is a mixture of glucose ($C_6H_{12}O_6$) and glutamic acid [$HOOCCH_2CH_2CH(NH_2)COOH$]. A mixture of 150 mg/liter of each is recommended. Pure glucose has an exceptionally high oxidation rate with relatively simple seeds. When used with glutamic acid, the oxidation rate is stabilized and is similar to that of most municipal wastewaters. BOD of the standard glucose–glutamic acid solution is 220 ± 11 mg/liter. Any appreciable divergence from these values raises questions concerning quality of the distilled water or viability of the seeding material. If a variation greater than ± 20–22 mg/liter occurs more frequently than 5% of the time, this indicates a faulty technique.

e. Determination of Dissolved Oxygen (DO)

The BOD dilution method requires determinations of the amount of dissolved oxygen. These determinations are performed by either titration or instrumental methods. The basic titration method is that of Winkler. Wastewaters may contain several ions and compounds which interfere with the original DO determination. To eliminate these interferences, several modifications of the basic method have been proposed [13]. A brief description follows of the azide modification of Winkler's method, which effectively removes interference caused by nitrites. This is the most common interference found in practice. Other modifications to remove interferences are described in Ref. [13].

Winkler's method is based on oxidation of iodide ion (I^-), which is contained in the alkali–iodide–azide reagent, to iodine (I_2) by dissolved oxygen of the sample, and titration of the iodine by sodium thiosulfate ($Na_2S_2O_3$), utilizing starch as indicator. Oxidation is performed in acid medium (H_2SO_4) in the presence of manganese sulfate ($MnSO_4$). The alkali–iodide–azide reagent is a solution of NaOH, NaI, and NaN_3 (sodium azide).

Equation (2.13) corresponds to the oxidation of I^- to I_2.

$$2I^- \rightarrow I_2 + 2e \tag{2.13}$$

Interference of nitrites is due to their oxidation to NO with formation of I_2 [Eq. (2.14)].

$$2NO_2^- + 2I^- + 4H^+ \rightarrow 2NO + I_2 + 2H_2O \tag{2.14}$$

Titration of I_2 by thiosulfate corresponds to Eq. (2.15) [thiosulfate ($S_2O_3^{2-}$) is oxidized to tetrathionate $(S_4O_6^2)^-$].

$$2S_2O_3^{2-} + I_2 \rightarrow S_4O_6^{2-} + 2I^- \tag{2.15}$$

Starch yields a blue color in the presence of iodine. Titration with sodium thiosulfate is continued until the blue color disappears.

A variation of this procedure utilizes a new reagent (phenylarsine oxide, PAO) instead of sodium thiosulfate. This reagent has the advantage of being stable, whereas sodium thiosulfate deteriorates rapidly and should be restandardized before each determination. A description of this improved procedure is found in Ref. [8].

Instrumental determination of dissolved oxygen is performed by DO analyzers. A diagram of a typical model of the instrument is shown in Fig. 2.4. The DO analyzer is a galvanic system which utilizes a cylinder-shaped lead anode surrounding a rod-shaped silver cathode. Both electrodes are covered by a layer of KOH electrolyte contained in a thin electrolytic pad. A plastic membrane covers the electrodes and electrolyte and serves as a selective diffusion barrier which is permeable to all gases, including molecular oxygen, but is virtually impermeable to ionic species which may be present in the wastewaters. To measure DO the probe is dipped into the sample. A cell current which is proportional to the oxygen concentration in the sample is measured directly in terms of mg/liter of dissolved oxygen by the needle in the oxygen

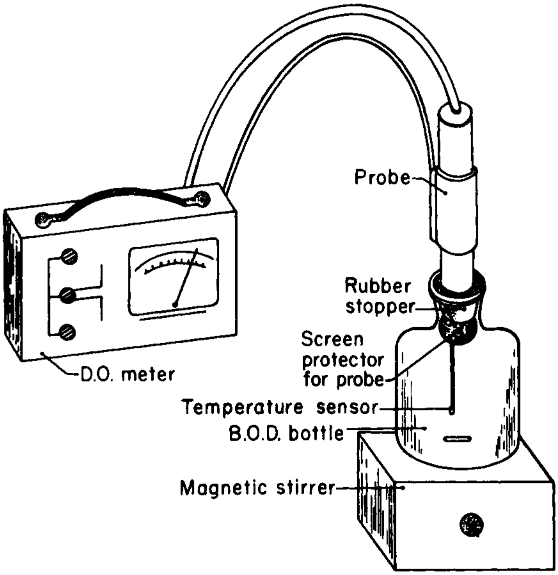

Fig. 2.4. Dissolved oxygen analyzer.

meter. The sample is constantly stirred during measurement, since only under these conditions is the current directly proportional to the oxygen concentration in the bulk of the test sample. Calibration of the DO analyzer is performed by measuring the DO of a sample of known oxygen content, which is determined by standard analytical methods (namely, the Winkler method) [13].

2.3.2. BOD Manometric Methods

The manometric apparatus described in this section is the Hach Model 2173 [7]. The Hach BOD apparatus has been compared with the standard dilution method under controlled laboratory conditions. In routine analysis it gives nearly equivalent results and precision. Since a physical change is observed, chemical laboratory analysis is not required. A diagram showing only one bottle is depicted in Fig. 2.5. The principle of operation is as follows: A measured sample of sewage or wastewater is placed in a bottle on the apparatus, which is connected to a closed-end mercury manometer. Above the sewage or water sample is a quantity of air (which contains approximately 21% oxygen by volume). Over a period of time bacteria in the sewage utilizes the oxygen to oxidize organic matter present in the sample, and thus dissolved oxygen is consumed. Air in the closed sample bottle replenishes the utilized oxygen, thus resulting in a drop of air pressure in the sample bottle. Mercury in the leg of the manometer connected to the bottle moves upward, as indicated by the arrow in Fig. 2.5. Thus, the pressure drop is registered on the mercury

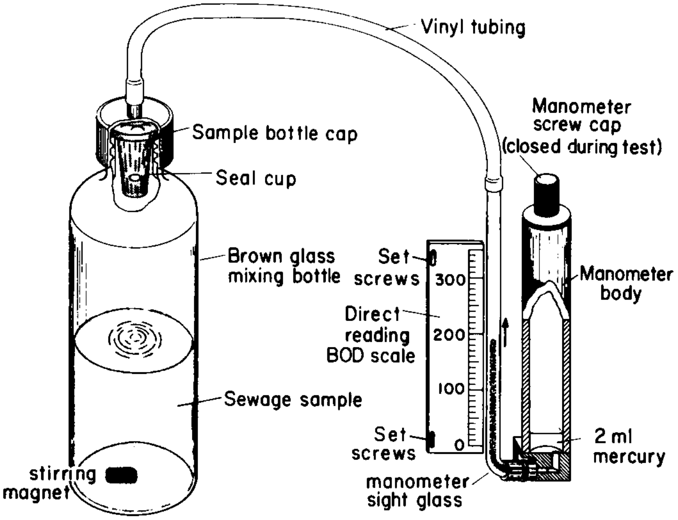

Fig. 2.5. Manometric BOD Apparatus (showing one cell) [7]. (*Courtesy of Hach Chemical Company.*)

manometer and read directly in mg/liter BOD. Prior to starting the test, set screws on the manometer scale are loosened and the zero mark is set at the top of the mercury column.

During the test period (5 days for BOD_5), the system is incubated at 20°C and the sample continually agitated by a magnetic stirring bar, which is rotated by a pulley system connected to a motor. Carbon dioxide is produced by oxidation of organic matter, and must be removed from the system so that it does not develop a positive gas pressure which would result in an error. This is accomplished by addition of a few drops of potassium hydroxide solution in the seal cup of each sample bottle. BOD readings are periodically checked by utilizing the standard glucose–glutamic acid solution.

When high oxygen demands are encountered the sample must be diluted. Accuracy of the manometric test is claimed as comparable to that of the dilution test.

2.4. TOTAL OXYGEN DEMAND (TOD) [6, 9, 17]

Usefulness of the standard COD method is due to the fact that results are obtained in 2 hr, rather than the 5 days taken for the common BOD measurement. However, the COD method is known *not* to oxidize contaminants as pyridine, benzene, and ammonia, although for many organic compounds oxidation has been reported as 95–100% of the theoretical.

Therefore, the search for improved analytical methods for determination of oxygen demand has focused on techniques [6] which are (1) meaningful and correlate with the accepted parameters for control and surveillance; (2) rapid, so results are known in minutes, not hours or days; and (3) truly adaptable to automation and continuous monitoring.

The Ionics model 225 Total Oxygen Demand (TOD) Analyzer determines total oxygen demand within 3 min. Figure 2.6 shows the functional elements of the system which includes the injection system, the combustion unit, the oxygen sensor assembly, and the recorder.

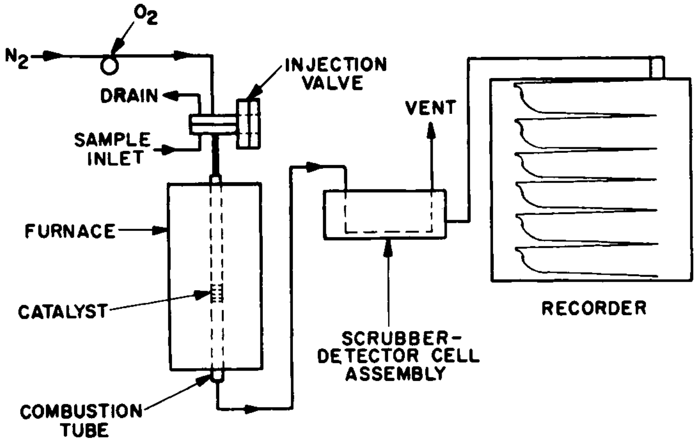

Fig. 2.6. ***Flow diagram for the TOD analyzer*** [6]. (*Reprinted with permission. Copyright by The American Chemical Society.*)

The wastewater sample is transmitted by an air-operated aspirator to the liquid injection valve. Upon actuation, the valve delivers a 20-μl (0.02 cm^3) sample into the combustion chamber. The sampling system is controlled by an adjustable program timer or by a manual pushbutton. A carrier gas (nitrogen) containing a small amount of oxygen of the order of 200 ppm is introduced simultaneously with the wastewater sample into the combustion chamber. The sample is vaporized and the combustible components are oxidized in a combustion tube. The tube, containing a platinum screen catalyst, is mounted in an electric furnace which is maintained at 900°C. As a result of the oxygen utilization in the combustion process, a momentary depletion of oxygen occurs in the inert gas stream. This depletion is accurately measured by passing the effluent through a platinum-lead fuel cell. Before entering the cell, the gas is scrubbed and humidified. Scrubbing is done by passing the gas through an aqueous caustic solution which removes carrier gas impurities harmful to the detector cell and humidifies the gaseous sample. The fuel cell and scrubber are located in a thermostatically controlled and insulated chamber.

Fuel cell current output is a function of oxygen concentration. This is graphically monitored on a potentiometer recorder, with changes in current taking the form of recorder peaks. The recorder system includes an automatic zero circuit to maintain a constant baseline. Peaks recorded are linearly proportional to the reduced oxygen concentration in the carrier gas and the sample total oxygen demand. TOD measurement for unknown samples is determined by comparison of the recorded peak heights with a standard calibration curve. A typical calibration curve for standard solution analysis is shown in Fig. 2.7, which demonstrates the linearity of peak height vs. TOD.

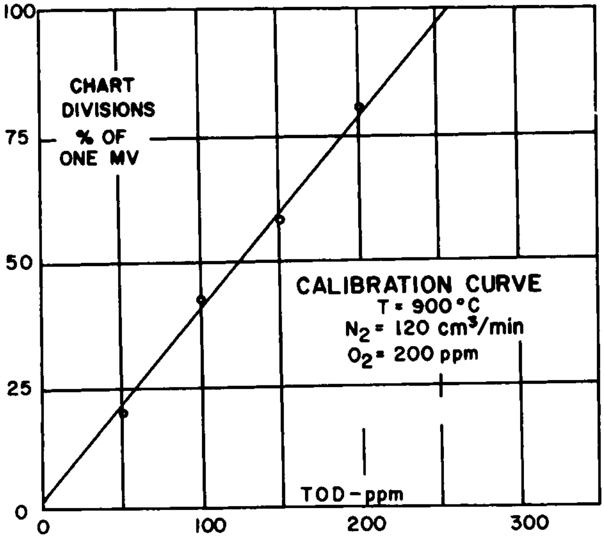

Fig. 2.7. Typical calibration curve for TOD analyzer [9]. (*Courtesy of Ionics Incorporated.*)

The TOD method measures the amount of oxygen consumed based on the following chemical reactions for the catalytic combustion process [Eqs. (2.16–2.18)].

$$C + O_2 \rightarrow CO_2 \tag{2.16}$$

$$H_2 + \tfrac{1}{2}O_2 \rightarrow H_2O \tag{2.17}$$

$$\text{N(combined)} + \tfrac{1}{2}O_2 \rightarrow NO \tag{2.18}$$

Sulfurous compounds are oxidized to a stable condition consisting of a fixed ratio of SO_2 to SO_3. Molecular nitrogen, normally used as the carrier gas, does not react in the combustion process.

Equation (2.19) corresponds to a typical theoretical oxidation (for the case of urea).

$$2NH_2CONH_2 + 5O_2 \rightarrow 2CO_2 + 4NO + 4H_2O \qquad (2.19)$$

Results of TOD analysis for a number of different compounds indicate that

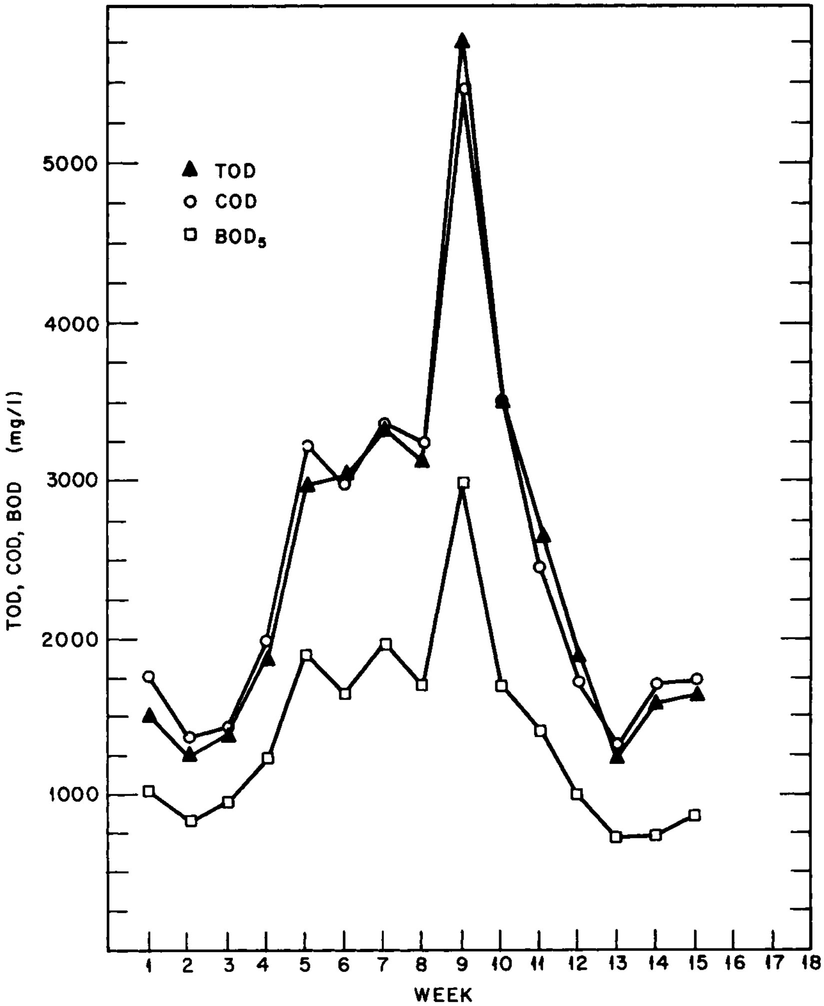

Fig. 2.8. ***Weekly analyses of a raw wastewater*** [17]. (*Reprinted with permission. Copyright by The American Chemical Society.*)

measured oxygen demand is usually closer to the theoretically calculated than is the case for chemical methods. These results are presented in Goldstein *et al.* [6]. None of the common ions normally found in water and wastewaters causes serious interference with TOD analyses [6].

Correlation of TOD analyses with COD has been checked for a number of typical waste streams [2, 3]. Figure 2.8 shows correlations of TOD, COD, and BOD_5 for a raw wastewater. Values of COD vs. TOD from Fig. 2.8 are plotted in Fig. 2.9, which shows a linear relationship. The relationship of TOD to COD or BOD_5 depends entirely on composition of the wastewater. Consequently, these ratios vary depending on the degree of biological treatment to which the wastewater is subjected.

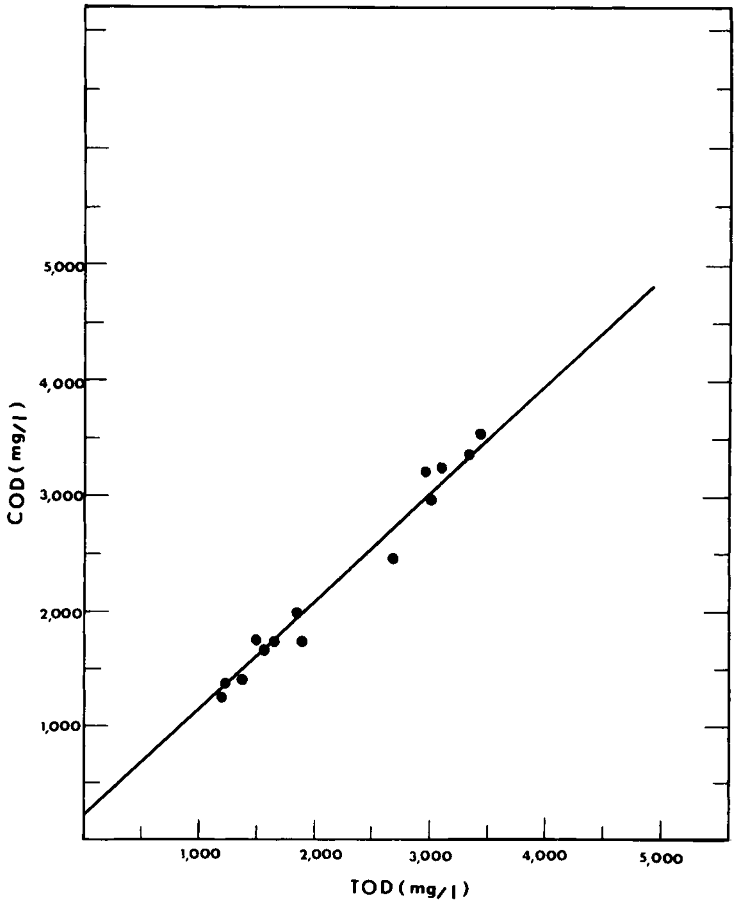

Fig. 2.9. The COD and TOD relationship of a raw wastewater [17]. (*Reprinted with permission. Copyright by The American Chemical Society.*)

3. Measurement of Organic Content: Group 2—Carbon Parameter Methods [2, 3]

Total organic carbon (TOC) tests are based on oxidation of the carbon of the organic matter to carbon dioxide, and determination of CO_2 either by absorption in KOH or instrumental analysis (infrared analyzer). Since theoretical oxygen demand (ThOD) measures O_2 and theoretical organic carbon (ThOC) measures carbon, the ratio of ThOD to ThOC is readily calculated from the stoichiometry of the oxidation equation.

Equation (2.20) corresponds to total oxidation of sucrose.

$$\underset{(12\times 12)}{C_{12}H_{22}O_{11}} + \underset{(12\times 32)}{12O_2} \rightarrow 12CO_2 + 11H_2O \tag{2.20}$$

$$\therefore \text{ThOD/ThOC} = (12\times 32)/(12\times 12) = 2.67 \tag{2.21}$$

The ratio of molecular weights of oxygen to carbon is 2.67.

Thus, the theoretical ratio of oxygen demand to organic carbon corresponds to the stoichiometric ratio of oxygen to carbon for total oxidation of the organic compound under consideration. The actual ratio obtained from COD (or BOD) tests and TOC determinations varies considerably from this theoretical ratio (Section 3.3). Experimental determination of TOD is performed by either manual (wet oxidation) or instrumental methods.

3.1. WET OXIDATION METHOD FOR TOC

The manual or wet oxidation method for TOC consists of oxidation of the sample in a solution of potassium dichromate ($K_2Cr_2O_7$), fuming sulfuric acid (H_2SO_4), potassium iodate (KIO_3), and phosphoric acid (H_3PO_4). Oxidation products are passed through a tube containing KOH, where the carbon dioxide collected is determined by weighing the absorption tube before and after the experiment.

3.2. CARBON ANALYZER DETERMINATIONS [1]

The fundamental operating principle of TOC analyzers is combustion of organic matter to carbon dioxide and water. Combustion gases are then passed through an infrared analyzer, sensitized for carbon dioxide, and the response is recorded on a strip chart. A diagram of the Beckman model 915-A Total Organic Carbon (TOC) Analyzer is shown in Fig. 2.10. This instrument permits separate measurements for total carbon and inorganic carbon. Total carbon includes the carbon of organic materials and inorganic carbon in the form of carbonates (CO_3^{2-}), bicarbonates (HCO_3^-), and CO_2 dissolved in the sample. There are two separate reaction tubes: one operated

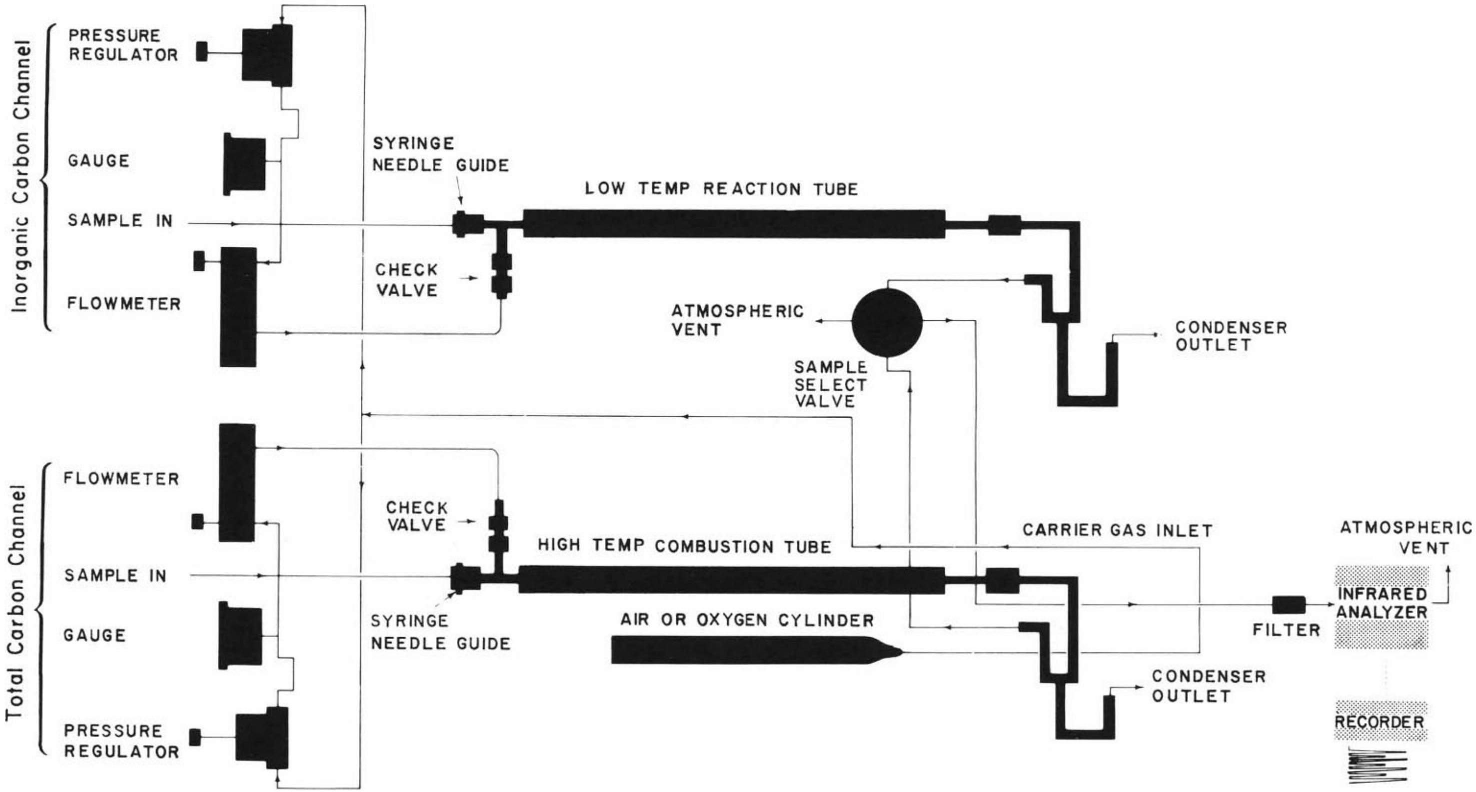

Fig. 2.10. Flow diagram of Beckman model 915-A Total Organic Carbon (TOC) Analyzer [1]. (*Courtesy of Beckman Instruments, Inc.*)

at high temperature (950°C) for measurement of total carbon and another operated at low temperature (150°C) for measurement of inorganic carbon.

Depending on range of analysis, a 20–200 μl water sample is syringe injected into a flowing stream of air and swept into a catalytic combustion tube containing a cobalt oxide-impregnated packing. The source of air which is used as carrier/oxidizer should be a low hydrocarbon, low CO_2 content cylinder. The combustion tube (high temperature combustion tube) is enclosed in an electric furnace thermostated at 950°C. Water is vaporized and all carbonaceous material is oxidized to CO_2 and steam. Airflow carries this cloud out of the furnace where the steam is condensed and removed. The CO_2 is swept into the nondispersive infrared analyzer.

Transient CO_2 is indicated as a peak on a strip chart recorder. Peak height is a measure of CO_2 present, which is directly proportional to the concentration of total carbon in the original sample and includes organic carbon, inorganic carbon, and CO_2 dissolved in the sample. By using standard solutions, the chart is calibrated in milligrams total carbon per liter of sample.

In a second operation, a sample of similar size is also syringe injected into a stream of air and swept into the second reaction tube (low temperature reaction tube), containing quartz chips wetted with 85% phosphoric acid. This tube is enclosed in an electric heater thermostated at 150°C, which is below the temperature at which organic matter is oxidized. The acid-treated packing causes release of CO_2 from inorganic carbonates, and the water is vaporized. Airflow carries the cloud of steam and CO_2 out of the furnace, where steam is condensed and removed. By previous repositioning of a dual channel selector valve, the CO_2 is swept into the infrared analyzer.

This quantity of CO_2 is also indicated on the strip chart recorder as a transient peak. Peak height is a measure of the CO_2 present, which is proportional to the concentration of inorganic carbonates plus CO_2 dissolved in the original sample. By using standard solutions, the chart is calibrated in milligrams inorganic carbon per liter of sample. Subtracting results obtained in the second operation from those in the first yields total organic carbon in milligrams TOC per liter of sample.

3.3. OXYGEN DEMAND–ORGANIC CARBON CORRELATION

The ratio ThOD/ThOC, which theoretically is equal to the stoichiometric ratio of oxygen to carbon for total oxidation of the organic compound under consideration, ranges in practice from nearly zero, when the organic matter is resistant to dichromate oxidation (e.g., pyridine), to values of the order of 6.33 for methane or even slightly higher when inorganic reducing agents are present. Table 2.2 presents relationships between oxygen demand and total carbon for several organic compounds.

Table 2.2
Relationships between Oxygen Demand and Total Carbon for Organic Compounds [3]

Substance	ThOD/ThOC (calculated)	COD/TOC (measured)
Acetone	3.56	2.44
Ethanol	4.00	3.35
Phenol	3.12	2.96
Benzene	3.34	0.84
Pyridine	3.33	—
Salicylic acid	2.86	2.83
Methanol	4.00	3.89
Benzoic acid	2.86	2.90
Sucrose	2.67	2.44

Correlation of BOD with TOC for industrial wastewaters is difficult because of their considerable variation in chemical composition. For domestic wastewaters a relatively good correlation has been obtained, which is represented by the straight line relationship given by Eq. (2.22).

$$\mathrm{BOD}_2 = 1.87(\mathrm{TOC}) - 17 \tag{2.22}$$

4. Mathematical Model for the BOD Curve

It is desirable to represent the BOD curve (Fig. 2.2) by a mathematical model. From kinetic considerations (Chapter 5, Section 3), the mathematical model utilized to portray the rate of oxygen utilization is that of a first-order reaction. Figure 2.2 reveals that the rate of oxygen utilization, given by the tangent to the curve at a given incubation time, decreases as concentration of organic matter remaining unoxidized becomes gradually smaller. Since there is a proportionality between the rate of oxygen utilization and that of destruction of organic matter by biological oxidation, rate equation [Eq. (2.23)] is written in terms of organic matter concentration (L; mg/liter).

$$dL/dt = -k_1 L \tag{2.23}$$

where L is concentration of organic matter (mg/liter) at time t; dL/dt, rate of disappearance of organic matter by aerobic biological oxidation ($dL/dt < 0$); t, time of incubation (days); and k_1, rate constant (day^{-1}).

Separating variables L and t, and integrating from time zero corresponding to initial concentration of organic matter, L_0, to a time t corresponding to concentration L [Eq. (2.24)]:

$$\ln(L/L_0) = -k_1 t \tag{2.24}$$

Changing to decimal logarithms [Eq. (2.25)]

$$\log(L/L_0) = -k_1 t/2.303 \tag{2.25}$$

let $k_1/2.303 = k$. Then [Eq. (2.26)]

$$L/L_0 = 10^{-kt} \tag{2.26}$$

or

$$L = L_0 \cdot 10^{-kt} \tag{2.27}$$

Let y be the organic matter oxidized up to time t, i.e.,

$$y = L_0 - L \tag{2.28}$$

Conversely, y also measures the oxygen consumption up to time t, i.e., the ordinate of the BOD curve in Fig. 2.2 at time t. Combining Eqs. (2.28) and (2.27),

$$y = L_0(1-10^{-kt}) \tag{2.29}$$

which is the mathematical model for the BOD curve. From Eq. (2.29) it follows that for a very long oxidation period (i.e., $t \rightarrow \infty$), $y = L_0$. Therefore, k and L_0 measure, respectively, the rate of biochemical stabilization and the total amount of putrescible matter present.

From Eq. (2.27)

$$\log L = \log L_0 - kt \tag{2.30}$$

Equation (2.30) indicates that constants k and L_0 can be obtained from a semilogarithmic plot of L vs. t. Typical values of the rate constant k are presented in Chapter 5 for several types of wastewaters (Table 5.2, Section 5).

5. Determination of Parameters k and L_0

In application of Eq. (2.29) one usually has available a series of BOD measurements (y) at a sequence ($n = 1, 2, 3, \ldots, x$) days. It is desired to determine the optimum values of parameters k and L_0 which satisfy Eq. (2.29) for the set of data. Thus, it is fundamentally a curve-fitting problem.

Several methods for calculating parameters k and L_0 have been proposed. Three of these, recommended by Eckenfelder [3], are (1) log-difference method, (2) method of moments [10], and (3) Thomas' graphical method.

5.1. LOG-DIFFERENCE METHOD

This method is based on the following considerations. Differentiating Eq. (2.29) with respect to t:

$$dy/dt = r = L_0(-10^{-kt})(\ln 10)(-k) \tag{2.31}$$

or

$$dy/dt = r = 2.303 L_0 k \cdot 10^{-kt} \tag{2.32}$$

where r is the rate of oxygen utilization. Taking decimal logarithms

$$\log r = \log(2.303 L_0 k) - kt \tag{2.33}$$

Equation (2.33) indicates that k and L_0 can be obtained from a semilog plot of r vs. t.

Step 1. Plot y (oxygen utilization) vs. t on cartesian coordinate paper. Draw a smooth best-fit curve through the points, discarding data which seem to be in error.

Step 2. Plot daily differences, $\Delta y/\Delta t$ vs. time (on semilog graph paper). Time intervals are usually taken as $0, 1, 2, 3, \ldots$ days, so that $\Delta t = 1$. Values of Δy's are conventionally plotted vs. the time t corresponding to the middle of each interval (e.g., the value of Δy corresponding to interval 0–1 is plotted vs. $t = 0.5$).

Draw the best-fit straight line through these points.

Step 3. Calculation of k and L_0. From the straight line drawn in Step 2, Eq. (2.33) yields:

$$k = -(\text{slope}) \tag{2.34}$$

$$\text{Intercept} = 2.303 L_0 k \tag{2.35}$$

$$\therefore\ L_0 = \text{intercept}/(2.303k) = \text{intercept}/(2.303)(-\text{slope}) \tag{2.36}$$

Therefore, k and L_0 are calculated from Eqs. (2.34) and (2.36), respectively.

Example 2.1

The data in Table 2.3 on oxygen utilization are available from BOD tests of a wastewater.

Obtain the values of k and L_0 in the BOD equation.

TABLE 2.3
BOD Tests of Wastewater

t (days)	y (mg/liter of BOD)
0	0.0
1	9.2
2	15.9
3	20.9
4	24.4
5	27.2
6	29.1
7	30.6

SOLUTION

Step 1. Data are plotted in Fig. 2.11. The curve is fairly smooth and the raw data are used in Step 2. If needed, smoothing is done by the best straight-line fit.

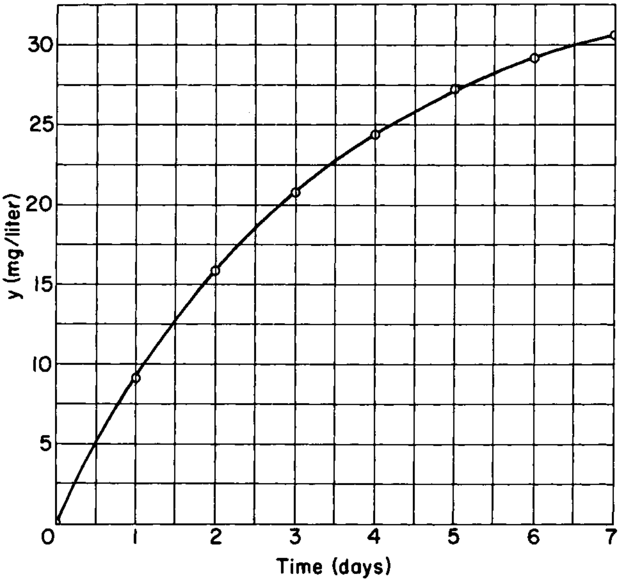

Fig. 2.11. Plot y vs. t (Example 2.1).

Step 2. Construct the difference table (Table 2.4). Values in column (*3*) are plotted vs. those in column (*4*) on semilog paper. This plot is shown in Fig. 2.12.

TABLE 2.4
Log-Difference Values from Table 2.3

(*1*) t (days)	(*2*) y (mg/liter)	(*3*) Δy (mg/liter) = $\Delta y/\Delta t$; since $\Delta t = 1$	(*4*) Mid-interval values of t
0	0	—	—
1	9.2	9.2	0.5
2	15.9	6.7	1.5
3	20.9	5.0	2.5
4	24.4	3.5	3.5
5	27.2	2.8	4.5
6	29.1	1.9	5.5
7	30.6	1.5	6.5

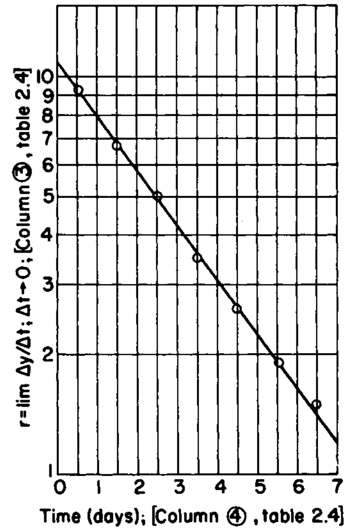

Fig. 2.12. Calculation of k and L_0 by the log-difference method.

Step 3. Calculate k and L_0.

(a) *Calculation of k.* Base calculations on the coordinates of two points on the straight line (7, 1.2; 0, 10.9) and Eq. (2.34).

$$\text{Slope} = (\log 10.9 - \log 1.2)/(0-7) = -0.137$$

$$\therefore\ k = 0.137\ \text{day}^{-1}$$

(b) *Calculation of L_0.* From Eq. (2.36)

$$L_0 = 10.9/(2.303 \times 0.137) = 34.5\ \text{mg/liter}$$

5.2. METHOD OF MOMENTS [10]

This method is of simple application once diagrams of $\sum y/L_0$ vs. k and $\sum y/\sum ty$ vs. k are constructed for an n-day sequence of BOD measurements. Equations are derived next for construction of Moore's diagrams for an n-day sequence of BOD measurements. These equations are applied to 7-, 5-, and 3-day sequences, yielding Figs. 2.13–2.15, respectively.

Consider BOD measurements taken over an n-day sequence, as indicated in Table 2.5. First, calculate ratio $\sum y/L_0$. The summation of entries in column (*2*) of Table 2.5 is [Eq. (2.37)]

$$\sum y = L_0[(1+1+1+\cdots+1) - (10^{-k}+10^{-2k}+10^{-3k}+\cdots+10^{-nk})] \tag{2.37}$$

or

$$\sum y = L_0[n - (10^{-k}+10^{-2k}+10^{-3k}+\cdots+10^{-nk})] \tag{2.38}$$

TABLE 2.5
BOD Measurements, *n*-Day Sequence

(*1*) *t* (days)	(*2*) Eq. (2.29); $t = 1, 2, 3, \ldots, n$; $y = L_0(1 - 10^{-kt})$	(*3*) = (*1*) × (*2*) *ty*
1	$L_0(1-10^{-k})$	$L_0(1-10^{-k})$
2	$L_0(1-10^{-2k})$	$2L_0(1-10^{-2k})$
3	$L_0(1-10^{-3k})$	$3L_0(1-10^{-3k})$
4	$L_0(1-10^{-4k})$	$4L_0(1-10^{-4k})$
5	$L_0(1-10^{-5k})$	$5L_0(1-10^{-5k})$
6	$L_0(1-10^{-6k})$	$6L_0(1-10^{-6k})$
7	$L_0(1-10^{-7k})$	$7L_0(1-10^{-7k})$
—	—	—
—	—	—
—	—	—
n	$L_0(1-10^{-nk})$	$nL_0(1-10^{-nk})$
	$\sum y$	$\sum ty$

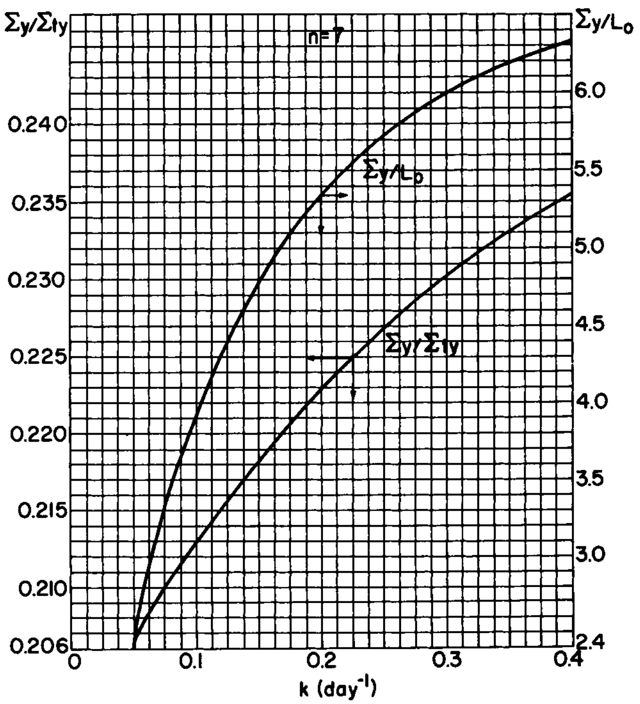

Fig. 2.13. Moore's method (7-day sequence).

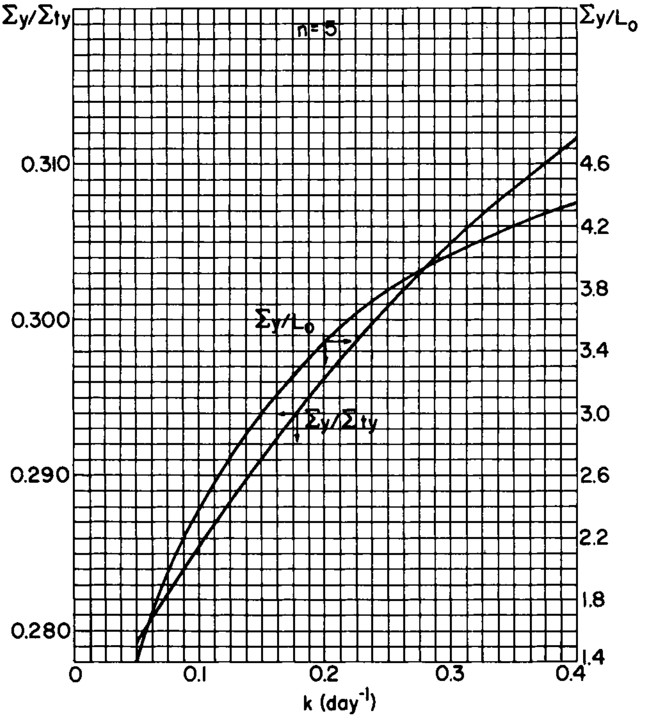

Fig. 2.14. Moore's method (5-day sequence).

The terms within parentheses in Eq. (2.38) form a geometrical progression for which the sum of terms is [Eq. (2.39)]

$$S = [(10^{-k})(10^{-nk}-1)]/(10^{-k}-1) \tag{2.39}$$

Substituting this value in Eq. (2.38) and solving for the ratio $\sum y/L_0$:

$$\sum y/L_0 = n - [10^{-k}(10^{-nk}-1)/(10^{-k}-1)] \tag{2.40}$$

From Eq. (2.40) it follows that for a given sequence of n days, the ratio $\sum y/L_0$ is only a function of k. Thus for a fixed n, one assumes values of k and plots a curve of $\sum y/L_0$ vs. k.

Now calculate ratio $\sum y/\sum ty$. $\sum y$ is obtained from Eq. (2.40), and $\sum ty$ corresponds to summation of entries in column (*3*) of Table 2.5.

$$\sum ty = L_0[(1+2+3+\cdots+n) - (10^{-k}+2\times 10^{-2k}+3\times 10^{-3k}+\cdots+n\times 10^{-nk})] \tag{2.41}$$

or

$$\sum ty = L_0\left(\sum_{i=1}^{i=n} i - \sum_{i=1}^{i=n} i\times 10^{-ik}\right) \tag{2.42}$$

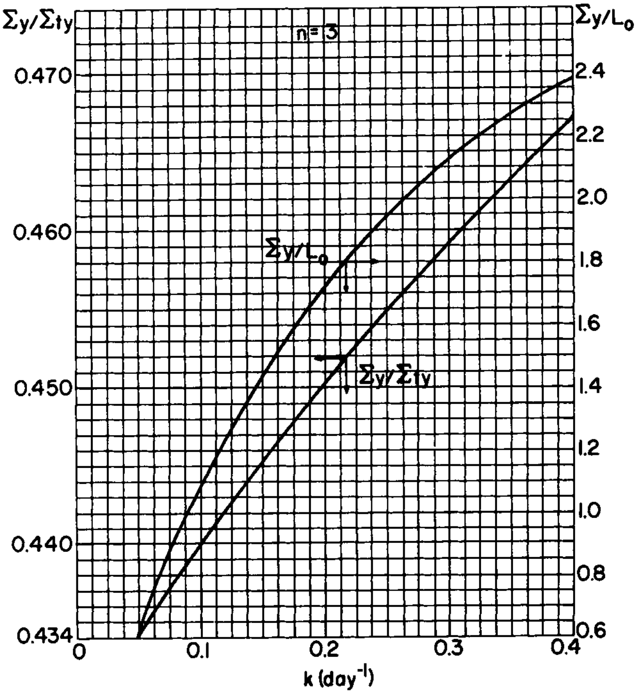

Fig. 2.15. Moore's method (3-day sequence).

Therefore, from Eqs. (2.40) and (2.42), ratio $\sum y/\sum ty$ is

$$\sum y/\sum ty = \frac{n - [10^{-k}(10^{-nk} - 1)/(10^{-k} - 1)]}{\sum_{i=1}^{i=n} i - \sum_{i=1}^{i=n} i \times 10^{-ik}} \tag{2.43}$$

From Eq. (2.43) it follows that for a given sequence of n days, ratio $\sum y/\sum ty$ is only a function of k. Thus for a fixed n, one assumes values of k and plots a curve of $\sum y/\sum ty$ vs. k from Eq. (2.43).

For specific cases such as the 7-day sequence, Eqs. (2.40) and (2.43) yield Eq. (2.44).

For $n = 7$

$$\sum y/L_0 = 7 - [10^{-k}(10^{-7k} - 1)/(10^{-k} - 1)] \tag{2.44}$$

$$\sum y/\sum ty = \frac{7 - [10^{-k}(10^{-7k} - 1)/(10^{-k} - 1)]}{28 - \sum_{i=1}^{i=7} i \times 10^{-ik}} \tag{2.45}$$

where, in Eq. (2.45)

$$\sum_{i=1}^{i=n} i = \sum_{i=1}^{i=7} i = 1 + 2 + 3 + 4 + 5 + 6 + 7 = 28$$

and

$$\sum_{i=1}^{i=7} i \times 10^{-ik} = 10^{-k} + 2 \times 10^{-2k} + 3 \times 10^{-3k} + 4 \times 10^{-4k} + 5 \times 10^{-5k} + 6 \times 10^{-6k} + 7 \times 10^{-7k}$$

Figures 2.13–2.15 present graphs of $\sum y/L_0$ vs. k and $\sum y/\sum ty$ vs. k for 7-, 5-, and 3-day sequences, respectively. These figures are constructed from Eqs. (2.40) and (2.43), respectively, by assuming values of n (7, 5, 3) and k and calculating the corresponding ratios.

Application of Moore's diagram for calculation of parameters k and L_0 is illustrated by Example 2.2.

Example 2.2

Determine values of k and L_0 from the set of BOD determinations of Example 2.1.

SOLUTION

Step 1. Construct Table 2.6.

TABLE 2.6
Application of Moore's Method (Example 2.2)

t (days)	y (mg/liter BOD)	ty
0	0.0	0.0
1	9.2	9.2
2	15.9	31.8
3	20.9	62.7
4	24.4	97.6
5	27.2	136.0
6	29.1	174.6
7	30.6	214.2
	$\sum y = 157.3$	$\sum ty = 726.1$

Step 2. Calculate ratio $\sum y/\sum ty$.

$$\sum y/\sum ty = 157.3/726.1 = 0.217$$

Step 3. From Fig. 2.13 ($n = 7$) read for $\sum y/\sum ty = 0.217$.

$$k = 0.140 \text{ day}^{-1} \text{ (Abscissa of lower curve)}$$

From the ordinate of upper curve read

$$\sum y/L_0 = 4.62$$

$$\therefore\ L_0 = \sum y/4.62 = 157.3/4.62 = 34.05 \text{ mg/liter}$$

These values of k and L_0 agree closely with those calculated by the log-difference method in Example 2.1 ($k = 0.137$ day^{-1} and $L_0 = 34.5$ mg/liter).

5.3. THOMAS' GRAPHICAL METHOD [16]

This is an approximate method which is justified since precision of the experimental results is often limited. The method is based on the similarity of the function

$$(1-10^{-kt}) \tag{2.46}$$

which is a factor of Eq. (2.29), and the function

$$2.3kt[1+(2.3/6)kt]^{-3} \tag{2.47}$$

This similarity is seen in their respective series expansions, which are

$$(1-10^{-kt}) = (2.3kt)[1-(1/2)(2.3kt)+(1/6)(2.3kt)^2-(1/24)(2.3kt)^3+\cdots] \tag{2.48}$$

and

$$2.3kt[1+(2.3/6)kt]^{-3} = (2.3kt)[1-(1/2)(2.3kt)+(1/6)(2.3kt)^2 -(1/21.6)(2.3kt)^3+\cdots] \tag{2.49}$$

Comparison of the right-hand members of Eqs. (2.48) and (2.49) reveals that the first three terms in the two series within brackets are identical, and that the difference between the fourth terms is small. Replacing the function between parentheses in Eq. (2.29) by its approximation given by Eq. (2.47) yields Eq. (2.50).

$$y = L_0(2.3kt)[1+(2.3/6)kt]^{-3} \tag{2.50}$$

from which, taking the inverse and rearranging,

$$t/y = [1+(2.3/6)kt]^3/2.3kL_0 \tag{2.51}$$

Taking the cube root of both members of Eq. (2.51) and rearranging,

$$(t/y)^{1/3} = 1/(2.3kL_0)^{1/3} + [(2.3k)^{2/3}/6L_0{}^{1/3}]t \tag{2.52}$$

From Eq. (2.52), a plot of $(t/y)^{1/3}$ vs. t yields a straight line (Fig. 2.16 for Example 2.3) from which

$$\text{Slope} = B = (2.3k)^{2/3}/6L_0{}^{1/3} \tag{2.53}$$

$$\text{Intercept} = A = 1/(2.3kL_0)^{1/3} \tag{2.54}$$

From Eqs. (2.53) and (2.54) one obtains Eqs. (2.55) and (2.56).

$$k = 6B/2.3A = 2.61(B/A) \tag{2.55}$$

$$L_0 = 1/(2.3kA^3) \tag{2.56}$$

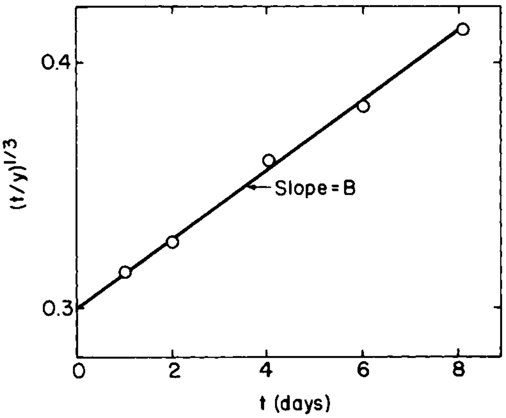

Fig. 2.16. Application of Thomas' method **[14].**

Application of this method is illustrated by Example 2.3.

Example 2.3 [16]

The BOD results tabulated below are observed on a sample of raw sewage at 23°C. Calculate parameters k and L_0.

t (days)	y (BOD, mg/liter)
0	0
1	32
2	57
4	84
6	106
8	111

SOLUTION

Step 1. Construct Table 2.7.

TABLE 2.7
Application of Thomas' Method (Example 2.3)

(*1*) t	(*2*) y	(*3*) = (*1*)÷(*2*) t/y	(*4*) = $[(3)]^{1/3}$ $(t/y)^{1/3}$
0	0	—	—
1	32	0.03125	0.315
2	57	0.03509	0.327
4	84	0.04762	0.362
6	106	0.05660	0.384
8	111	0.07207	0.416

Step 2. Plot $(t/y)^{1/3}$ vs. t. The plot is shown in Fig. 2.16. From Fig. 2.16 obtain

$$A = 0.30 \quad \text{(intercept)}$$

$$B = (0.416-0.300)/(8.0-0.0) = 0.0145 \quad \text{(slope)}$$

Step 3. From Eqs. (2.55) and (2.56), obtain k and L_0.

$$k = 2.61(0.0145/0.30) = 0.13 \text{ day}^{-1}$$

$$L_0 = 1/[2.3 \times 0.13(0.30)^3] = 124 \text{ mg/liter}$$

6. Relationship between *k* and Ratio BOD_5/BOD_u

Equation (2.29) is written as in Eq. (2.57) for $t = 5$ days, letting $y = BOD_5$ and $L_0 = BOD_u$.

$$BOD_5 = BOD_u(1-10^{-5k}) \tag{2.57}$$

from which

$$BOD_5/BOD_u = 1 - 1/10^{5k} \tag{2.58}$$

Assuming values of k, a curve of BOD_5/BOD_u vs. k is plotted from Eq. (2.58). This curve rises with increasing k's, reaching a plateau corresponding to an ordinate BOD_5/BOD_u approaching unity for values of k beyond 0.3 [3].

From Eq. (2.58) for large values of k, BOD_5/BOD_u approaches unity. This means that for a given substrate, if the rate of biochemical oxidation is very high, the value of BOD_5 is essentially equal to that of the ultimate BOD.

7. Environmental Effects on the BOD Test

The BOD test is affected by temperature and pH.

7.1. EFFECT OF TEMPERATURE

The reaction rate constant k is directly affected by temperature. The temperature dependence of k is given by the van't Hoff-Arrhenius equation [Eq. (2.59)].

$$d\ln k/dT = E/RT^2 \tag{2.59}$$

where k is reaction rate constant; T, absolute temperature; R, universal gas constant; and E, activation energy for the reaction (common values for wastewater treatment processes are in the range of 2000–20,000 cal/g mole). Integrating between limits [Eq. (2.60)]:

$$\ln(k_2/k_1) = [E(T_2-T_1)]/(RT_1T_2) \tag{2.60}$$

Since most wastewater treatment processes take place at nearly room temperature, the term E/RT_1T_2 is nearly constant. Let it be denoted as C. Then [Eq. (2.61)]

$$\ln(k_2/k_1) = C(T_2-T_1) \tag{2.61}$$

$$\therefore\ k_2/k_1 = e^{C(T_2-T_1)} \tag{2.62}$$

Let $e^C = \theta$ = temperature coefficient. Then

$$k_2/k_1 = \theta^{(T_2-T_1)} \tag{2.63}$$

The most usual application consists of estimation of constant k at a temperature T from its value determined experimentally at 20°C. From Eq. (2.63) we obtain Eq. (2.64).

$$k_T = k_{20}\theta^{(T-20)} \tag{2.64}$$

where k_T is reaction rate at T°C; k_{20}, reaction rate at 20°C; and T, temperature (°C). Although θ is approximately constant, it varies slightly with temperature and its appropriate value should be selected. Values given below are those recommended by Schroepfer [12].

$$\theta = 1.135\ (4°\text{–}20°\text{C})$$

$$\theta = 1.056\ (20°\text{–}30°\text{C})$$

From Eq. (2.64) it follows that for a 10° rise in temperature the reaction rate nearly doubles.

7.2. EFFECT OF pH

The standard BOD test specifies a pH of 7.2. If the pH is not 7.2, values of BOD_5 obtained are lower. It is recommended, therefore, to adjust the pH to 7.2. A typical curve of percentage of normal 5-day BOD vs. pH is presented by Eckenfelder and Ford [4]; its maximum (100%) corresponding to pH 7.2.

8. Nitrification

Equation (2.29) describes the oxidation of carbonaceous matter. Oxidation of nitrogenous material also contributes to oxygen demand if incubation is carried out for a sufficiently long period of time. This oxidation (referred to as nitrification) takes place in two steps:

1. Ammonium ion, NH_4^+, is oxidized to nitrites in the presence of *Nitrosomonas* microorganisms [Eq. (2.65)].

$$2NH_4^+ + 3O_2 \xrightarrow{\textit{Nitrosomonas}} 2NO_2^- + 2H_2O + 4H^+ \tag{2.65}$$

2. Nitrites are then oxidized to nitrates in the presence of *Nitrobacter* microorganisms [Eq. (2.66)].

$$2NO_2^- + O_2 \xrightarrow{\textit{Nitrobacter}} 2NO_3^- \tag{2.66}$$

Rate constants, k, for nitrification are much lower than those for oxidation of carbonaceous matter. Although oxidation of carbonaceous and nitrogenous matter may occur simultaneously, nitrification normally does not begin until the carbonaceous oxygen demand is partially satisfied.

A typical BOD curve for a wastewater showing carbonaceous oxidation and nitrification phases is shown in Fig. 2.17. Nitrification is suppressed by addition of certain chemicals (e.g., methylene blue, thiourea). If this is done, the BOD curve thus obtained approaches a limiting ordinate L_0 (ultimate carbonaceous demand), as indicated in Fig. 2.17. Beyond time t_c, the carbonaceous oxygen demand is essentially satisfied, so the ordinate value becomes constant at L_0. If, on the other hand, nitrification is not suppressed beyond $t = t_c$, the effect of nitrification is superimposed on the carbonaceous oxygen demand to yield the combined oxygen demand curve (carbonaceous + nitrification demand).

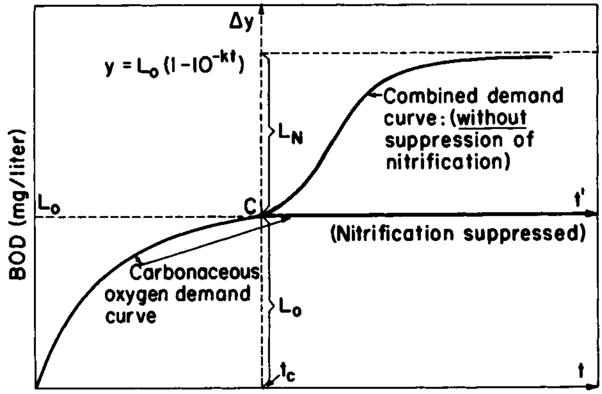

Fig. 2.17. Carbonaceous and nitrogenous BOD.

The carbonaceous oxygen demand curve is described by Eq. (2.29). If a translation of coordinate axes is performed so that the origin of the new system of coordinates coincides with point C (system Δy vs. t'), the equation for the nitrification oxygen demand curve [Eq. (2.67)] is written as

$$\Delta y = L_N(1-10^{-k_N t'}) \qquad (t > t_c) \tag{2.67}$$

where $t' = t - t_c$. Thus

$$\Delta y = L_N[1-10^{-k_N(t-t_c)}] \tag{2.68}$$

where k_N is the rate constant for nitrogenous demand and L_N the ultimate nitrogenous demand. The abscissa axis of the coordinate system Δy vs. t' essentially coincides with the carbonaceous oxygen demand curve beyond $t = t_c$.

Equations (2.29) and (2.69) for combined oxygen demand curve are

For $t < t_c$ (carbonaceous oxygen demand only)

$$y = L_0(1-10^{-kt}) \tag{2.29}$$

For $t > t_c$

$$y = \underbrace{L_0(1-10^{-kt})}_{\substack{\text{for } t > t_c, \\ L_0(1-10^{-kt}) \to L_0}} + \underbrace{L_N[1-10^{-k_N(t-t_c)}]}_{\substack{\text{increment } \Delta y \\ \text{due to nitrogenous} \\ \text{demand}}} \tag{2.69}$$

Values of parameters k_N and L_N are determined by any of the methods previously discussed with reference to the new system of coordinates [i.e., Eq. (2.68)].

9. Evaluation of Feasibility of Biological Treatment for an Industrial Wastewater

9.1. INTRODUCTION

Frequently, it is necessary to conduct treatability studies for streams of industrial wastewaters, since they may contain toxic substances which have an adverse effect on biological systems. The problem of acclimation of microorganism seed to toxic substances is discussed in Section 2.3.1. Two types of tests to evaluate the feasibility of biological treatment for industrial wastewater [4] are (1) manometric techniques (*Warburg respirometer*), and (2) batch reactor evaluation.

9.2. WARBURG RESPIROMETER

A schematic diagram of the Warburg respirometer is shown in Fig. 2.18. The principle of operation, which consists in respiring a wastewater sample in a closed air atmosphere at constant temperature, is identical to that of the BOD manometric apparatus (Section 2.3.2). Oxygen utilized is measured with respect to time by noting the decrease in pressure of the system at constant volume. The CO_2 evolved is absorbed by a solution of KOH; thus the decrease in pressure is a measure of oxygen consumption only.

Steps in the operational procedure are given below [4].

1. The wastewater sample is placed in the sample flask with the required volume of biological seed. The sample flask is immersed in a constant temperature bath and agitated by a shaking mechanism.

2. A 20% solution of KOH is placed in the center well (about one-quarter full). Insert a strip of folded filter paper inside the center well to enhance the alkali absorption of carbon dioxide. The paper soaks up KOH solution and in this way a larger alkali surface becomes available for absorption of carbon dioxide.

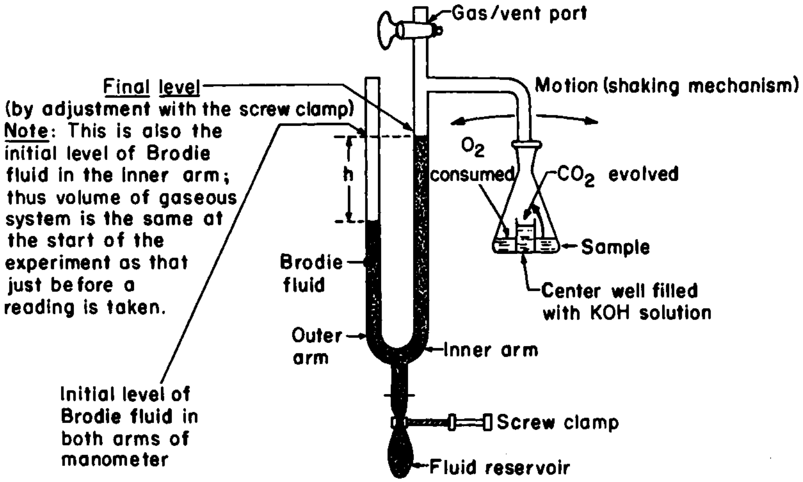

Fig. 2.18. Schematic diagram of Warburg respirometer.

3. Set up a reference flask ("thermobarometer") by adding to a sample flask only distilled water. The volume of distilled water equals the total waste seed volume in each of the test flasks. This reference flask is used for correction due to changes of atmospheric pressure during the time of the experiment, hence the name "thermobarometer."

4. Shake the system* with the gas vent port open for approximately 5 min. The level of the manometric fluid (Brodie's fluid) is the same in both arms of the manometer when equilibrium is reached. Adjust the manometer fluid to the reference mark in the inner arm of the manometer, with the gas vent port open. Adjustment of the level of the manometric fluid is made by means of a screw clamp, thus permitting adjustment of the height of manometric liquid within the two arms of the manometer. Stop the shaking and check all fittings.

5. Close the gas vent port, turn on the shaking assembly, and take readings at selected time intervals. Prior to a reading, turn the shaker off and adjust level of Brodie's fluid in the inner arm to the reference mark. Thus, the volume of the gaseous system is the same at the start of the experiment as that just before a reading is taken.

6. The reference flask readings serve the purpose of correcting for atmospheric pressure changes during the test. If the fluid in the outer arm of the

---

* Shaking is necessary because a film with a depleted oxygen concentration forms at the interface between the gas phase and the liquid sample if there is no agitation. This slows down the rate of oxygen utilization. Shaking provides for film renewal so that the liquor is always in contact with a gas phase rich in oxygen.

manometer attached to the thermobarometer flask rises, there has been a decrease in atmospheric pressure, and the observed reading must be added to the test value. If, on the other hand, the fluid in the outer arm of the manometer falls, there has been an increase in atmospheric pressure, and the observed reading must be subtracted from the test value.

$$\text{Sample reading } (h) = P_{\mathrm{at}} - P_{\mathrm{system}} \tag{2.70}$$

where P_{at} is the value of atmospheric pressure at start of experiment. During an experiment if the atmospheric pressure (P_{at}) rises, the calculated h [Eq. (2.70)] would be higher than the true value unless the appropriate correction is subtracted.

7. Once the substrate has been utilized, oxygen uptake stabilizes and the test series is terminated.

The cumulative oxygen uptakes (milligrams of oxygen per liter of solution) are then plotted vs. time (hr). A typical graph obtained for a toxic wastewater stream is shown in Fig. 2.19. This stream is added to domestic sewage (indicated as "seed" in Fig. 2.19) in increasingly larger proportions.

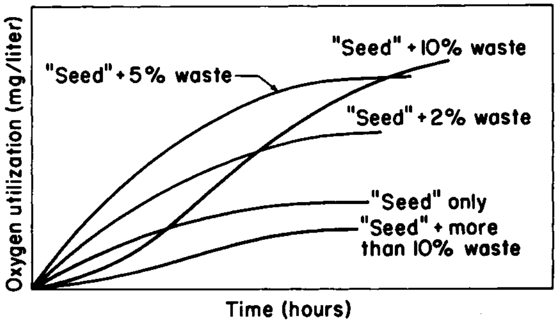

Fig. 2.19. Oxygen uptakes at different wastewater concentrations.

Figure 2.19 indicates that this specific wastewater is toxic or inhibitory when its concentration exceeds 10% in volume, in which case the oxygen uptake suffers a large drop.

Calculation of Oxygen Uptake (mg O_2/Liter of Sample) from the Reading h (cm)

In this calculation procedure, it is assumed that atmospheric pressure has not changed during the time of experiment [if it does, correct as indicated in Step 6, Eq. (2.70)]. At the beginning of the experiment, the ideal gas law is applied to the air in the closed system, i.e.,

$$P_1 V = N_1 RT \tag{2.71}$$

where P_1 is the atmospheric pressure at start of experiment (cm of Brodie's fluid)*; V the gas volume in closed system (ml); T the temperature of constant temperature bath (°K); N_1 the g moles of air at the beginning of experiment in closed system; and R the universal gas constant.†

From Eq. (2.71) we derive Eq. (2.72).

$$N_1 = P_1 V/RT \tag{2.72}$$

At the time a reading (h) is taken (Fig. 2.18)

$$P_2 V = N_2 RT \tag{2.73}$$

where P_2 is the system pressure ($P_2 < P_1$); V the volume of gas phase in system (kept constant by adjustment with the screw clamp); and N_2 the g moles of gas phase in closed system at time of reading. For $N_2 < N_1$ due to oxygen absorption, then

$$N_2 = N_1 - x \tag{2.74}$$

where x is g moles of O_2 adsorbed.

From Eq. (2.73)

$$N_2 = P_2 V/RT \tag{2.75}$$

Equating Eqs. (2.74) and (2.75) and solving for x:

$$x = N_1 - (P_2 V/RT) \tag{2.76}$$

Substituting in Eq. (2.76) N_1 by its value given by Eq. (2.72):

$$x = (V/RT)(P_1 - P_2) \tag{2.77}$$

where $(P_1 - P_2)$ equals the height h of Brodie's fluid (Fig. 2.18). Therefore,

$$x = (V/RT)h \qquad (\text{g moles } O_2) \tag{2.78}$$

If V_s is the volume of the wastewater sample in ml, oxygen utilization in mg/liter is

$$\text{Oxygen utilization} = (Vh/RT)\text{ g moles } O_2 \times \frac{1}{(V_s \times 10^{-3})\text{ liter}} \times \frac{32\text{ g}}{\text{g mole } O_2} \times 10^3 \frac{\text{mg}}{\text{g}}$$

---

* Specific gravity of Brodie's fluid is 1.001 at 0°C (with respect to water at 4°C). Therefore, normal atmospheric pressure is equivalent to a column of Brodie's fluid (at 0°C) of height equal to

$$76.0\text{ cm Hg} \times 13.6\text{ cm water/cm Hg} \times \text{cm Brodie's fluid}/1.001\text{ cm water}$$
$$= (76.0 \times 13.6)/(1.001) = 1032.6\text{ cm Brodie's fluid at } 0°\text{C}$$

† $R = P_0 V_0/T_0 = \dfrac{(1032.6\text{ cm Brodie})(22{,}412\text{ ml/g mole})}{273.2°\text{K}}$

$$= 84{,}709\text{ (cm Brodie)(ml)/(g mole)(°K)}$$

or

$$\text{Oxygen utilization} = 32 \times 10^6 (V/V_s)(h/RT) \qquad \text{(mg/liter)} \qquad (2.79)$$

where V is the gas volume in closed system (ml); V_s the volume of wastewater sample (ml); h the reading (cm of Brodie's fluid) (Fig. 2.18); R the universal gas constant [84,709 (cm Brodie)(ml)/(g mole)(°K)]; and T the temperature of bath (°K).

9.3. BATCH REACTOR EVALUATION

A series of batch biological reactors (bench scale) are used to accomplish essentially the same objective as the Warburg respirometer. A battery of batch reactors recommended for this type of work is shown in Fig. 5.2 and described in Section 3.1 of Chapter 5.

An acclimated seed is added to the series of reactors. Various concentrations of a wastewater are then added to each reactor. The mixed contents are aerated for 2–3 days. The apparent toxicity is evaluated, samples are withdrawn at the end of 1, 2, 4, 8, 12, and 24 hr of aeration, and COD or BOD removal tests are performed. Typical BOD curves obtained in this manner are similar to the ones shown in Fig. 2.19.

10. Characteristics of Municipal Sewage

Municipal sewage is composed mainly of organic matter, either in soluble or colloidal form or as suspended solids. Eckenfelder [3] reports of analysis of data on municipal sewage for a survey which included 73 cities in 27 states of the United States. Some average per capita values from this survey are

Flow: 135 gal/(capita)(day)

BOD_5: 0.2 lb/(capita)(day) = 90.7 g/(capita)(day)

Suspended solids: 0.23 lb/(capita)(day) = 104 g/(capita)(day)

For a city of one million people, the following values are obtained by prorating this per capita data.

Flow: 135 Mgal/day × 8.34 lb/gal = 1126 Mlb/day

BOD_5: 200,000 lb/day

or in terms of mg/liter

BOD_5: 200,000 lb/day × day/1126 Mlb = 178 lb/Mlb = 178 ppm $\approx$ 178 mg/liter*

---

* Since most wastewaters contain small concentrations of soluble (and/or insoluble) matter, wastewater density is taken hence as equal to that for pure water, i.e., approximately 1 mg/liter. Consequently, mg/liter becomes essentially equivalent to parts per million (ppm), since 1.0 mg/liter $\approx$ 1.0 mg/10^3 g = 1.0 g/10^6 g = 1.0 ppm.

Presence of industrial wastes in a municipal sewage system may change these values considerably.

11. Industrial Wastewater Surveys

The procedure to be followed in industrial wastewater surveys has been described in Chapter 1 (Section 2.5.2, Step 1). A complete sewer map of the plant is developed. For accomplishing this objective, sampling and measuring stations are located in the plant, including all significant sources of wastewaters. Analyses to be run are selected and sampling and analyses schedules carefully planned. Material balances, including both process and sewer lines, are written.

Statistical plots for all significant characteristics are prepared. Whenever possible, these statistical plots are related to production, that is, gal/ton of product or lb BOD/ton of product. This permits extrapolation to other production schedules. Sources for wastewater segregation, reuse, and recirculation are identified.

Flow measurements of wastewater streams are performed by a variety of methods, which are summarized by Eckenfelder [3].

1. Installation of weirs for flow in open channels and partially filled sewers
2. Bucket and stopwatch method, suitable for low flow rates and/or intermittent discharges. In the latter case, flow rate and duration of operation are determined
3. Pumping duration and rate. Flow is estimated from the characteristic curves of the pump
4. Timing a floating object between two fixed points along the course. This method is applied to partially filled sewers. Depth of flow in the sewer is also measured. Average velocity is estimated from surface velocity, which is the one directly measured. For laminar flow the average velocity is approximately 0.8 times the surface velocity. Flow is then evaluated from the knowledge of this average velocity
5. Examination of plant water use records. Taking into account water losses in product or due to evaporation, this method leads to approximate estimates
6. Timing change of level in tanks or reactors, used primarily for batch operation discharges

12. Statistical Correlation of Industrial Waste Survey Data

Industrial wastewater discharges are highly variable in volume and composition and are appropriately treated by statistical analysis. Probability plots are used when dealing with statistics of events which fall into the bell-shaped probability curve so familiar to statisticians. A plot of data on probability-type graph paper straightens out the probability curve, leading to a straight line plot (linearization of the data). The straight line thus obtained is referred to as Henry's line. Therefore, if a series of experimental data is plotted on this paper and the result is a straight line, this indicates a random distribution of experimental data.

Probability graph paper utilized in this work is illustrated in Fig. 2.20. The abscissa is a probability scale and the ordinate is a logarithmic one.

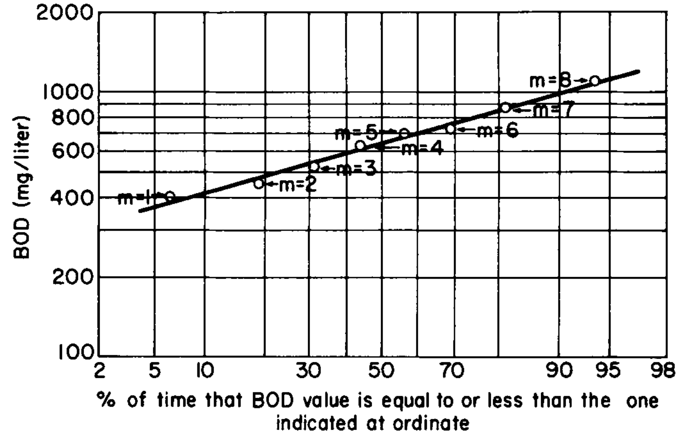

Fig. 2.20. Probability plot for Example 2.4 (Method 1).

Two methods for statistical correlation of industrial waste survey data are recommended by Eckenfelder [3].

Method (1) is recommended for small amounts of data (i.e., less than 20 datum points).

Step 1. Arrange data in increasing order of magnitude.

Step 2. Let n be the total number of points and m the assigned serial number from 1 to n. Tabulate data (in increasing order of magnitude) vs. m.

Step 3. Plotting positions (abscissas of the probability plot) are determined from

$$\text{Frequency} = (100/n)(m-0.5)$$

This quantity is equivalent to the percent occurrence of the value plotted in the ordinate, i.e., percent of time that the value in question is equal to or less than the reading of the ordinate. The position of the best-fit line is judged by eye or the least-squares method is used. Application of method (1) is illustrated by Example 2.4.

Example 2.4

The following BOD data (mg/liter) arranged in increasing order of magnitude was obtained for an industrial stream [column (*1*) of Table 2.8]. Plot Henry's line by the method described.

TABLE 2.8
Calculations for Example 2.4

(*1*) Step 1, BOD (mg/liter)	(*2*) Step 2 ($n = 8, m = 1, 2, \ldots, 8$), values of m	(*3*) Frequency = $(100/n)(m - 0.5)$, % time equal to or less than
400	1	6.25
450	2	18.75
520	3	31.25
630	4	43.75
700	5	56.25
730	6	68.75
860	7	81.25
1100	8	93.75

SOLUTION The procedure is indicated in Table 2.8 and Fig. 2.20. The probability of occurrence of any value is now estimated. For example, from Fig. 2.20 the BOD is equal to or less than 1000 mg/liter 90% of the time.

A statistical analysis of the various waste characteristics provides a basis for choice of design values. For example, the hydraulic capacity of a plant is selected in excess of the 99% frequency (here the ordinate is flow rate). On the other hand, sludge-handling facilities are usually designed on the basis of the 50% frequency.

Method (2) is employed when a large number of data (more than 20 datum points) have to be analyzed. Calculate the plotting position [column (*3*) of Table 2.8] from

$$\text{Frequency} = m/(n+1)$$

Otherwise, the procedure is the same as in method (1).

Problems

I. The following BOD data are given:

t (days)	BOD (mg/liter)
1	6.5
2	11.0
3	15.0
4	18.0
5	20.0
6	22.0
7	23.0
8	24.0
9	25.0
10	26.0

1. Plot the BOD curve.
2. Calculate the values of parameters k and L_0 by the following methods:
   (a) Log-difference method
   (b) Moore's method of moments, utilizing 3-, 5-, and 7-day sequences
   (c) Thomas' graphical method

Tabulate values obtained under (a), (b), and (c). Compare values of L_0 with that obtained by visual extrapolation of the curve.

II. For a wastewater, $k = 0.1$ (decimal log basis) and the 5-day BOD is 200 mg/liter. Estimate the 1-day BOD and the ultimate demand (L_0). What is the 5-day BOD if the incubation is at 30°C instead of the conventional temperature of 20°C?

III. Determine ThOD for alanine [$CH_3CH(NH_2)COOH$] (in g O_2/g mole of alanine) using the following assumptions:

1. Carbon atoms are oxidized to CO_2 while nitrogen is converted to ammonia.
2. Ammonia is then oxidized to HNO_2 in the presence of nitrite-forming bacteria.
3. Finally, HNO_2 is oxidized to HNO_3 in the presence of nitrate-forming bacteria.

ThOD is the sum of the oxygen required for these three steps.

IV. Nine determinations of suspended solids (ppm) in a waste stream yield the following results, arranged in order of increasing magnitude: 48, 83, 85, 102, 130, 134, 153, 167, and 180.

1. Linearize the distribution by a probability plot.
2. What is the probability of occurrence of a suspended solid (SS) value equal to or less than 200 ppm?

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3

Pretreatment and Primary Treatment

1. Introduction

Selection of a wastewater treatment process or sequence of processes depends on a number of factors, i.e., (1) characteristics of the wastewater, e.g., BOD, % of suspended solids, pH, presence of toxic materials; (2) required effluent quality; (3) cost and availability of land, e.g., certain biological processes (stabilization ponds) are only economically feasible if low cost land is available; and (4) consideration of a possible future upgrading of water quality standards, necessitating design of a more sophisticated type of treatment for future use.

Pretreatment of wastewater implies removal of suspended solids or conditioning of wastewater for discharge into either a receiving body of water or a secondary treatment facility through neutralization and/or equalization. Types of primary treatment discussed in this chapter are (1) screening, (2) sedimentation, (3) flotation, and (4) neutralization and equalization.

2. Screening

Screening is employed for removal of suspended solids of various sizes. Screen openings range in size depending on their purpose, and cleaning of screens is done either manually or mechanically. Screenings are disposed of by burial, incineration, or anaerobic digestion. Screens are classified as fine and coarse.

Fine screens have openings of 3/16 in. or smaller. They are usually made of steel mesh or perforated steel plates and sometimes used instead of sedimentation tanks. However, whereas they remove from 5 to 25% of suspended solids, 40–60% is removed by sedimentation. For this reason, and also because clogging is frequently a problem, use of fine screens is not very common.

Coarse screens have openings ranging from 1.5 to 3.0 in. They are used as protecting devices so that large suspended solids do not damage pumps and other equipment.

Sometimes shredders are used instead of coarse screens. These devices tear down suspended solids, which are then removed by sedimentation.

3. Sedimentation

3.1. INTRODUCTION

Sedimentation is utilized in wastewater treatment to separate suspended solids from wastewaters. Removal by sedimentation is based on the difference in specific gravity between solid particles and the bulk of the liquid, which results in settling of suspended solids.

In some cases, sedimentation is the only treatment to which the wastewater is subjected. Sedimentation is also utilized in one or more steps of a treatment sequence. In a typical activated sludge plant, sedimentation is utilized in three of the treatment steps: (1) in grit chambers, in which inorganic matter (e.g., sand) is removed from the wastewater; (2) in the primary clarifier, which precedes the biological reactor, solids (organic and others) are separated; and (3) in the secondary clarifier, which follows the biological reactor, the biological sludge is separated from the treated effluent.

3.2. TYPES OF SETTLING

Three types of settling are recognized depending on the nature of solids present in the suspension.

1. *Discrete settling.* Particles being settled keep their individuality, i.e., they do not coalesce with other particles. Thus, the physical properties of the particles (size, shape, specific gravity) are unchanged during the process. The settling of sand particles in grit chambers is a typical example of discrete settling.
2. *Flocculent settling.* Agglomeration of the settling particles is accompanied by changes in density and settling velocity. The sedimentation occurring in primary clarifiers is an example.
3. *Zone settling.* Particles form a lattice (or blanket) which settles as a mass exhibiting a distinct interface with the liquid phase. Examples include sedimentation of activated sludge in secondary clarifiers and that of alum flocs in water treatment processes.

3.3. THEORY OF DISCRETE SETTLING

The fundamental relationship for settling of discrete particles is Newton's law, which is based on the assumption that particles are spherical with a uniform diameter. When a particle settles, it accelerates until the forces promoting settling, i.e., the particle effective weight, are balanced by the drag or frictional resistance of the liquid. When this equality is achieved, the particle reaches a constant settling velocity called the terminal or settling velocity of the particle.*

Consider the particle in Fig. 3.1, which has reached its terminal velocity, and write the appropriate force balance. The force promoting sedimentation, i.e., the effective weight of the particle, is the difference between its actual weight and the hydrostatic lift:

$$F_S = v\rho_S g - v\rho_L g = (\rho_S - \rho_L)gv \tag{3.1}$$

where F_S is the particle effective weight; ρ_S the particle density; ρ_L the liquid density; g the acceleration of gravity; and v the particle volume, $\frac{1}{6}\pi d^3$, where d is the diameter of the spherical particle.

The drag force impeding sedimentation is

$$F_D = C_D A(\rho_L V^2/2) \tag{3.2}$$

where F_D is the drag force; C_D the drag coefficient; A the projected area of the particle, $A = \frac{1}{4}\pi d^2$; and V the relative velocity between particle and fluid.

For the condition defining the terminal velocity, equate Eqs. (3.1) and (3.2).

$$(\rho_S - \rho_L)gv = C_D A(\rho_L V_S^2/2)$$

where

$$V = V_S = \text{settling velocity}$$

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* This results from force = (mass) (acceleration). Thus zero acceleration corresponds to a net force of zero, i.e., a perfect balance of forces.

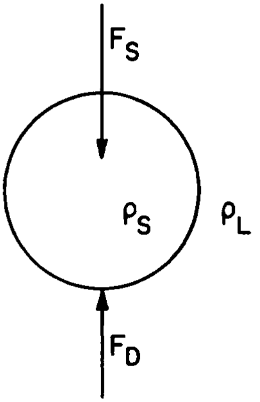

Fig. 3.1. Discrete settling of a particle.

Substituting $v = \frac{1}{6}\pi d^3$, $A = \frac{1}{4}\pi d^2$, and solving for the terminal velocity, V_S [Eq. (3.3)]:

$$V_S = \left[\tfrac{4}{3}(g/C_D)\frac{\rho_S - \rho_L}{\rho_L} \cdot d\right]^{1/2} \tag{3.3}$$

which is Newton's law.

For spherical particles, the drag coefficient C_D is related to the Reynolds number N_R defined in Eq. (3.4).

$$N_R = dV_S\rho_L/\mu_L \tag{3.4}$$

where d is the diameter of sphere, V_S the terminal velocity (settling velocity), and ρ_L and μ_L the density and viscosity of liquid. This relationship is shown in Fig. 3.2.*

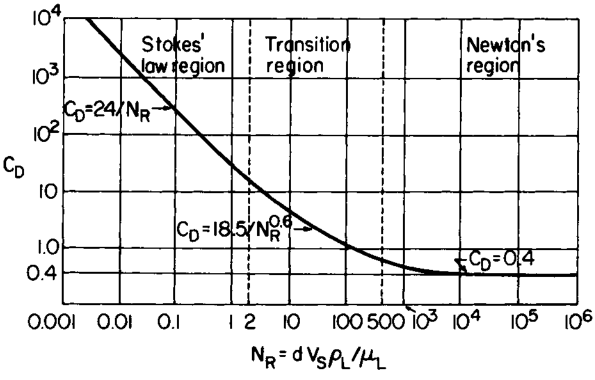

Fig. 3.2. Correlation for drag coefficient for spherical particles.

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* For nonspherical particles Fig. 3.2 is plotted as a family of curves, each curve corresponding to specified value of a parameter defined as sphericity [sphericity, ψ = (surface area of a sphere having same volume as particle)/(surface area of particle)]. See Waddel [8].

In general, the drag coefficient C_D is approximated by

$$C_D = b/N_R^n \tag{3.5}$$

where coefficients b and n for the different regions of Fig. 3.2 are indicated in Table 3.1. The approximate relationship between C_D and N_R given by Eq. (3.5) is written in logarithmic form for the three regions indicated in Fig. 3.2.

TABLE 3.1
Drag Coefficient C_D

Region	b	n	$C_D = b/N_R^n$
Stokes' law			
$N_R < 2$	24	1.0	$C_D = 24/N_R$
Transition			
$2 < N_R < 500$	18.5	0.6	$C_D = 18.5/N_R^{0.6}$
Newton's			
$N_R > 500$	0.4	0.0	$C_D = 0.4$

Stokes' region: $C_D = 24/N_R$

$$\therefore \log C_D = -\log N_R + \log 24$$

Transition region: $C_D = 18.5/N_R^{0.6}$

$$\therefore \log C_D = -0.6 \log N_R + \log 18.5$$

Newton's region: $C_D = 0.4$

$$\therefore \log C_D = \log 0.4 = 0.0 \log N_R + \log 0.4$$

Thus the curve in Fig. 3.2 is approximately replaced by three segments of straight line with slopes of, respectively, -1, -0.6, and 0.0.

Many sedimentation problems in wastewater treatment occur in Stokes' region. Substituting $C_D = 24/N_R = 24\mu_L/dV_S\rho_L$ in Eq. (3.3) and simplifying, Stokes' law is obtained.

$$V_S = 1/18 \cdot \frac{\rho_S - \rho_L}{\mu_L} \cdot gd^2 \tag{3.6}$$

For a specific problem in Stokes' region (ρ_S, ρ_L, and μ_L fixed) Eq. (3.6) is written as [Eq. (3.7)]

$$V_S = Kd^2 \tag{3.7}$$

(where K is a constant), which in logarithmic form becomes [Eq. (3.8)]

$$\log V_S = 2 \log d + \log K = 2 \log d + C \qquad (C \text{ is a constant}) \tag{3.8}$$

Thus, a logarithmic plot of V_S vs. d yields a straight line of slope equal to 2.0 for Stokes' region.

For a specific problem in Newton's region, since C_D is a constant ($C_D = 0.4$), Eq. (3.3) yields Eq. (3.9).

$$V_S = K'd^{1/2} \tag{3.9}$$

(where K' = a constant) which in logarithmic form becomes Eq. (3.10).

$$\log V_S = \tfrac{1}{2}\log d + \log K' = \tfrac{1}{2}\log d + C' \tag{3.10}$$

where C' is a constant. Thus, a logarithmic plot of V_S vs. d yields a straight line of slope equal to $\frac{1}{2}$ for Newton's region.

For the transition region, $C_D = 18.5/N_R^{0.6} = (18.5\mu_L^{0.6})/(d^{0.6}V_S^{0.6}\rho_L^{0.6})$. Substituting this value in Eq. (3.3) and simplifying, one obtains Eq. (3.11).

$$V_S = [(4g/55.5)(\rho_L{}^{0.6}/\mu_L{}^{0.6})(\rho_S-\rho_L)/\rho_L]^{1/4}d^{1.143} = K''d^{1.143} \tag{3.11}$$

where K'' is a constant. In logarithmic form this becomes [Eq. (3.12)]

$$\log V_S = 1.143\log d + \log K'' = 1.143\log d + C'' \tag{3.12}$$

where C'' is a constant. Thus, a logarithmic plot of V_S vs. d yields a straight line of slope 1.143 for the transition region. The logarithmic plot of V_S vs. d for the three regions is shown in Fig. 3.3. Even in the case of grit chambers, the theory just outlined suffers from two serious limitations: (1) grit particles are seldom spherical, and (2) grit particles do not have uniform density.

A graph corresponding to Eq. (3.3) is plotted in Fig. 3.4, giving the relationship between particle diameter and velocity V_S. Particles of specific gravities 1.001, 1.01, and 2.65 are considered in plotting Fig. 3.4. Value 2.65 corresponds to the specific gravity of typical sand. The liquid used is water at temperatures indicated, corresponding to respective values of ρ_L, μ_L. Values of C_D are obtained from Fig. 3.2 by a trial and error procedure: (1) for specified particle

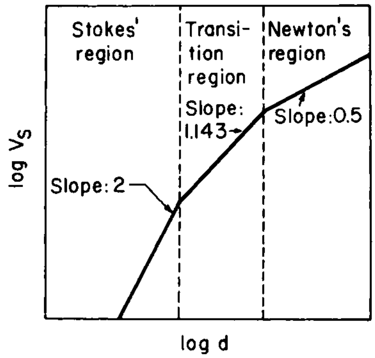


Fig. 3.3. Logarithmic plot of V_S vs. d.

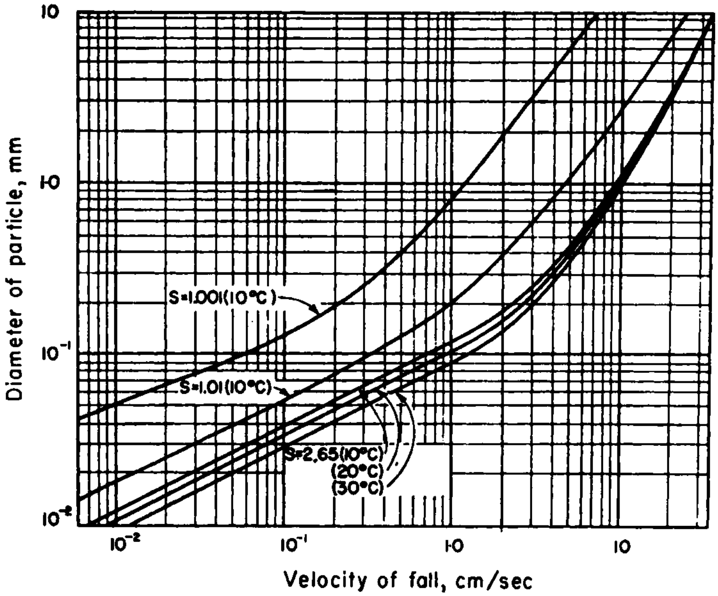


Fig. 3.4. Relation between settling velocity and particle diameter [4].

diameter and temperature (ρ_L and μ_L fixed) assume a settling velocity V_S; (2) Calculate N_R based upon this assumed velocity; (3) From Fig. 3.2 read C_D; and (4) From Eq. 3.3 recalculate V_S. If it agrees with the value assumed in (1) calculations are consistent. Otherwise, iteration is continued until agreement is obtained.

Figure 3.4 is constructed in this manner. Since construction is based upon the actual curve of C_D vs. N_R (i.e., Fig. 3.2), the lines in Fig. 3.4 exhibit some curvature, by contrast with the three straight line segments in Fig. 3.3 [construction of which is based on the approximate relationships given by Eqs. (3.8), (3.10), and (3.12)]. As an approximation, however, curves in Fig. 3.4 are replaceable by three straight line segments.

3.4. THE IDEAL SEDIMENTATION TANK CONCEPT

This concept, developed by Hazen [5] and Camp [1], is the basis for arriving at relationships utilized in the design of sedimentation tanks. The model chosen for a sedimentation tank consists of four zones (Figs. 3.5 and 3.6).

1. Inlet zone. Here the flow becomes quiescent. It is assumed that at the limit of this zone (i.e., along vertical line *xt*) particles are uniformly distributed across the influent cross section.

2. Sedimentation zone. A particle is assumed to be removed from suspension once it hits the bottom of this zone (horizontal line *ty*).

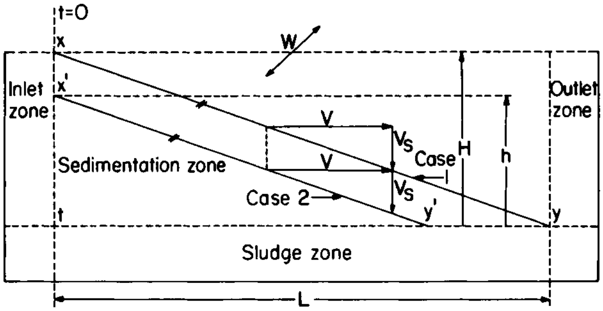


Fig. 3.5. Model of a sedimentation tank with discrete settling particles (Cases 1 and 2).

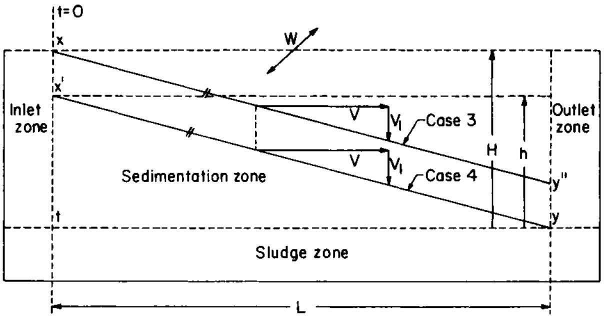


Fig. 3.6. Model of a sedimentation tank with discrete settling particles (Cases 3 and 4).

3. Outlet zone. Wastewater is collected here prior to transfer to the next treatment.

4. Sludge zone. This zone is provided for sludge removal.

Settling paths of particles entering the sedimentation zone at points x and x' for discrete settling are indicated by lines xy and $x'y'$ in Fig. 3.5 and lines xy'' and $x'y$ in Fig. 3.6. These settling paths are the net result of two velocity vector components:

1. Flow-through velocity V [Eq. (3.13)]

$$V = Q/A' = Q/WH \tag{3.13}$$

where V is the flow-through velocity (ft/sec); Q the flow rate (ft^3/sec); A' the vertical cross-sectional area of sedimentation zone, namely $A' = WH$ (ft^2) (refer to Fig. 3.7); W the width of sedimentation zone (ft); and H the depth of sedimentation zone (ft).

2. Settling velocity, indicated by either vectors V_S or V_l in Figs. 3.5 and 3.6, respectively.

For discrete settling the settling velocity is constant for any specific settling

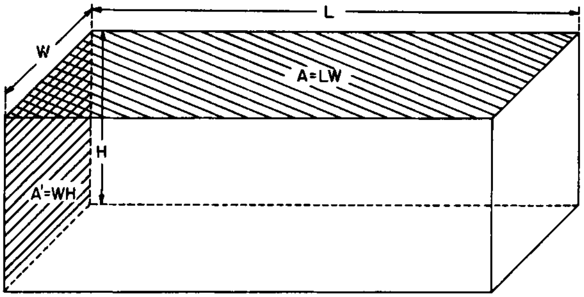


Fig. 3.7. Geometry of the sedimentation zone.

path, i.e., V_S and V_1 do not vary along their respective paths. This is due to the fact that a discrete particle is unhindered by neighboring ones (no coalescence), so it settles with a uniform velocity, read from Fig. 3.4, as a function of the particle diameter.

For flocculent settling the situation is different. Figure 3.8 illustrates a typical sedimentation path of flocculent settling (Section 3.5). As coalescence with neighboring particles takes place, the effective diameter of the particle increases, and thus its settling velocity V_S also increases. The net result is that settling paths are curved in contrast with straight line paths for discrete settling.

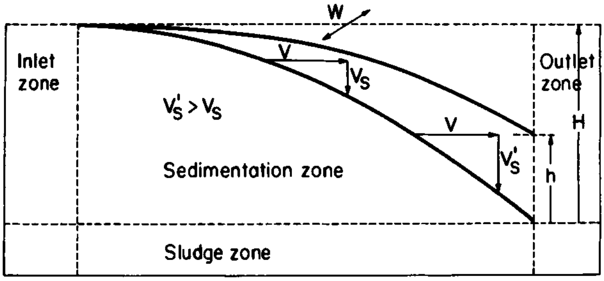


Fig. 3.8. Model of a sedimentation tank for flocculent settling.

Consider the following cases with reference to Figs. 3.5 and 3.6, keeping in mind that a particle is assumed to be removed from the suspension once it hits the bottom of the sedimentation zone.

Case 1 (Fig. 3.5). A particle which at time zero ($t = 0$) is located at point x and possesses a settling velocity V_S (and diameter d_S read from Fig. 3.4). This particle is removed, since it touches the bottom of the sedimentation zone at y (path xy).

Case 2 (Fig. 3.5). A particle which at $t = 0$ is located at x' on the same vertical line as x but *below* x, and has a settling velocity V_S (or greater than V_S). This particle is also removed since it hits the bottom of the sedimentation zone to the left of point y (at y'). If the settling velocity is greater than V_S the particle touches the bottom of the sedimentation zone to the left of y'. Figure 3.6 portrays the case of particles with a diameter d_1 ($d_1 < d_S$), possessing a settling velocity V_1 ($V_1 < V_S$). (Settling velocity V_1 is read from Fig. 3.4 for $d = d_1$.) These particles are shown separately in Fig. 3.6 simply to avoid overcrowding of the diagram. In fact they are together in the slurry with particles of settling velocity V_S.

Case 3 (Fig. 3.6). A particle which at $t = 0$ is located at x and has a settling velocity V_1 (where $V_1 < V_S$). This particle is not removed since it does not reach the bottom of the sedimentation zone (i.e., line ty) in time for removal (sedimentation path xy'').

Case 4 (Fig. 3.6). A particle with settling velocity V_1 (where $V_1 < V_S$) and situated at x' at $t = 0$. This particle is removed (sedimentation path $x'y$). Consider now the settling velocity V_S. From Eq. (3.13)

$$Q = VA' = VWH \tag{3.14}$$

From consideration of similar triangles in Fig. 3.5

$$V = V_S(L/H) \tag{3.15}$$

Substitution of Eq. (3.15) in Eq. (3.14) and simplification leads to

$$Q = V_S LW = V_S A \tag{3.16}$$

where $A = LW =$ horizontal cross-sectional area of the sedimentation zone (ft^2) (Fig. 3.7).

From Eq. (3.16)

$$V_S = Q/LW = Q/A \tag{3.17}$$

From Eq. (3.17) it follows that the settling efficiency is a function of the horizontal cross-sectional area, rather than of the depth H. Thus, in principle, it is advisable to utilize sedimentation tanks of high surface area A and low depths. The only reasons for using a reasonable depth are (1) to satisfy depth requirements in order to provide for mechanical rakes utilized for removal of settled sludge, and (2) the horizontal component of velocity (flow-through velocity V) must be kept within certain limits to prevent scouring the particles which have settled.

From Eq. (3.13) it follows that H should not be too low since V would rise above the scour velocity. The subject of scour velocity is discussed in this section. Scour occurs when flow-through velocity V is sufficient to suspend previously settled particles. Scour is not usually a problem in large settling tanks, but it can be an important factor in grit chambers and narrow channels.

The two fundamental premises of the ideal tank concept are (1) uniform distribution of particles across the influent cross section (i.e., along vertical line xt) and (2) assumption that a particle is considered removed when it reaches the bottom of the sedimentation zone (i.e., horizontal line ty).

Keeping in mind these two premises, two corollaries follow: (1) All particles with a settling velocity equal to or greater than V_S are removed; and (2) all particles with a settling velocity less than V_S (such as V_1 in Fig. 3.6) are removed in a proportion given by the ratio V_1/V_S.

From geometric consideration of the particle path (Figs. 3.5 and 3.6) touching point y at the bottom of the sedimentation tank (particles with settling velocity V_S in Fig. 3.5 following path xy, and particles with settling velocity V_1 in Fig. 3.6 following path $x'y$), one writes [Eq. (3.18)]

$$V_1/V_S = h/H \tag{3.18}$$

For example, if $H = 100$ in. and $h = 75$ in., then

$$V_1/V_S = 75/100 = 0.75$$

or 75% of the particles with a settling velocity V_1 are removed, i.e., those which at time $t = 0$ are at a height x' or below. The remaining particles with settling velocity V_1, i.e., those situated between x and x' at time $t = 0$, are *not* removed.

The overflow rate defined as

$$Q/A = \mathrm{ft^3/(ft^2)(hr) = ft/hr}$$

is defined as the settling velocity V_S of a particle that settles through a distance exactly equal to the effective depth of the tank during the theoretical detention period. This results from the definition of detention period:

$$t = \text{detention period} = (\text{volume of tank})/Q = HA/Q \qquad (A = LW) \tag{3.19}$$

From Eq. (3.19) it follows that a settling velocity V_S defined as $V_S = H/t$ is equivalent to the overflow rate, since

$$V_S = H/t = H/(HA/Q) = Q/A$$

which is Eq. (3.17).

The *scour velocity* V_C is the value of the flow-through velocity V [Eq. (3.20)],

$$V = V_C = Q/A' = Q/WH \tag{3.20}$$

for which "previously settled" particles are scoured away.*

---

* The words "previously settled" are placed in quotes because a particle which is scoured away never actually settles. Mentally one separates the processes of settling and scouring and imagines that a particle settles and subsequently is scoured away. This reasoning is compatible with the hypothetical resolution of the velocity trajectory into vectors V (flow-through velocity) and V_S (settling velocity).

The scour velocity is estimated by the following empirical equation [Eq. (3.21)] [1]:

$$V_C = [8\beta g d(s-1)/f]^{1/2} \tag{3.21}$$

where V_C is the velocity of scour (mm/sec); i.e., flow-through velocity required to scour all particles of diameter d or smaller; β the constant (0.04 for unigranular sand, 0.06 for nonuniform sticky material); f the Weisbach–D'Arcy friction factor (0.03 for concrete); g the acceleration of gravity (mm/sec^2) (normal: 9800 mm/sec^2); d the particle diameter (mm) (particles with diameter d or less than d are scoured away); and s the specific gravity of particle.

Example 3.1

Consider a suspension of sand ($s = 2.65$) in water at 20°C with a uniform particle size ($d = 0.07$ mm). Flow is 1.0 Mgal/day.

1. Calculate the settling tank surface (horizontal cross section) for obtaining removal of 70% of the particles.

2. Suppose that instead of a uniform particle diameter, there is, besides particles of $d = 0.07$ mm, another set with a uniformly larger diameter, which are completely removed in the settling tank designed for 70% removal of the particles with $d = 0.07$ mm. Determine what is the minimum particle diameter for total removal.

3. For case (2), determine the flow-though velocity V_C so that all particles of lower settling velocity than those completely removed are scoured away. What combination of length, width, and depth for the settling tank meets these requirements?

SOLUTION: Part 1

Step 1. From Fig. 3.4 [for $d = 0.07$ mm and $s = 2.65$ (at 20°C)], read

$$V_S = 0.45 \text{ cm/sec}$$

or

$$V_S = 0.45 \text{ cm/sec} \times \text{ft}/30.48 \text{ cm} \times 3600 \text{ sec/hr}$$

$$V_S = 53.1 \text{ ft/hr} = 53.1 \text{ ft}^3/(\text{ft}^2)(\text{hr})$$

The overflow rate in gal/(day)(ft^2) is

$$53.1 \text{ ft}^3/(\text{ft}^2)(\text{hr}) \times 7.48 \text{ gal/ft}^3 \times 24 \text{ hr/day}$$

$$\therefore V_S = 9533 \text{ gal/(day)(ft}^2)$$

Step 2. The horizontal cross-sectional area is (for 100% removal)

$$A = Q/V_S = \frac{1{,}000{,}000 \text{ gal/day}}{9533 \text{ gal/(day)(ft}^2)} = 105 \text{ ft}^2$$

With this area, 100% removal is obtained. For 70% removal the residence time (and thus the cross-sectional area) is reduced by 30%. The cross-sectional area is

$$A = 105 \times 0.7 = 73.5\ \text{ft}^2$$

Under these circumstances, the settling velocity for 100% removal is

$$1{,}000{,}000/73.5 = 13{,}605\ \text{gal/(day)(ft}^2\text{)}$$

Note: Check on percent removal. Since the settling velocity is 9533 gal/(day)(ft^2), fixed by the particle diameter as determined from Fig. 3.4, percentage removal is $9533/13{,}605 = 0.70$ (70%). Referring to Fig. 3.6 the particles removed are those which at the end of inlet zone, along vertical line *xt*, are already at distance *h* (or less than *h*) from the bottom of the sedimentation zone, where [from Eq. (3.18)]

$$V_S = 13{,}605\ \text{gal/(day)(ft}^2\text{)} \quad (100\%\ \text{removal})$$

$$V_1 = 9533\ \text{gal/(day)(ft}^2\text{)} \quad (70\%\ \text{removal})$$

$$\therefore\ h/H = 0.70$$

SOLUTION: Part 2 If there is a distribution of particle diameters instead of uniform diameter $d = 0.07$ mm, one reads from Fig. 3.4 the diameter, which is larger than 0.07 mm, for which the settling velocity corresponds to 13,605 gal/(day)(ft^2). Therefore, the abscissa in Fig. 3.4 is

$$(13{,}605/9533) \times 0.45\ \text{cm/sec} = 0.45/0.7 = 0.642\ \text{cm/sec}$$

From Fig. 3.4 [for $V_S = 0.642$ cm/sec and $s = 2.65$ ($t = 20°\text{C}$)], read $d = 0.085$ mm (100% removal). If distribution of particle diameter in the influent is known, one can calculate the % removal corresponding to each group of particles for a given diameter (Example 3.2).

SOLUTION: Part 3 The scour velocity to sweep all particles of lower settling velocity than those to be completely removed is calculated from Eq. (3.21).

$$V_C = [8 \times 0.04 \times 9800 \times 0.07(2.65 - 1)/0.03]^{1/2} = 110\ \text{mm/sec}$$

Assuming that the sand contains only two particle sizes, e.g., 0.07 and 0.085 mm, the scour velocity $V_C = 110.0$ mm/sec sweeps away all particles of $d = 0.07$ mm, leaving behind those of $d = 0.085$ mm. The value of V_C in practical units is

$$V_C = 110\ \text{mm/sec} \times \text{ft}/304.8\ \text{mm} = 0.36\ \text{ft/sec}$$

The vertical cross-sectional area is calculated from Eq. (3.13).

$$A' = \frac{1{,}000{,}000\ \text{gal/day} \times \text{day}/86{,}400\ \text{sec} \times \text{ft}^3/7.48\ \text{gal}}{0.36\ \text{ft/sec}} = 4.3\ \text{ft}^2$$

Any practical combination of length, width, and depth is used to satisfy the requirements $A = 73.5\ \text{ft}^2 = LW$ and $A' = 4.3\ \text{ft}^2 = WH$.

Example 3.2

Suppose that for Example 3.1 instead of a uniform particle size, there is a distribution of diameters. Assume the same specific gravity and temperature as in Example 3.1, i.e., $s = 2.65$ and $t = 20°\text{C}$. Assume that for each 100 lb of grit the following distribution of particle sizes applies (see tabulation below)

(*1*) Group no.	(*2*) lb of each particle size	(*3*) Particle size, *d* (mm)
1	50	0.085
2	20	0.070
3	20	0.060
4	10	0.050
	100	

The settling velocities for each group of particles are read from Fig. 3.4. This is indicated in column (*4*) of Table 3.2. Percent removals are then calculated [column (*5*)] and expressed as fractions of unity in column (*6*). The weight of sand removed by settling for each group of particles is computed in column (*7*). Therefore 78.3% of the weight of the original particles is removed by settling.

TABLE 3.2
Calculations for Example 3.2

(*1*) Group no.	(*2*) lb of each particle size	(*3*) Particle size, *d* (mm)	(*4*) V_S (Fig. 3.4)	(*5*) $\% \text{ removal} = \frac{(4)}{0.642} \times 100$	(*6*) Fraction removed = (*5*) ÷ 100	(*7*) lb removed (*7*) = (*2*) × (*6*)
1	50	0.085	0.642	(0.642/0.642) 100 = 100[a]	1.00	50.0
2	20	0.070	0.450	(0.45/0.642) 100 = 70[a]	0.70	14.0
3	20	0.060	0.350	(0.35/0.642)100 = 54.5	0.545	10.9
4	10	0.050	0.220	(0.22/0.642)100 = 34.3	0.343	3.43
	100					78.3

[a] Already calculated in Example 3.1.

In this example, if the vertical cross section A' is taken equal to 4.3 ft^2 (value calculated in Example 3.1), all particles of $d = 0.07$ mm and smaller are scoured away (groups 2, 3, and 4). Therefore, the net removal is of only 50 lb per 100 lb of total grit, i.e., the particles with $d = 0.085$ (group 1). This indicates a net removal by weight of 50%.

If A' is taken larger than 4.3 ft^2 the net removal is greater, since there is less scouring. Examining calculations for Example 3.1, it follows that if A' is taken as 8.6 ft^2 (twice 4.3 ft^2) the value of V_C is 0.18 ft/sec (instead of 0.36 ft/sec). This corresponds to

$$V_C = 55 \text{ mm/sec} \qquad \text{(instead of 110.0 mm/sec)}$$

Since from Eq. (3.21), V_C is proportional to $d^{1/2}$, it follows that d is $0.07/4 = 0.0175$ mm, so as to yield $V_C = 55$ mm/sec (half of 110.0 mm/sec). Therefore, only particles with $d = 0.0175$ mm or smaller are removed by scouring. Since for the given distribution the smallest particle diameter is 0.05 mm, there is no removal by scouring. Consequently, the net removal is 78.3 lb from every 100 lb of sand, or 78.3% by weight.

If this value is adopted,

$$A = 73.5 \text{ ft}^2 = LW$$

$$A' = 8.6 \text{ ft}^2 = WH$$

Selecting $H = 4$ ft, then

$$W = 8.6/4 = 2.15 \text{ ft}$$

$$L = 73.5/2.15 = 34.2 \text{ ft}$$

This indicates specification of a narrow settling channel 34.2 ft long, 2.15 ft wide, and 4 ft deep.

3.5. FLOCCULENT SETTLING

Flocculent settling takes place when settling velocity of the particles increases due to coalescence with other particles. A diagram of flocculent sedimentation profiles is shown in Fig. 3.8. The settling paths of the particles are curves, rather than straight lines as for discrete settling.

Design criteria for systems exhibiting flocculent settling are established by a laboratory settling analysis. A typical laboratory settling column is shown in Fig. 3.9.

Concentration of suspended solids is kept uniform throughout the column at the beginning of the test by means of a portable stirrer. The depth of the column is approximately the same as that of the settling tank to be designed. Temperature is kept constant during the test. A practical design of a settling column (Fig. 3.9) is 8 ft deep, with sampling ports at depths of 2, 4, 6, and 8 ft.

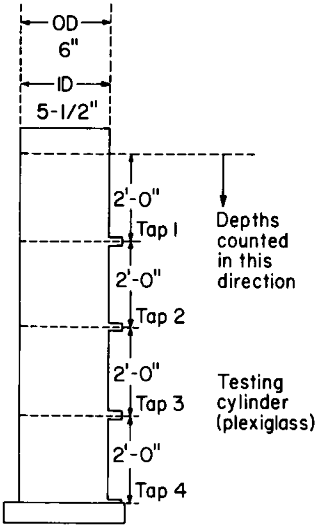


Fig. 3.9. Laboratory settling column.

Data taken at 2-, 4-, and 6-ft depths are utilized to determine settling velocity and detention time relationships. Data from the 8-ft port are used for sludge concentration and compaction determinations.

Step 1. Fill the column with wastewater, maintaining a uniform concentration of suspended solids throughout. A portable stirrer is used for this purpose.

Step 2. Remove the stirrer from the cylinder. At each sampling port, the concentrations of suspended solids are measured at predetermined times.

Examples 3.3 and 3.4 illustrate the procedure from data treatment to the design of a clarifier.

Example 3.3

The suspension being tested has an initial suspended solids concentration of 430 mg/liter (SS_0). The suspended solids (hence abbreviated as SS) concentrations in Table 3.3 are measured at the indicated times at the 2-, 4-, and 6-ft sampling ports. Perform treatment of the data arriving at curves for (a) % SS removal vs. detention time (min), (b) % SS removal vs. overflow rate [gal/(day)(ft^2)], and (c) % SS remaining (fraction of particles with less than stated velocity) vs. settling velocity (ft/hr).

Step 1. Calculate fraction of solids remaining in suspension for each sample [Eq. (3.22)].

$$x = SS/SS_0$$

or in %

$$y = SS/SS_0 \times 100 \tag{3.22}$$

TABLE 3.3
Laboratory Sedimentation Data (Example 3.3)

Time (min)	SS concentrations at indicated depths 2 ft (Tap 1)	4 ft (Tap 2)	6 ft (Tap 3)
5	356.9	387.0	395.6
10	309.6	346.2	365.5
20	251.6	298.9	316.1
30	197.8	253.7	288.1
40	163.4	230.1	251.6
50	144.1	195.7	232.2
60	116.1	178.5	204.3
75	107.5	143.2	180.6

Then calculate for each sample the fraction of solids removed

$$1 - x$$

or in %

$$z = 100 - y \tag{3.23}$$

A sample of these calculations (for a 2-ft depth) is shown in Table 3.4. Similar calculations are performed for 4- and 6-ft depths.

TABLE 3.4
Calculation of Fraction of Solids Remaining and Removed for a 2-ft Depth

(*1*) Time (min)	(*2*) SS remaining (mg/liter) (Table 3.3)	(*3*) Solids remaining (%) $y = (\mathrm{SS}/\mathrm{SS}_0) \times 100$	(*4*) Solids removed (%) $z = 100 - y$
5	356.9	83.0	17.0
10	309.6	72.0	28.0
20	251.6	58.5	41.5
30	197.8	46.0	54.0
40	163.4	38.0	62.0
50	144.1	33.5	66.5
60	116.1	27.0	73.0
75	107.5	25.0	75.0

Step 2. In order to smooth the experimental data construct a graph of % SS removed vs. time. This plot is shown in Fig. 3.10 for the 2-, 4-, and 6-ft depths.

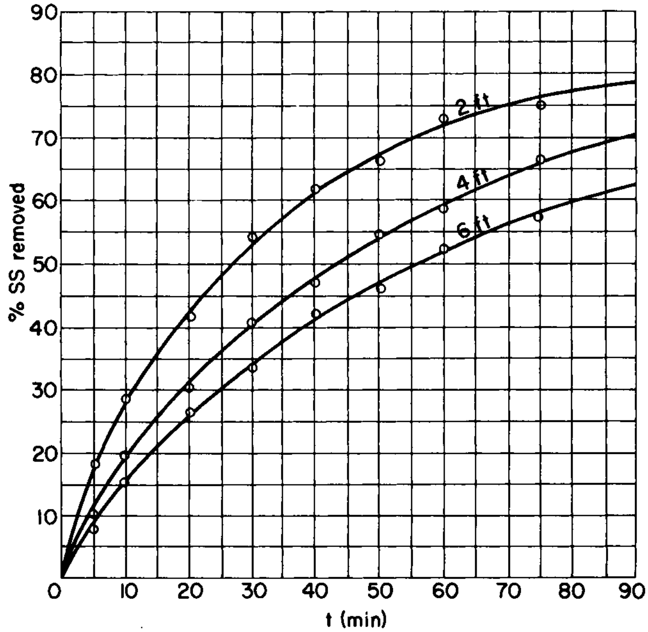


Fig. 3.10. Suspended solids (% SS) removed vs. time.

Step 3. From Fig. 3.10 construct a settling profile graph (Fig. 3.11). This is done by reading from the smoothed curves of Fig. 3.10 the abscissas (t, min) corresponding to selected values of % SS removed (e.g., 5, 10, 20, ..., 70, 75%) for each one of the three depths. These values are tabulated (Table 3.5) and utilized for construction of Fig. 3.11.

Step 4. Calculate % removal of SS and overflow rate [gal/(day)(ft^2)]. Before the procedure described here is fully understood, some preliminary considerations must be made. An effective settling velocity V_S is defined as the effective depth (6 ft in this example) divided by the time (detention time, t) required for a given particle to travel this distance, i.e. [Eq. (3.24)],

$$V_S = H/t \tag{3.24}$$

If a suspension contains particles with different settling velocities, the efficiency of removal by sedimentation is obtained by performing a settling column test as just described. Let SS be the concentration of solids remaining for one specific sample and time. Thus

$$x_0 = \mathrm{SS}/\mathrm{SS}_0 = \text{fraction of solids remaining}$$

and

$$1 - x_0 = 1 - \mathrm{SS}/\mathrm{SS}_0 = y_0 = \text{fraction of solids removed}$$

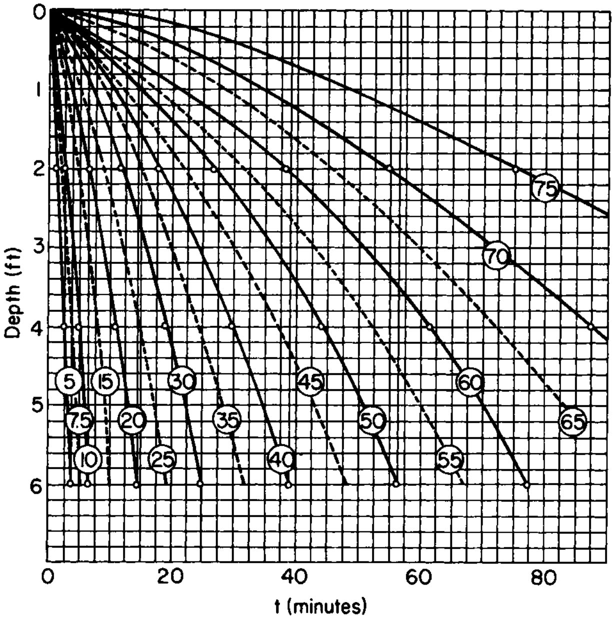


Fig. 3.11. Settling profile. Encircled numbers are % SS removed.

TABLE 3.5
Values for Plotting Fig. 3.11

	t (min)		
% SS removed	2 ft	4 ft	6 ft
5	1.2	2.5	3.7
10	2.5	5.0	6.5
20	6.7	11.0	14.5
30	11.7	19.0	25.0
40	18.0	30.0	39.0
50	27.0	44.0	56.5
60	38.5	61.5	77.5
70	55.0	87.5	—
75	75.0	—	—

Particles with a settling velocity V_S or higher (where $V_S = H/t$) are completely removed. Particles with a lower settling velocity V_1 ($V_1 < V_S$) are removed at a ratio given by Eq. (3.18).

A typical graph like the one in Fig. 3.12 is plotted by analysis of data obtained with the sedimentation column. The details for construction of such

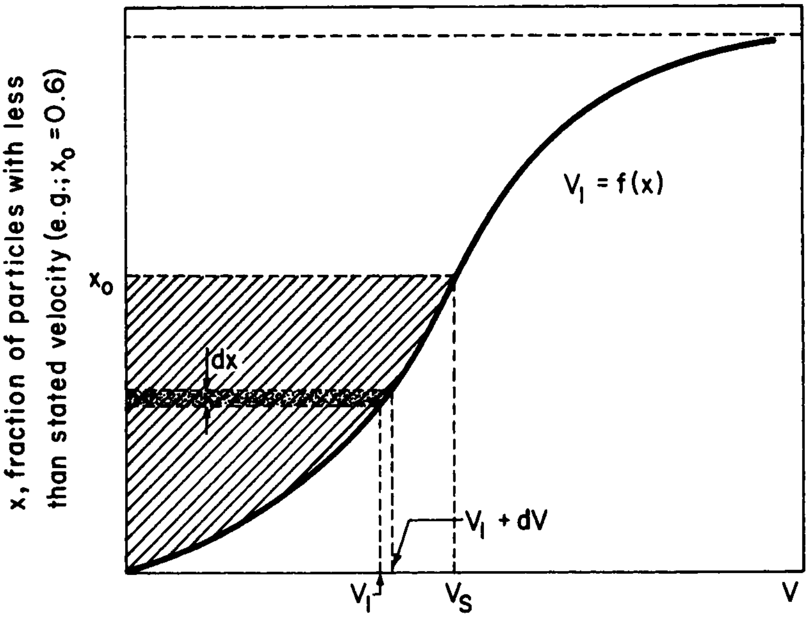


Fig. 3.12. Determination of overall removal.

a graph from experimental data are discussed later in Step 4(d). However, for understanding the calculation procedure described next, it is convenient to assume that this graph is already available.

Figure 3.12 is a plot of the fraction of particles with *less* than the stated velocity vs. the settling velocity in question. Notice that the fraction of particles with less than the stated velocity (if $V_S = H/t$) corresponds to the fraction of particles *not* completely removed. For example, if 40% of the particles in a specific case are completely removed, then $x_0 = 0.6$ is the ordinate corresponding to the settling velocity $V_S = H/t$.

For particles with settling velocities between V_1 and $V_1 + dV$ (where $V_1 < V_S$), the fraction removed is V_1/V_S. Similarly, for particles with settling velocities between $V_1 + dV$ and $V_1 + 2\,dV$, the fraction removed is $(V_1 + dV)/V_S$.

The overall removal of suspended solids is

$$\text{Overall removal} = (1 - x_0) + \int_0^{x_0} (V_1/V_S)\,dx \qquad (3.25)$$

In Eq. (3.25) V_1 is a variable ($0 \leqslant V_1 \leqslant V_S$) with $V_1 = f(x)$ portrayed by the curve in Fig. 3.12. Term $(1 - x_0)$ is the fraction completely removed, corresponding to particles with velocities $\geqslant V_S$. The second term in Eq. (3.25), i.e.,

$$\int_0^{x_0} (V_1/V_S)\,dx = 1/V_S \int_0^{x_0} V_1\,dx$$

which is the fraction of removal corresponding to particles with velocities less than V_S (calculated by graphical integration as indicated by the hatched area in Fig. 3.12). The differential area of width dx, indicated in Fig. 3.12, corresponds to particles with settling velocities between V_1 and V_1+dV.

Combining Eqs. (3.18) and (3.25) the final expression for the overall removal is obtained [Eq. (3.26)].

$$\text{Overall removal} = (1-x_0) + \int_0^{x_0} (V_1/V_S)\,dx = (1-x_0) + (1/V_S)\int_0^{x_0} V_1\,dx$$

$$= (1-x_0) + \int_0^{x_0} (h/H)\,dx \tag{3.26}$$

After these considerations return to discussion of Step 4.

Step 4(a). From Fig. 3.11 for a depth of 6 ft read the values t (min) corresponding to 5, 10, 20, 30, 40, 50, and 60% removal, and calculate the corresponding settling velocities V_S (ft/hr). These values are tabulated in Table 3.6.

TABLE 3.6
Settling Velocities ($H = 6$ ft)

Constant % removal	t (min) $H = 6$ ft	Settling velocity (ft/hr) $V_S = H/t = 6/(t/60) = 360/t$
5	3.7	97.2
10	6.5	55.2
20	14.5	24.8
30	25.0	14.4
40	39.0	9.2
50	56.5	6.35
60	77.5	4.64

Step 4(b). Calculate % removal of SS. Calculations for % removal of SS and overflow rate for a 25-min settling time (fourth entries in Table 3.6) are illustrated next. Similar calculations are also performed for the other settling times listed in the second column of Table 3.6. For $t = 25$ min for the settling depth $H = 6$ ft, 30% of the suspended solids are completely removed. Consider next the particles in each additional 10% range. Start with those in the range 30–40% removal in Fig. 3.11. Particles in this range are removed in the proportion V_1/V_S or in the proportion of average settled depth (h_1) to the total settling depth (H). The average settled depth (h_1) is estimated by drawing (by interpolation) a curve corresponding to 35% constant removal in Fig. 3.11, and reading from it the depth h_1 corresponding to $t = 25$ min. Therefore for this first interval, the % solids removal is $(h_1/H) \times 10 = (4.2/6) \times 10 = 7.0\%$.

In a similar manner for succeeding 10% intervals, the curves for constant % removal of 45, 55, 65, and 75% are drawn and the average settled depths of 2.4, 1.4, 0.84, and 0.28 ft are read for $t = 25$ min. The calculations for $t = 25$ min are indicated below.

Settling velocity: $V_S = H/t = 6.0/(25/60) = 14.4$ ft/hr

Percent solids removal (for $t = 25$ min)

100% removal @ 30%		30.00%
1st interval (35%):	$(4.2/6.0) \times 10 =$	7.00%
2nd interval (45%):	$(2.4/6.0) \times 10 =$	4.00%
3rd interval (55%):	$(1.4/6.0) \times 10 =$	2.33%
4th interval (65%):	$(0.84/6.0) \times 10 =$	1.40%
5th interval (75%):	$(0.28/6.0) \times 10 =$	0.46%
		45.19%
Total removed after 25 min		45.2%

Beyond the fifth interval the % removals are negligible, so calculations are stopped at that point. In general, if 10% intervals are selected, the total % removal is given by

$$\text{Total \% removal} = X_{\text{total}} + (h_1/H) \times 10 + (h_2/H) \times 10 + (h_3/H) \times 10 + \cdots \tag{3.27}$$

Equation (3.27) is simply an approximation of Eq. (3.26). $(1 - x_0)$ corresponds to X_{total} and the integral $\int_0^{x_0} (h/H)\,dx$ is replaced by a finite summation of terms,

$$\sum (h_{\text{ave}}/H) \times \Delta x$$

where h_{ave} is the average settled depth for each selected interval. The Δx's in this example are selected arbitrarily as a 10% range. The smaller the Δx selected, the closer the approximation between the finite summation and the integral.

Similar calculations are performed for the other residence times listed in the second column of Table 3.6. The final results are summarized in Table 3.7.

TABLE 3.7
SS (%) Removed vs. Detention Time

(*1*) t (min)	(*2*) % SS removal
3.7	13.4
6.5	20.1
14.5	33.9
25.0	45.2
39.0	55.0
56.5	64.3
577.5	71.1

From Table 3.7 a graph of % SS removed vs. detention time is prepared (Fig. 3.13).

Step 4(c). Prepare a plot of % SS removed vs. overflow rate. Calculations needed to prepare this plot are presented in Table 3.8.

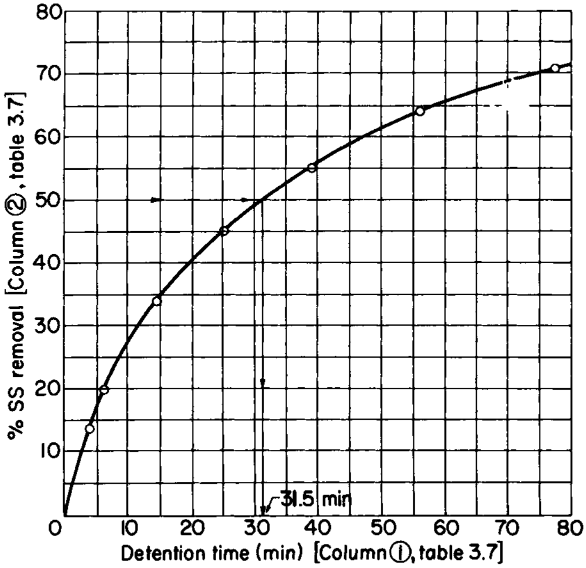


Fig. 3.13. Suspended solids removal (% SS) vs. detention time.

TABLE 3.8
SS (%) Removed vs. Overflow Rate

(*1*) t (min)	(*2*) Settling velocity, V_S (ft/hr) (Table 3.6)	(*3*) Overflow rate $\left[\frac{\text{gal}}{(\text{day})(\text{ft}^2)}\right]$ $V_S\left[\frac{\text{ft}}{\text{hr}} \times \frac{24\ \text{hr}}{\text{day}} \times \frac{7.48\ \text{gal}}{\text{ft}^3} = V_S \times 24 \times 7.48 = 179.5 V_S\right]$	(*4*) % SS removal [column (*2*) of Table 3.7]
3.7	97.2	17,450	13.4
6.5	55.2	9,908	20.1
14.5	24.8	4,452	33.9
25.0	14.4	2,585	45.2
39.0	9.2	1,651	55.0
56.5	6.35	1,140	64.3
77.5	4.64	833	71.1

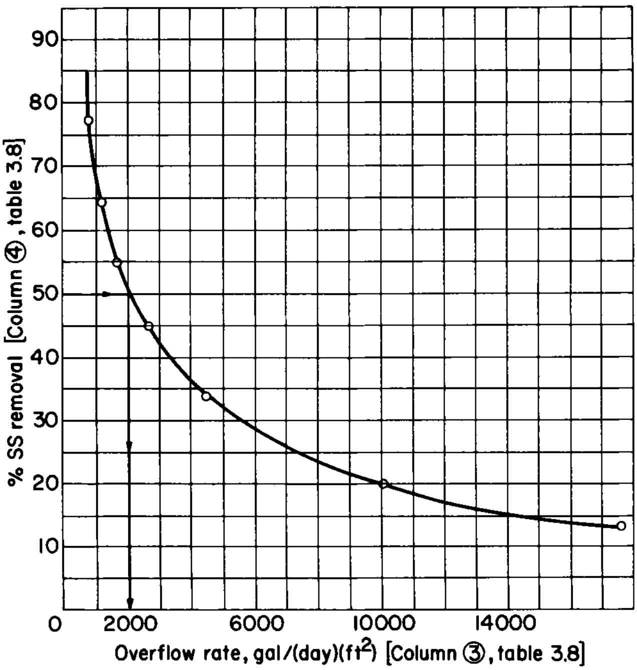


Fig. 3.14. Suspended solids removal (% SS) vs. overflow rate.

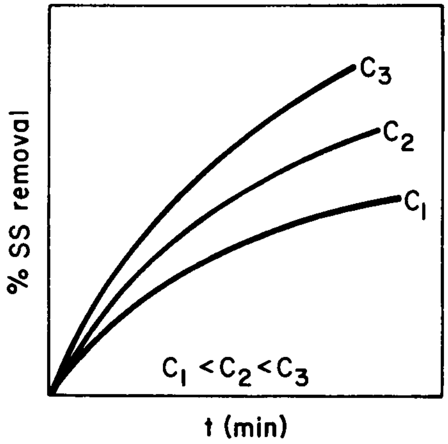


Fig. 3.15. Suspended solids removal (% SS) vs. detention time for different initial SS concentrations.

The plot of % SS removed vs. overflow rate is presented in Fig. 3.14. All calculations are performed for an initial suspended solids concentration SS_0 of 430 mg/liter. If similar calculations are performed for other values of these concentrations $(C_1, C_2, C_3, \ldots)$, the data plotted in Figs. 3.13 and 3.14 yield families of curves, as indicated in Figs. 3.15 and 3.16.

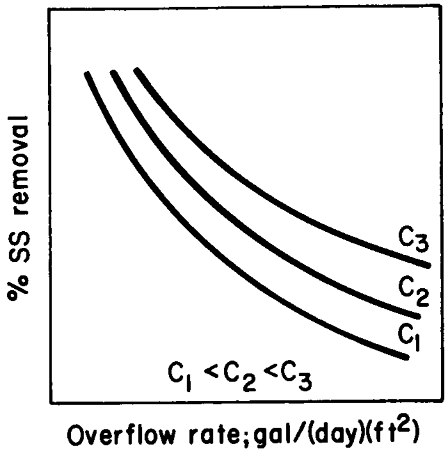


Fig. 3.16. Suspended solids removal (% SS) vs. overflow rate for different initial SS concentrations.

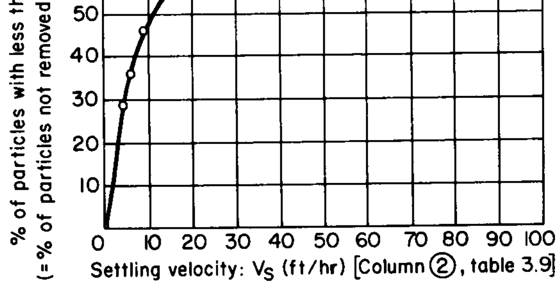


Fig. 3.17. Percentage of particles with less than stated velocity vs. settling velocity.

Step 4(d). Prepare a plot of percentage of particles with less than stated velocity (percentage not removed) vs. settling velocity (ft/hr). Calculations needed to prepare Fig. 3.17 are presented in Table 3.9. Figure 3.17 is not required for the design calculations; it is shown because a typical graph of this type was utilized in developing Eq. (3.26). The plot is presented in Fig. 3.17.

TABLE 3.9

Percentage of Particles with Less Than Stated Velocity vs. Settling Velocity

(*1*) t (min)	(*2*) V_S (ft/hr) (Table 3.6)	(*3*) % SS removal (Table 3.7)	(*4*) Percentage not removed: 100−(% SS removal)
3.7	97.2	13.4	86.6
6.5	55.2	20.1	79.9
14.5	24.8	33.9	66.1
25.0	14.4	45.2	54.8
39.0	9.2	55.0	45.0
56.5	6.35	64.3	35.7
77.5	4.64	71.1	28.9

3.5.1. Design Calculations from Laboratory Data

For purposes of scale-up, the fact that the efficiency of the process in an actual settling tank is reduced owing to the effect of parameters such as turbulence, short circuiting, and interference of the inlet and outlet must be taken into account. The net effect of these factors results in a decrease of the overflow rate and an increase in the detention time over values derived from the laboratory analysis. For design purposes, it is customary to divide the overflow rate obtained from the laboratory analysis by a factor ranging between 1.25 and 1.75, and to multiply the detention time by a factor in the same range [3]. Table 3.10 presents some commonly used design values.

TABLE 3.10

Design Values (Primary Clarifiers)

Depth: 7–12 ft
Detention time: 1–2 hr
Flow-through velocity, $V = 1$–5 ft/min
Overflow rate: 900–1200 gal/(day)(ft^2)
Efficiencies
  SS removal: 40–60%
  BOD removal: 30–50%

Design procedure of a primary clarifier is illustrated by Example 3.4.

Example 3.4

It is determined by field observation that a raw wastewater has an average of 430 mg/liter suspended solids at a flow of 1.0 Mgal/day. Data shown in Table 3.3 are obtained from laboratory settling tests.

1. Design a settling tank of circular cross section, i.e., calculate its diameter and effective depth, to remove 50% of the suspended solids at the flow rate of 1.0 Mgal/day.
2. For the tank designed in (1), what is the removal if flow is doubled to 2.0 Mgal/day?
3. For the flow of 1.0 Mgal/day, calculate the daily accumulation of sludge in lb/day and the average pumping rate in gal/min. Sludge concentration is estimated as 1.5% solids from tests made with samples withdrawn from Tap 4 of the laboratory settling column (Fig. 3.9). A plot of % solids for the compacted sludge vs. settling time is constructed from data obtained from samples withdrawn from Tap 4 (Fig. 3.9). A typical plot of this type is shown in Fig. 3.18.

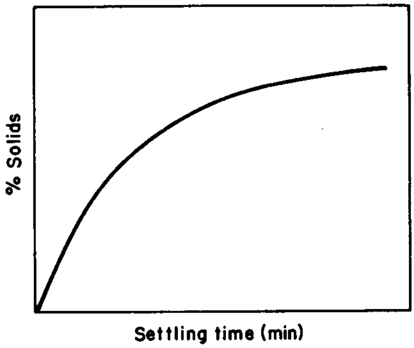


Fig. 3.18. Typical plot of % solids in the sludge vs. detention time.

SOLUTION: Part 1

Step 1. Determine the material balance for SS (see Fig. 3.19).

Influent: 430 mg/liter
Removal: (0.50)(430) = 215 mg/liter
Effluent: 430 − 215 = 215 mg/liter

Step 2. Determine the overflow rate. From Fig. 3.14 read overflow rate corresponding to a 50% removal, 2000 gal/(day)(ft^2). Using a 1.75 scale-up factor, take a design overflow rate of 2000/1.75 = 1143 gal/(day)(ft^2).

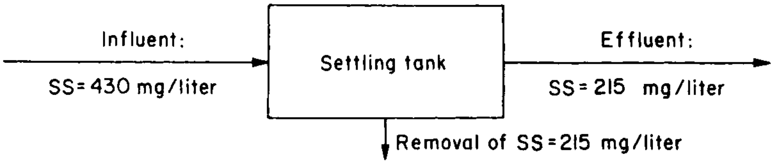


Fig. 3.19. Material balance for primary clarifier (Example 3.4).

Step 3. Determine the detention time. From Fig. 3.13 read detention time corresponding to a 50% removal value, $t = 31.5$ min.* Using a 1.75 scale-up factor, take $t = (31.5)(1.75) = 55.1$ min or $t = 55.1/60 = 0.92$ hr.

Step 4. Calculate required horizontal cross section of clarifier and its diameter. Horizontal cross section of clarifier is

$$A = \frac{1 \times 10^6 \text{ gal/day}}{1143 \text{ gal/(day)(ft}^2\text{)}} = 874.9 \text{ ft}^2$$

and diameter is

$$D = (4A/\pi)^{1/2} = (874.9/0.785)^{1/2} = 33.4 \text{ ft}$$

Step 5. Calculate effective depth of the clarifier.

$$H = \text{volume}/A = Qt/A$$

$$H = \frac{1 \times 10^6 \text{ gal/day} \times \text{ft}^3/7.48 \text{ gal} \times 0.92/24 \text{ day}}{874.9 \text{ ft}^2} = 5.9 \text{ ft}$$

SOLUTION: Part 2 Removal for a flow rate of 2 MGD with the same clarifier

This amounts to doubling the design overflow rate, i.e.,

New design overflow rate $(2)(1143) = 2286$ gal/(day)(ft^2)

From Fig. 3.14 this corresponds to a removal of 47.5% of the suspended solids.

SOLUTION: Part 3 Daily accumulation of sludge and average pumping rate for flow of 1.0 MGD

Step 1. Determine the daily accumulation of sludge.

Removal of SS 215 mg/liter → 215×10^{-6} lb SS/lb liquor

Therefore, the daily accumulation of sludge in lb/day is

$$1 \times 10^6 \text{ gal liquor/day} \times 8.34 \text{ lb liquor/gal liquor} \times 215 \times 10^{-6} \text{ lb SS/lb liquor}$$

$$= 1793 \text{ lb SS/day}$$

---

* From this value of the residence time ($t = 31.5$ min), the % solids in the sludge is estimated as 1.5% from a curve of the type in Fig. 3.19.

Step 2. Calculate the average pumping rate. Notice that 1.5% solids corresponds to 1.5 g SS/100 g of liquor = 15 g SS/1000 g of liquor ≈ 15 g SS/liter = 15,000 mg/liter = ppm = $15{,}000 \times 10^{-6}$ lb SS/lb liquor = $15{,}000 \times 8.34 \times 10^{-6}$ lb SS/gal liquor. Since accumulation is 1793 lb SS/day, pumping rate in gal/day is

$$\frac{1793 \text{ lb SS/day}}{(15{,}000 \times 8.34 \times 10^{-6}) \text{ lb SS/gal liquor}} = 0.0143 \times 10^{6} \text{ gal/day}$$

or

$$(0.0143 \times 10^{6})/(24 \times 60) = 9.93 \text{ gal/min}$$

Since pumping rate is low, intermittent pumping is used.

3.6. ZONE SETTLING

Zone settling occurs in clarifiers of activated or chemically coagulated sludge when the concentration exceeds 500 mg/liter. The sludge blanket exhibits several distinct zones. Each zone is characterized by a specific sludge concentration and settling velocity. Consider what happens when a suspension which initially has a uniform sludge concentration C_0 (mg/liter) is placed in a settling cylinder (Fig. 3.20).

Sludge begins to settle out and an interface (interface 1) is established between the surface of the blanket of settling sludge and the clarified liquid above. The zone below the clarified liquid is called the interfacial zone. Concentration of the sludge in this zone is uniform, and it settles as a blanket with a constant velocity (V_S). Simultaneously with formation of interface 1

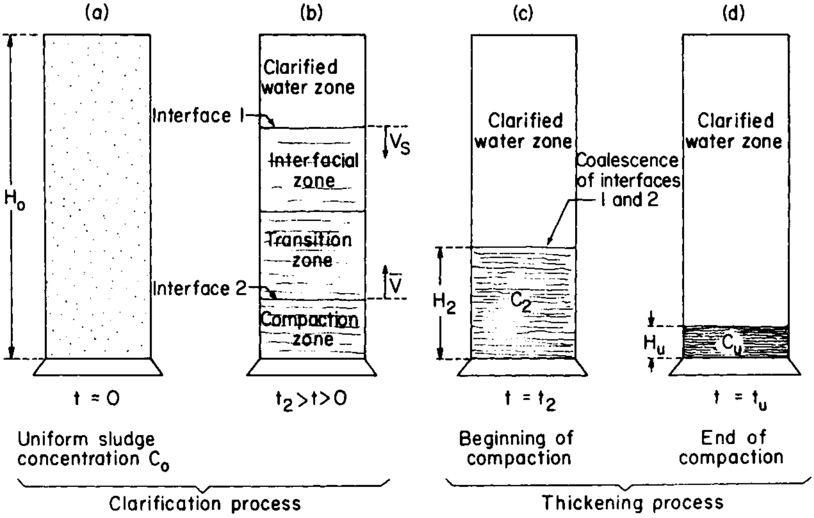


Fig. 3.20. Zone settling.

and interfacial zone, compaction of suspended solids starts at the bottom of the cylinder (compaction zone). In this zone concentration of SS is also uniform, and the interface bordering this zone (interface 2) rises in the cylinder with a constant velocity ($\bar{V}$).

Between the interfacial and compaction zones there is a transition zone. There, the velocity of settling solids decreases owing to increase of viscosity and density of the suspension. In this same zone, sludge changes gradually in concentration from that of the interfacial zone to that of the compaction zone.

Consider interfaces 1 and 2 in Fig. 3.20(b). Interface 1 moves downward with a constant velocity V_S, whereas interface 2 moves upward with a constant velocity $\bar{V}$. Eventually, interfacial and compaction zones meet, at which time (t_2) the transition zone fades away [Fig. 3.20(c)]. At this time, the settled sludge exhibits a uniform concentration C_2, which is termed the critical concentration. Compaction starts and the sludge begins to thicken, eventually reaching an ultimate concentration C_u [Fig. 3.20(d)]. Sedimentation velocity at time t_2 corresponds to a value V_2, which is given by the slope of the tangent to the settling curve at C_2, as indicated in Fig. 3.21 where $V_2 < V_S$.

Procedure for designing clarifiers operating under conditions of zone settling:

1. Calculate the minimum surface area required to allow for clarification of sludge.
2. Calculate the minimum surface area required to provide for thickening of sludge to the desired underflow concentration.
3. Take the larger of these two areas as the design area for the clarifier.

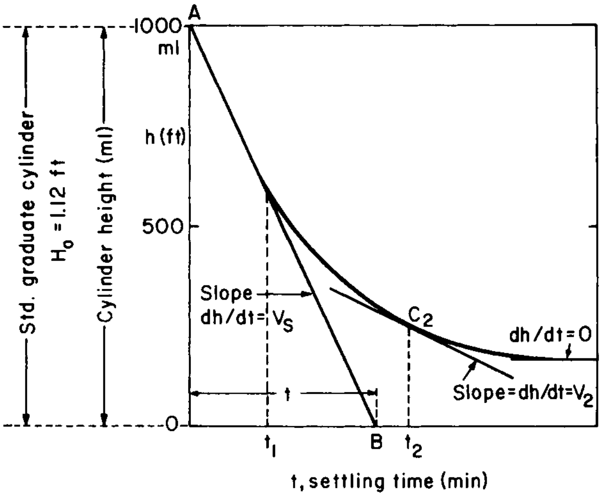


Fig. 3.21. Sludge settling curve.

3.6.1. Laboratory Measurements

To obtain the parameters necessary for design of the clarifier, a settling test for the sludge is performed in the laboratory using a 1000-ml graduated cylinder (a standard graduate cylinder has a height of 1.12 ft). The cylinder is filled with the slurry to be studied. At the beginning of the experiment ($t = 0$), slurry concentration is uniform throughout the cylinder. Height of interface 1 is recorded at selected time intervals. This yields the type of settling curve shown in Fig. 3.21.

It is important to stir the suspension at a rate of about 5 rph. This stirring simulates the action of the mechanical rakes utilized in sludge removal and prevents stratification of the sludge.

Figure 3.21 shows that from the start of the experiment up to a time t_1, interface 1 falls with a constant velocity V_S given by the slope of the tangent, which essentially coincides with the settling curve from $t = 0$ to $t = t_1$. After $t = t_1$ this velocity decreases appreciably. At time $t = t_2$, the velocity is V_2 given by the slope of the tangent at C_2. At t_2 compaction starts and the velocity is further reduced until it becomes essentially zero, the tangent being parallel to the abscissa.

Zone settling velocity (ZSV) corresponds to the velocity at which the suspension settles prior to reaching the critical concentration C_2, and is given by the slope of the tangent AB in Fig. 3.21 [Eq. (3.28)].

$$V_S = \overline{\mathrm{OA}}/\overline{\mathrm{OB}} = H_0/t = 1.12\ \mathrm{ft/min} \tag{3.28}$$

3.6.2. Determination of Minimum Surface Area Required to Allow Clarification of the Sludge

Minimum surface area A_c required for clarification depends on velocity V_S at which the suspension settles before reaching the interfacial critical concentration C_2. Under continuous flow conditions, velocity of the liquor over the overflow weir cannot exceed V_S if clarification is to take place. This follows directly from the basic concept of the ideal sedimentation tank [Section 3.4, Eqs. (3.17) and (3.19)].

Therefore,

$$A_c = Q/V_S \tag{3.29}$$

where Q is the flow rate (ft^3/min); V_S the settling velocity (ft/min); and A_c the minimum surface area required for clarification (ft^2).

The value of the zone settling velocity V_S is determined from Fig. 3.21 and Eq. (3.28). Value of t is read directly from the abscissa of Fig. 3.21 (point B). V_S is then calculated from Eq. (3.28) and A_c obtained from Eq. (3.29).

3.6.3. Determination of Minimum Surface Area Required for Thickening of the Sludge

Consider settling of a sludge under zone settling conditions in a cylinder (Fig. 3.20). At start of the experiment, let C_0 be the uniform sludge concentration throughout the cylinder. Total weight of solids in the cylinder is C_0AH_0, where A is the cross-sectional area of the cylinder. Let t_2 be the time counted from the beginning of the experiment, when interfacial and compaction zones merge together [Fig. 3.20(c)]. Let C_2 be the critical concentration which is uniform throughout this sludge zone formed by the merging of interfacial and compaction zones.

A graphical procedure has been proposed [7] for determining t_2. Consider the settling curve in Fig. 3.22. Draw two tangents (AB and CD) to the two branches of the curve. Tangent AB corresponds to the constant velocity of settling V_S for the interfacial zone (zone settling velocity, ZSV), and tangent CD corresponds to the settling velocity for the compacted sludge. Point C_2 (corresponding to time t_2) is obtained by bisecting the angle formed by tangents AB and CD. The abscissa of the point where the bisecting line cuts the settling curve corresponds to the desired value of t_2.

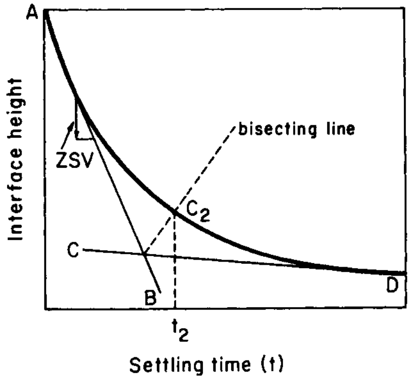


Fig. 3.22. Determination of t_2.

Consider now the thickening process.

1. Start of thickening [Fig. 3.20(c)]. Time, t_2; concentration of SS in sludge zone, C_2; height of sludge zone, H_2.

2. End of thickening [Fig. 3.20(d)]. The compacted sludge reaches desired underflow concentration C_u. The time at which this occurs is designated as t_u. Height of the sludge zone is H_u.

Consider separately the sludge zone at the start and end of thickening (Fig. 3.23). Since the total mass of sludge in the cylinder is constant, the

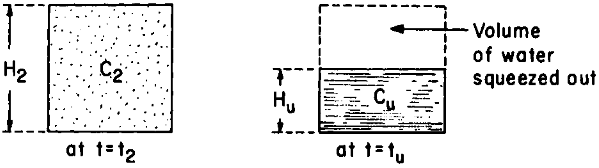


Fig. 3.23. Thickening process.

following material balance equation [Eq. (3.30)] is written, neglecting the amount of suspended solids in the clarified water zone.

$$C_0 A H_0 = C_2 A H_2 = C_u A H_u \tag{3.30}$$

or

$$C_0 H_0 = C_2 H_2 = C_u H_u \tag{3.31}$$

Consider Fig. 3.23. The volume of water which is squeezed out and discharged over the overflow weir is calculated from Eq. (3.32).

$$V = A(H_2 - H_u) \tag{3.32}$$

The time interval required to discharge this volume of water is $t_u - t_2$. Average rate of flow Q' (ft^3/min) over the weir is [Eq. 3.33)]

$$Q' = V/(t_u - t_2) = A(H_2 - H_u)/(t_u - t_2) \tag{3.33}$$

Solving for $t_u - t_2$,

$$t_u - t_2 = A(H_2 - H_u)/Q' \tag{3.34}$$

Consider now the settling curve and determine graphically the settling velocity V_2 at time t_2 (tangent at point C_2). This is shown in Fig. 3.24.

Let H_1 be the intercept of this tangent. Settling velocity at t_2 is shown in Eq. (3.35).

$$V_2 = \tan\alpha = (H_1 - H_2)/t_2 \tag{3.35}$$

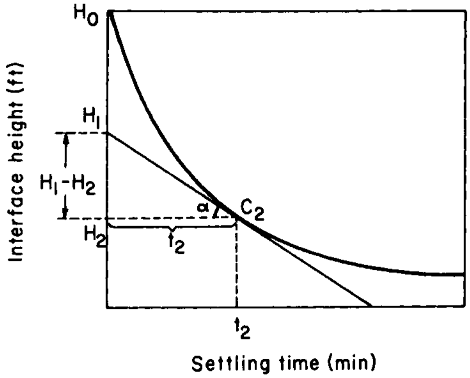


Fig. 3.24. Determination of velocity V_2.

Under continuous flow conditions the velocity of the liquor over the weir cannot be greater than V_2 if thickening is to take place. Therefore, the flow rate Q' at time t_2 when thickening starts is

$$Q' = AV_2 = A[(H_1 - H_2)/t_2] = \text{ft}^3/\text{min} \tag{3.36}$$

Substitution of Q' given by Eq. (3.36) in Eq. (3.34) yields after simplification and rearrangement Eq. (3.37).

$$(H_2 - H_u)/(t_u - t_2) = (H_1 - H_2)/t_2 \tag{3.37}$$

This equation is the basis for the graphical procedure for determination of t_u illustrated by Fig. 3.25.

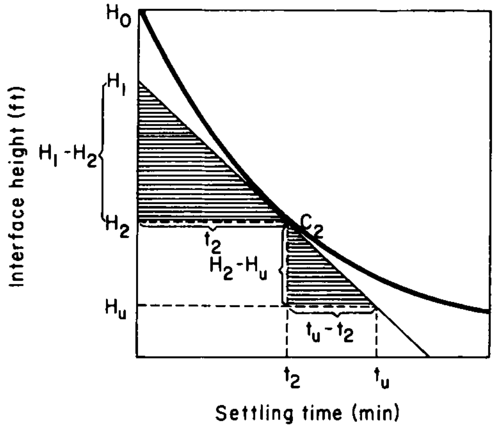


Fig. 3.25. Determination of t_u.

To summarize the steps in the graphical procedure for determination of t_u:

1. Draw the tangent to the settling curve at C_2.
2. From material balance [Eq. (3.31)]

$$H_u = H_0 C_0 / C_u \tag{3.38}$$

Calculate H_u from Eq. (3.38).

3. Mark distance H_u on the ordinate axis of Fig. 3.25. Draw the horizontal dotted line from H_u until its intersection with the tangent to C_2. The abscissa of this intersection is the value of t_u. This can be seen by inspection of Eq. (3.37) and consideration of the two cross-hatched similar triangles in Fig. 3.25.

Minimum surface area required for thickening (A_t) is obtained from the following considerations. Average rate at which the layer of concentration C_u forms (in lb/min) is

$$C_u H_u A_t / t_u \tag{3.39}$$

Since from Eq. (3.31) $C_u H_u = C_0 H_0$, then Eq. (3.39) is rewritten as

$$C_0 H_0 A_t / t_u \tag{3.40}$$

Under conditions of continuous flow and steady state, the rate at which the layer of concentration C_u is formed must equal that at which suspended solids enter in the influent (QC_0). Therefore

$$QC_0 = C_0 H_0 A_t / t_u \tag{3.41}$$

Solving for A_t

$$A_t = Q t_u / H_0 \tag{3.42}$$

where $H_0 = 1.12$ ft.

The design procedure for clarifiers under zone settling conditions is illustrated by Example 3.5.

Example 3.5

Design a secondary settling tank to produce an underflow concentration of 10,900 mg/liter from an influent containing 2510 mg/liter of suspended solids. Wastewater flow is 1.2 MGD. Calculate the clarifier area required.

The data tabulated below are obtained in a laboratory test of the slurry.

t (min)	Interface height H (ml)
0	1000
1	850
2	725
3	600
5	450
8	350
12	280
16	240
20	220
25	210

SOLUTION

Step 1. Settling curve is plotted from available data (Fig. 3.26).

Step 2. Minimum surface area required for clarification (A_c) is determined.

1. Draw tangent AB. Read $t = 7.5$ min. Then $V_S = H_0/t = 1.12/7.5 = 0.149$ ft/min.
2. Area required for clarification:

$$A_c = \frac{1{,}200{,}000 \text{ gal/day} \times \text{ft}^3/7.48 \text{ gal} \times \text{day}/1440 \text{ min}}{0.149 \text{ ft/min}} = 748 \text{ ft}^2$$

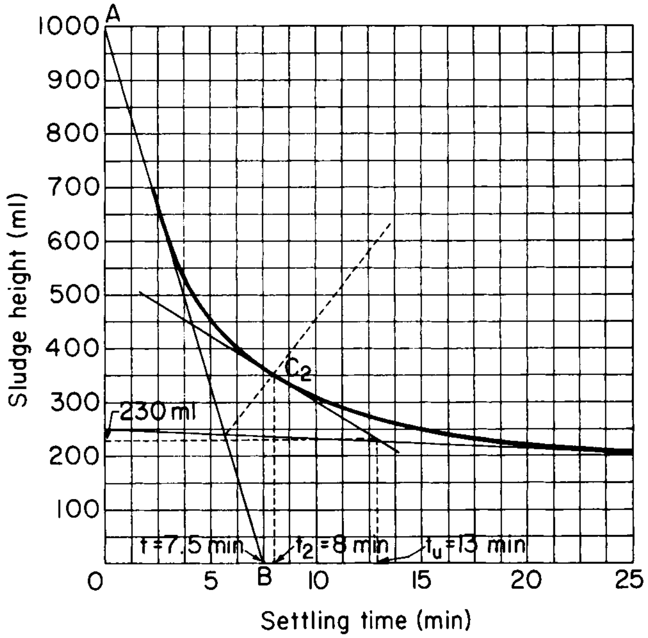


Fig. 3.26. Graph for Example 3.5.

Step 3. Minimum surface area required for thickening (A_t) is calculated.

1. Determine t_2 by the graphical procedure studied. From Fig. 3.26 read $t_2 = 8.0$ min.

2. Determine time t_u. Draw the tangent to the settling curve at C_2 and calculate H_u from Eq. (3.28).

$$H_u = (1000 \times 2510)/10{,}900 = 230 \text{ ml}$$

3. Determine t_u by the graphical procedure described. From Fig. (3.26) read $t_u = 13$ min.

4. Calculate A_t from Eq. (3.42).

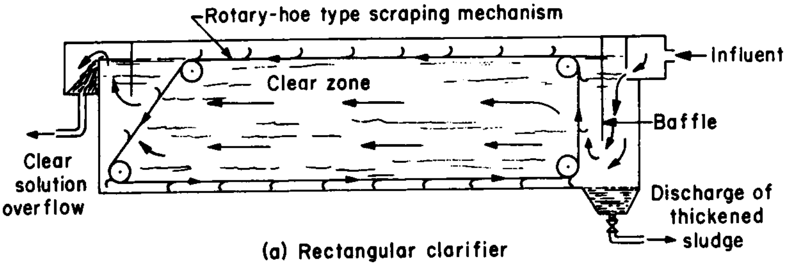
$$A_t = \frac{1{,}200{,}000 \text{ gal/day} \times \text{ft}^3/7.48 \text{ gal} \times \text{day}/1440 \text{ min} \times 13 \text{ min}}{1.12 \text{ ft}} = 1293 \text{ ft}^2$$

Step 4. Take $A = A_t = 1293$ ft^2. Required diameter for a sedimentation tank of circular cross section is

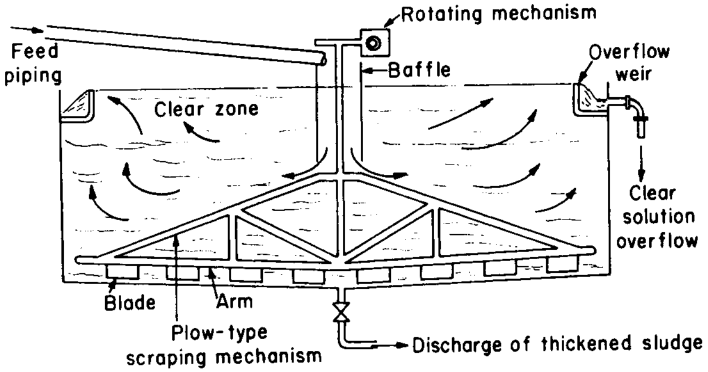
$$d = (4A/\pi)^{1/2} = (1293/0.785)^{1/2} = 40.6 \text{ ft}$$

3.7. TYPES OF CLARIFIERS

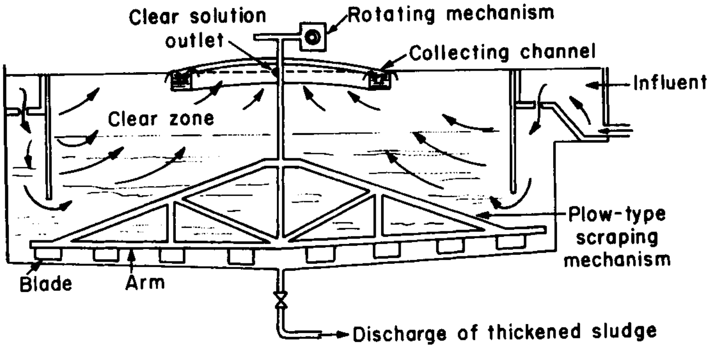
Clarifiers are classified according to geometry of their horizontal cross section as (1) rectangular and (2) circular: (a) center and (b) peripheral feed. Sketches of typical clarifiers are shown in Figs. 3.27(a), (b), and (c).



(a) Rectangular clarifier



(b) Circular clarifier (center feed)



(c) Circular clarifier (peripheral feed)

Fig. 3.27 Types of clarifiers (a), (b), and (c).

1. Rectangular clarifier [Fig. 3.27(a)]

In the type shown in Fig. 3.27(a), scraped sludge is moved toward the inlet end of the tank. Some other designs move sludge toward the effluent end of the tank.

Scraping mechanism shown is of rotary-hoe type, consisting of a series of short scrapers mounted on an endless chain, which make contact with the bottom of the tank. It moves slowly at speeds of approximately 1 ft/min.

2a. Circular clarifier with center feed [Fig. 3.27(b)]

Feed is at the center and clear solution overflows to a collecting channel at the periphery. The bottom of the clarifier has a minimum slope of 1 in./ft. Scraping mechanism is of plow type to overcome inertia and prevent adherence of sludge to the bottom of the tank.

2b. Circular clarifier with peripheral feed [Fig. 3.27(c)]

The feed is at the periphery and the clear solution overflows to a collecting channel at the center. The other details are similar to those for the type shown in Fig. 3.27(b).

The inlet section should be carefully designed for a uniform flow distribution across the width and depth of the tank. Similarly, the outlet section should be designed to collect the effluent uniformly.

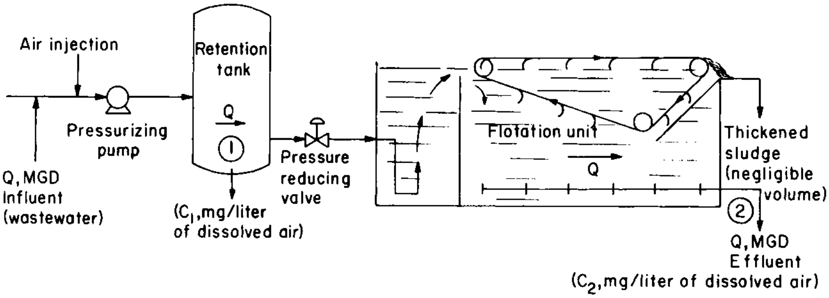
A good design of inlet and outlet sections reduces possibilities of flow short circuiting, which lead to poor performance of the clarifier. Proper positioning of weirs and baffles, as indicated in Fig. 3.27, prevents short circuiting.

4. Flotation

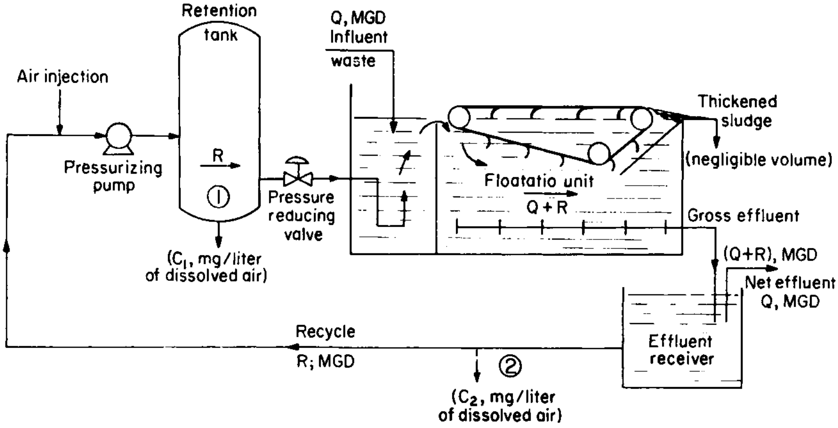
4.1. INTRODUCTION

Flotation is a process for separating low density solids or liquid particles from a liquid phase. Separation is brought about by introduction of gas (usually air) bubbles into the liquid phase. The liquid phase is pressurized to an operating pressure ranging from 30 to 60 psia (2–4 atm) in presence of sufficient air to promote saturation of air in the water. Then, this air-saturated liquid is depressurized to atmospheric pressure by passage through a pressure-reducing valve. Minute air bubbles are released from the solution because of depressurization. Suspended solids or liquid particles, e.g., oil, are floated by these minute air bubbles, causing them to rise to the surface of the tank. Concentrated suspended solids are skimmed off by mechanical means from the tank surface. Clarified liquor is withdrawn near the bottom, and part of it may be recycled [Fig. 3.28(b)]. A flotation system without recycle is shown diagrammatically in Fig. 3.28(a).

In the field of wastewater treatment, flotation is used for the following purposes: (1) separation of greases, oils, fibers, and other low density solids from wastewaters; (2) thickening of the sludge from the activated sludge



(a) Flotation system without recycle



(b) Flotation system with recycle

Fig. 3.28. Flotation systems (a) and (b).

process; and (3) thickening of flocculated chemical sludges resulting from chemical coagulation treatment. Superior effluent quality, i.e., effluent containing lower percentage of suspended solids, and economy in power are achieved by flotation systems with recycle.

Basic components of a flotation system are (1) pressurizing pump; (2) air injection facilities; (3) retention tank (to provide air–liquid contact); (4) pressure-reducing valve; and (5) flotation tank.

4.2. EVALUATION OF FLOTATION VARIABLES FOR PROCESS DESIGN

4.2.1. Parameter *A/S*

For design of flotation systems, a fundamental parameter commonly utilized is a dimensionless air to solids ratio (A/S) defined by Eq. (3.43).

$$A/S = \frac{\text{lb/day of air released by depressurization}}{\text{lb/day of solids in the influent}} \tag{3.43}$$

This parameter is estimated from studies with a laboratory-scale flotation cell of pilot-plant data. The value of A is obtained from determinations of dissolved air (mg/liter) at sampling locations indicated as (1) and (2) in Fig. 3.28(a) and (b). Thus [Eq. (3.44)]

$$A = A_1 - A_2 \tag{3.44}$$

where A is the lb/day of air released by depressurization; A_1 the lb/day of dissolved air at (1) [Fig. 3.28(a) and (b)]; and A_2 the lb/day of dissolved air at (2) [Fig. 3.28(a) and (b)].

For flotation systems without recycle [Eq. (3.45)],*

$$A_1 = Q\frac{\text{Mgal liquor}}{\text{day}} \times C_1\frac{\text{lb air}}{\text{Mlb liquor}} \times 8.34\frac{\text{lb liquor}}{\text{gal liquor}} = 8.34QC_1 \qquad \text{(lb air/day)} \tag{3.45}$$

Similarly [Eq. (3.46)]

$$A_2 = 8.34QC_2 \qquad \text{(lb air/day)} \tag{3.46}$$

Therefore

$$A = A_1 - A_2 = 8.34Q(C_1 - C_2) \qquad \text{(lb/day of air released)} \tag{3.47}$$

For flotation systems with recycle the corresponding equation is

$$A = A_1 - A_2 = 8.34R(C_1 - C_2) \qquad \text{(lb/day of air released)} \tag{3.48}$$

If S_t is the concentration of suspended solids (mg/liter) in the influent, the value of S [denominator of Eq. (3.43)] is

$$S = Q\frac{\text{Mgal liquor}}{\text{day}} \times \frac{8.34 \text{ lb liquor}}{\text{gal liquor}} \times S_t\frac{\text{lb SS}}{\text{Mlb liquor}} = 8.34QS_t \qquad \text{(lb SS/day)} \tag{3.49}$$

Substitution of Eqs. (3.47) [or Eq. (3.48)] and (3.49) in Eq. (3.43) leads to

Flotation systems without recycle

$$A/S = (C_1 - C_2)/S_t \tag{3.50}$$

Flotation systems with recycle

$$A/S = (R/Q)(C_1 - C_2)/S_t \tag{3.51}$$

4.2.2. Correlation of Flotation Variables to Parameter *A*/*S*

By use of a laboratory flotation cell or pilot-plant data, it is possible to correlate mg/liter of suspended solids contained in liquid effluent to parameter A/S. A typical correlation curve for a wastewater has the shape indicated in Fig. 3.29.

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* Notice that C_1 mg of air/liter of liquor = C_1 lb of air/Mlb of liquor.

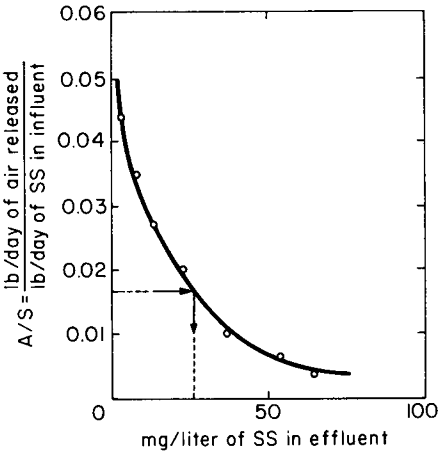

Fig. 3.29. Typical correlation of parameter A/S vs. concentration of SS in effluent.

For a given influent, the lb/day of suspended solids (term S in ratio A/S) is fixed. Ratio A/S increases by operating at higher air rates, which results in increase of air released (A). From Fig. 3.29 it follows that a higher quality effluent is obtained.

Graphs like Fig. 3.29 permit selection of the A/S ratio for a required degree of effluent clarification. These curves indicate that increasing the A/S ratio beyond an optimum value does *not* result in substantial reduction in effluent suspended solids.

Judicious selection of the A/S ratio involves an economical balance between equipment and maintenance costs and desired effluent quality. Typical range of A/S ratios for thickening of sludges in wastewater treatment is 0.005–0.060.

4.2.3. Alternative Expressions for Parameter A/S [2]

Consider Eq. (3.46) for A_2, where C_2 is the solubility of air in water in mg/liter. Frequently, the solubility of air is expressed in terms of cm^3 of air/liter of water. It is assumed that conditions at (2) [Figs. 28(a) and (b)] are atmospheric pressure and ambient temperature. Solubility of air in water in cm^3 of air/liter of water [hence denoted as S_a] is presented in Table 3.11 for atmospheric pressure at several temperatures.

This concentration of dissolved air in cm^3 of air/liter of water is converted to the value C_2 (mg air/liter of water) [Eq. (3.52)].

$$S_a \text{ cm}^3 \text{ air/liter water} \times \rho_a \text{ mg air/cm}^3 \text{ air} = C_2 \quad \text{(mg air/liter water)} \tag{3.52}$$

TABLE 3.11
Solubility of Air in Water at Atmospheric Pressure at Several Temperatures [6]

Temperature (°C)	S_a (cm^3/liter)
0	29.2
10	22.8
20	18.7
30	15.7

Utilize for the density of air ρ_a a mean value of 1.2 mg/cm^3. (This corresponds to the value at 1 atm and 20°C.) Then [Eq. (3.53)]

$$C_2 = 1.2S_a \tag{3.53}$$

Consider now Eq. (3.45), where C_1 is the solubility of air in mg/liter. The cm^3 of air/liter of water at the retention tank [point (1)] is shown in Eq. (3.54).

$$\varphi S_a \quad (\text{cm}^3 \text{ air/liter water}) \tag{3.54}$$

where φ is the ratio of solubility of air in water at the pressure in the retention tank [at (1)] to the solubility at atmospheric pressure [at (2) in Figs. 28(a) and (b)] [Eq. (3.55)].

$$\varphi = C_1/C_2 \qquad (\varphi > 1.0) \tag{3.55}$$

It is found experimentally that for a specific design of the retention tank, and within ordinary pressure ranges utilized in flotation operations, ratio φ is proportional to the pressure

$$\varphi = fP \tag{3.56}$$

where f is the proportionality factor and P the pressure in retention tank in atmospheres.

At 20°C for pressure range 30–60 psia utilized in most flotation systems and for baffled retention tanks, the value of f in Eq. (3.56) is approximately 0.5. Consequently, values of φ vary from

$$30 \text{ psia } (= 30/14.7 = 2.04 \text{ atm}) = 0.5 \times 2.04 = 1.02$$

$$60 \text{ psia } (= 60/14.7 = 4.08 \text{ atm}) = 0.5 \times 4.08 = 2.04$$

Therefore for the pressure range from 30 to 60 psia at 20°C, solubility of air in baffled retention tanks varies from 1.02 to 2.04 times its saturation value at 20°C and 1 atm. Since from Eq. (3.55)

$$C_1 = \varphi C_2 \tag{3.57}$$

substitution of φ and C_2 by their values given by Eqs. (3.56) and (3.53) yields

$$C_1 = fP(1.2S_a) \tag{3.58}$$

Substitution of values of C_1 and C_2 given, respectively, by Eqs. (3.58) and (3.53) in Eqs. (3.50) and (3.51) leads to

Flotation systems without recycle

$$A/S = 1.2S_a(fP-1)/S_i \tag{3.59}$$

Flotation systems with recycle

$$A/S = (R/Q)1.2S_a(fP-1)/S_i \tag{3.60}$$

4.3. DESIGN PROCEDURE FOR FLOTATION UNITS WITHOUT AND WITH RECYCLE

4.3.1. Flotation Systems without Recycle

From Eq. (3.59) it follows that if a ratio A/S is selected, f being fixed for a selected type of retention tank and S_i from characteristics of the influent, this amounts to specification of operating pressure P, which is calculated from Eq. (3.59) as

$$P = (1/f)[(A/S)S_i/1.2S_a+1] \tag{3.61}$$

Design of flotation systems without recycle involves calculating the required operating pressure [Eq. (3.61)] and determining the cross-sectional area of the flotation unit. This area is calculated from a selected value of the overflow rate, usually a value between 2 and 4 gal/(min)(ft^2) (Example 3.6).

Example 3.6

Laboratory flotation tests for a given wastewater indicate optimum air/solids ratio (A/S) as 0.04 lb air/lb solids. Flow of wastewater is 1.0 MGD and it contains 250 mg/liter of suspended solids. Laboratory flotation tests (without recycle) indicate for a ratio $A/S = 0.04$ an optimum effluent containing 25 mg/liter of suspended solids. Take $f = 0.50$ for retention tank and an operating temperature of 20°C. Design a flotation system without recycle for this service.

SOLUTION

Step 1. Select $A/S = 0.04$, as stated.

Step 2. Calculate P from Eq. (3.61).

$$P = (1/0.5)(0.04 \times 250/1.2 \times 18.7+1) = 2.9 \text{ atm}$$

Step 3. Select an overflow rate,

$$OR = 3 \text{ gal/(min)(ft}^2\text{)}$$

Step 4. Calculate required surface area,

$$Q = 1.0\ \text{MGD}$$

or

$$Q = 1{,}000{,}000\ \text{gal/day} \times \text{day/24 hr} \times \text{hr/60 min} = 695\ \text{gal/min}$$

$$\text{Surface area (ft}^2) = Q/OR = \frac{695\ \text{gal/min}}{3.0\ \text{gal/(min)(ft}^2)} = 232\ \text{ft}^2$$

4.3.2. Flotation Systems with Recycle

It follows from Eq. (3.60) that for a specific application (i.e., for fixed values of f and S_i) one must specify not only the A/S ratio but also the recycle before the operating pressure P becomes fixed. The usual procedure is to specify an operating pressure P and an A/S ratio and calculate the required recycle from Eq. (3.60), which solved for R yields Eq. (3.62).

$$R = (A/S)\,QS_i/1.2S_a(fP-1) \tag{3.62}$$

The design procedure for flotation systems with recycle is illustrated by Example 3.7.

Example 3.7

For the application in Example 3.6 design a flotation system with recycle, taking an operating pressure of 2.9 atm.

SOLUTION

Step 1. Select $A/S = 0.04$, as stated.
Step 2. Calculate R from Eq. (3.62).

$$R = (0.04)(1.0)(250)/1.2 \times 18.7(0.5\times 2.9-1) = 0.99\ \text{MGD} \approx 1.0\ \text{MGD}$$

This means that recycle ratio R/Q is approximately unity.

Step 3. Select an overflow rate

$$OR = 3\ \text{gal/(min)(ft}^2)$$

Step 4. Required surface area is defined as

$$Q + R \approx 2.0\ \text{MGD}$$

or

$$2{,}000{,}000\ \text{gal/day} \times \text{day/24 hr} \times \text{hr/60 min} = Q + R = 1390\ \text{gal/min}$$

$$\text{Surface area (ft}^2) = (Q+R)/OR = 1390/3.0 = 464\ \text{ft}^2$$

Surface area is twice as large as for the unit without recycle. However, an effluent of superior quality (i.e., SS < 25 mg/liter) is obtained.

5. Neutralization (and Equalization)

5.1. NEUTRALIZATION IN THE FIELD OF WASTEWATER TREATMENT

Neutralization treatment is often utilized in the following cases arising in wastewater treatment:

1. Prior to discharge of the wastewater into a receiving water. The justification for neutralization is that aquatic life is sensitive to pH variations beyond a narrow range around pH 7.

2. Prior to discharge of industrial wastewaters to the municipal sewer system. Specification of the pH of industrial discharges into municipal sewer systems is frequently made. It is more economical to neutralize industrial wastewater streams prior to the discharge into the municipal sewer, rather than attempting to perform neutralization of the larger volume of combined domestic and industrial sewage.

3. Prior to chemical or biological treatment. For biological treatment, pH of the system is maintained within the range 6.5–8.5 to ensure optimum biological activity. The biological process itself provides a neutralization and buffer capacity as a result of production of CO_2, which forms carbonates and bicarbonates in solution. The degree of preneutralization required for biological treatment depends on two factors: (1) the alkalinity or acidity present in the wastewater and (2) the mg/liter BOD to be removed in the biological treatment. The latter is related to the production of CO_2, which may provide for partial neutralization of alkaline wastes.

5.2. METHODS FOR NEUTRALIZATION OF WASTEWATERS

Methods for neutralization of wastewaters include (1) equalization, which consists of mixing acidic and alkaline waste streams available in the plant and (2) direct pH control methods, which consist of addition of acids (or bases) for neutralization of alkaline (or acidic) wastewater streams.

5.3. EQUALIZATION

When utilized for purpose of neutralization, equalization involves mixing wastewater streams of acidic and alkaline nature in an equalization basin. Equalization is often used for purposes other than neutralization such as (1) to smooth out individual wastewater stream flow variations, so that a composite stream of relatively constant flow rate is fed to the treatment plant; and (2) to smooth out variations in influent BOD to the treatment facilities. Constant and variable level equalization basins are utilized.

1. *Constant level equalization basins.* This arrangement is illustrated in Fig. 3.30. The level in the equalization basin is held constant. Therefore as

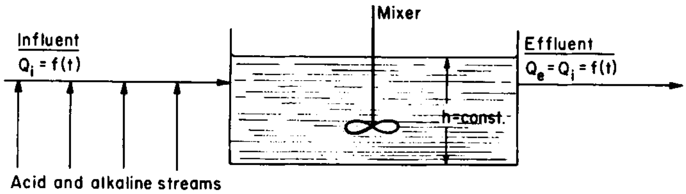

Fig. 3.30. Constant level equalization basin.

the rate of flow of influent varies, that of effluent is equally affected. Consequently, this is not a technique of flow equalization, simply a method of neutralization. If fluctuations in the flow rate are too great, the effluent from the constant level equalization basin is fed to another equalization basin having as objective flow equalization.

2. *Variable level equalization basins.* In this method of equalization, the effluent is taken out at a constant rate, and since the flow rate of influent varies with time, the level of the equalization basin is variable. This method is also utilized for the purpose of flow equalization, as well as providing for neutralization. A diagram of a variable level equalization basin is shown in Fig. 3.31.

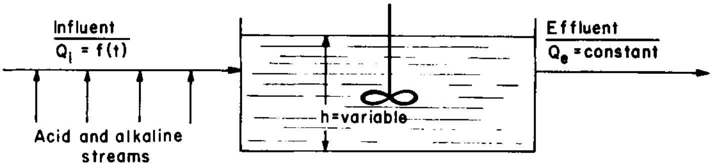

Fig. 3.31. Variable level equalization basin.

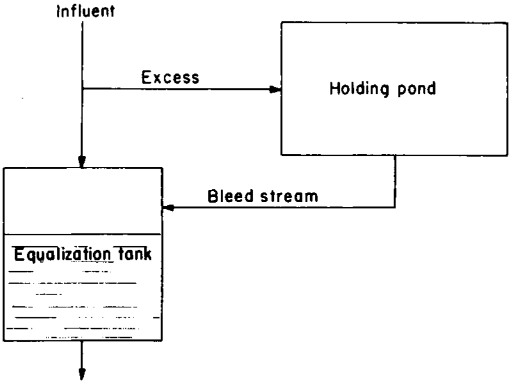

Fig. 3.32. "Holding pond" method of equalization.

Another method of equalization consists of diverting the "excess" of the incoming stream to a holding pond, from which a bleed stream is fed to the equalization tank. This method is not used for neutralization purposes, but only for equalization of BOD content or flow rate. This is illustrated by Fig. 3.32.

5.4. DIRECT pH CONTROL METHODS: NEUTRALIZATION OF ACIDIC WASTES BY DIRECT pH CONTROL METHODS

The following methods of direct neutralization of acidic wastes are the most commonly employed: (1) limestone beds, (2) slurried lime neutralization, (3) caustic soda (NaOH) neutralization, (4) sodium carbonate neutralization, and (5) ammonia neutralization.

A few specific comments about these methods are as follows:

Slurried lime neutralization is the most common method and is discussed in Section 5.6. Design of limestone beds is described in Section 5.5.

Caustic soda (NaOH) is more expensive than lime. It offers an advantage with respect to uniformity of the reagent, ease of storage and feeding, rapid reaction rate, and the fact that the end products of neutralization (sodium salts) are soluble.

Sodium carbonate (Na_2CO_3) is not as reactive as caustic soda and presents frothing problems owing to release of carbon dioxide.

Ammonia (NH_4OH) presents the disadvantage of being a contaminant; consequently its use may be ruled out by pollution control standards.

Factors guiding selection of a neutralization reagent are (1) purchase cost, (2) neutralization capacity, (3) reaction rate, and (4) storage and disposal of neutralization products.

5.5. LIMESTONE BEDS

5.5.1. Types of Equipment

Both upflow and downflow types of limestone beds are employed. For wastewaters containing H_2SO_4, limestone beds should not be used if concentration of H_2SO_4 exceeds 0.6%. The reason for this limitation is that the limestone becomes covered with an insoluble coat of $CaSO_4$, rendering it ineffective. In addition, evolution of CO_2 causes frothing problems.

Upflow type arrangement is preferable to downflow type since in upflow units, $CaSO_4$ tends to be flushed out before precipitation on the limestone. Also, escape of CO_2 generated by the neutralization reaction is easier in upflow type units. For these reasons, maximum hydraulic rate for downflow systems is limited to approximately 50 gal/(hr)(ft^2).

Presence of metallic ions (e.g., Al^{3+}, Fe^{3+}) in the wastewater reduces effectiveness of the limestone bed owing to coating of limestone with precipitated hydroxides. Finally, if dilution of the acid in the wastewater is increased, higher residence times are required for neutralization.

5.5.2. Design Procedure for Limestone Beds

In this section, the laboratory procedure recommended by Eckenfelder and Ford [3] for obtaining the basic design data is described. A numerical example is presented to illustrate design of an actual limestone column. A model of a laboratory limestone neutralization column is shown in Fig. 3.33. Bench-scale columns operate with heights of limestone of 1.0–5 ft, which is the actual range of heights for plant-scale units. Column diameter is approximately 6 in., and rates of flow [gal/(hr)(ft^2)] are comparable to those for plant operation.

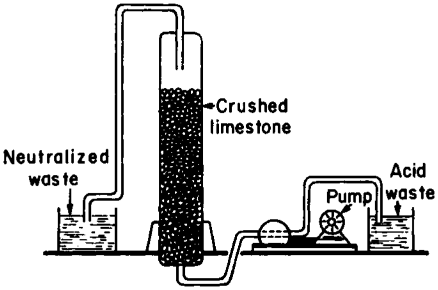

Fig. 3.33. Laboratory model of limestone neutralization column.

Step 1. Fill neutralization columns with limestone (after washing and screening) to depths of 1, 2, 3, 4, and 5 ft (5 columns).

Step 2. Adjust upward flow rate of acid wastewater for each column. Flow rates varying from 50 to 1000 gal/(hr)(ft^2) are used.

Step 3. Check effluent pH from each column at each flow rate utilized until it is stabilized.

Step 4. After each run replace limestone used in the columns.

Step 5. Plot terminal pH as a function of rate of flow [gal/(hr)(ft^2)] for each depth of limestone. A typical plot of this type is shown in Fig. 3.34.

Purposes of the design procedure are (1) to select the most economical height of column for a specified pH of the effluent. This is the column height corresponding to a maximum allowable flow rate, expressed in terms of volume of limestone utilized, i.e., gal of liquor/(hr)(ft^3 of limestone); and (2) to calculate annual requirement of limestone under these conditions, which corresponds to a minimum requirement of limestone.

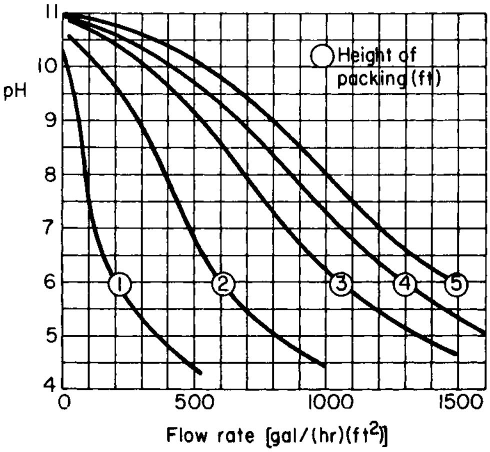

Fig. 3.34. Limestone neutralization data.

Example 3.8

It is desired to neutralize a wastewater acid stream containing 0.1 *N* HCl to a pH of 7.0 through a limestone bed. Laboratory tests with the wastewater yield data plotted in Fig. 3.34. Design a neutralization system for 100 gal/min (6000 gal/hr) of wastewater and estimate annual limestone requirement for the most economical operation.

SOLUTION

Step 1. For pH 7 read from Fig. 3.34 rates of flow [gal/(hr)(ft^2)] corresponding to each column depth. Then calculate cross-sectional area, volume of limestone bed required, and flow rate in gal/(hr)(ft^3 of limestone bed) (Table 3.12).

TABLE 3.12
Calculations for Example 3.8

(*1*) Depth (ft)	(*2*) Flow rate [gal/(hr)(ft^2)] [From Fig. 3.34 for pH 7]	(*3*) Cross section $(3) = \frac{6000\ \text{gal/hr}}{(2)}$	(*4*) Volume of limestone (ft^3) $(4) = (1) \times (3)$	(*5*) Flow rate [gal/(hr)(ft^3)] $(5) = \frac{6000\ \text{gal/hr}}{(4)}$
1.0	118	51	51	118
2.0	492	12.2	24.4	246
3.0	845	7.1	21.3	282
4.0	1047	5.73	22.9	262
5.0	1200	5.0	25.0	240

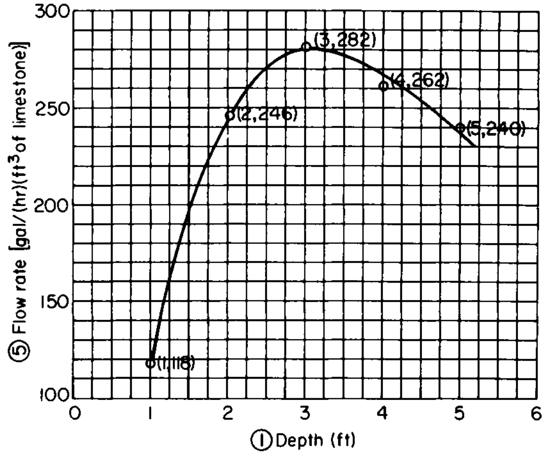

Fig. 3.35. Determination of optimum bed depth.

Step 2. Plot flow rates [gal/(hr)(ft^3)] from column (*5*) of Table 3.12 vs. depths [column (*1*)]. This plot is shown in Fig. 3.35 and indicates that a 3-ft limestone bed is the most economical, corresponding to a maximum on the curve. The cross-sectional area in this case (see Table 3.12 for a depth of 3.0 ft) is 7.1 ft^2, corresponding to a diameter of 3.0 ft.

Step 3. Calculate the lb of acid to be neutralized per day (acid contained in the 6000 gal/hr of the 0.1 *N* solution of HCl). Since this is a dilute acid solution, calculation is based on the density of water, taken as 8.34 lb/gal. A 0.1 *N* solution of HCl contains 3.65 g/liter of HCl, or approximately 3.65 lb of acid per 1000 lb of solution. Therefore

$$\text{Mass flow rate: } 6000 \text{ gal/hr} \times 24 \text{ hr/day} \times 8.34 \text{ lb/gal} = 1.2 \text{ Mlb/day}$$

$$\text{Acid content: } 3.65 \times (1{,}200{,}000/1000) = 4380 \text{ lb/day}$$

Step 4. Calculate limestone ($CaCO_3$) required. The neutralization equation is

$$2HCl + CaCO_3 \rightarrow CaCl_2 + CO_2 + H_2O$$

Molecular weight: $(2 \times 36.5 = 73)$ (100)

Therefore limestone required is

$$4380 \times 100/73 = 6000 \text{ lb/day of limestone}$$

This is the theoretical amount of limestone assuming 100% reactivity. For design purposes, assume an 80% reactivity. Limestone required is then

$$6000/0.8 = 7500 \text{ lb/day or } 2.738 \times 10^6 \text{ lb/year}$$

5.6. SLURRIED LIME TREATMENT

5.6.1. Equipment for Slurried Lime Systems

Slurried lime is the most commonly used reagent for neutralization of acid wastewaters, the low cost of lime being the main reason for its widespread utilization. A flowsheet of a two-stage slurried lime neutralization system is shown in Fig. 3.36.

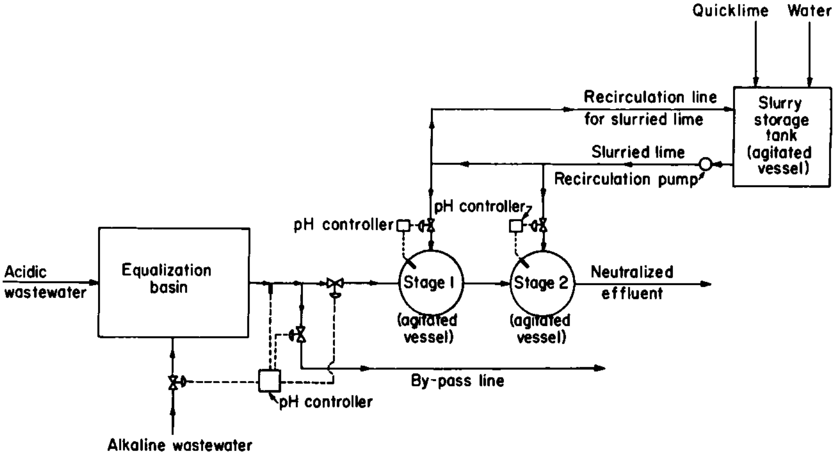

Fig. 3.36. Flow diagram of a two-stage slurried lime neutralization system.

Stepwise addition of lime is recommended. For highly acidic wastes a minimum of two stages is desirable, the first (bulk neutralization) to raise pH to a value of 3.0–3.5, and the second (fine tuning) to adjust pH to desired effluent value. Sometimes a third stage is desirable.

Automatic control of this process is not simple because the relationship between pH and amount of lime added is highly nonlinear, particularly in the vicinity of the neutralization point (pH 7). This is appreciated by inspection of a typical neutralization curve of an industrial wastewater, shown in Fig. 3.37.

In the vicinity of the neutralization point, the pH becomes exceedingly sensitive to small additions of lime, varying in an ordinary operation of slurried lime systems at a rate as fast as one pH unit per minute. Also, fluctuation in flow rate of influent complicates operation of the process. Use of an equalization tank is indicated to dampen fluctuations, as shown in Fig. 3.36. A relatively small amount of reagent is thoroughly mixed with a large liquid volume in a short time interval. Mechanical mixers are provided for this purpose.

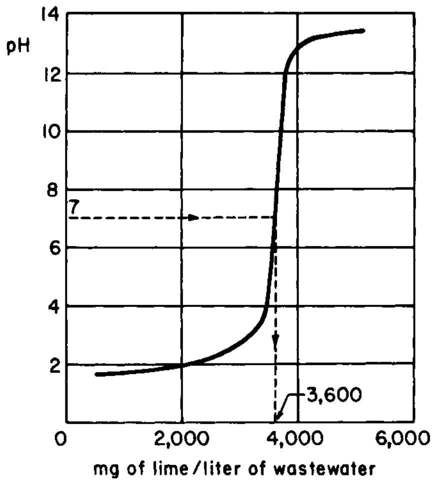

Fig. 3.37. Typical neutralization curve for an industrial wastewater.

5.6.2. Design Procedure for Slurried Lime Neutralization Systems

The procedure recommended by Eckenfelder and Ford [3] is summarized in this section and illustrated by a numerical example. Basic information required is (1) neutralization curve for the wastewater (see Fig. 3.37), and (2) power consumption data, i.e., a curve of level of agitation vs. detention time for a desired terminal pH (see Fig. 3.38). This basic information is obtained by simple laboratory procedures [3].

Purposes of the design procedure illustrated by Example 3.9 are (1) to select the number of stages of neutralization and to size neutralization reactors, and (2) to select appropriate mixing equipment.

Example 3.9

200 gal/min of an acidic industrial wastewater are neutralized to pH 7.0. From laboratory tests, the neutralization curve is plotted (Fig. 3.37). Also, a curve of level of agitation vs. detention time is obtained for this neutralization (Fig. 3.38). Design a lime slurry neutralization system.

SOLUTION

Step 1. From the neutralization curve in Fig. 3.37, lime slurry requirement for neutralizing the wastewater to a pH of 7 is read as 3600 mg/liter.

Step 2. Calculate lime slurry requirement for 200 gal/min wastewater flow. Since

$$3600 \text{ mg lime/liter waste} = 3600 \times 10^{-6} \text{ lb lime/lb waste}$$

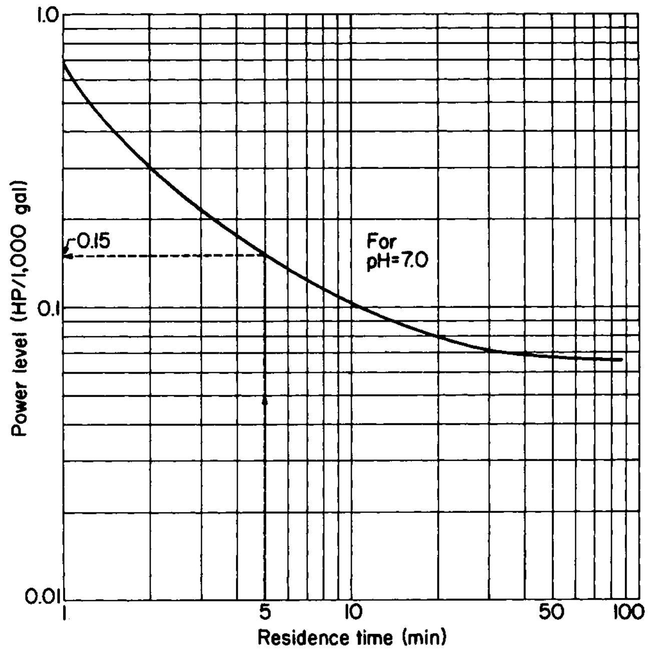

Fig. 3.38. Level of agitation vs. detention time.

then

$$200 \text{ gal waste/min} \times 60 \text{ min/hr} \times 24 \text{ hr/day} \times 8.34 \text{ lb waste/gal waste}$$
$$\times\, 3600 \times 10^{-6} \text{ lb lime/lb waste} = 8647 \text{ lb lime/day}$$

Step 3. Neutralize acid in two steps as indicated by flow diagram in Fig. 3.36, the first stage for bulk neutralization, and the second for fine tuning.

Step 4. Select a detention time and size the reactors.

$$\text{Volume of reactor (gal)} = Q \text{ (gal/min)} \times t \text{ (min)} \tag{3.63}$$

Selection of the optimum detention time is arrived at by an economical balance.

1. Assume a residence time (usually a value between 5 and 10 min).
2. Size the reactor [Eq. (3.63)].
3. From Fig. 3.38 for the assumed residence time, determine power level required.
4. Select mixers (as shown in Step 5) from knowledge of power level [item (3)].
5. Estimate total costs (capital and operating) corresponding to this assumed residence time. Main items in the column of capital costs are the reactors themselves (and auxiliary equipment) and the mixers. Energy requirement is the variable item among operating costs.

6. Repeating steps (1)–(5), a curve of total cost per year vs. a series of selected residence times is plotted. This curve passes through a minimum which corresponds to the optimum detention time. Capital costs are expressed on a yearly basis by estimating equipment life and utilizing the current value for interest rate.

Assume for Example 3.9 that optimum residence time is estimated by this procedure as 5 min. Then

$$\text{Volume of each reactor (gal)} = 200 \text{ gal/min} \times 5 \text{ min} = 1000 \text{ gal}$$

or

$$1000 \text{ gal} \times \text{ft}^3/7.48 \text{ gal} = 134 \text{ ft}^3$$

Selecting a reactor depth of 5 ft, required cross-sectional area is

$$\text{Area} = 134 \text{ ft}^3/5 \text{ ft} = 26.8 \text{ ft}^2$$

corresponding to a diameter of 5.84 ft.

Step 5. Select mixers. From Fig. 3.38, power level required for 5-min detention time is

$$0.15 \text{ HP}/1000 \text{ gal}$$

Since each tank has a volume of 1000 gal, specify one 0.15-HP mixer for each tank.

5.7. NEUTRALIZATION OF ALKALINE WASTES

In principle, any strong acid can be used to neutralize alkaline wastewaters. Cost considerations limit choice to H_2SO_4 (the most common) and HCl. Reaction rates are essentially instantaneous. The basic design procedure for alkaline wastes is similar to that for acidic wastes described in Section 5.6.

Flue gases containing 14% or more of CO_2 are used for neutralization of alkaline wastewaters. When bubbled through the wastewater the CO_2 forms carbonic acid, which reacts with the base. Reaction rate is slow but sufficient if pH need not be adjusted below 7 or 8. Either bubbling through a perforated pipe or using spray towers is satisfactory.

Problems

I. *Sedimentation (discrete settling).* A particle size distribution is obtained from a sieve analysis of sand particles. For each weight fraction an average settling velocity is calculated. Data [6] are presented in the following tabulation.

Settling velocity (ft/min)	Weight fraction remaining
10.0	0.55
5.0	0.46
2.0	0.35
1.0	0.21
0.75	0.11
0.50	0.03

1. Prepare a plot of fraction of particles with less than stated velocity vs. settling velocity (ft/min).

2. For an overflow rate of 100,000 gal/(day)(ft^2), calculate overall removal utilizing Eq. (3.26).

II. *Sedimentation* (*flocculent settling*). A laboratory settling analysis gave the results tabulated below.

Time	% suspended solids removed at indicated depth		
Time (min)	2 ft	4 ft	6 ft
10	40	25	16
20	54	37	28
30	62	47	37
45	71	56	46
60	76	65	53

1. Perform analysis of the data and arrive at curves for % SS removal vs. detention time (min), and % SS removal vs. overflow rate [gal/(day)(ft^2)].

2. If the initial concentration of the slurry is 430 ppm, design a settling tank (i.e., calculate diameter and effective depth of the tank) to remove 70% of the suspended solids for a 1 Mgal/day flow.

3. What removal is attained if flow if increased to 2 Mgal/day?

4. For the flow of 1 Mgal/day calculate daily accumulation of sludge in lb/day and average pumping rate in gal/min. Assume sludge concentration to be 1.5% solids ($\approx$ 15,000 mg/liter).

III. *Sedimentation* (*zone settling*). It is desired to design a secondary settling tank to produce an underflow concentration of 15,000 mg/liter from a mixed liquor solids content of 3750 mg/liter in the influent. Wastewater flow is 2.0 Mgal/day. Calculate clarifier area required. Data below are obtained in a laboratory test of the slurry.

t (min)	Interface height, H (ml)
0	1000
2	920
4	840
6	760
8	690
10	600
15	400
20	300
25	280
30	270

IV. *Flotation*. A pilot-plant flotation operation indicated optimum air/solid ratio to be 0.04 lb air/lb of solids.

1. If a wastewater to be treated has 250 ppm suspended solids, compute the % recycle to be pressurized to 60 psia at 20°C. Take $f = 0.68$.

2. For a wastewater flow of 1.0 Mgal/day and an overflow rate of 4.0 gal/(min)(ft^2) compute surface area required.

V. *Neutralization.* For the 3-ft limestone bed designed in Example 3.8 prepare a plot of volume of limestone required vs. a range of selected values of pH for the effluent (select pH = 5, 6, 7, 8, 9, and 10).

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4

Theory and Practice of Aeration in Wastewater Treatment

1. Introduction

This chapter is concerned with the transfer of oxygen from air to a wastewater subjected to biological aerobic treatment. Knowledge of the rate of oxygen transfer is essential for specification of aerators to be utilized in the process. Theory for oxygen transfer is discussed, and the determination of transfer coefficients from laboratory experiments described. Common types of aerators and the procedure for specifying aerator systems for aerobic wastewater treatment processes are also described.

The best known explanation for the mechanism of gas transfer to a liquid is given by the two-film theory. According to this theory, it is the presence of two films, one liquid and one gas, at the gas–liquid interface which provides the resistance to the passage of gas molecules from the bulk of the gas phase to that of the liquid phase.

For gases of high solubility in the liquid phase, e.g., absorption of SO_2 by water, the major resistance to absorption is that offered by the gas film. For gases of low solubility in the liquid phase, e.g., absorption of oxygen by an aqueous liquor, the controlling resistance resides in the liquid film. For intermediate solubilities, both films offer significant resistance.

Oxygen saturation values (C_s) for distilled water at standard conditions (1 atm) are presented in Table 4.1. For a more complete table with 1°C increments, consult Ref. [7].

TABLE 4.1
Oxygen Saturation Values (C_s) for Distilled Water at Standard Conditions (1 atm) [7]

Temperature (°C)	Temperature (°F)	O_2 (mg/liter)
0	32.0	14.6
5	41.0	12.8
10	50.0	11.3
15	59.0	10.2
20	68.0	9.2
25	77.0	8.4
30	86.0	7.6
35	95.0	7.1
40	104.0	6.6
45	113.0	6.1
50	122.0	5.6

2. Steps Involved in the Oxygen-Transfer Process

The process of oxygen transfer from a gaseous to an aqueous phase occurs in three steps.

Step 1. Saturation of the liquid surface between the two phases (let C_s be this saturation concentration of oxygen). This rate of oxygen transfer is very rapid since the resistance of the gas film is negligible, and thus Step 1 is never the controlling one.

Step 2. Passage of the oxygen molecules through the liquid interface film by molecular diffusion. At very low mixing levels the rate of oxygen absorption is controlled by Step 2. At higher turbulence levels, the interface film is broken up and the rate of renewal of the film controls the absorption of oxygen. Surface renewal rate is the frequency at which liquid with an oxygen concentration C_L (oxygen concentration in the bulk of the liquid phase) replaces that from the interface with an oxygen concentration equal to C_s.

Step 3. Oxygen is transferred to the bulk of the liquid by diffusion and convection.

3. Oxygen-Transfer Rate Equation

The basic equation for oxygen-transfer rate is

$$N = K_L A(C_s - C_L) \tag{4.1}$$

where N is the mass of oxygen transferred per unit time (lb O_2/day); K_L the liquid film coefficient [lb O_2/(day)(ft^2)(unit ΔC)]; A the interfacial area for transfer (ft^2); C_s the saturation concentration of oxygen (mg/liter); and C_L the concentration of oxygen in the body of the liquid (mg/liter).

Equation (4.1) is usually rewritten in concentration units by dividing by volume V of the system. Then [Eq. (4.2)]

$$N/V = dC_L/dt = K_L(A/V)(C_s - C_L) = K_L a(C_s - C_L) \tag{4.2}$$

where $a = A/V$ = interfacial area per unit volume (ft^2/ft^3); and $K_L a$ is the overall coefficient of oxygen transfer [lb O_2/(day)(ft^3)(unit ΔC)].

In the determination of the mass-transfer coefficient, the overall coefficient $K_L a$ is obtained without attempting to separate the factors K_L and a. It is admittedly impossible to measure the interfacial area A.

The difference $(C_s - C_L)$ between saturation value and actual concentration of oxygen (C_L) in the body of the liquid phase is called oxygen deficit (hence denoted as OD). For aerobic treatment processes designed for removal of organic BOD, the range for operating dissolved oxygen level C_L is between 0.5 and 1.5 mg/liter. When nitrification is to be achieved, dissolved oxygen level is in excess of 2.0 mg/liter.

4. Determination of the Overall Mass-Transfer Coefficient K_La by Unsteady State Aeration of Tap Water

Studies of the transfer coefficient are usually made on tap water and then corrected for the wastewater, as described in Section 11 of this chapter. The procedure more commonly used for determination of K_La is the unsteady state aeration of tap water. The four steps involved in this determination are given below.

Step 1. Deoxygenate the water to an essentially zero concentration of dissolved oxygen. This is done by addition of deoxygenation chemicals, the most commonly used being sodium sulfite (Na_2SO_3). Cobalt chloride ($CoCl_2$) is added as a catalyst for the deoxygenation reaction.

$$Na_2SO_3 + \tfrac{1}{2}O_2 \xrightarrow{CoCl_2} Na_2SO_4$$

The stoichiometric ratio is

$$Na_2SO_3/\tfrac{1}{2}O_2 = 126/16 = 7.9$$

This means that theoretically 7.9 ppm of Na_2SO_3 are required to remove 1 ppm of DO. Based on the DO of the test tap water, the approximate Na_2SO_3 requirements are estimated (a 10–20% excess is used). Sufficient cobalt chloride is added to provide a minimum Co^{2+} concentration of 1.5 ppm. An alternative deoxygenation procedure consists of removal of dissolved oxygen by purging with nitrogen gas.

Step 2. After DO concentration becomes essentially zero, start aeration, measuring the increasing concentrations of DO at selected time intervals. Since DO concentration increases with time, this method is termed unsteady state aeration. Steady state methods, in which DO concentration is kept constant, are discussed in Section 7. Dissolved oxygen measurements are preferably performed by instrumental methods. A properly calibrated galvanic cell oxygen analyzer and probe is the most reliable method. Experimental determination of DO by this technique is described in Section 2.3.1 of Chapter 2. Chemical analysis of dissolved oxygen (Winkler method) is also employed [7]. The aeration device is located at the center of the test basin. When circular basins are employed, baffles are placed at the quarter points of the basin, as indicated in Fig. 4.1, in order to prevent vortexing.

When testing is performed in a circular tank for pilot or full scale tests, sampling depths for the DO determinations are 1 ft from the surface and 1 ft from the bottom, at the mid- and end-points of the radii trisecting the basin.

This yields a total of 12 sampling points, as illustrated by Fig. 4.2. Samples from various test locations are analyzed for DO, and the results are averaged and recorded for the particular sampling time.

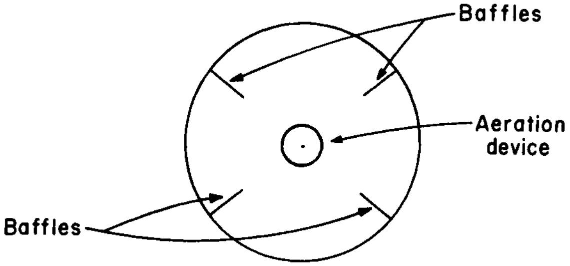

Fig. 4.1. Baffle arrangement.

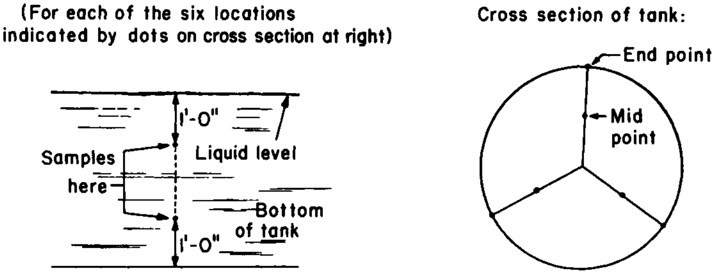

Fig. 4.2. Location of sampling points.

Step 3. Results are tabulated as shown by sample data in Table 4.2.

TABLE 4.2
Data for Example 4.1

(*1*) Test time (min)	(*2*) C_L (mg/liter)	(*3*) $C_s - C_L$ (mg/liter)[a]
0	0.2	10.0
10	2.6	7.6
20	4.8	5.4
30	6.0	4.2
40	7.1	3.1
50	7.8	2.4
60	8.5	1.7

[a] $C_s = 10.2$ mg/liter.

Step 4. From Eq. (4.2) it follows that a plot of $(C_s - C_L)$ vs. time in semilog scale yields a straight line, the slope of which equals $(-K_La)$. Take Eq. (4.2),

$$dC_L/dt = K_La(C_s - C_L)$$

Separating variables, integrating, and assuming K_La to be independent of the time of sampling [Eq. (4.3)]:

$$\ln(C_s - C_L) = -K_Lat + \text{const.} \tag{4.3}$$

The plot of $\ln(C_s - C_L)$ vs. time is shown in Fig. 4.3 for data in Table 4.2. For accuracy the straight line is plotted by the least-squares method.

Determination of the overall mass-transfer coefficient K_La by the method of unsteady state aeration of tap water is illustrated by Example 4.1.

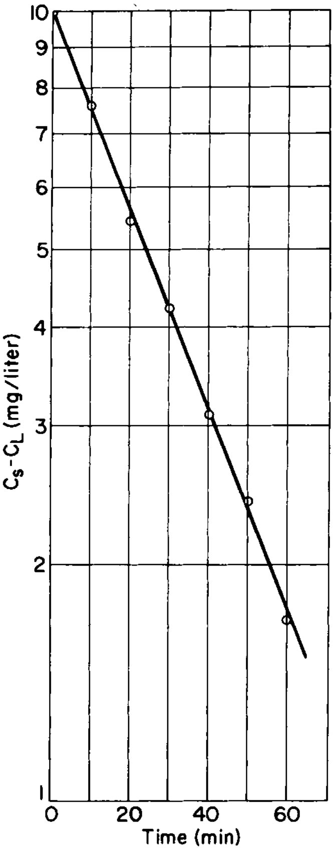

Fig. 4.3. Determination of K_La (Example 4.1).

Example 4.1

Data presented in Table 4.2 are obtained by utilizing an 8-HP surface aerator in a 150,000 gal circular test tank under the following conditions:

Water temperature: 15°C

Atmospheric pressure: 28 in. Hg

C_s: 10.2 mg/liter (at 15°C, $P = 28$ in. Hg)

Test water is deoxygenated using sodium sulfite and a cobalt catalyst. Calculate

1. Chemical requirements (lb Na_2SO_3/lb liquor) to deoxygenate test water with 9 ppm of DO and total lb Na_2SO_3
2. $CoCl_2$ requirements (lb)
3. Value of K_La lb O_2/(hr)(ft^3)Δ(mg/liter)

SOLUTION

Step 1. Establish the chemical requirements. Theoretically, 7.9 ppm of Na_2SO_3 are required to remove 1 ppm DO. Thus

$$7.9 \times 9.0 = 71.1 \text{ ppm of } Na_2SO_3$$

Utilizing a 20% excess

$$20/100 \times 71.1 = 14.22 \text{ ppm (take 15 ppm excess)}$$

Requirements

Theoretical	71.1 ppm
Excess	15.0 ppm
	86.1 ppm of Na_2SO_3

$$\text{Requirements} = 86.1 \text{ ppm} = 86.1 \times 10^{-6} \text{ lb } Na_2SO_3/\text{lb liquor}$$

$$\text{Total required: } 86.1 \times 10^{-6} \text{ lb } Na_2SO_3/\text{lb liquor} \times 8.34 \text{ lb liquor/gal liquor}$$
$$\times\ 150{,}000 \text{ gal liquor} = 108 \text{ lb of } Na_2SO_3$$

Step 2. Determine the $CoCl_2$ requirements (lb).

Basis: 1.5 ppm of Co^{2+}

Molecular weight of $CoCl_2$: 130

Atomic weight of Co: 59

Thus

$$1.5 \text{ ppm } Co^{2+} \text{ or } 1.5 \times 130/59 = 3.3 \text{ ppm } CoCl_2 = 3.3 \times 10^{-6} \text{ lb } CoCl_2/\text{lb liquor}$$

Therefore, the lb of $CoCl_2$ required are

$$3.3 \times 10^{-6} \text{ lb } CoCl_2/\text{lb liquor} \times 8.34 \text{ lb liquor/gal liquor} \times 150{,}000 \text{ gal liquor}$$

$$= 4.13 \text{ lb } CoCl_2 \text{ (minimum)}$$

Take 5 lb $CoCl_2$.

Step 3. Ascertain the value of $K_L a(\text{hr}^{-1})$. By plotting in semilog scale graph column (*3*) vs. column (*1*) of Table 4.2 one obtains Fig. 4.3. Then at 15°C

$$\begin{aligned} K_L a &= -(\text{slope}) = -2.303\,[(\log 10 - \log 3.1)/(0-40)] \times 60 \\ &= 1.76 \text{ lb } O_2/(\text{hr})(\text{ft}^3)\,\Delta(\text{mg/liter}) \end{aligned}$$

5. Integration of the Differential Equation for Oxygen Transfer between Limits

Integration of Eq. (4.2) between times t_1 and t_2, corresponding to DO concentrations $C_{L,1}$ and $C_{L,2}$, yields

$$\ln[(C_s - C_{L,2})/(C_s - C_{L,1})] = -K_L a(t_2 - t_1)$$

or

$$K_L a = 2.303 \log[(C_s - C_{L,1})/(C_s - C_{L,2})]/(t_2 - t_1) \tag{4.4}$$

From Eq. (4.4) the value of $K_L a$ is calculated from only two experimental determinations of DO. However, it is preferable to utilize the semilog linear plot method with several experimental points, since this permits statistical averaging of errors.

6. Unsteady State Aeration of Activated Sludge Liquor

In aeration of activated sludge liquor, oxygen utilization (respiration rate) by the microorganisms is taken into account. Equation (4.2) is modified as follows:

$$dC_L/dt = K_L a(C_{SW} - C_L) - r \tag{4.5}$$

where r is the rate of oxygen utilization by the microorganisms; C_{SW} the saturation concentration of oxygen for the wastewater; and C_L the operating concentration of dissolved oxygen in the aerator liquor. Values of dC_L/dt are obtained by plotting C_L (measured by DO tests) vs. time and determining slopes at selected time intervals (Fig. 4.4).

Equation (4.5) is rearranged to yield

$$dC_L/dt = (K_L aC_{SW} - r) - K_L aC_L \tag{4.6}$$

Equation (4.6) indicates that a plot of dC_L/dt (values of slopes obtained from Fig. 4.4) vs. C_L yields a straight line, as indicated in Fig. 4.5. The slope of this line yields K_La, and respiration rate r is determined from the intercept.

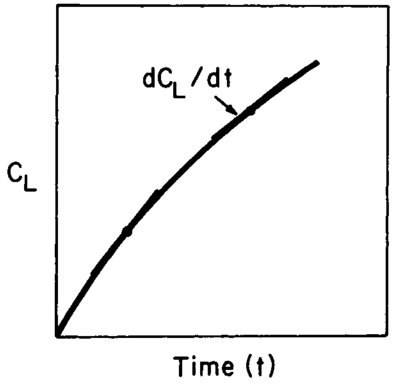

Fig. 4.4. Determination of dC_L/dt.

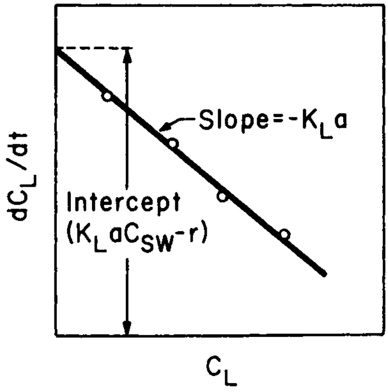

Fig. 4.5. Determination of K_La (unsteady state aeration of activated sludge liquor).

7. Steady State Determination of K_La for the Activated Sludge Liquor

Activated sludge liquor is aerated at a rate just sufficient to supply the oxygen required for respiration of the microorganisms. In Eq. (4.5) $dC_L/dt = 0$, and therefore

$$K_La = r/(C_{SW}-C_L) \tag{4.7}$$

Respiration rate is measured (e.g., by the Warburg respirometer) when the dissolved oxygen concentration becomes stabilized. Equation (4.7) yields K_La.

8. Oxygenation Capacity (OC)

In evaluation of an aerator the oxygen transferred is estimated under standard conditions (SC), corresponding to a temperature of 20°C and standard atmospheric pressure. Rate of oxygen transferred by the aerator is reported as its oxygenation capacity (OC), which is defined as the rate of oxygen transfer dC/dt at an initial oxygen concentration $C_L = 0$ and standard conditions. From Eq. (4.2)

$$\text{OC [lb O}_2\text{/(hr)(unit volume)]} = dC_L/dt = (K_La)_{20^\circ\text{C}}(C_s - 0) = (K_La)_{20^\circ\text{C}}C_s \tag{4.8}$$

or if V is the volume of aeration basin,

$$\text{OC (lb O}_2\text{/hr)} = (K_La)_{20^\circ\text{C}}C_sV \tag{4.9}$$

Since the test is performed under conditions other than standard, the value of K_La obtained is corrected for temperature and pressure before application of Eq. (4.9). These corrections are discussed in Section 9.

9. Corrections for K_La and Oxygenation Capacity (OC) with Temperature and Pressure

9.1. TEMPERATURE CORRECTION

The oxygen-transfer coefficient K_La increases with temperature. The following temperature correction is used to determine K_La at 20°C:

$$K_La_{(T)} = K_La_{(20^\circ\text{C})} \times 1.024^{(T-20)} \tag{4.10}$$

where T is the temperature in °C.

9.2. PRESSURE CORRECTION

A barometric correction for oxygen saturation value C_s in Eq. (4.8) or (4.9) is applied as indicated by Eq. (4.11), which assumes that C_s is directly proportional to the barometric pressure.

$$C_s\text{ (corrected)} = C_s\text{ (test)} \times 29.92\text{ in. Hg/(in. Hg at test conditions)} \tag{4.11}$$

Since oxygen saturation is related to partial pressure of oxygen in the gas phase (Henry's law), a correction is made for saturation in submerged aeration devices (bubble aeration), where partial pressure at the point of discharge exceeds atmospheric pressure due to hydrostatic pressure. Oldshue [5] proposed the following correction:

$$C_{s,m} = C_{s,s}[(P_b/29.4) + (O_t/42)] \tag{4.12}$$

where $C_{s,m}$ is the saturation of oxygen at aeration tank mid-depth (mg/liter); $C_{s,s}$ the saturation of oxygen at standard conditions (mg/liter); P_b the pressure (psia) at the depth of air release; and O_t the oxygen in exit gas (%).

For aeration, $O_t = 21\%$ of O_2, and Eq. (4.12) yields

$$C_{s,m} = C_{s,s} \times [(P_b/29.4) + 0.5] \tag{4.13}$$

Thus, the value for the oxygenation capacity for surface aerators (no hydrostatic correction required) is given by Eq. (4.9), which is modified as follows [Eq. (4.14)]:

$$\mathrm{OC} = (K_La)_{20^\circ\mathrm{C}} \times C_s\,(\text{corrected}) \times V \tag{4.14}$$

where $(K_La)_{20^\circ\mathrm{C}}$ is calculated from Eq. (4.10), and C_s (corrected) is given by Eq. (4.11). Therefore, for surface aerators

$$\mathrm{OC} = K_La_{(T)} \times 1.024^{(20-T)} \times C_s\,(\text{test}) \times \frac{29.92 \text{ in. Hg}}{(\text{in. Hg at test conditions})} \times V \tag{4.15}$$

For bubble aerators, one utilizes Eq. (4.16).

$$\mathrm{OC} = (K_La)_{20^\circ\mathrm{C}} \times C_{s,s} \times V \tag{4.16}$$

where $(K_La)_{20^\circ\mathrm{C}}$ is calculated from Eq. (4.10) and $C_{s,s}$ from Eq. (4.12) [or Eq. (4.13)]. Therefore, for bubble aerators

$$\mathrm{OC} = K_La_{(T)} \times 1.024^{(20-T)} \times \frac{C_{s,m}}{[(P_b/29.4)+0.5]} \times V \tag{4.17}$$

Example 4.2

For the surface aerator in Example 4.1, calculate

1. Value of K_La corrected to 20°C
2. Value of C_s corrected to normal atmospheric pressure
3. Oxygenation capacity (lb O_2/hr)

SOLUTION

Step 1. $(K_La)_{20^\circ\mathrm{C}}$ is calculated from Eq. (4.10).

$$(K_La)_{20^\circ\mathrm{C}} = 1.76 \times (1.024)^{(20-15)} = 1.98 \text{ hr}^{-1}$$

Step 2. C_s (corrected) is calculated from Eq. (4.11).

$$C_s\,(\text{corrected}) = 10.2(29.92/28.0) = 10.9 \text{ mg/liter} = 10.9 \times 10^{-6} \text{ lb } O_2/\text{lb liquor}$$

Step 3. Oxygenation capacity is calculated from Eq. (4.14).

$$\mathrm{OC} = 1.98\ \mathrm{l/hr} \times 10.9 \times 10^{-6}\ \mathrm{lb\ O_2/lb\ liquor} \times 150{,}000\ \mathrm{gal} \times 8.34\ \mathrm{lb\ liquor/gal}$$
$$= 27.0\ \mathrm{lb\ O_2/hr}$$

10. Transfer Efficiency of Aeration Units

Transfer efficiency (TE) of aeration units is commonly expressed in terms of mass of oxygen actually transferred per (HP × hr) of work expenditure, i.e. [Eq. (4.18)],

$$\mathrm{TE} = \mathrm{lb\ O_2\ transferred/(HP \times hr)} \tag{4.18}$$

Sometimes, nominal HP (nameplate HP) of the aerator is utilized for simplicity in evaluating TE. It is more accurate to base calculation upon actual HP (blade HP) measured during the test by a watt meter or an energy counter. When the power factor (cos PF) is known, blade HP is calculated from Eq. (4.19) [2].

$$\begin{aligned}\text{Blade HP} &= (\text{line voltage}) \times (\text{line amperage}) \times [\cos \mathrm{PF}\,(3)^{1/2}] \\ &\quad \times (1/746) \times (\text{motor efficiency}) \times (\text{gear efficiency})\end{aligned} \tag{4.19}$$

where 1/746 is the conversion factor HP/W. Values of TE up to 7 lb O_2/(HP × hr) are reported for surface aerators, although for most units the values of TE range from 2 to 4 lb O_2/(HP × hr). For turbine aerators the usual range is 2–3 lb O_2/(HP × hr).

Example 4.3

For the aerator in Examples 4.1 and 4.2, report aerator efficiency in terms of nameplate HP and blade HP. The following data are available in addition to those from Examples 4.1 and 4.2:

Drawn voltage: 225 V (average)
Amperage: 20 A (average)
cos PF (measured): 0.85
Motor efficiency (estimated): 90%
Gear efficiency (estimated): 90%

SOLUTION Oxygen transferred had been determined in Example 4.2 as 27.0 lb O_2/hr. Therefore

$$\mathrm{TE} = 27.0/8.0 = 3.38\ \mathrm{lb\ O_2/HP \times hr} \qquad \text{(nameplate)}$$

From Eq. (4.19)

$$\text{Blade HP} = (225)(20)[0.85 \times (3)^{1/2}](1/746)(0.9)(0.9) = 7.19$$

Then

$$\mathrm{TE} = (27.0/7.19) = 3.76\ \mathrm{lb\ O_2/HP \times hr} \qquad \text{(blade)}$$

11. Effect of Wastewater Characteristics on Oxygen Transfer

When oxygen is supplied for aerobic biological treatment of wastewater, it is necessary to define a correction factor which relates oxygen transfer to the nature of the waste. This correction factor α relates the overall mass-transfer coefficient ($K_L a$) of the wastewater to that of the tap water [Eq. (4.20)].

$$\alpha = K_L a \,(\text{wastewater})/K_L a \,(\text{tap water}) \tag{4.20}$$

There are many variables which affect the magnitude of α. These include (1) temperature of the mixed liquor; (2) nature of the dissolved organic and mineral constituents; (3) level of agitation of aeration basin, usually expressed in terms of HP per 1000 gal of basin volume; (4) characteristics of the aeration equipment; and (5) liquid depth and geometry of aeration basin.

The temperature effect is attributable to temperature dependence of the liquid film coefficient K_L. Figure 4.8 illustrates typical temperature effect on values of α.

Since the nature of dissolved organic and mineral constituents affects α, its value is expected to increase during the course of biological oxidation, because dissolved organic materials affecting the transfer rate are removed in the biological process. A typical situation is shown in Fig. 4.6. As the final effluent approaches purity of tap water, the value of α approaches unity asymptotically.

Effect of mixing intensity in aeration basin (usually expressed in terms of HP/1000 gal) is illustrated by Fig. 4.7, which is a typical curve for a wastewater containing surface-active agents. As explained in Section 2, at low mixing intensities the rate of oxygen transfer is controlled by the passage of the oxygen molecules through the liquid interface film by molecular diffusion. The presence of surface-active agents inhibits molecular diffusion through

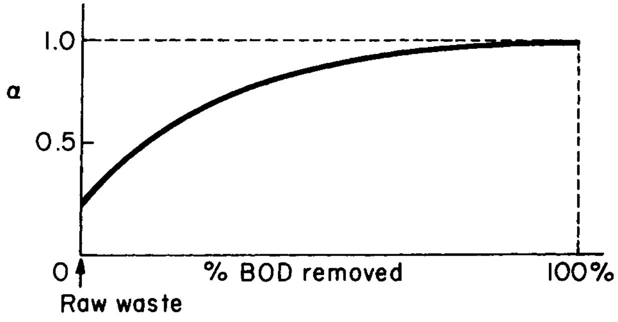

Fig. 4.6. Plot of α *vs. % BOD removed.*

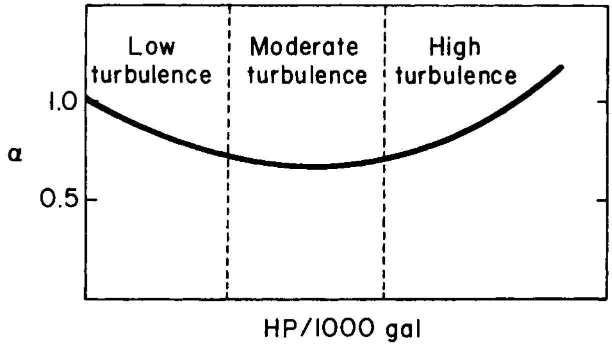

Fig. 4.7. Plot of α *vs. mixing intensity.*

the interface, and thus α decreases. At high mixing intensities, however, the oxygen transfer is controlled by the rate of surface renewal, and thus at conditions of high turbulence, α increases with the degree of mixing intensity. Figure 4.8 shows simultaneously the effect of temperature and mixing intensity on the value of α for a typical wastewater.

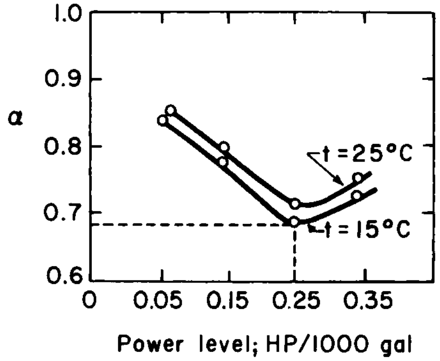

Fig. 4.8. Plot of α *vs. power level at two different temperatures* [3].

An interesting illustration of the effect of characteristics of aeration equipment on the value of α is bubble aeration (air diffusion or turbine units) in the presence of surface-active agents. Presence of these agents decreases bubble size, and thus increases interfacial area per unit volume.* Under these conditions the $K_L a$ value of the wastewater usually increases, because increase in a exceeds decrease of K_L caused by the surface barrier. This increase in $K_L a$ of the wastewater results in a corresponding increase of α [Eq. (4.20)].

---

* Since $a = A/V = k_1 r^2/k_2 r^3 = K(1/r)$, thus as radius ($r$) of the bubble decreases, a (and therefore $K_L a$) increases.

12. Laboratory Determination of Oxygen-Transfer Coefficient α

This determination is based directly on the definition of α given by Eq. (4.20). Values of $K_L a$ of wastewater and tap water are determined as described in Section 4. It is interesting to make parallel determinations for wastewater and tap water at different mixing intensities and at different temperatures, in order to obtain curves such as the ones shown in Fig. 4.8.

13. Classification of Aeration Equipment—Oxygen-Transfer Efficiency

Aeration equipment commonly employed in wastewater treatment is classified into three categories: (1) air diffusion units, (2) turbine aeration units, and (3) surface aeration units. When comparing aeration devices or evaluating absorption of oxygen in various wastewaters, it is useful to consider the oxygen-transfer efficiency, which is defined as [Eq. (4.21)]

$$\varepsilon = \frac{\text{weight of } O_2 \text{ absorbed/unit time}}{\text{weight of } O_2 \text{ supplied/unit time}} \times 100 \qquad (4.21)$$

This definition is not applicable to surface aeration units since the oxygen supplied comes from surrounding air, and thus it is not possible to determine the weight of oxygen supplied per unit time. In aerobic biological processes, aerators perform two basic functions: (1) provision of the required oxygen transfer needed for oxidation of organic matter in the wastewater; and (2) maintenance of an adequate level of agitation in the biological reactor, in order to yield relatively uniform concentrations of dissolved oxygen and biological mass throughout.

For the activated sludge process, most of the power expenditure by the aerators is for the purpose of providing oxygen transfer. For large volume biological units (namely, aerated lagoons), the larger share of the power expenditure is for maintenance of an adequate level of agitation. Characteristics and specifications for the three categories of aeration equipment are discussed individually in the next three sections.

14. Air Diffusion Units

14.1. TYPE 1. FINE BUBBLE DIFFUSERS

Small orifice diffusion units such as porous media, plates, or tubes are constructed of silicon dioxide (SiO_2) or aluminum oxide (Al_2O_3) grains, held in a porous mass with a ceramic binder. Other units employed consist of Saran, Dacron, or nylon-wrapped tubes.

Small bubbles, having a high surface area per unit volume, provide good oxygen–liquid contact, leading to relatively high values of the oxygen-transfer efficiency. Diameter of the bubbles released from these diffusers are 2.0–2.5 mm, the oxygen-transfer efficiency depending on bubble size. (ε's from 5 to 15% are common.) Standard porous diffuser units are designed to deliver 4–15 SCFM of air per unit.

A disadvantage of small orifice diffusion units is the high maintenance costs in some applications owing to orifice clogging. Air filters are commonly used to clean and eliminate dust particles that might clog the diffusers. A sketch of a fine bubble air diffusion system is shown in Fig. 4.9.

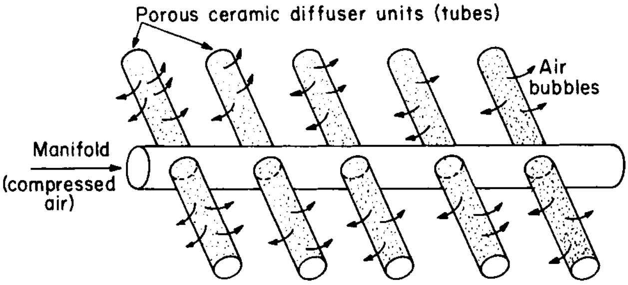

Fig. 4.9. Sketch of a fine bubble air diffusion system consisting of a series of porous ceramic diffusers.

14.2. TYPE 2. LARGE BUBBLE DIFFUSERS

These units employ large orifice or hydraulic shear devices. Large bubble units have lower oxygen-transfer efficiency than fine bubble units, since the interfacial area for oxygen transfer is considerably less. They have the advantage, however, of not requiring air filters and of generally requiring less maintenance. Sketches of two typical large bubble air diffuser units are shown in Fig. 4.10.

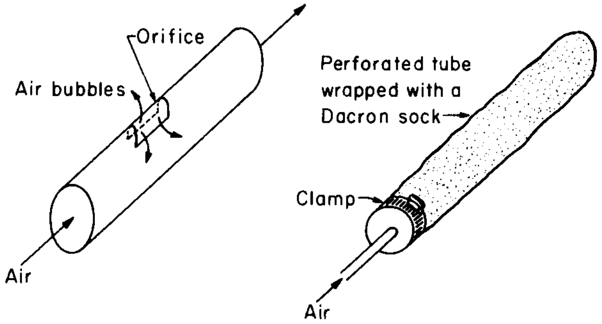

Fig. 4.10. Sketches of typical large bubble diffuser units.

14.3. PERFORMANCE OF AIR DIFFUSION UNITS

Performance data for air diffusion units are available as graphs for the lb of O_2 transferred/hr per aeration unit vs. the air flow per unit. A sample of typical data for a Saran-wrapped tube (small bubble) is shown in Fig. 4.11.

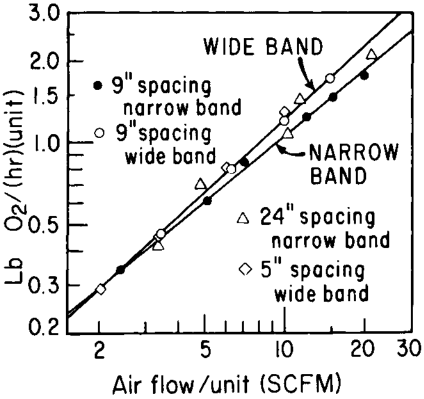

Fig. 4.11. Oxygen-transfer data from Saran tubes in water **[2].** *(Reprinted with permission, copyright by the University of Texas Press.)*

Eckenfelder [1] recommends the following empirical equation for correlating performance of air diffusion units, and tabulates values of characteristic constants for several types of diffusers.

$$N = CG_s^{(1-n)}(H^m/W^p)(C_{SW}-C_L) \times 1.024^{(T-20)} \times \alpha \qquad (4.22)$$

where N is the lb of O_2 transferred/(hr)(aeration unit); G_s the air flow (SCFM/aeration unit; SCFM are measured at 1 atm and 60°F); H the liquid depth (ft); W the aeration tank width (ft); C_{SW} the saturation concentration of DO in wastewater (mg/liter, at tank mid-depth); C_L the operating concentration of DO (mg/liter, usually 0.5–1.5 mg/liter); T the temperature (°C); α the oxygen-transfer coefficient of the wastewater [defined by Eq. (4.20)]; and C, n, m, p the constants characteristic of the aeration device.

From Eq. (4.2),

$$N = K_L a V(C_{SW}-C_L) \qquad (4.23)$$

By comparing Eqs. (4.22) and (4.23) the result is

$$K_L a V = CG_s^{(1-n)}(H^m/W^p) \times 1.024^{(T-20)} \times \alpha \qquad (\text{ft}^3/\text{hr}) \qquad (4.24)$$

Term $K_L a$ has units of hr^{-1} and V is the volume of the aeration tank in ft^3. If $(C_{SW}-C_L)$ is given in mg/liter it is multiplied by a conversion factor to express it in lb/ft^3, so that N is obtained in lb/hr.

$$(C_{SW} - C_L) = \text{mg/liter} \times \text{g/1000 mg} \times \text{lb/454 g} \times 28.3 \text{ liter/ft}^3$$

$$(C_{SW} - C_L) \times 6.23 \times 10^{-5} = \text{lb/ft}^3$$

Factor 6.23×10^{-5} is conveniently included in constant C in Eq. (4.22), so that $(C_{SW} - C_L)$ enters the equation directly in mg/liter, and N is obtained in lb/hr.

14.4. DESIGN PROCEDURE FOR AERATION SYSTEMS UTILIZING AIR DIFFUSION UNITS

Fundamental information required is as follows:

1. Volume of aeration tank (V), calculated from biological reactor requirements (Chapter 5, Section 7)
2. Oxygen requirements (lb O_2/hr), also calculated from biological reactor requirements (Chapter 5, Section 7)
3. Operating temperature
4. Operating DO (C_L, mg/liter), usually 0.5–1.5 mg/liter except for nitrification units, when values above 2 mg/liter are employed
5. Oxygen-transfer coefficient α
6. Performance data for the air diffuser units [available as graphs, see Fig. 4.11, or expressed in terms of values for constants C, n, m, and p in Eq. (4.22)]

Step 1. Select a tank depth H usually between 10 and 15 ft.

Step 2. Cross-sectional area is then $A = V/H$.

Step 3. For aeration tanks with rectangular cross section, select a width W of approximately twice the tank depth. This is necessary in order to maintain adequate mixing. Then tank length $L = A/W$. For aeration tanks with circular cross section, calculate diameter from $D = (4A/\pi)^{1/2}$.

Step 4. Select air flow rate G_s per air diffusion unit. Usual range for values of G_s is 4–8 SCFM/unit and 4–16 SCFM/unit for fine and large bubble diffusers, respectively.

Step 5. Value C_{SW} [used in Eq. (4.22)] is computed at tank mid-depth from Eq. (4.12) [or Eq. (4.13)], i.e., $C_{SW} = C_{s,m}$.

Step 6. Oxygenation capacity per aeration unit [N = lb O_2 transferred/(hr)(unit)] is estimated from manufacturer's data (e.g., Fig. 4.11) or computed from Eq. (4.22).

Step 7. From oxygen requirements (lb O_2/hr) calculated in Chapter 5, Section 7 and value of N calculated in Step 6, calculate the number of aeration units required to transfer required amount of oxygen.

$$\text{No. of units} = \frac{\text{lb } O_2\text{/hr (required)}}{N}$$

Step 8. Prepare a layout of the aeration tank and determine the spacing between the aeration units. Minimum spacing is about 6 in. and maximum between 24 and 30 in. This is necessary in order to maintain solids in suspension and to minimize coalescence of air bubbles. If spacings calculated fall outside this range, double rows or adjustment in the number of units (selection of different air flow rate G_s) are made.

Step 9. Compute total air flow.

$$\text{Total air flow} = G_s \times (\text{no. units}) \qquad (\text{SCFM})$$

Step 10. Compute required horsepower of the blower.

$$\text{HP} = [(\text{pressure drop, psi}) \times (\text{SCFM}) \times 144]/(33{,}000)(\varepsilon_m)$$

where 33,000 and 144 are the conversion factors for (ftlbf/min)/HP and in.^{2}/ft^2, respectively, and ε_m is the mechanical efficiency (estimated). Usually a 6–10 psi pressure drop is adopted for the blower.

Step 11. Compute oxygenation efficiency from Eq. (4.21), where the numerator was calculated in Step 6 and the denominator obtained from values of G_s selected in Step 4 (it equals approximately 23.2% of the weight of air corresponding to G_s). The weight of air corresponding to G_s is calculated from the ideal gas equation. From

$$PV = NRT = (\text{weight/molecular weight}) \times RT$$

then

$$\text{Weight of air} = (\text{molecular weight})(PV)/RT$$

where molecular weight = 29 lb/lb mole (average molecular weight of air) $P = 1$ atm, $V = G_s$, $R = 0.73$(atm)(ft^3)/(lb mole)(°R), and $T = 520$°R (60°F). Therefore

$$\text{Weight of air} = (29 \times 1 \times G_s)/(0.73 \times 520) = 0.076G_s \text{ lb/min}$$

and [Eq. (4.25)]

$$\text{Weight of } O_2/\text{min} = 0.232 \times 0.076G_s = 0.0176G_s \qquad (4.25)$$

15. Turbine Aeration Units

15.1. DESCRIPTION OF UNIT

A sketch of a typical turbine aeration unit is shown in Fig. 4.12. These units entrain atmospheric oxygen by surface aeration and disperse compressed air by a shearing action employing a rotating turbine or agitator. Air bubbles discharged from a pipe or sparger below the agitator are broken down by the shearing action of the high speed rotating blades of the agitator. For systems of low oxygen utilization rate, oxygen is supplied by the flow of air self-induced from a negative head produced by the rotor (suction effect). For systems of higher oxygen utilization rate a blower or compressor is needed.

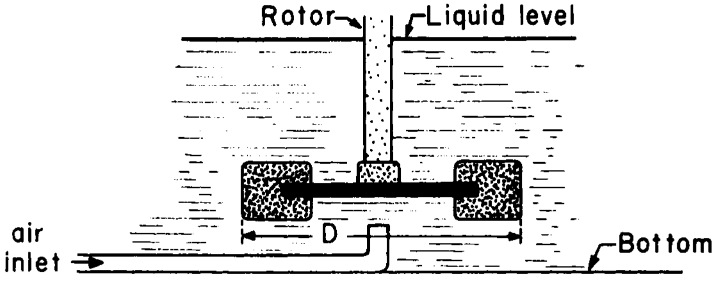

Fig. 4.12. Turbine aeration unit.

15.2. PERFORMANCE OF TURBINE AERATION UNITS

Main variables to be considered are air flow, diameter (D), and speed of impeller. These variables determine bubble size and degree of agitation in the tank, thus affecting the overall oxygen-transfer coefficient K_La. Performance data for turbine aeration units are available from manufacturer's information taking these variables into consideration. Eckenfelder [1] recommends the following empirical equation for correlating performance of turbine aeration units:

$$N = CG_s^nS^xD^y(C_{SW} - C_L)1.024^{(T-20)}\alpha \tag{4.26}$$

where N is the lb O_2 transferred/(hr)(aeration unit); G_s the air flow (SCFM/aeration unit); S the peripheral speed of the impeller (ft/sec); D the impeller diameter (ft, see Fig. 4.12); C_{SW} the saturation DO concentration in wastewater (mg/liter); C_L the operating DO concentration (mg/liter); and C, n, x, y the constants characteristic of the aeration device.*

Comparing Eqs. (4.23) and (4.26) the result is Eq. (4.27).

$$K_LaV = CG_s^nS^xD^y \times 1.024^{(T-20)}\alpha \quad (\text{ft}^3/\text{hr}) \tag{4.27}$$

where term K_La has units of hr^{-1}, and V is the volume of the aerator tank in ft^3. If $(C_{SW} - C_L)$ in Eq. (4.26) is given in mg/liter, its value is multiplied by factor 6.23×10^{-5} to obtain N in lb/hr, as shown in Section 14.3.

15.3. POWER REQUIREMENTS FOR TURBINE AERATORS

Power is required for two purposes: (1) operation of rotor (corresponding horsepower is designated hence as HP_r), and (2) operation of compressor or

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* Typical values for n, x, and y (depending on impeller geometry) [4]: n, 0.4–0.9; x, 1.2–2.4; and y, 0.6–1.8.

blower (corresponding horsepower is designated hence as HP_c). Power drawn by the rotor is computed from the relationship [4]

$$HP_r = C'D^m S^p \tag{4.28}$$

C', m, and p are constants characteristic of the aeration device. Actual drawn horsepower decreases as air flow is increased under the impeller due to decreased density of the aerated mixture. For this reason, horsepower calculated from Eq. (4.28) is referred to as ungassed horsepower. Equation (4.28) is rewritten as

$$HP_r = C''D^m n^p \tag{4.29}$$

where n is expressed in revolutions/sec. Since S is the peripheral speed in ft/sec, n and S are related as [Eq. (4.30)]

$$n = S/\pi D = (\text{ft/sec})/(\text{ft/rev}) = \text{rev/sec} \tag{4.30}$$

where πD is the perimeter of the circumference described by the rotation of the impeller.

Typical values of exponents m and p are [4]

$$4.8 \leqslant m \leqslant 5.3$$

$$2.0 \leqslant p \leqslant 2.5$$

Ungassed horsepower is correlated to actual horsepower. A correlation is presented in Fig. 4.13. Power drawn by the compressor is calculated from

$$HP_c = (\text{pressure drop, psi}) \times (\text{SCFM}) \times 144/(33{,}000)(\varepsilon_m) \tag{4.31}$$

where ε_m is the estimated turbine efficiency.

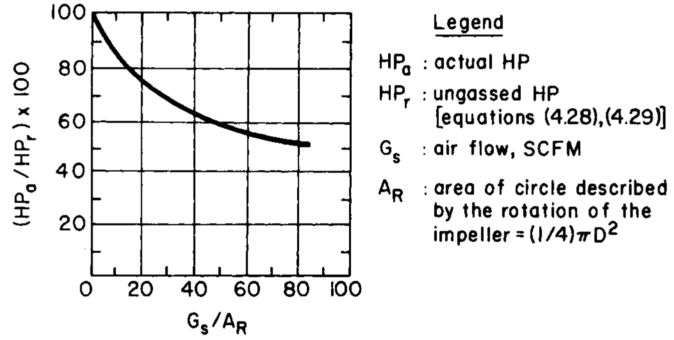

Fig. 4.13. ***Effect of air rate on turbine horsepower*** **[4].**

Next determine the optimum power split between rotor and compressor. A correlation between oxygen-transfer efficiency [expressed as (lb O_2 transferred)/(HP × hr)] and a factor P_d defined as [Eq. (4.32)]

$$P_d = HP_r/HP_c \tag{4.32}$$

has been developed by Quirk [6], and its utilization is summarized by Eckenfelder [1]. P_d represents the power split between rotor and compressor. A typical correlation curve is shown in Fig. 4.14.

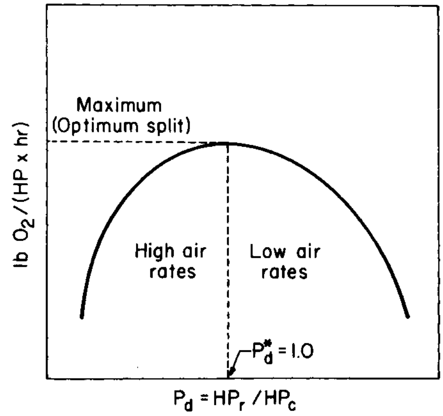

Fig. 4.14. Correlation for power split for turbine aeration units.

The value indicated in Fig. 4.14 as P_d^* is the optimum power split corresponding to the maximum oxygen-transfer efficiency. In most cases P_d^* is approximately unity, this implying an equal power expenditure by the turbine and the compressor. At extremely high air rates (high values of HP_c), values of P_d are less than 1.0, i.e., $HP_c > HP_r$ and $P_d < 1.0$. Under these conditions, large air bubbles and flooding of the impeller yield poor oxygenation efficiencies. On the other hand, at very low air rates $P_d > 1.0$ and too much turbine horsepower is spent in mixing the liquor.

15.4. DESIGN PROCEDURE FOR AERATION SYSTEMS UTILIZING TURBINE AERATION UNITS

For fundamental information required see items 1–5 for air diffusion units (Section 14.4), then obtain performance data for the turbine aeration units available from manufacturer's information, or expressed in terms of values for the constants in Eq. (4.26).

Step 1. Select a tank depth H, usually between 15 and 20 ft. In some cases deeper liquid depths are employed.

Step 2. Cross-sectional area is then $A = V/H$.

Step 3. Select a ratio $r = D/T$, where D is diameter of the turbine and T the "diameter" of the tank. For tanks of circular cross section, the meaning of ratio D/T is straightforward. For tanks with rectangular or square cross sections, select a value for T based on geometry of the system. Selection of T

for a rectangular tank with two turbine aerators is illustrated by Fig. 4.15 (T equals the diameter of influence of the aeration unit). Typical D/T ratios are 0.1–0.2.

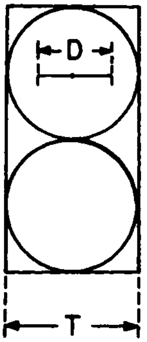

Fig. 4.15. Ratio $r = D/T$ for a rectangular tank with two turbine aerators.

Step 4. Select a tank width T approximately twice the tank depth. Therefore turbine diameter is $D = T \times r$.

Step 5. From Fig. 4.14 (manufacturer's data) determine optimum power split P_d^*.

Step 6. Select air flow rate per unit G_s (SCFM/aeration unit). Typical values are between 200 and 1500 SCFM.

Step 7. Value C_{SW} [to be used in Eq. (4.26)] is computed at tank mid-depth from Eq. (4.12) [or Eq. (4.13)], i.e., $C_{SW} = C_{s,m}$.

Step 8. Oxygenation capacity per aeration unit [N = lb O_2 transferred/(hr)(unit)] is estimated from manufacturer's data or computed from Eq. (4.26).

Step 9. From oxygen requirements (lb O_2/hr) calculated in Chapter 5, Section 7 and value of N calculated in Step 8, calculate the number of aeration units needed to transfer required amount of oxygen.

$$\text{No. of units} = \text{lb } O_2/\text{hr (required)}/N$$

There should be one turbine unit for every 900–2500 ft^2. By varying air rate per unit G_s, one adjusts calculations so that spacing falls within this range.

Step 10. Compute total air flow.

$$\text{Total air flow} = G_s \times (\text{no. of units}) \qquad (\text{SCFM})$$

Step 11. Compute operating compressor horsepower from Eq. (4.31).

Step 12. Determine turbine horsepower from optimum power split established in Step 5.

$$HP_r = (P_d^*)(HP_c)$$

Step 13. Compute oxygenation efficiency from Eq. (4.21), where numerator was obtained in Step 8. Calculate denominator from Eq. (4.25).

16. Surface Aeration Units

16.1. DESCRIPTION OF UNIT

Surface aeration units are based solely on entrainment of oxygen from atmospheric air. Unlike air diffusion and turbine aerators there is *no* stream of air involved in this system.

Improved design of surface aerators has resulted in improvement of oxygen-transfer capacity, and their use has increased rapidly in the past few years. They are widely used in activated sludge plants and aerated lagoons.

The principle of operation of surface aerators is illustrated by the sketch in Fig. 4.16. Liquid is drawn from underneath the unit and sprayed upward and outward by a propeller inside a vertical tube.

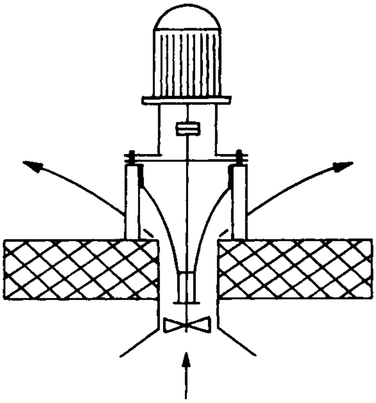

Fig. 4.16. Cross-sectional diagram of a surface aerator.

Most conventional surface aerators are fixed to piers mounted across the aerating tanks. Floating units are also available, the whole unit being supported on a reinforced fiberglass float filled with plastic foam to make it unsinkable.

Oxygen transfer in surface aerators occurs according to two mechanisms: (1) transfer at the turbulent liquid surface, and (2) transfer to droplets sprayed by the blades of the unit.

16.2. CORRELATION BETWEEN TRANSFER EFFICIENCY AND LEVEL OF AGITATION

A correlation has been developed [1] between transfer efficiency [expressed as lb of O_2 transferred/(HP × hr)] and level of agitation of the basin (in HP/1000 gal basin). There is an approximate linear relationship between these two parameters, as indicated by the straight line in Fig. 4.17, which is a

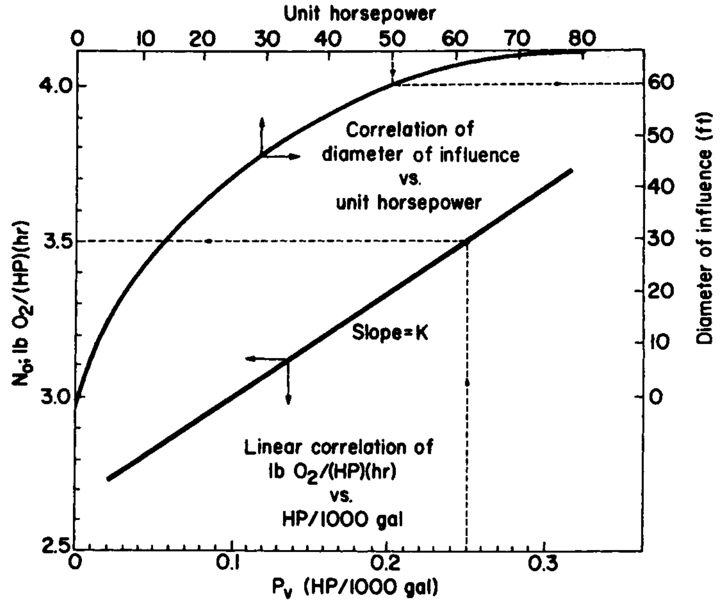

Fig. 4.17. Characteristics of surface aerator.

typical example of this correlation for a specific surface aerator unit. Correlation between diameter of influence and unit horsepower, which is also plotted in Fig. 4.17, is discussed in this section. Ordinate N_0 equals the lb of O_2 transferred to tap water at standard conditions (20°C and 1 atm, with initial zero dissolved oxygen) per (HP × hr). A correction to obtain oxygen transfer (N) for a wastewater under operating conditions is presented in this section [Eq. (4.34)].

The linear correlation indicated in Fig. 4.17 is expressed by the relationship [1]

$$N_0 = KP_v + N_s \tag{4.33}$$

where N_0* is the total oxygen transferred to tap water under standard conditions per unit [lb O_2/(HP × hr)]; P_v the HP per 1000 gal of basin liquid; K the constant characteristic of the aeration device (in Fig. 4.17 this corresponds to slope of the straight line); and N_s the oxygen transferred to tap water at standard conditions per unit horsepower × hr at zero turbulence (in Fig. 4.17 this corresponds to ordinate of the straight line at the origin).

In Eq. (4.33), N_s corresponds to oxygen transferred at standard conditions (N_0) for conditions of zero turbulence (i.e., $P_v = 0$). For a given aerator this

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*For the specific surface aerator unit corresponding to Fig. 4.17, this relationship is

$$N_0 = 3.4P_v + 2.65$$

corresponds to its operation in a basin of infinite volume. In such cases, all oxygen transfer is accomplished by the spray mechanism alone, since turbulence is negligible.

Performance of surface aerators is related to the following factors: (1) submergence of impeller, and (2) diameter and speed of rotor. Values for transfer efficiency [lb O_2 transferred/(HP × hr)] are 2–4 for most surface aerators, although values as high as 7 are reported. Transfer efficiency remains essentially constant at an optimum submergence regardless of size of the unit.

For design of surface aerator systems, standard transfer efficiency N_0 obtained from Fig. 4.17, for example, is corrected for actual wastewater conditions and temperature. This is done by application of the following relationship [1]:

$$N = N_0\left[\frac{C_{SW} - C_L}{9.2} \times 1.024^{(T-20)}\alpha\right] \tag{4.34}$$

where N is the oxygen-transfer efficiency under field conditions [lb O_2/(HP × hr)]; N_0 the oxygen-transfer efficiency at standard conditions [tap water at 20°C with initial zero dissolved oxygen at atmospheric pressure; lb O_2/(HP × hr)]; C_{SW} the saturation concentration of dissolved oxygen in the wastewater; C_L the operating DO level in aeration basin; T the temperature of the basin (°C); and $\alpha = K_La$ (wastewater)/K_La (tap water).

In Eq. (4.34), $C_{SW} - C_L$ is the actual driving force for oxygen transfer under field conditions. Driving force at standard conditions with initial zero dissolved oxygen is $9.2 - 0.0 = 9.2$, where 9.2 is the oxygen saturation value at 20°C in mg/liter (Table 4.1). Thus, in Eq. (4.34) a proportionality between N and N_0 and the corresponding driving forces is assumed.

16.3. DESIGN PROCEDURE FOR AERATION SYSTEMS UTILIZING SURFACE AERATION UNITS

For fundamental information required see items 1–5 for air diffusion units (Section 14.4). Obtain characteristics for the aerator. This includes (1) correlation of N_0 vs. HP/1000 gal, and (2) correlation between unit horsepower and diameter of influence for solids in suspension (ft).

For the specific design illustrated by Example 4.5, these two correlations are presented in Fig. 4.17. Depths of aerator basins for surface aerators are usually lower than for diffusion or turbine aeration, ranging from 8 to 12 ft.

Step 1. Take Eq. (4.34) and calculate the term between brackets for summer and winter conditions to determine which is the controlling one. Note that $[C_{SW}]_{\text{summer}} < [C_{SW}]_{\text{winter}}$ [thus $(C_{SW} - C_L)$ is larger for the winter conditions] whereas $T_{\text{summer}} > T_{\text{winter}}$ [thus $1.024^{(T-20)}$ is larger for summer conditions]. Let the results of this calculation be

$$N = K_{\text{summer}} \times N_0 \tag{4.35}$$

$$N = K_{\text{winter}} \times N_0 \tag{4.36}$$

where K's are values of the term between brackets in Eq. (4.34). The *lower K* corresponds to controlling condition (lower transfer of oxygen).

Step 2. Since power level (abscissa of Fig. 4.17) is not known, a trial and error solution is necessary for determination of N_0 (and N) based on correlation of N_0 vs. power level.

(1) Assume a power level HP/1000 gal.

(2) From Fig. 4.17 read N_0.

(3) Calculate N from Eq. (4.35) [or Eq. (4.36)], whichever is the controlling one.

(4) Power requirements are calculated for assumed power level from

$$\text{Power requirements} = \frac{\text{O}_2 \text{ required (lb O}_2\text{/hr)}}{N \text{ (lb O}_2\text{/HP} \times \text{hr)}} = \text{HP}$$

where the oxygen requirement has been previously calculated from biological reactor requirements (item 2 on "Fundamental information required," Section 14.4)

(5) Select HP per unit and calculate number of units.

(6) Recalculate power level.

$$\text{Power level} = \text{HP [Step 2(4)]/volume of aeration basin}$$

where volume of the aeration basin is calculated from biological reactor requirements (item 1 on "Fundamental information required," Section 14.4). Express recalculated power level in terms of HP/1000 gal and compare it with the value assumed in Step 2(1). If agreement is within 5%, calculations are stopped. Otherwise, iterate Steps 2(1)–(6) until agreement is obtained.

Step 3. Spacing between agitators is determined from the correlation indicated in Fig. 4.17. The procedure for the aerator basin layout is illustrated in Example 4.5.

Example 4.5

Surface aerators are specified for an activated sludge plant treating an industrial wastewater. Oxygen requirements and volume of the biological reactor are calculated by the procedure described in Chapter 5, Section 7 (Example 5.7), yielding the following results.

Oxygen requirements: 665 lb O_2/hr
Volume of reactor: 1,200,000 gal

The following additional information is available.

Wastewater temperature (summer): 30°C, $C_{SW} = 7.4$ mg/liter
Wastewater temperature (winter): 18°C, $C_{SW} = 10.3$ mg/liter

Take the operating DO level at the basin as $C_L = 1.0$ mg/liter, and $\alpha = 0.72$.

Characteristics of the surface aerator selected are given by Fig. 4.17. Design the aeration system for this application.

SOLUTION

Step 1. Utilize Eq. (4.34).

Summer: $t = 30°\text{C}$, $C_{SW} = 7.4$ mg/liter
Winter: $t = 18°\text{C}$, $C_{SW} = 10.3$ mg/liter

Thus for summer conditions

$$N = N_0\left[\frac{7.4-1.0}{9.2} \times 0.72 \times 1.024^{(30-20)}\right] = 0.635N_0$$

and for winter conditions

$$N = N_0\left[\frac{10.3-1.0}{9.2} \times 0.72 \times 1.024^{(18-20)}\right] = 0.694N_0$$

Therefore, summer conditions control design (*lower* oxygen-transfer efficiency).

Step 2.

1. Assume a power level, e.g., 0.25 HP/1000 gal.
2. From Fig. 4.17 read $N_0 = 3.5$ lb O_2/(HP × hr).
3. Then $N = 0.635$; $N_0 = 0.635 \times 3.5 = 2.22$ lb O_2/(HP × hr).
4. Power requirements are then calculated.

   Oxygen requirement: 665 lb O_2/hr

   Power requirements: 665 lb O_2/hr × (HP × hr)/ 2.22 lb O_2 = 299.5 HP

5. Select six units of 50 HP each (total HP = 6 × 50 = 300 HP).
6. Power level is then

$$300 \text{ HP}/1200 \text{ thousands of gal} = 0.25 \text{ HP}/1000 \text{ gal}$$

which agrees with assumed value. Thus design is satisfactory.

Step 3. Diameter of influence for 50 HP units (see Fig. 4.17) is 60 ft (or radius of influence of 30 ft). Spacing distance of 56 ft is selected to provide a minimum overlap. Arrange aerators according to layout in Fig. 4.18.

Cross-sectional area of the basin is $168 \times 112 = 18{,}816$ ft^2, and its volume in ft^3 is

$$1{,}200{,}000 \text{ gal} \times \text{ft}^3/7.48 \text{ gal} = 161{,}000 \text{ ft}^3$$

Therefore depth is

$$161{,}000/18{,}816 = 8.6 \text{ ft}$$

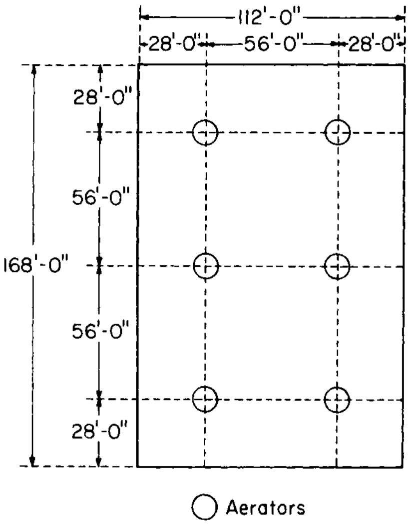

Fig. 4.18. Layout of aerator tank (Example 4.5).

Several selections of aerator units are possible for a given application, leading to various layouts. Engineering judgment and economic considerations determine final selection of aerator units and their layout.

Problems

I. The following results are obtained in an unsteady state aeration test utilizing a 5-HP surface aerator. $C_s = 9.2$ ppm (measured at 20°C; $P = 27$ in Hg). Aerator is a 100,000-gal circular test tank.

Time (min)	C_L (ppm)
0	0
12	2.6
24	4.5
36	5.8
48	6.7
60	7.4

Calculate

1. Chemical requirements (lb Na_2SO_3/lb liquor) for deoxygenation of test water with 8 ppm of DO
2. Value of K_La(hr^{-1})
3. If aeration tank has a capacity of 100,000 gal, calculate the lb/hr of oxygen transferred at standard conditions

4. If aerator has a nominal HP of 5, report aerator efficiency [lb O_2/(hr)(HP)] in terms of nameplate HP and blade HP.

The following information is available for the aerator:

Drawn voltage: 220 V (average)
Amperage: 13.5 A (average)
cos PF (measured): 0.8
Motor efficiency (estimated): 85%
Gear efficiency (estimated): 85%

II. Unsteady state aeration data is obtained in a diffused aerator system for water at 6.5°C and a wastewater at 0°C, and is tabulated below. Calculate coefficient α (at 20°C).

TABLE 1a
Water at 6.5°C, $C_s = 12.3$ mg/liter

Time (min)	C_L (mg/liter)
3	0.6
6	1.6
9	3.1
12	4.3
15	5.4
18	6.0
21	7.0

TABLE 1b
Wastewater at 0°C, $C_s \simeq 14.3$ mg/liter

Time (min)	C_L (mg/liter)
3	0.9
6	1.7
9	2.5
12	3.2
15	3.9
18	4.6
21	5.2

III. A tubine aerator in an aeration tank 30 × 50 × 15 ft transfers oxygen according to the relationship

$$K_L a V = 25 G_s^{0.45} S^{1.5} D^{1.8} \qquad (\mathrm{ft}^3/\mathrm{hr})$$

where $K_L a$ is in liter/hr; V is the tank volume (ft^3); G_s the air flow (SCFM); S the peripheral speed of impeller (ft/sec); and D the impeller diameter (ft). Power drawn by the turbine is defined by the relationship

$$HP_r = 0.02 D^{5.25} n^{2.75}$$

where D is the impeller diameter (ft) and n the revolutions/sec of rotor. For calculation of compressor horsepower, take a pressure drop of 5.55 psi and an efficiency $\varepsilon_m = 0.65$ (65%).

1. Compute $K_La(\mathrm{hr}^{-1})$. Turbine is 40 in. in diameter, rotating at 15 ft/sec peripheral speed, with an air flow of 300 SCFM.
2. Calculate O_2 transfer (lb/hr) under standard conditions. Saturation solubility of oxygen in the sewage liquid at 20°C is 8.45 ppm.
3. Calculate turbine horsepower corrected from Fig. 4.13.
4. Calculate blower horsepower.
5. Calculate transfer efficiency in terms of lb of O_2 transferred per HP × hr.

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5

Secondary Treatment: The Activated Sludge Process

1. Introduction

The heading secondary treatment encompasses all biological treatment processes of wastewaters, both aerobic and anaerobic. In this chapter the activated sludge process is studied in detail, and the mathematical models developed are applicable, with minor changes, to all aerobic processes described in Chapter 6.

The activated sludge process has been utilized for treatment of both domestic and industrial wastewaters for approximately half a century. Design of activated sludge plants was carried out to a large extent in an empirical manner. It was only after the early 1960's that a more rational approach to the design of activated sludge systems was developed. This process originated from the observation made a long time ago that whenever wastewater, either domestic or industrial, is aerated for a period of time, the content of organic matter is reduced, and at the same time a flocculent sludge is formed.

Microscopic examination of this sludge reveals that it is formed by a heterogeneous population of microorganisms, which changes continually in nature in response to variation in the composition of the wastewater and environmental conditions. Microorganisms present are unicellular bacteria, fungi, algae, protozoa, and rotifers. Of these, bacteria are possibly the most important, being found in all types of biological treatment processes.

The purpose of this chapter is to discuss the design principles for the activated sludge process and to apply them to design of treatment plants. This involves development of fundamental design information from laboratory scale reactors. The approach utilized is based mainly on the work of Eckenfelder and associates.

The activated sludge process has been developed as a continuous operation by recycling the biological sludge. A flow diagram of this continuous process

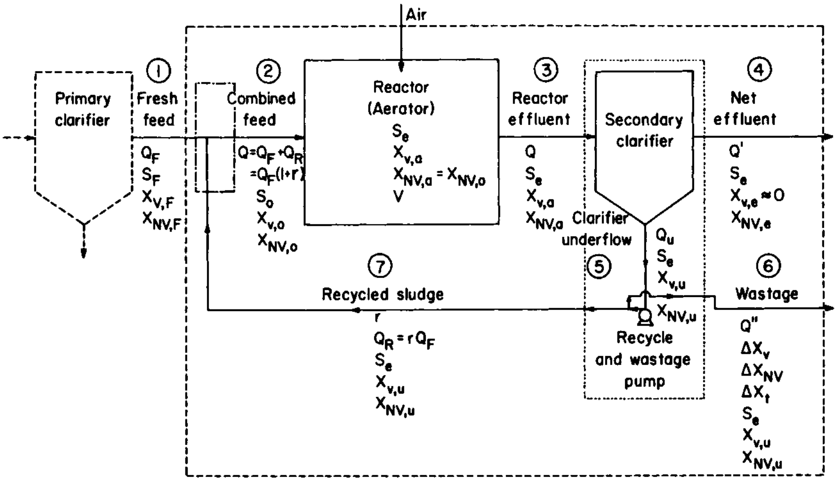

Fig. 5.1. Conventional activated sludge process. (See Table 5.1 for a definition of symbols.)

is shown in Fig. 5.1. All important process variables are indicated in Fig. 5.1 and defined in Table 5.1. These should be carefully examined by the reader.

In Fig. 5.1, compositions of the different streams (numbered 1–7) are characterized by three types of concentrations:

1. *Concentration of soluble BOD.* Denoted by the symbol S_i, where subscript i refers to the specific stream under consideration, as indicated in Table 5.1. Soluble BOD comprises mainly carbonaceous materials in solution.

2. *Concentrations of volatile suspended solids* (*VSS*). These are denoted by symbol $X_{v,i}$, where subscript v stands for volatile, and subscript i refers to the specific stream in question (Table 5.1). VSS correspond to the biological sludge, constituted by the heterogeneous population of microorganisms. Experimental determination of VSS is performed by measuring the loss of weight of total suspended solids (TSS), after incineration in a laboratory oven at 600°C. This loss of weight corresponds mainly to volatilization of biological sludge. Remaining solids after incineration at 600°C are referred to as nonvolatile suspended solids. Their nature is distinct from those in the biological sludge, being constituted of nonliving matter of both inorganic and organic nature.

3. *Concentrations of nonvolatile suspended solids* (*NVSS*). These are denoted by symbol $X_{NV,i}$, where NV stands for nonvolatile, and i refers to the specific stream in question.

Therefore

$$\text{TSS} = \text{VSS} + \text{NVSS}$$

Total suspended solids = Volatile suspended solids + Nonvolatile suspended solids

TABLE 5.1
Definition of Symbols Used in Fig. 5.1

Key

For suspended solids double subscripts are utilized, e.g., $X_{v,i}$, $X_{NV,i}$.

The first subscript (v or NV) designates volatile and nonvolatile suspended solids, respectively. The second subscript (i) refers to the specific stream in question:

F, fresh feed [stream 1]
o, combined feed [stream 2]
a, reactor effluent [stream 3]
e, net effluent [stream 4]
u, underflow from secondary clarifier [stream 5]

Symbols

1. Flow rates
   - Q_F, fresh feed; MGD (million gallons per day) [stream 1]
   - Q_R, recycle; MGD [stream 7]
   - r, recycle ratio; dimensionless ($r = Q_R/Q_F$)
   - Q, combined feed; MGD; $Q = Q_F + Q_R = Q_F(1+r)$ [stream 2] (MGD of combined feed = MGD of reactor effluent, i.e., Q [stream 2] = Q [stream 3])
   - Q', net effluent; MGD [stream 4]
   - Q'', wastage; MGD [stream 6] (Notice that $Q_F = Q' + Q''$)
   - Q_u, clarifier underflow; MGD; $Q_u = Q'' + Q_R = Q'' + rQ_F$ [stream 5]
2. Concentrations (mg/liter) of soluble BOD
   - S_F, soluble BOD of fresh feed
   - S_o, soluble BOD of combined feed
   - S_e, soluble BOD of effluent
3. Concentrations (mg/liter) of volatile suspended solids (VSS)
   - $X_{V,F}$, VSS in fresh feed
   - $X_{v,o}$, VSS in combined feed
   - $X_{v,a}$, VSS in reactor. This also is equal to concentration of VSS in reactor effluent
   - $X_{v,u}$, VSS in secondary clarifier underflow
   - $X_{v,e}$, VSS in net effluent (take $X_{v,e} \approx 0$)
4. Concentrations (mg/liter) of nonvolatile suspended solids (NVSS)
   - $X_{NV,F}$, NVSS in fresh feed
   - $X_{NV,o}$, NVSS in combined feed
   - $X_{NV,a}$, NVSS in reactor ($X_{NV,a} = X_{NV,o}$). This also equals concentration of NVSS in reactor effluent
   - $X_{NV,u}$, NVSS in secondary clarifier underflow
   - $X_{NV,e}$, NVSS in net effluent
5. Wastage (lb/day)
   - ΔX_v, net yield of MLVSS in reactor (wastage of MLVSS)
   - ΔX_{NV}, wastage of NVSS
   - ΔX_t, total sludge yield: $\Delta X_t = \Delta X_v + \Delta X_{NV} + Q_F X_{V,F}$
6. Reactor volume
   - V, reactor volume, MG (million gallons)

A description of the flowsheet in Fig. 5.1 follows, with emphasis on concentrations of (1) soluble BOD, (2) volatile suspended solids, and (3) nonvolatile suspended solids for the different streams.

1. *Soluble BOD*. Fresh feed, i.e., the wastewater to be treated [stream 1], enters the process with a value of soluble BOD denoted as S_F. Purpose of the treatment is to reduce this value to S_e (effluent BOD in stream 4) by oxidation through aerobic biological degradation of organic matter in the wastewater.

In the conventional activated sludge process, a reduction of soluble BOD to 5–10% of its value in the fresh feed is usually accomplished, i.e., $S_e = 5\text{–}10\%$ of S_F. This means a soluble BOD removal efficiency of 90–95%.

Fresh feed is combined with recycled sludge [stream 7] and enters the aerator (combined feed, stream 2). Biological sludge is continuously formed in the aerator. It is usually desirable to operate the reactor at steady state and under complete mixing conditions. These two assumptions are made in most mathematical models hence. Concentration of soluble BOD in the reactor liquor is denoted as S_e. Under steady state and complete mixing conditions the concentration of soluble BOD in reactor effluent [stream 3] also equals S_e.

Reactor effluent enters the secondary clarifier as indicated in Fig. 5.1. Concentration of soluble BOD is the same in clarifier underflow [stream 5] and net effluent [stream 4], i.e., S_e. Clarifier underflow is split into two streams: wastage [stream 6] and recycled sludge [stream 7]. For both these streams, the concentration of soluble BOD has the same value, S_e. The recycled sludge stream is then combined with fresh feed to form the combined feed. Concentration of soluble BOD in combined feed, designated as S_o, is calculated by a material balance at the junction point of streams 1, 2, and 7. This balance is written in Section 7.3.

2. *Volatile suspended solids* (*VSS*). At steady state, concentration of biological sludge in the reactor is kept constant at all times. In the conventional activated sludge process this concentration, designated as $X_{v,a}$, where the second subscript a refers to the aerator, is usually selected between 2000 and 3000 mg/liter. Since complete mixing conditions are postulated to exist in the reactor, volatile suspended solids in it are referred to as MLVSS (mixed liquor volatile suspended solids). Similarly, nonvolatile suspended solids in the reactor, being also completely mixed, are referred to as MLNVSS (mixed liquor nonvolatile suspended solids). Total suspended solids in the reactor are designated as MLTSS (mixed liquor total suspended solids).
Therefore

$$\text{MLTSS} = \text{MLVSS} + \text{MLNVSS}$$

$$\text{Mixed liquor total suspended solids} = \text{mixed liquor volatile suspended solids}$$
$$+ \text{ mixed liquor nonvolatile suspended solids}$$

Concentration of VSS in fresh feed ($X_{V,F}$) is negligible in many cases, since no appreciable amount of aeration has taken place at this stage. VSS is produced continuously in the aerator, owing to synthesis of biological matter, and withdrawn continuously with reactor effluent.

In order to maintain a constant concentration of MLVSS in the reactor, most of the clarifier underflow is recycled back. Recycle ratio r is calculated by material balance (Section 7.2) in order to maintain a constant selected concentration $X_{v,a}$ of MLVSS within the reactor at all times. Owing to synthesis of biological matter there is a *net* yield of MLVSS in the reactor (ΔX_v, lb/day). Therefore to maintain constant concentration of MLVSS in the reactor at all times, it is necessary to remove from the system a mass of MLVSS (lb/day) equal to this net yield ΔX_v. This is done by wastage of sludge [stream 6]. Although continuous wastage is indicated in Fig. 5.1, in practice it is usually an intermittent operation. It is simpler to write material balances for a steady state operation; thus continuous wastage is assumed in the remainder of this chapter. Intermittent wastage implies the assumption of unsteady state operation. Since the wastage stream is usually small by comparison with the recycle, assumption of continuous wastage does not introduce, in general, an appreciable error in the overall material balance. Concentration of VSS in the reactor effluent [stream 3] is also $X_{v,a}$, since complete mixing and steady state conditions are assumed.

Reactor effluent flows into the secondary clarifier. Underflow from the latter [stream 5] is a slurry containing a concentration of VSS designated as $X_{v,u}$ ($X_{v,u} > X_{v,a}$). The value of $X_{v,u}$ is selected by the designer, with clarifier being sized to yield this specified value. Usually $X_{v,u}$ is selected between 10,000 and 15,000 mg/liter of MLVSS. Concentrations of VSS in wastage and recycled sludge are also equal to $X_{v,u}$. In the net effluent from the secondary clarifier, concentration of VSS ($X_{v,e}$) is neglected in development of material balances in this chapter. This implies that complete separation of VSS is assumed to take place in the secondary clarifier (i.e., $X_{v,e} \approx 0$). This is usually a good assumption. Concentration of VSS in combined feed, $X_{v,o}$, is calculated by a material balance at the junction point of streams 1, 2, and 7. This balance is written in Section 4.5.

3. *Nonvolatile suspended solids* (*NVSS*). Concentration of MLNVSS in the aerator is denoted as $X_{NV,a}$ and is equal to those in both combined feed and reactor effluent. This is so because complete mixing is assumed and there is no production of NVSS in the aerator (unlike the net yield of VSS). Thus

$$X_{NV,a} = X_{NV,o}$$

Concentration of NVSS in fresh feed is designated as $X_{NV,F}$ and that in the recycled sludge as $X_{NV,u}$ (same as in underflow from secondary clarifier). In the combined feed this concentration is denoted as $X_{NV,o}$ and is calculated by a material balance which is written in Section 4.5.

Some of the NVSS in the reactor effluent is also separated by sedimentation in the secondary clarifier. Concentration of NVSS in clarifier underflow is denoted as $X_{NV,u}$ and that in net effluent as $X_{NV,e}$.

In wasted sludge, besides the ΔX_v lb/day of VSS there is also some nonvolatile sludge (ΔX_{NV} lb/day) resulting from partial sedimentation of NVSS in the secondary clarifier. In addition, there is the biological sludge introduced continuously with the fresh feed ($Q_F X_{V,F}$). Frequently, term $Q_F X_{V,F}$ is negligible since $X_{V,F}$ is usually very small. Total sludge wasted, ΔX_t lb/day, is [Eq. (5.1)]

$$\Delta X_t = \Delta X_v + \Delta X_{NV} + Q_F X_{V,F} \tag{5.1}$$

Respective concentrations of VSS, NVSS, and soluble BOD are the same for clarifier underflow, wastage stream, and recycle sludge, being denoted, respectively, as $X_{v,u}$, $X_{NV,u}$, and S_e.

In summary, concentrations of VSS, NVSS, and soluble BOD in combined feed ($X_{v,o}$, $X_{NV,o}$, and S_o, respectively) are obtained by material balances around the junction point of fresh feed and recycled sludge streams. These material balances are written in Sections 4.5 and 7.3.

From an overall balance for the wastewater [Eq. (5.2)],

$$Q_F = Q' + Q'' \tag{5.2}$$

Wastewater flows are usually expressed in millions of gallons per day (MGD).

Recycle ratio r is defined as

$$r = Q_R/Q_F = \text{recycle wastewater, MGD/fresh wastewater, MGD} \tag{5.3}$$

$$\therefore\ Q_R = rQ_F \tag{5.4}$$

Since combined feed Q is equal to fresh feed *plus* recycle,

$$Q = Q_F + Q_R = Q_F(1+r) \tag{5.5}$$

Hence, the density of all liquor streams in Fig. 5.1 is assumed equal to that of water at ambient temperature (8.34 lb/gal).* This is a good approximation since relatively dilute aqueous solutions are involved.

2. Mathematical Modeling of Activated Sludge Process

It is desirable to portray this process by a mathematical model and then to determine parameters utilized in mathematical equations from experimental data obtained utilizing a series of bench scale laboratory reactors. Relationships which are pertinent to the development of this mathematical model fall

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* This value is approximately 10.0 lb/gal when imperial gallons are utilized.

into three groups: (1) kinetics relationships; (2) material balance relationships—material balance for determination of oxygen utilization and of net yield of MLVSS; and (3) relationship for optimum settling conditions of sludge.

3. Kinetics Relationships

3.1. INTRODUCTION

Study of kinetics of aerobic biological treatment yields the rate at which microorganisms degrade a specific waste, and therefore provides the basic information required for sizing biological aerobic reactors. This study is conveniently performed in a laboratory scale batch reactor. Figure 5.2 shows a diagram of four units operating in parallel, each with a capacity of approximately 2.0 liters [3]. Reactors are built of plexiglass. Wastewater containing a seed of microorganisms* is introduced into the reactors, and compressed air is blown into the system. The biological sludge (MLVSS) is kept in a state of complete mixing due to agitation provided by air blown into the system.

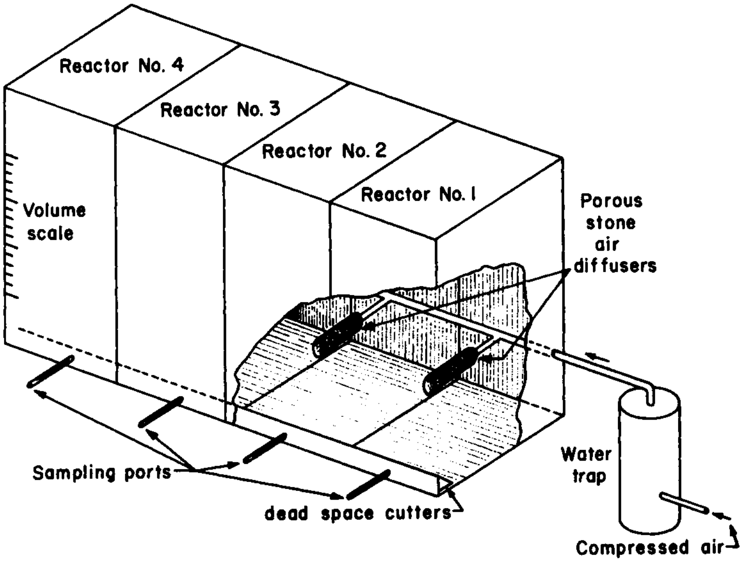

Fig. 5.2. Batch reactor.

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* Seed is either a mass of biological sludge taken from an operating activated sludge plant, or settled sewage.

BOD of wastewater (or COD, TOD, TOC) is determined at selected time intervals by withdrawing samples for the analysis. The mass of accumulated biological sludge (MLVSS) is also determined at these same time intervals by measuring the concentration of MLVSS in withdrawn samples and reading the volume of liquor in the reactor as indicated by the volume scale. Typical curves for decrease of BOD and variation of the amount of MLVSS with time are presented in Fig. 5.3.

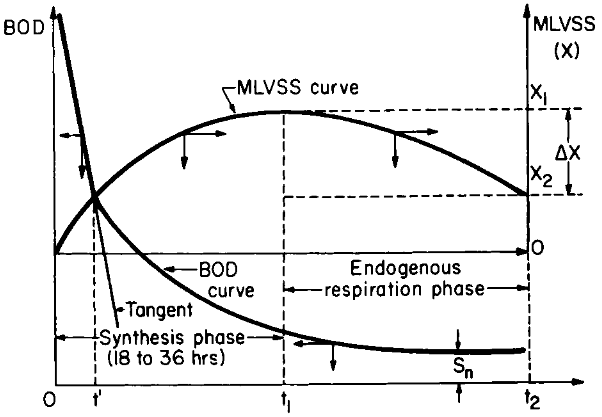

Fig. 5.3. Typical BOD and MLVSS curves for a batch reactor.

BOD of the wastewater, which is a measure of organic biodegradable matter concentration, decreases with time as the organic matter is oxidized. A plateau is eventually reached corresponding to the amount of nonbiodegradable matter (S_n).

Concentration of MLVSS increases at first (from time 0 to time t_1) during the period when a substantial concentration of substrate (relativley high BOD) is present to provide abundant food to sustain growth of microorganisms. This growth corresponds to the synthesis of new microorganism cells, indicated in Fig. 5.3 as "synthesis phase." After time t_1 when substrate concentration is considerably depleted, there is not enough food left to sustain growth of microorganisms. At this time, microorganisms start consuming their "fellow microorganisms" as food. As this "cannibalistic feast" proceeds, concentration of MLVSS drops when the rate of destruction of microorganism cells exceeds that of synthesis of new cells. This corresponds to the "endogenous respiration phase." The maximum on the MLVSS curve corresponds to time t_1, when these two rates are exactly equal. Distance ΔX corresponds to the *net* reduction of MLVSS concentration from t_1 to t_2.

There are two fundamental differences between operation of continuous (Fig. 5.1) and batch reactors (Fig. 5.2): (1) Contrary to what happens in the batch reactor, BOD of the wastewater in the continuous reactor operating at steady state conditions remains constant (S_e). This corresponds generally to a low substrate concentration, since the biological reactor is usually designed for removing most of the influent BOD. (2) Contrary to what happens in the batch reactor, concentration of MLVSS in the continuous reactor operating at steady state is kept constant ($X_{v,a}$) at a selected value. Maintenance of this constant $X_{v,a}$ is obtained by providing the calculated amount of concentrated return sludge. The material balance for MLVSS, necessary to arrive at required recycle ratio for this purpose, is presented in Section 7.2.

Kinetic data obtained from the batch reactor is portrayed by the Michaelis–Menten relationship, which is studied in Section 8. Two important corollaries of this relationship are postulated next, the second one being utilized for design of the continuous biological reactor.

1. At high substrate concentrations, BOD removal follows zero-order kinetics. This means that the rate of removal is essentially constant, independent of substrate concentration. This situation is found in early stages of the batch reactor operation, when substrate concentration is still very high (high BOD). This corresponds to the section of the BOD curve (Fig. 5.3) from time zero to approximately time t'. In this region, the tangent to the BOD curve, which equals the rate of substrate removal, coincides essentially with the curve itself (constant slope).

2. BOD removal at low substrate concentrations (corresponding to BOD values below 500 mg/liter) follows first-order kinetics. This means that rate of removal is proportional to remaining substrate concentration. This corresponds to the section of the BOD curve beyond time t'. Slope of the BOD curve (which equals rate of substrate removal) decreases with time as the BOD value is lowered. A plot of these slopes vs. corresponding BOD values yields a straight line relationship, which is discussed in Section 3.2. Thus in this region, rate of substrate removal is directly proportional to its concentration (first-order kinetics).

3.2. FORMULATION OF THE CONTINUOUS REACTOR

Since for the continuous reactor operating substrate concentrations (S_e) are considerably below 500 mg/liter (BOD_5), first-order kinetics is assumed in the formulation. Consider the continuous reactor operating under steady state and complete mixing conditions. This situation is illustrated by Fig. 5.4.

Assuming that rate of substrate removal dS/dt follows first-order kinetics,*

$$dS/dt = -KS \tag{5.6}$$

It is customary to express substrate removal rate per mg/liter of MLVSS

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* Minus sign in Eq. (5.6) is required since $dS/dt < 0$, whereas $S > 0$.

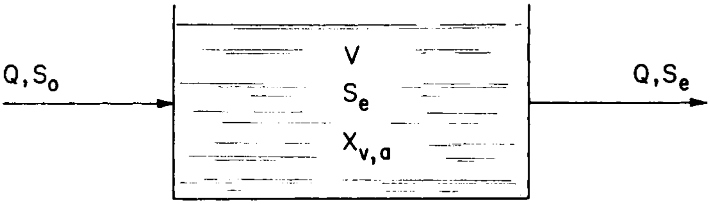

Fig. 5.4. Simplified diagram for continuous reactor.

present in the reactor. Let $X_{v,a}$ be this MLVSS concentration. Equation (5.6) is then rewritten

$$(1/X_{v,a})(dS/dt) = -kS \tag{5.7}$$

The relationship between K and k is

$$K = kX_{v,a} \tag{5.8}$$

From Eq. (5.7)

$$dS/dt = -kX_{v,a}S \tag{5.9}$$

k is the substrate removal rate constant. For time t equal to residence time in the continuous reactor, concentration S corresponds to S_e, and Eq. (5.9) becomes

$$(dS/dt)_{\text{cont. reactor}} = -kX_{v,a}S_e \tag{5.10}$$

The following material balance for substrate is written for the reactor in Fig. 5.4.

$$\begin{aligned}\text{Change of substrate in reactor} &= \text{increase due to influent flow}\\ &\quad - \text{decrease due to effluent flow}\\ &\quad - \text{decrease due to reaction}\end{aligned} \tag{5.11}$$

Under steady state conditions,

$$\text{Change of substrate in reactor} = 0 \tag{5.12}$$

$$\text{Increase due to influent flow} = QS_o \tag{5.13}$$

and

$$\text{Decrease due to effluent flow} = QS_e \tag{5.14}$$

According to Eq. (5.10), the decrease in the amount of substrate due to the reaction is $kX_{v,a}S_e$ [minus sign already included in Eq. (5.11)]. Before substituting in Eq. (5.11) this value is multiplied by reactor volume V, since $kX_{v,a}S_e$ represents decrease per unit volume.

$$\text{Decrease due to reaction} = kX_{v,a}S_eV \tag{5.15}$$

Substitution of values given by Eqs. (5.12)–(5.15) in Eq. (5.11) yields after manipulation

$$(Q/V)[(S_o - S_e)/X_{v,a}] = kS_e \tag{5.16}$$

However,

$$t = V/Q = \frac{\text{Mgal}}{(\text{Mgal/day})} = \text{day} = \text{residence time } (t) \text{ in the reactor} \tag{5.17}$$

Consequently, Eq. (5.16) is

$$(S_o - S_e)/X_{v,a}t = kS_e \tag{5.18}$$

Term $(S_o - S_e)/X_{v,a}t$ which also appears in other formulations is the substrate removal rate. It corresponds to rate of removal of substrate in the continuous reactor per mg/liter of MLVSS present. Units are

$$(S_o - S_e)/X_{v,a}t = \frac{\text{mg/liter of BOD removed}}{(\text{mg/liter of MLVSS})(\text{day})}$$

$$= \text{mg BOD removed/(day)(mg MLVSS)}$$

$$= \text{lb BOD removed/(day)(lb MLVSS)}$$

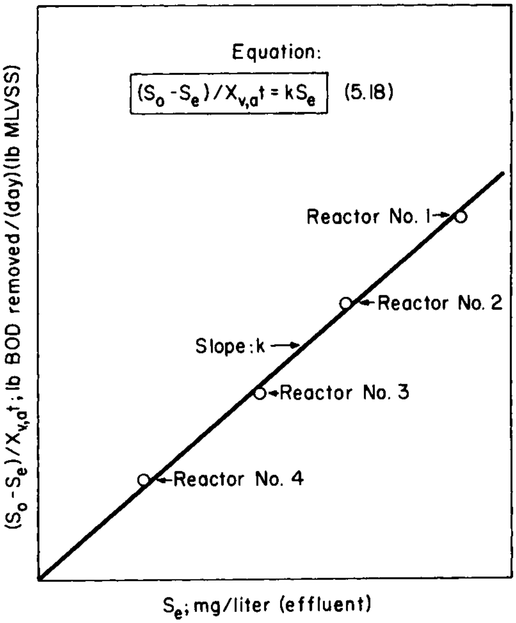

Fig. 5.5. Graphical determination of k (four continuous lab scale reactors).

Equation (5.18) indicates that the substrate removal rate is proportional to substrate concentration S_e (first-order kinetics). Substrate removal rate constant k (day^{-1}) is determined according to Eq. (5.18) from a plot of $(S_o - S_e)/X_{v,a}t$ vs. S_e. Figure 5.5 shows a graph of data obtained from four continuous laboratory reactors operating at steady state conditions. A numerical application is presented in Section 6.4 (Example 5.5).

Data plotted in Fig. 5.5 yield a straight line passing through the origin, assuming applicability of the mathematical model in Eq. (5.18). The left-hand member, $(S_o - S_e)/X_{v,a}t$, vanishes as t approaches infinity (infinite residence time). Consequently, term S_e in the right-hand member approaches zero since $k \neq 0$. This corresponds to *complete* removal of substrate, which is not always the case since some substrates cannot be completely degraded by the aerobic biological process, even at infinite residence time. In these cases, the straight line cuts the abscissa at a value of $S_e > 0$ corresponding to the concentration of nonbiodegradable matter. An example of this situation is shown in Fig. 5.14 (Section 6.4, Example 5.5).

When nonbiodegradable matter is present, Eq. (5.18) is modified to Eq. (5.19).

$$(S_o - S_e)/X_{v,a}t = k(S_e - S_n) \tag{5.19}$$

where S_n is the concentration of nonbiodegradable matter.

4. Material Balance Relationships

4.1. DESIGN PARAMETERS CORRESPONDING TO NET YIELD OF MLVSS AND OXYGEN REQUIREMENTS FOR AEROBIC BIOLOGICAL DEGRADATION OF WASTES

4.1.1. Introduction: Mechanism of Aerobic Biological Degradation

Accumulation of MLVSS and utilization of oxygen are two important elements needed for design of aerobic biological reactors. To obtain mathematical models which yield these two values, several design parameters designated by symbols a', $\bar{a}$, a, b, and b' are defined in this section. The approach followed is that proposed by Eckenfelder and associates [1–3].

Evaluation of these parameters is accomplished by using bench scale continuous biological reactors (Section 6). In the discussion which follows, numerical values for these parameters are utilized for clarification of some concepts. These values are obtained by techniques discussed in Section 6.

To arrive at the definition of these parameters, the basic mechanism of aerobic degradation of a substrate must be understood. Consider that a substrate is charged to a batch reactor (Fig. 5.2), and that curves for BOD removal and MLVSS concentration are obtained (Fig. 5.3). For clarification, take the hypothetical case of pure lactose as substrate. Assume that a lactose solution is charged to the batch reactor with a seed of microorganism, and that compressed air is bubbled through the solution. Let initial concentration of lactose be equal to 1050 mg/liter. Suppose that after a time t this concentration is reduced to 50 mg/liter. Thus substrate removed is $1050-50 = 1000$ mg/liter. Assume that ThOD is utilized as a measure of lactose concentration.* The chemical equation corresponding to ThOD for lactose is [Eq. (5.20)]†

$$(CH_2O) + O_2 \rightarrow CO_2 + H_2O \tag{5.20}$$

Molecular weight: 30 32

Thus, the initial ThOD of the solution is $(32/30)\times 1050 = 1120$ mg/liter. After time t, remaining ThOD is $(32/30)\times 50 = 53.3$ mg/liter. Therefore, ThOD removed is

$$1120 - 53.3 = 1066.7 \text{ mg/liter}$$

or

$$(32/30)(1050-50) = 1066.7 \text{ mg/liter} \tag{5.21}$$

Thus, ThOD and substrate removed are proportional, the proportionality constant being $32/30 = 1.07$. Since ThOD is correlated to COD, BOD, etc., one may also express substrate removal in terms of these parameters.

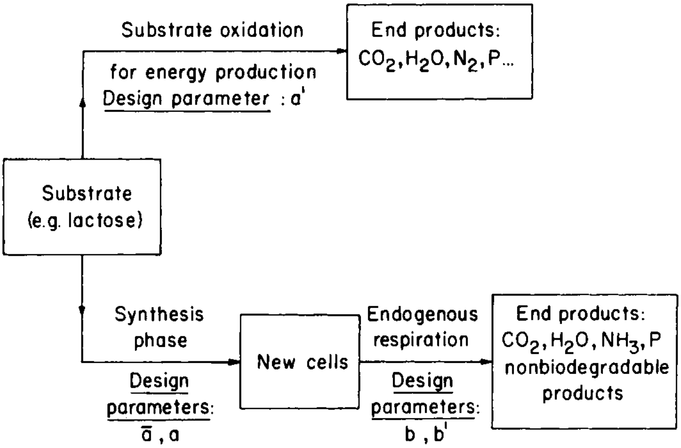

Fig. 5.6. Mechanism of aerobic biological degradation.

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* As discussed in Chapter 2, ThOD is only utilized in rare cases when complete analysis of the wastewater is known.

† For simplicity in Eq. (5.20), lactose is represented by one sugar unit (CH_2O). Multiplying this unit by a factor of 12 one obtains

$$12(CH_2O) = C_{12}H_{22}O_{11} \cdot H_2O$$

which is the molecular formula for lactose.

Mechanism of aerobic biological degradation of a substrate is represented diagrammatically by Fig. 5.6. Design parameters (a', $\bar{a}$, a, b, and b') indicated in Fig. 5.6 are defined in Sections 4.1.2–4.1.9. These values are calculated from laboratory data (Section 6).

Figure 5.6 indicates that substrate is removed during the biological process in two ways.

1. Part of the substrate, after being consumed as food by microorganisms, is utilized to synthesize new microorganism cells. This corresponds to the synthesis phase. For the lactose example, this synthesis corresponds to‡

$$5(CH_2O) \xrightarrow{\text{synthesis}} C_5H_7NO_2 \tag{5.22}$$

$$\text{Molecular weight: } 5 \times 30 = 150 \qquad \underset{(\text{MLVSS})}{113}$$

Intermediate steps in Eq. (5.22) are complicated and irrelevant. The empirical formula $C_5H_7NO_2$ corresponds to the average composition of MLVSS cells. Nitrogen is needed for synthesis and must be provided. From the approximate empirical formula $C_5H_7NO_2$ it follows that % of nitrogen in the MLVSS cells is $(14/113) \times 100 = 12.4\%$.

2. The remainder of the substrate is oxidized, terminal products being CO_2 and H_2O. In the lactose example, this substrate oxidation corresponds to Eq. (5.20). This terminal oxidation process is extremely important in the production of cellular energy utilized by the cells to maintain their normal functions, such as synthesis, reproduction, and mobility. Assume that 65% of the lactose removed (i.e., 65% of 1000 mg/liter = 650 mg/liter) is dioxized to provide energy requirements, and that 35% (i.e., 350 mg/liter) is utilized in the synthesis of new cell matter. Since there is a proportionality constant relating substrate and ThOD removals [factor (32/30) in Eq. (5.20) for lactose], it follows that 65% of the ThOD removed is utilized for energy generation and 35% for synthesis of new cells. Similar statements are valid in terms of COD and other parameters defined in Chapter 2 (Sections 2 and 3).

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‡ Phosphorus is also utilized in the synthesis and becomes a constituent of cell matter. The % of phosphorus in the MLVSS cells is approximately 2%, so a more accurate empirical formula for the MLVSS cells is $C_5H_7NO_2P_n$ where n is given by (atomic weight of phosphorus = 31)

$$31n/(113+31n) = 2/100$$

$$\therefore n = 0.074$$

or $C_5H_7NO_2P_{0.074}$. Nitrogen and phosphorus needed are provided by addition of ammonium phosphate to the wastewater, if it does not already contain the nitrogen and phosphorus required.

4.1.2. Definition of Parameter $\bar{a}$ (Synthesis Phase)

Let $\bar{a}$ be the fraction of substrate removed that is utilized for synthesis (namely, $\bar{a} = 0.35$ in lactose example). Due to the proportionality between removal of substrate and those of ThOD, COD, or BOD, $\bar{a}$ also represents fractions of ThOD (or COD, BOD) utilized for synthesis of new cells. Therefore,

$$\begin{aligned} \bar{a} &= \text{lb of substrate removed utilized for synthesis/lb of total substrate removed} \\ &= \text{lb ThOD removed for synthesis/lb of total ThOD removed} \qquad (5.23) \\ &= \text{lb COD removed for synthesis/lb total COD removed} \\ &= \text{lb BOD removed for synthesis/lb total BOD removed} \end{aligned}$$

The numerical value of $\bar{a}$ is independent of parameters utilized for expressing substrate removal, since $\bar{a}$ represents the fraction of substrate removed utilized for synthesis, and is therefore a dimensionless quantity. The same conversion factor for changing parameters in which substrate removal is to be expressed appears simultaneously in the numerator and denominator of Eq. (5.23), and therefore cancels out.

Parameter $\bar{a}$ *does not* appear in the final formulation of aerobic processes developed in Section 6. Instead parameter a, which is related to $\bar{a}$, is utilized.

4.1.3. Definition of Parameter a' (Oxidation)

Let a' be the fraction of substrate removed utilized for energy production (namely, $a' = 0.65$ in lactose example).
Therefore,

$$\bar{a} + a' = 1.0 \qquad (5.24)$$

where

$$\begin{aligned} a' &= \text{lb of substrate removed utilized for energy/lb of total substrate removed} \\ &= \text{lb ThOD removed for energy/lb total ThOD removed} \qquad (5.25) \\ &= \text{lb COD removed for energy/lb total COD removed} \\ &= \text{lb BOD removed for energy/lb total BOD removed} \end{aligned}$$

The numerical value of a' defined by Eq. (5.25) is independent of the parameters utilized for expressing substrate removal. The same observations made for $\bar{a}$ are applicable here.

SUMMARY For the lactose example

Total substrate removed: 1000 mg/liter
Total ThOD removed: $32/30 \times 1000 = 1066.7$ mg/liter. These removals take place in two ways:

(1) Synthesis:

$$5(CH_2O) \rightarrow C_5H_7NO_2$$

Substrate removed utilized for synthesis:

$$(0.35)(1000) = 350 \text{ mg/liter}$$

ThOD removed for synthesis:

$$(0.35)(1066.7) = 373.3 \text{ mg/liter}$$

[Ratios, $350/1000 = 373.3/1066.7 = 0.35 = \bar{a}$]

(2) Energy production:

$$(CH_2O) + O_2 \rightarrow CO_2 + H_2O$$

Substrate removed utilized for energy production:

$$(0.65)(1000) = 650 \text{ mg/liter}$$

ThOD removed for energy production:

$$(0.65)(1066.7) = 693.4 \text{ mg/liter}$$

[Ratios, $650/1000 = 693.4/1066.7 = 0.65 = a'$]

From Eq. (5.20) ThOD removed for energy production equals the lb of oxygen utilized for oxidation of substrate. Therefore the definition of a' (in terms of ThOD) given by Eq. (5.25) is reformulated as

$$a' = a'_{\text{ThOD}} = \text{lb of } O_2 \text{ utilized in oxidation of substrate/lb of total ThOD removed} \quad (5.26)$$

i.e., a' is equal to the lb of oxygen utilized in energy production per lb of total ThOD removed.

Therefore from Eq. (5.26),

$$\begin{aligned} \text{lb } O_2 \text{ (for energy)} &= a'(\text{lb total ThOD removed}) \\ &= a'_{\text{ThOD}}(\text{lb total ThOD removed}) \quad (5.27) \end{aligned}$$

Writing the right-hand member of Eq. (5.27) in terms of COD, BOD, and TOC by utilizing ratios ThOD/COD, ThOD/BOD, etc., yields

$$\begin{aligned} \text{lb } O_2 \text{ (for energy)} &= a'(\text{lb total COD removed})(\text{ThOD/COD}) \\ &= a'(\text{lb total BOD removed})(\text{ThOD/BOD}) \quad (5.28) \end{aligned}$$

Define subscript values of a' as

$$a'_{\text{COD}} = a'(\text{ThOD/COD}) \quad (5.29)$$

$$a'_{\text{BOD}} = a'(\text{ThOD/BOD}) \quad (5.30)$$

(where a' without the subscript stands for value $a' = a'_{\text{ThOD}}$).

Combining Eqs. (5.27) and (5.28) with Eqs. (5.29) and (5.30),

$$\begin{aligned}\text{lb } O_2 \text{ (for energy)} &= a'_{\text{ThOD}}(\text{lb total ThOD removed})\\ &= a'_{\text{COD}}(\text{lb total COD removed})\\ &= a'_{\text{BOD}}(\text{lb total BOD removed})\end{aligned} \tag{5.31}$$

Hence, whenever parameter a' is utilized for calculation of oxygen requirements, no subscripts are indicated. An appropriate value of a' is chosen to be compatible with parameters for expressing substrate removal. From Eq. (5.31) it follows that a' equals the lb of oxygen utilized for energy production per lb of substrate removed (removal in terms of ThOD, COD, and TOD).

Utilization of subscripts COD and BOD for a' may seem inconsistent since a' is thought of as a ratio, and therefore its numerical value should be *independent* of parameters utilized for expressing removal. However, this independence applies only to values of a' as defined by Eq. (5.25). In Eq. (5.25) the same conversion factor for parameters expressing removal appears simultaneously in the numerator and denominator, and therefore cancels out. From Eq. (5.31), however, it follows that a modified definition of a' is being utilized, i.e.,

$$a'_{\text{ThOD}} = a' = \text{lb } O_2 \quad \text{(for energy)/lb total ThOD removed} \tag{5.32}$$

$$a'_{\text{COD}} = \text{lb } O_2 \quad \text{(for energy)/lb total COD removed} \tag{5.33}$$

$$a'_{\text{BOD}} = \text{lb } O_2 \quad \text{(for energy)/lb total BOD removed} \tag{5.34}$$

The numerical value of the numerators in Eqs. (5.32), (5.33), and (5.34) is the same (lb of oxygen utilized for energy requirements). Values of denominators, however, vary depending on choice of parameters for expressing substrate removal. Consequently, numerical values of a' from Eqs. (5.32), (5.33), and (5.34) are different from each other. Therefore, utilization of subscripts is justified.

Furthermore, *only* the value of a' given by Eq. (5.32) is numerically equal to the ratios defined by Eq. (5.25), i.e., $a'_{\text{ThOD}} = a'$. Values of a' given by Eqs. (5.33) and (5.34) are not only different from each other, but also neither equals the fraction of substrate removed utilized in energy production.

4.1.4. Definition of Parameter *a* (Synthesis Phase)

Parameter a, related to $\bar{a}$, is defined as

$$a = \text{lb of MLVSS produced/lb of total substrate removed} \tag{5.35}$$

Consequently, a represents yield of biological sludge per lb of total substrate removed.

The relationship between parameters $\bar{a}$ and a is arrived at by consideration of the lactose example [Eq. (5.22)]. It is assumed that 350 mg/liter (35% of the total 1000 mg/liter of lactose removed) are utilized for the synthesis indicated by Eq. (5.22). Yield of MLVSS is calculated as

$$\begin{aligned}&\text{MLVSS produced per 1000 mg of total substrate removed}\\&\quad = [(0.35)(1000)](113/150)\\&\quad = 263.7 \text{ mg/liter} \qquad (5.36)\end{aligned}$$

Therefore, from Eq. (5.36) one obtains

$$\begin{aligned}a &= \text{lb MLVSS produced/lb of total substrate removed}\\&= [(0.35)(1000)](113/150)/1000 = 263.7/1000 = 0.2637 \qquad (5.37)\end{aligned}$$

i.e., 263.7 mg/liter of MLVSS are produced per 1000 mg/liter of lactose removed; thus $a = 263.7/1000 = 0.2637$.

The relationship between a and $\bar{a}$ from Eq. (5.37) for the lactose example is

$$a = \bar{a}(113/150)$$

$$\therefore \bar{a} = a(150/113)$$

where 113/150 is the stoichiometric ratio for Eq. (5.22). Substitution of this value of $\bar{a}$ in Eq. (5.24) yields

$$(150/113)a + a' = 1.0 \qquad (5.38)$$

Parameter a may be written in terms of total ThOD removed. Let a_{ThOD} be the numerical value of a expressed in this manner.

$$a_{\text{ThOD}} = \text{lb MLVSS produced/lb of total ThOD removed} \qquad (5.39)$$

Ratio a/a_{ThOD} from Eqs. (5.35) and (5.39), taking into account the stoichiometric ratio 32/30 in Eq. (5.20), is

$$\begin{aligned}a/a_{\text{ThOD}} &= \text{Eq. (5.35)/Eq. (5.39)}\\&= \text{lb total ThOD removed/lb total substrate removed}\\&= 32/30 \qquad (5.40)\end{aligned}$$

or

$$a = a_{\text{ThOD}}(32/30) \qquad (5.41)$$

Equation (5.38) written in terms of a_{ThOD} by utilizing Eq. (5.41) is

$$\underbrace{(150/113)(32/30)a_{\text{ThOD}}}_{\bar{a}} + a' = 1.0$$

or

$$\underbrace{1.42a_{\text{ThOD}}}_{\bar{a}} + a' = 1.0 \qquad (5.42)$$

MLVSS yield (synthesis) is obtained from Eq. (5.39).

$$\text{lb MLVSS produced} = a_{\text{ThOD}}(\text{lb total ThOD removed}) \tag{5.43}$$

Equation (5.43) may be rewritten expressing substrate removal in terms of COD, BOD, etc., by utilizing ratios ThOD/COD, ThOD/BOD, etc.:

$$\begin{aligned}\text{lb MLVSS produced} &= a_{\text{ThOD}}(\text{lb total COD removed})(\text{ThOD/COD}) \\ &= a_{\text{ThOD}}(\text{lb total BOD removed})(\text{ThOD/BOD})\end{aligned} \tag{5.44}$$

Define

$$a_{\text{COD}} = a_{\text{ThOD}}(\text{ThOD/COD}) \tag{5.45}$$

$$a_{\text{BOD}} = a_{\text{ThOD}}(\text{ThOD/BOD}) \tag{5.46}$$

Therefore

$$\begin{aligned}\text{lb MLVSS produced} &= a_{\text{COD}}(\text{lb total COD removed}) \\ &= a_{\text{BOD}}(\text{lb total BOD removed})\end{aligned} \tag{5.47}$$

No subscripts are utilized for the parameter *a* hence. It is understood that the appropriate value of parameter *a* is chosen to be compatible with the parameters for expressing substrate removal.

4.1.5. An Observation Concerning Factor 1.42

Although factor 1.42 in Eq. (5.42) is obtained in Section 4.1.4 from consideration of the specific lactose example, it is shown next that it applies to all substrates, provided the average empirical formula for the MLVSS is taken as $C_5H_7NO_2$. Consider the specific lactose example. Write Eqs. (5.20) and (5.22), multiplying the first one by a factor of 5, i.e.,

$$5(CH_2O) + 5O_2 \rightarrow 5CO_2 + 5H_2O \tag{5.48}$$

Molecular weight: 5×30 $\quad 5 \times 32$

Recall that factor 1.42 originated from [Eq. (5.42)].

$$(150/113)(32/30) = 1.42$$

or (5.49)

$$[5(30)]/113 \times 32/30 = 1.42$$

The "molecule" of substrate is defined here as a sugar group (CH_2O) containing *one* carbon atom, which corresponds to a "molecular weight" of 30. Notice that in Eq. (5.49), the molecular weight of substrate (30 in this case) is canceled out. For any substrate of molecular weight *M*, Eq. (5.49) is

$$5M/113 \times 32/M = (5 \times 32)/113 = 1.42 \tag{5.50}$$

Thus, Eq. (5.42) is an *approximate* equation for most substrates, the only restriction being the assumption that the average empirical formula for MLVSS is $C_5H_7NO_2$. In Section 4.1.9, it is shown that value 1.42 corresponds to lb of oxygen required to oxidize 1 lb of MLVSS during the process of endogenous respiration.

4.1.6. Summary

Note: Approximate values of the ratio between parameters for expressing oxygen demand are taken from Table 2.1.

a. Parameter a′ in Different Units (See Tabulation Below)

a'_{ThOD}	lb $O_2 = a'_{\mathrm{ThOD}}$ (lb total ThOD removed) (energy) where $a'_{\mathrm{ThOD}} = a'$ = fraction of substrate removed utilized for energy production
a'_{COD} (standard COD test)	lb $O_2 = a'_{\mathrm{COD}}$(lb total COD removed) (energy) where $a'_{\mathrm{COD}} = a'(\mathrm{ThOD/COD}) = a'(100/83) = 1.20a'$
a'_{BOD} (5-day BOD)	lb $O_2 = a'_{\mathrm{BOD}}$(lb total BOD removed) (energy) where $a'_{\mathrm{BOD}} = a'(\mathrm{ThOD/BOD}) = a'(100/58) = 1.72a'$

Relationships for other oxygen and carbon parameters studied in Chapter 2 are readily written.

b. Parameter a in Different Units (See Tabulation Below)

a_{ThOD}	lb MLVSS produced = a_{ThOD}(lb total ThOD removed) where $a_{\mathrm{ThOD}} = \bar{a}/1.42$; $\bar{a}$ = fraction of substrate removed utilized for synthesis
a_{COD} (standard COD test)	lb MLVSS produced = a_{COD}(lb total COD removed) where $a_{\mathrm{COD}} = a_{\mathrm{ThOD}}(\mathrm{ThOD/COD}) = a_{\mathrm{ThOD}}(100/83) = (\bar{a}/1.42)(100/83) = 0.85\bar{a}$
a_{BOD} (5-day BOD)	lb MLVSS produced = a_{BOD}(lb total BOD removed) where $a_{\mathrm{BOD}} = a_{\mathrm{ThOD}}(\mathrm{ThOD/BOD}) = a_{\mathrm{ThOD}}(100/58) = (\bar{a}/1.42)(100/58) = 1.21\bar{a}$

Relationships for other oxygen and carbon parameters defined in Chapter 2 are readily written.

c. *Equation (5.24) Written with Different Units for the Parameters (See Tabulation Below)*

a' = fraction of the total substrate removed utilized for energy $= a'_{\text{ThOD}}$; $\bar{a}$ = fraction of total substrate removed utilized for synthesis. Then $\bar{a}+a' = 1.0$.

ThOD	$1.42a_{\text{ThOD}}+a'_{\text{ThOD}} = 1.0$ or $1.42a_{\text{ThOD}}+a' = 1.0$
COD (standard COD test)	$a_{\text{ThOD}} = a_{\text{COD}}(\text{COD}/\text{ThOD}) = a_{\text{COD}}(83/100)$ $a'_{\text{ThOD}} = a'_{\text{COD}}(\text{COD}/\text{ThOD}) = a'_{\text{COD}}(83/100)$ $\therefore\ 1.42(83/100)a_{\text{COD}}+(83/100)a'_{\text{COD}} = 1.0$ $1.18a_{\text{COD}}+0.83a'_{\text{COD}} = 1.0$
BOD (5-day BOD)	$a_{\text{ThOD}} = a_{\text{BOD}}(\text{BOD}/\text{ThOD}) = a_{\text{BOD}}(58/100)$ $a' = a'_{\text{ThOD}} = a'_{\text{BOD}}(\text{BOD}/\text{ThOD}) = a'_{\text{BOD}}(58/100)$ $\therefore\ 1.42(58/100)a_{\text{BOD}}+(58/100)a'_{\text{BOD}} = 1.0$ $0.82a_{\text{BOD}}+0.58a'_{\text{BOD}} = 1.0$

Equation (5.24) is readily written in terms of other oxygen and carbon parameters defined in Chapter 2.

4.1.7. Design Parameters Corresponding to Endogenous Respiration: Introduction

Two design parameters, b and b', are defined corresponding to the endogenous respiration phase. Endogenous respiration involves oxidation of cellular matter in order to provide food for the microorganisms when the concentration of substrate has decreased considerably. It corresponds to the "cannibalistic feast" described in Section 3.1.

Assuming that the chemical formula for the MLVSS is $C_5H_7NO_2$, oxidation of cells corresponding to endogenous respiration is given by Eq. (5.51).

$$\begin{array}{lccl} & C_5H_7NO_2 + & 5O_2 & \rightarrow 5CO_2 + NH_3 + 2H_2O \qquad (5.51) \\ \text{Molecular weight:} & 113 & 5\times 32 = 160 & \end{array}$$

4.1.8. Definition of Parameter *b* (Endogenous Respiration)

Parameter b is defined as fraction of MLVSS per unit time (day^{-1}, hour^{-1}, etc.) oxidized during process of endogenous respiration. For example, a value of $b = 0.1\ \text{day}^{-1}$ means that 10% of the total lb of MLVSS present in the reactor at any time is oxidized per day. Therefore, endogenous respiration

$$b = \text{lb MLVSS oxidized/(day)(lb MLVSS in reactor)} \qquad (5.52)$$

Consequently, the lb of MLVSS oxidized per day are

$$\begin{array}{l}\text{lb MLVSS oxidized/day} = b(\text{lb MLVSS in reactor}) \\ \text{(endogenous respiration)}\end{array} \qquad (5.53)$$

MLVSS present in reactor at any time assuming steady state operation is constant, being given by

$$\text{lb MLVSS in reactor} = X_{v,a}V \tag{5.54}$$

where $X_{v,a}$ is the concentration of MLVSS, i.e., lb MLVSS per unit volume of reactor; and V the reactor volume.

Thus Eqs. (5.53) and (5.54) yield

$$\underset{\text{(endogenous respiration)}}{\text{lb MLVSS oxidized/day}} = bX_{v,a}V \tag{5.55}$$

4.1.9. Definition of Parameter *b′*

Parameter b' is defined as the lb of oxygen utilized per day per lb of MLVSS in the reactor for the process of endogenous respiration, i.e., [Eq. (5.56)]

$$b' = \text{lb O}_2/(\text{day})(\text{lb MLVSS inreactor}) \tag{5.56}$$

Thus, oxygen utilization for endogenous respiration is

$$\underset{\text{(endogenous respiration)}}{\text{lb O}_2/\text{day}} = b'(\text{lb MLVSS in reactor}) \tag{5.57}$$

or from Eq. (5.54)

$$\underset{\text{(endogenous respiration)}}{\text{lb O}_2/\text{day}} = b'X_{v,a}V \tag{5.58}$$

The approximate relationship between b and b' is written assuming that average empirical formula for MLVSS is $C_5H_7NO_2$, and that endogenous respiration corresponds to chemical equation (5.51). From Eqs. (5.52) and (5.56) ratio b'/b is [Eq. (5.59)]

$$b'/b = \text{lb O}_2/\text{lb MLVSS oxidized} \tag{5.59}$$

From Eq. (5.51) this ratio is

$$b'/b = 1.42 \tag{5.60}$$

Consequently, it takes approximately 1.42 lb of oxygen to oxidize 1 lb of MLVSS. This value is used as an approximation for aerobic degradation of most substrates.

Whereas parameters a and a' are ratios [Eqs. (5.25), (5.32), (5.33), and (5.34) for a'; and Eqs. (5.35) and (5.39) for a], b and b' are rates. Time is not involved in the definitions of a and a', but it is in those of b and b'.

4.2. MATERIAL BALANCE FOR DETERMINATION OF OXYGEN UTILIZATION

Knowledge of oxygen requirements to effect a specified BOD removal is necessary for specification of aeration equipment. From discussions in Sections 4.1.3 and 4.1.9 it follows that oxygen is required for two purposes:

(1) to oxidize substrate in order to provide energy requirements for cells [Eq. (5.20)] and (2) for the endogenous respiration process [Eq. (5.51)].

1. *Oxygen required for energy.* The lb of oxygen required per day are calculated from Eq. (5.31). Referring to Fig. 5.1 and symbols defined in Table 5.1,

$$\underset{\text{(for energy)}}{\text{lb } O_2/\text{day}} = a'(S_o - S_e)Q \tag{5.61}$$

Appropriate values of a' compatible with parameters in which total substrate removal $(S_o - S_e)$ is expressed are utilized in Eq. (5.61).

Example 5.1

Calculate the oxygen required for energy.

$a'_{\text{BOD}} = 0.79$ lb O_2 (for energy)/lb total BOD_5 removed*

$S_o = 893$ mg/liter

$S_e = 40$ mg/liter

$Q = 2.04$ MGD $\quad (2.04 \times 10^6$ gal/day)

Then

$$S_o - S_e = 893 - 40 = 853 \text{ mg/liter} = S_r \qquad \text{(total substrate removed)}$$

Therefore

$$S_o - S_e = S_r = 853 \frac{\text{mg BOD}_r}{\text{liter liquor}} = 853 \frac{\text{mg BOD}_r}{10^3\text{g liquor}}$$

$$= 853 \frac{\text{g BOD}_r}{10^6\text{g liquor}} = 853 \text{ ppm} = 853 \times 10^{-6} \frac{\text{g BOD}_r}{\text{g liquor}}$$

$$= 853 \times 10^{-6} \frac{\text{lb BOD}_r}{\text{lb liquor}} = 853 \frac{\text{lb BOD}_r}{\text{Mlb liquor}}$$

From Eq. (5.61),

$$\underset{\text{(for energy)}}{\text{lb } O_2/\text{day}} = 0.79 \frac{\text{lb } O_2}{\text{lb BOD}_r} \times 853 \times 10^{-6} \frac{\text{lb BOD}_r}{\text{lb liquor}}$$

$$\times 2.04 \times 10^6 \text{gal} \frac{\text{liquor}}{\text{day}} \times 8.34 \frac{\text{lb liquor}}{\text{gal liquor}}$$

$$= 11{,}500 \text{ lb } O_2/\text{day}$$

---

* Experimental determination of parameter a' is described in Section 6.3.2. Example 5.1 is simply an illustration of unit conversion. Value $a' = 0.79$ is determined experimentally (Example 5.5, Section 6.4).

If S_r is in mg/liter and Q in MGD owing to cancellation of factors 10^{-6} and 10^6, Eq. (5.61) becomes Eq. (5.62).

$$\underset{\text{(for energy)}}{\text{lb O}_2\text{/day}} = a'S_rQ \times 8.34 \tag{5.62}$$

2. *Oxygen required for endogenous respiration.* Equation (5.58) is utilized for this calculation, illustrated by Example 5.2.

Example 5.2

Calculate the oxygen required for endogenous respiration. Let

$b' = 0.15$ lb O_2/(day)(lb MLVSS in reactor)*

$X_{v,a} = 300$ mg/liter (of MLVSS)

$V = 1.2$ MG (1.2×10^6 gal) (reactor volume)

By a similar procedure to that in Example 5.1 it follows that [Eq. (5.63)]

$$\underset{\text{(endogenous respiration)}}{\text{lb O}_2\text{/day}} = b'X_{v,a}V \times 8.34 \tag{5.63}$$

where b' is the lb O_2/(day)(lb MLVSS in reactor), $X_{v,a}$ the mg/liter of MLVSS, and V the reactor volume (MG).
Consequently,

$$\underset{\text{(endogenous respiration)}}{\text{lb O}_2\text{/day}} = 0.15 \times 3000 \times 1.2 \times 8.34 = 4500 \text{ lb O}_2\text{/day}$$

SUMMARY Total oxygen utilization is given by the sum of Eqs. (5.61) and (5.58) as

$$\text{lb O}_2\text{/day} = a'(S_o - S_e)Q + b'X_{v,a}V = a'S_rQ + b'x_{v,a}V \tag{5.64}$$

For Examples 5.1 and 5.2,

$$\text{lb O}_2\text{/day} = 11{,}500 + 4500 = 16{,}000 \text{ lb O}_2\text{/day}$$

4.3. MATERIAL BALANCE FOR DETERMINATION OF NET YIELD OF BIOLOGICAL SLUDGE (MLVSS)

From Sections 4.1.4 and 4.1.8 it follows that (1) a fraction of the substrate removed is utilized in production of MLVSS, the lb of MLVSS produced being given by Eq. (5.47), and that (2) part of the sludge produced is destroyed by oxidation (endogenous respiration), the lb of sludge oxidized being given by Eq. (5.55).

---

* Experimental determination of parameter b' is described in Section 6.3.2. Example 5.2 is simply an illustration of unit conversion. Value $b' = 0.15$ is determined experimentally (Example 5.5, Section 6.4).

1. *Sludge produced from substrate removal.* Sludge produced in lb/day is calculated from Eq. (5.47), where total substrate removal refers to one-day production. Referring to Fig. 5.1 and symbols defined in Table 5.1,

$$\text{lb/day of MLVSS produced} = a(S_o - S_e)Q = aS_rQ \tag{5.65}$$

Appropriate values of a compatible with parameters in which total substrate removal $(S_o - S_e)$ is expressed are utilized in Eq. (5.65).

Example 5.3

Calculate MLVSS produced by substrate removal. Let

a = 0.575 lb MLVSS produced/lb total BOD_5 removed*

S_o = 893 mg/liter

S_e = 40 mg/liter

Q = 2.04 MGD (2.04 × 10^6 gal/day)

Conversion of units for Eq. (5.65) is similar to that for Eq. (5.61) (Example 5.1, Section 4.2). The final result is Eq. (5.66).

$$\text{lb/day MLVSS produced} = aS_rQ \times 8.34 \tag{5.66}$$

where S_r is in mg/liter and Q in MGD.
Therefore,

$$\begin{aligned}\text{lb/day MLVSS produced} &= 0.575(893-40) \times 2.04 \times 8.34 \\ &= 8342 \text{ lb/day of MLVSS}\end{aligned}$$

2. *Sludge destroyed by endogenous respiration.* Sludge destroyed by endogenous respiration is obtained from Eq. (5.55). This calculation is illustrated by Example 5.4.

Example 5.4

Calculate MLVSS destroyed by endogenous respiration. Let

b = 0.075 lb MLVSS oxidized/(day)(lb MLVSS in reactor) = day^{-1}†

$X_{v,a}$ = 3000 mg/liter

V = 1.2 MG (1.2 × 10^6 gal; reactor volume)

---

* Experimental determination of parameter a is described in Section 6.3.4. Example 5.3 is simply an illustration of unit conversion. Value $a = 0.575$ is determined experimentally (Example 5.5, Section 6.4).

† Experimental determination of the parameter b is described in Section 6.3.4. Example 5.4 is simply an illustration of unit conversion. Value $b = 0.075$ is determined experimentally (Example 5.5, Section 6.4).

Conversion of units for Eq. (5.55) is similar to that for Eq. (5.58) (Example 5.2, Section 4.2). The final result is Eq. (5.67).

$$\text{lb MLVSS oxidized/day} = bX_{v,a}V \times 8.34 \tag{5.67}$$

where $X_{v,a}$ is in mg/liter and V in MG.
Therefore

$$\text{lb MLVSS oxidized/day} = 0.075 \times 3000 \times 1.2 \times 8.34 = 2252 \text{ lb/day of MLVSS}$$

SUMMARY Net yield of MLVSS is obtained by the difference between MLVSS produced [Eq. (5.65)] and MLVSS oxidized (endogenous respiration), given by Eq. (5.55). This net yield in lb/day is denoted as ΔX_v [Eq. (5.68)].

$$\underset{\text{(net yield)}}{\text{lb MLVSS/day}} = \Delta X_v = a(S_o - S_e)Q - bX_{v,a}V = aS_rQ - bX_{v,a}V \tag{5.68}$$

For examples 5.3 and 5.4

$$\Delta X_v = 8342 - 2252 = 6090 \text{ lb/day}$$

4.4. TOTAL SLUDGE YIELD

So far, only the yield of biological sludge (MLVSS) has been considered. Now, examine the diagram for the reactor system in Fig. 5.1. The fresh feed may contain nonvolatile suspended solids (NVSS). Let $X_{NV,F}$ be the concentration (mg/liter) of these NVSS.

Reactor contents are under conditions of complete mixing, therefore no settling of MLNVSS (or MLVSS) takes place. Consequently, concentration of NVSS in reactor effluent is the same as that in combined feed ($X_{NV,o}$). In the secondary clarifier, however, part of the NVSS as well as most of VSS settles. Let $X_{NV,u}$ be the concentration of NVSS in underflow from the clarifier (same as in wastage Q and recycle Q_R). Concentration of NVSS in net effluent from clarifier (Q') is $X_{NV,e}$.

Wastage of sludge corresponds to

1. Net yield of biological sludge (MLVSS) from the reactor. This is ΔX_v [Eq. (5.68)]. Since the reactor operates at steady state, this wastage is equal to net yield of MLVSS, so that total lb of MLVSS in the reactor remain the same at all times. In addition, wastage includes volatile solids entering with fresh feed ($Q_F X_{V,F}$), as seen from an overall balance of volatile solids (loop --- in Fig. 5.1). Therefore, total wastage of MLVSS is shown in Eq. (5.69) [utilizing Eq. (5.68) for ΔX_v].

$$\Delta X_v + Q_F X_{V,F} = a(S_o - S_e)Q - bX_{v,a}V + Q_F X_{V,F} \tag{5.69}$$

2. Settled NVSS denoted as ΔX_{NV} (lb/day). This value is determined by

an overall material balance for NVSS over loop --- in Fig. 5.1.

NVSS, IN: $Q_F X_{NV,F}$

NVSS, OUT: $Q'X_{NV,e} + Q''X_{NV,u} = Q'X_{NV,e} + \Delta X_{NV}$ (since $\Delta X_{NV} = Q''X_{NV,u}$) (5.70)

Thus the overall balance is [Eq. (5.71)]

$$Q_F X_{NV,F} = Q'X_{NV,e} + \Delta X_{NV}$$

$$\therefore\ \Delta X_{NV} = Q_F X_{NV,F} - Q'X_{NV,e} \tag{5.71}$$

Eliminating Q' and utilizing Eq. (5.2),

$$\Delta X_{NV} = Q''X_{NV,u} = Q_F X_{NV,F} - (Q_F - Q'')X_{NV,e} = Q_F(X_{NV,F} - X_{NV,e}) + Q''X_{NV,e} \tag{5.72}$$

Substitution of ΔX_v and ΔX_{NV} in Eq. (5.1) by their values given by Eqs. (5.68) and (5.72) yields total sludge yield ΔX_t [Eq. (5.73)].

$$\Delta X_t = a(S_o - S_e)Q - bX_{v,a}V + Q_F X_{V,F} + Q_F(X_{NV,F} - X_{NV,e}) + Q''X_{NV,e} \tag{5.73}$$

where

$a(S_o - S_e)Q - bX_{v,a}V = \Delta X_v =$ net yield of MLVSS [Eq. (5.68)]

$Q_F X_{V,F} =$ MLVSS in fresh feed

$Q_F(X_{NV,F} - X_{NV,e}) + Q''X_{NV,e} = \Delta X_{NV} =$ net yield of sludge due to settling NVSS from influent [Eq. (5.72)]

4.5. MATERIAL BALANCES FOR $X_{NV,o}$ AND $X_{V,o}$

The value of $X_{NV,o}$, i.e., concentration of NVSS in combined feed, is established by a material balance around the junction of the fresh feed with the recycle to form combined feed (Fig. 5.1, loop -·-·-).

$$\text{NVSS, IN} = Q_F X_{NV,F} + Q_R X_{NV,u}$$

$$\text{NVSS, OUT} = QX_{NV,o}$$

Then

$$Q_F X_{NV,F} + Q_R X_{NV,u} = QX_{NV,o}$$

Utilizing Eqs. (5.4) and (5.5) and solving for $X_{NV,o}$,

$$X_{NV,o} = (X_{NV,F} + rX_{NV,u})/(1+r) \tag{5.74}$$

A similar material balance is written for $X_{v,o}$, the concentration of VSS in combined feed. Final result is

$$X_{v,o} = (X_{V,F} + rX_{v,u})/(1+r) \tag{5.75}$$

4.6. TYPICAL VALUES OF AEROBIC BIOLOGICAL WASTEWATER TREATMENT PARAMETERS FOR DIFFERENT TYPES OF WASTEWATERS

Typical values of these parameters are presented in Table 5.2.

TABLE 5.2
Aerobic Biological Waste—Treatment Parameters[a, b]

Wastewater	a	a'	b	b'[c]	k
Domestic	0.73	0.52	0.075	0.106	0.017–0.03
Refinery	0.49–0.62	0.40–0.77	0.10–0.16	0.142–0.227	0.074
Chemical and petrochemical	0.31–0.72	0.31–0.76	0.05–0.18	0.071–0.255	0.0029–0.018
Brewery	0.56	0.48	0.10	0.142	—
Pharmaceutical	0.72–0.77	0.46	—	—	0.018
Kraft pulping and bleaching	0.5	0.65–0.8	0.08	0.114	—

[a] Adapted from Ref. [2].

[b] Units: a, lb MLVSS produced/lb total BOD_5 removed; b, lb MLVSS oxidized/(day)(lb MLVSS in reactor) = day^{-1}; a', lb O_2 (for energy)/lb total BOD_5 removed; b', lb O_2/(day)(lb MLVSS in reactor) = day^{-1}; k, day^{-1}.

[c] Values of b' estimated from $b' = 1.42b$.

5. Relationship for Optimum Settling Conditions of Sludge

For adequate operation of the activated sludge process, MLVSS in the reactor effluent should be readily separated in the secondary clarifier. The condition occurring when sludge is light and fluffy and thus difficult to settle is termed bulking. Bulky sludge flakes over separating weirs and comes out with the secondary clarifier effluent. Since concentration of substrate in the effluent is small, there is not enough food material to sustain the growth of the microorganisms which constitute the sludge. Therefore the microorganisms are driven to endogenous respiration. Owing to the consumption of oxygen for endogenous respiration, the effluent has a relatively high BOD, which is undesirable.

Settling characteristics of sludge are evaluated from sedimentation tests performed in the laboratory. For this evaluation two parameters are utilized.

1. *Zone settling velocity* (*ZSV*). This parameter and its experimental determination are discussed in Chapter 3, Section 3.6. An easily settling sludge has a high ZSV of about 20 ft/hr.

2. *Sludge volume index* (*SVI*). Sludge volume index is defined as volume (in cm^3) occupied by 1 g of dry sludge solids after settling for 30 min. The smaller the SVI, the easier is the settling of the sludge.

Several authors have correlated settling characteristics of sludge (in terms of ZSV or SVI) with a parameter designated as food to microorganism ratio (hence denoted as F/M). This parameter is defined as [Eq. (5.76)]

$$F/M = \text{lb of substrate in influent/(day)(lb MLVSS in reactor)} \tag{5.76}$$

Values of F and M are given by

$$F = (QS_o) \times 8.34 \qquad \text{(lb/day)} \tag{5.77}$$

$$M = (X_{v,a}V) \times 8.34 \qquad \text{(lb)} \tag{5.78}$$

where Q is in MGD and $(S_o, X_{v,a})$ in mg/liter. Therefore

$$F/M = QS_o/X_{v,a}V \tag{5.79}$$

Since $V/Q = t =$ residence time,

$$F/M = S_o/X_{v,a}t \qquad (\text{day}^{-1}) \tag{5.80}$$

In order to arrive at correlations for settling characteristics of a sludge, a series of bench scale continuous reactors are operated, each at a selected F/M ratio. Sludge obtained in each reactor is subjected to settling tests (ZSV and SVI). If these two parameters, which are a measure of the ability of the sludge to settle, are plotted vs. the corresponding F/M ratios, curves like the ones shown in Fig. 5.7 are obtained.

Since for optimum settling the sludge should have a high ZSV and a low SVI, the optimum F/M ratio as indicated in Fig. 5.7 corresponds to the maximum for the ZSV curve and the minimum for the SVI curve. For most wastewaters this optimum value of the F/M ratio falls between the following limits [Eq. (5.81)]:

$$0.6 > F/M > 0.3 \tag{5.81}$$

where F/M is expressed in lb BOD_5 influent/(day)(lb MLVSS). An explanation for the correlation between F/M ratio and sedimentation characteristics of the sludge is given below.

1. At low F/M ratios (e.g., below $F/M = 0.3$) the amount of food (substrate) present in the system is insufficient to maintain the growth of the microorganisms. Therefore, they are driven to endogenous respiration. A typical bacterial cell is shown in Fig. 5.8. Cytoplasmic material is rich in proteins and ribonucleic acid (RNA), and it is the main portion of the cell which is metabolized during the process of endogenous respiration. The residue left from endogenous metabolism is constituted mainly by cell capsules, which are very light and resist sedimentation. This is why at low F/M ratios,

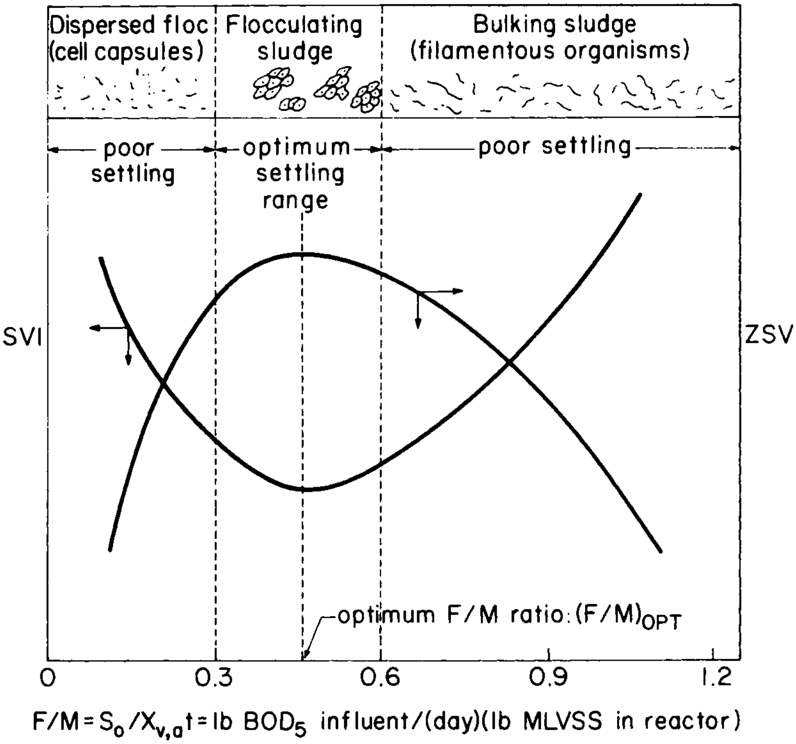

Fig. 5.7. Typical correlation of SVI and ZSV with F/M ratio.

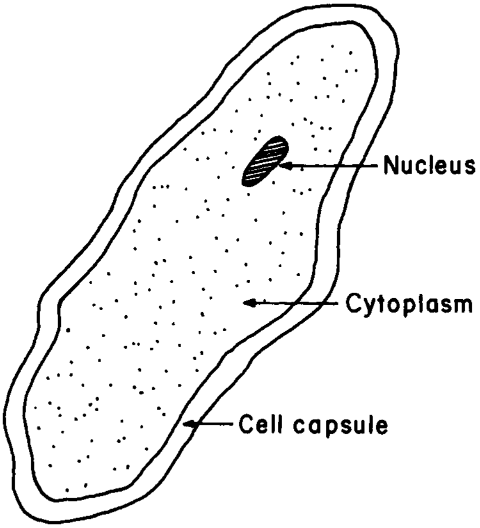

Fig. 5.8. Typical bacterial cell.

the sludge has poor settling characteristics. Sludge obtained under these conditions is referred to as dispersed floc, and a microscopic view of it is shown in Fig. 5.7 for the region of low F/M ratios.

2. At high F/M ratios (e.g., $F/M > 0.6$) there is predominance of a type of microorganism which is filamentous in nature (*Sphaerotilus*). This type of growth does not settle well, remaining in suspension almost indefinitely. Sludge under these conditions is referred to as a bulking sludge.

3. At values of the F/M ratio between these two extremes, sludge with good settling characteristics is obtained. Sludge under these conditions is referred to as flocculating sludge.

From Eq. (5.80) the residence time t to yield an optimum flocculating sludge is obtained. Written for the optimum F/M ratio as determined from Fig. 5.7, Eq. (5.80) is

$$(F/M)_{\mathrm{OPT}} = S_o/X_{v,a}t \tag{5.82}$$

Solving for t,

$$t = S_o/[X_{v,a}(F/M)_{\mathrm{OPT}}] \tag{5.83}$$

The geometry of the system and the manner in which wastewater is fed to the aerator have an effect on flocculating characteristics of the sludge. For example, if the aerator is a long rectangular tank with relatively poor mixing, MLVSS is initially contacted at the feed end with entering sewage, and therefore a high F/M ratio prevails at the entrance. Filamentous growth developed under these conditions persists throughout the aeration period, and sludge with poor settling characteristics is obtained (Fig. 5.9). The same situation occurs in a batch reactor, since a high F/M ratio prevails at the start of the operation. The reactor depicted in Fig. 5.9 is the plug flow continuous reactor. A general discussion of the kinetics of a continuous treatment system (plug flow, complete mix, and arbitrary flow reactors) is presented in Section 10.

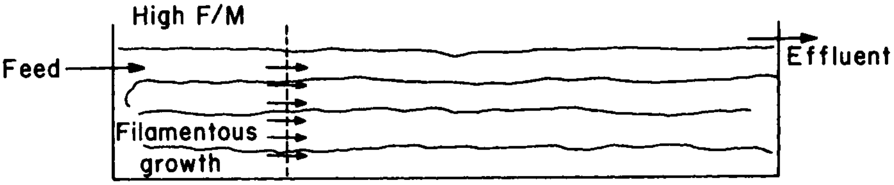

Fig. 5.9. Effect of geometry in settling characteristics of MLVSS (plug flow model).

If there is complete mixing in the system, the F/M ratio is uniform throughout, possibly falling within the optimum range. Under steady state and complete mix conditions, sludge is always in contact with a BOD concentration equal to that in the effluent. Therefore a dense sludge is likely to be obtained.

It is important to obtain experimentally the graph in Fig. 5.7 for the specific substrate under study, since considerable variation occurs depending on substrate characteristics. Substrates which are easily degradable (e.g., soluble sugars) become immediately available as food to the microorganisms, and therefore the result is a fast growth response. On the other hand, complex organic substrates (e.g., wastewaters from petroleum and petrochemical plants) must undergo chemical breakdown before being available as food to the microorganisms, growth response being therefore slower.

6. Experimental Determination of Parameters Needed for Design of Aerobic Biological Reactors

6.1. BENCH SCALE CONTINUOUS REACTORS

A bench scale continuous reactor utilized for these determinations is described in this section. Parameters to be determined are defined in Sections 4.1.2 to 4.1.9, i.e., for kinetic relationship: k; for material balance relationships: a, a', b, and b'. A diagram of the continuous flow reactor is shown in Fig. 5.10. This unit is designed and built by Bio-Development Associates, Austin, Texas. The reactor is made of plexiglass and divided into two sections: the aeration and settling chambers. These simulate the reactor and the secondary clarifier for an actual plant.

Capacity of the aeration chamber is approximately 7 liters. Air is supplied as indicated in the diagram. Bubbling air keeps the contents of the aeration chamber in a completely mixed condition. Wastewater is fed continuously from a constant head feed reservoir by means of a Sigmamotor pump, and overflows continuously into the effluent bottle. The aeration and sedimentation chambers are separated by a sliding baffle which can be completely removed if desired.

Start-up is performed by placing in the aeration chamber a seed of domestic activated sludge collected from an operating plant, and gradually acclimating it to the wastewater under study. For wastewaters of industrial origin containing compounds which are toxic to the microorganisms, mixtures of industrial wastewater and domestic sewage are fed to the reactor with a gradually increased proportion of industrial wastewater. Eventually, feed is 100% industrial wastewater without deleterious effects on the microorganisms.

Flow rate is varied by proper setting of the Sigmamotor pump, and by utilizing different internal diameters for the Tygon tubing. A Sigmamotor pump operates by "squeezing" the wastewater through the Tygon tubing by means of mechanical "fingers," the speed of which is set. One pump promotes wastewater flow through several reactor units in parallel, each one provided

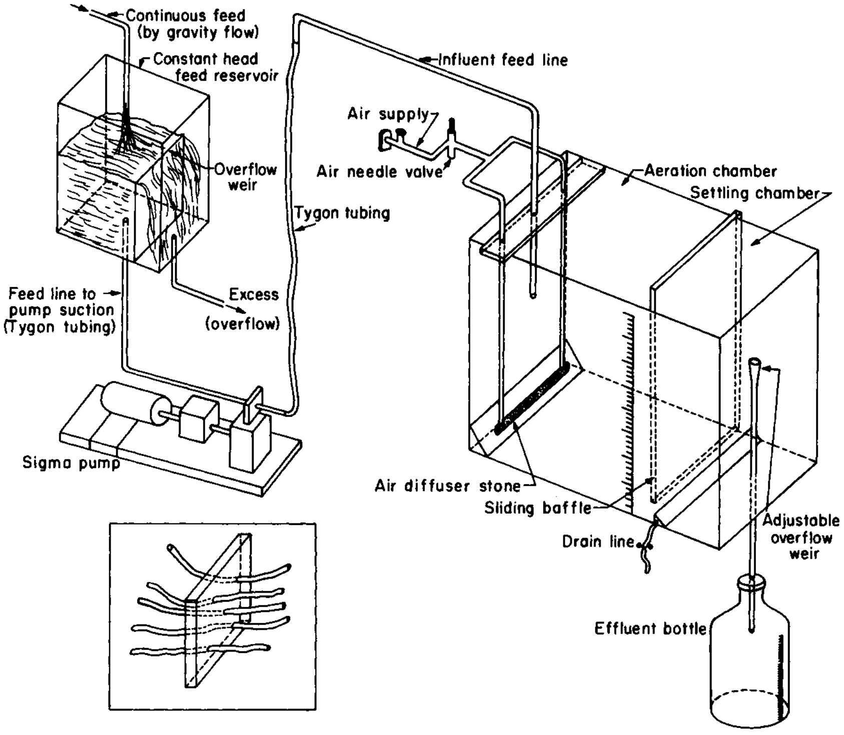

Fig. 5.10. Continuous flow reactor (bench scale model). Insert: detail of Sigma pump setup for operation of five reactors in parallel.

with its own Tygon feed line [Fig. 5.10 (insert)]. The "fingers" simultaneously squeeze these several Tygon tubings, promoting different flow rates for each line depending on the internal diameter of each tubing.*

Flow rates are determined by calibration, either weighing or measuring the volume of effluent obtained during a timed period corresponding to a selected settling of the pump and a chosen internal diameter of tubing. Flow rates are reproducible within less than 1% fluctuation.

Flow rates vary considerably, e.g., from 350 down to about 1.0 liter/day. For an aerator chamber volume of 7.0 liters, these rates correspond to residence times of

$$Q = 350 \text{ liters/day}, \quad t = V/Q = 7/(350/24) = 0.48 \text{ hr} \approx 30 \text{ min}$$

$$Q = 1 \text{ liter/day}, \quad t = V/Q = 7/1 = 7 \text{ days}$$

As the section of Tygon tubing subjected to this continuous squeezing

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* Sigmamotor pump model T-8 (manufactured by Sigmamotor Inc., Houston, Texas) an be used to operate five units in parallel.

action wears out, it softens and flow rates change. It is advisable to slide the tubing along at periodic intervals, so that a new section of it becomes exposed to the squeezing action. Frequent calibration is performed to ensure confidence in the results. Tubing is replaced after it is worn out.

The main difference in operating principle between this bench scale reactor and the one in plant scale (Fig. 5.1) is that no controlled recycle of sludge is provided in the bench scale unit. Sludge is returned to aeration chamber from the settling chamber through the opening between the baffle and the bottom of the unit. This rate of return cannot be controlled. It is desirable to maintain the concentration of MLVSS in the aeration chamber approximately constant (at a selected value usually between 2000 and 3000 mg/liter). In order to achieve this constant MLVSS concentration, the procedure is

1. Determine periodically the MLVSS concentration in the aerator liquor from samples withdrawn through the drain line.

2. Withdraw calculated weights of MLVSS in order to keep this concentration at the selected value for a given experiment. For a reactor operating with MLVSS under endogenous respiration conditions, it is necessary to add sludge instead of withdrawing it, in order to keep a constant MLVSS concentration.

When the sliding baffle is inserted, the bench scale reactor is utilized to simulate the activated sludge unit as described. By removing the sliding baffle, simulation of an aerated lagoon is obtained (Chapter 6, Section 5).

6.2. EXPERIMENTAL PROCEDURE

Each experiment requires 2–4 weeks before steady state conditions are achieved. For this reason it is convenient to operate simultaneously four or five reactors in parallel.

Steps in the experimental procedure are [3]

1. Each unit is filled with seed sludge up to a predetermined volume. Dilution is made with wastewater in order to obtain a MLVSS concentration of 2000–3000 mg/liter.

2. Air is turned on and contents of the aeration chamber are completely mixed by the turbulence thus produced. The sliding baffle is adjusted to leave an opening of $\frac{1}{4}$ to $\frac{1}{2}$ in. at the bottom. During operation of the reactor, further baffle adjustments are made in order to provide a desired blanket height of sludge in the settling chamber and an interchange of sludge between the two chambers (Fig. 5.11).

3. Start the Sigmamotor pump at a flow rate necessary to obtain the desired residence time in the aeration chamber. Acclimation of sludge, if required, is performed as previously described.

4. Operate the reactor until steady state conditions are achieved. Attainment of steady state is assumed when two criteria are satisfied: (a) oxygen

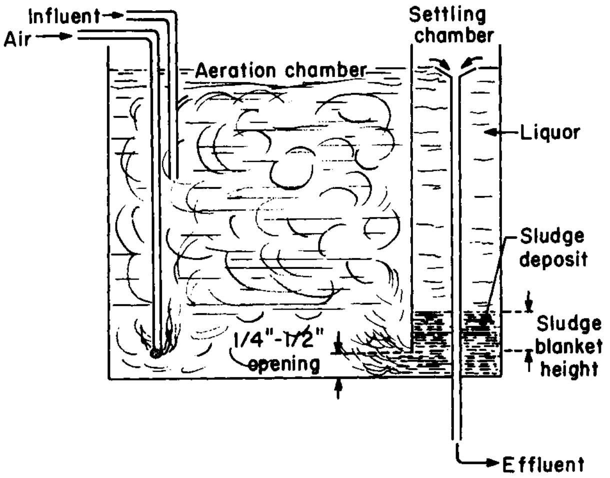


Fig. 5.11. Side view (section) of aeration and settling chambers.

uptake rate of reactor contents remains unchanged (determination of oxygen uptake rate is described in Section 6.3.3) and (b) BOD of effluent becomes stable.

5. Concentration of MLVSS is measured daily and adjusted to a nearly constant value for the duration of an experiment. To check net increase of MLVSS, plug overflow weir, raise baffle, and withdraw a sample from mixed tank contents. If V_t is the total volume (aeration chamber plus settling chamber) and two determinations of MLVSS are made, e.g., 24 hr apart yielding values X_1 and X_2, respectively, the net increase of MLVSS is

$$\Delta X_v = V_t X_2 - V_t X_1 = V_t(X_2 - X_1) \quad \text{(24-hr growth)} \qquad (5.84)$$

Values (X_1, X_2) represent "averaged" concentrations of MLVSS for the total volume of the tank, since the baffle has been raised and the contents of aeration and sedimentation chambers mixed. Essentially, no sludge growth occurs in the sedimentation chamber because there is no direct aeration there. Therefore, the value of ΔX_v calculated from Eq. (5.84) represents the net growth occurring in the aerator.

For application of Eq. (5.68), it is recommended to take $X_{v,a}$ as the MLVSS concentration determined after the baffle is raised. This may seem controversial since in Eq. (5.68) $X_{v,a}$ stands for MLVSS concentration in aeration chamber during the operation (V is the volume of aeration chamber). The concentration of MLVSS in the sedimentation chamber is probably different from that in the aerator. At the bottom of the sedimentation chamber there is a sludge blanket of very high MLVSS concentration, and at the top a supernatant liquid with negligible MLVSS concentration. After the baffle is raised, this heterogeneous mass in the sedimentation chamber is mixed with the contents of the aerator. The whole volume is thoroughly mixed by the bubbling air before the sample is taken. The designers of this laboratory reactor claim that there is no significant difference between MLVSS concentration in the aeration chamber during operation and that in the whole mixed content of the two chambers. In any event, it is practically impossible to withdraw representative samples from the aeration chamber during the operation for analysis of $X_{v,a}$. Recall also that the volume of the sedimentation chamber is much smaller than that of the aeration chamber (ratio of about 3/7). Therefore, MLVSS concentration in the mixed contents of the two chambers is not too different from that of the aeration chamber during operation.

6. Once steady state operation is attained, the sampling schedule presented in Table 5.3 is followed.

TABLE 5.3
Sampling Schedule [3]

Analysis	Frequency	Raw waste[a]	Mixed liquor[b]	Effluent[c]
1. COD, BOD, or TOC (mg/liter) (filtered and unfiltered composite samples)	3/week	x (S_o)	—	x (S_e)
2. pH	daily	x	x	x
3. SS, MLVSS (mg/liter) (also determine sludge settling curves and sludge volume index of mixed liquor at the end of test run)	3/week	—	x ($X_{v,a}$)	x (keep low)
4. Dissolved oxygen (DO) (mg/liter)	daily	—	x	—
5. Oxygen uptake rate	3/week	—	x	—
6. Microscopic analysis (gram stain)	1/week	—	x	—
7. Color, turbidity	3/week	—	—	x
8. Significant ions, compounds	3/week	x	—	x

[a] Sample withdrawn from influent feed line or raw waste containers.
[b] Sample withdrawn from the unbaffled tank.
[c] Sample withdrawn from effluent bottle.

6.3. CALCULATION OF DESIGN PARAMETERS

Calculation of parameters k, a, a', b, and b' is made from obtained data. Procedure is described in Sections 6.3.1–6.3.4.

6.3.1. Determination of Substrate Removal Rate (*k*)

This determination, based on Eq. (5.18) or Eq. (5.19), is described in Section 3.2.

6.3.2. Determination of Oxygen Utilization Parameters *a′* and *b′*

This determination is based on Eq. (5.64) in which the left-hand member is written as $R_r V$, i.e.,

$$R_r V = a'(S_o - S_e)Q + b'X_{v,a}V \tag{5.85}$$

where R_r is the oxygen uptake rate, i.e., oxygen utilized per day per unit volume of reactor; and V the reactor volume.

Experimental determination of R_r is discussed in Section 6.3.3. Dividing Eq. (5.85) by $X_{v,a}V$ and letting $V/Q = t$ (residence time) yields

$$R_r/X_{v,a} = a'[(S_o - S_e)/X_{v,a}t] + b' \tag{5.86}$$

Equation (5.86) is the basic relationship for determination of oxygen utilization parameters a' and b'. Notice the presence of term $(S_o - S_e)/X_{v,a}t$ (substrate removal rate), which also occurs in Eqs. (5.18) and (5.19) for determination of k.

Units for R_r obtained from laboratory scale determinations are metric, i.e., mg O_2/(day)(liter). Since

$$\text{mg O}_2\text{/liter liquor} = \text{lb O}_2\text{/Mlb liquor} \qquad \text{(Section 4.2, Example 5.1)}$$

then

$$R_r = \text{lb O}_2\text{/(day)(Mlb liquor)}$$

Similarly, for $X_{v,a}$

$$X_{v,a} = \text{mg MLVSS/liter liquor} = \text{lb MLVSS/Mlb liquor}$$

Therefore in Eq. (5.86)

$$R_r/X_{v,a} = \frac{\text{lb O}_2\text{/(day)(Mlb liquor)}}{\text{lb MLVSS/Mlb liquor}} = \text{lb O}_2\text{/(day)(lb MLVSS)}$$

Thus $R_r/X_{v,a}$ is a measure of utilization of oxygen per day and per lb of biological sludge present in the reactor.

As shown in Section 3.2,

$$(S_o - S_e)/X_{v,a}t = \text{lb BOD removed/(day)(lb MLVSS)}$$

According to Eq. (5.86) a plot of $R_r/X_{v,a}$ vs. $(S_o - S_e)/X_{v,a}t$ yields a straight line from the slope and intercept of which oxygen utilization parameters a'

and b' are obtained. A typical plot is shown in Fig. 5.16, and a numerical illustration of its construction from laboratory data is presented in Example 5.5 (Section 6.4).

6.3.3. Experimental Determination of the Oxygen Uptake Rate (R_r)

Possibly the simplest way to determine the oxygen uptake rate is by galvanic cell oxygen measurements. This is the only method described in this section. Other methods are polarographic and Warburg techniques and off-gas analysis. Of all these methods, galvanic cell measurement is the simplest, and its accuracy is usually adequate. The apparatus for this measurement is the dissolved oxygen analyzer (DO analyzer) described in Chapter 2 (Section 2.3.1) and shown in Fig. 2.4.

Experimental technique for measuring oxygen uptake rate (R_r) is [6]

1. Fill BOD bottle with aerated mixed liquor from test solution.

2. Insert probe into bottle, allowing displaced liquid to overflow. Care is taken to prevent accumulation of air bubbles inside bottle.

3. Mix the contents using a magnetic stirring apparatus.

4. Record galvonometer readings at various time intervals, usually every 30 sec.

5. Correct readings based on a predetermined sensitivity factor (for details refer to [6]), and plot dissolved oxygen level (ordinate) vs. time (abscissa) (Fig. 5.12).

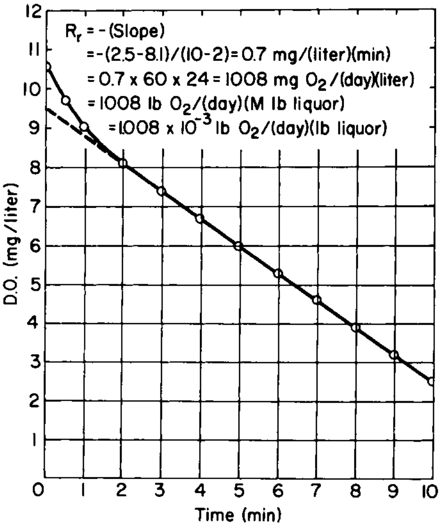


Fig. 5.12. Determination of oxygen uptake rate.

6. In Fig. 5.12, the slope of the line is oxygen uptake rate in mg/(liter)(min). A specific uptake rate ($R_r/X_{v,a}$) is then determined by dividing this value by MLVSS concentration in the test sample. In Fig. 5.12, the first data points immediately after $t = 0$ are *not* to be taken into account in evaluating the slope. The higher slope of this section of the line is due to loss of entrained air from the liquor. After a few minutes the slope becomes stabilized, and it is taken as the uptake rate.

A temperature correction available from nomographs furnished by the manufacturer is applied to the readings. Probe readings are inaccurate at DO concentrations below 0.5 mg/liter. Transfer of the mixed liquor from the reactor to the DO analyzer bottle should be rapid, and the test started as soon as possible following sample withdrawal. If oxygen depletion is too rapid, the sample is diluted in order to reduce MLVSS concentration. It is advisable to calibrate the probe in a sample of water similar to that in which the DO analyzer is used, in order to eliminate errors due to the salt effect.

6.3.4. Determination of Parameters for Sludge Yield (*a* and *b*)

Determination of parameters *a* and *b* is based on Eq. (5.73). For the bench scale reactor there is no recycle of sludge, contrary to what happens for the reactor in Fig. 5.1, for which Eq. (5.73) is written.

A simplified diagram of the bench scale reactor is shown in Fig. 5.13. By comparing Fig. 5.13 with Fig. 5.1, terms in Eq. (5.73) are modified for application to the laboratory unit.

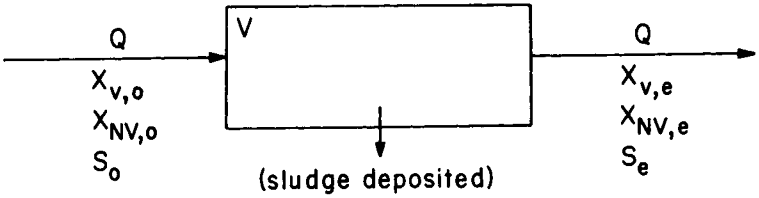


Fig. 5.13. Simplified diagram of the bench scale continuous reactor.

Fig. 5.1	Fig. 5.13
Q_F	Q
Q''	zero
$X_{V,F}$	$X_{v,o}$
$X_{NV,F}$	$X_{NV,o}$

Therefore, Eq. (5.73) for the laboratory reactor becomes

$$\Delta X_t = \underbrace{a(S_o - S_e)Q - bX_{v,a}V}_{\Delta X_v} + QX_{v,o} + \underbrace{Q(X_{NV,o} - X_{NV,e})}_{\Delta X_{NV}} \tag{5.87}$$

Equation (5.87) is rearranged as

$$\Delta X_t - Q(X_{NV,o} - X_{NV,e} + X_{v,o}) = \Delta X_v = a(S_o - S_e)Q - bX_{v,a}V \tag{5.88}$$

Dividing through by $X_{v,a}$ V and noticing that $V/Q = t$ (residence time),

$$\frac{\Delta X_t/V - (X_{NV,o} - X_{NV,e} + X_{v,o})/t}{X_{v,a}} = \frac{(\Delta X_v/V)}{X_{v,a}} = a[(S_o - S_e)/X_{v,a}t] - b \tag{5.89}$$

In the numerator of the left-hand member of Eq. (5.89), term $\Delta X_t/V$ equals the net yield of total sludge per unit volume [i.e., mg total sludge/(day)(liter)]. Term $\Delta X_v/V$ corresponds to the net yield of MLVSS per unit volume. If concentrations of NVSS and MLVSS in the influent are negligible (i.e., $X_{NV,o} \approx X_{NV,e} \approx X_{v,o} \approx 0$), this equation reduces to

$$\frac{\Delta X_t/V}{X_{v,a}} = \frac{\Delta X_v/V}{X_{v,a}} = a[(S_o - S_e)/X_{v,a}t] - b \tag{5.90}$$

Equation (5.89) [or Eq. (5.90)] is the basic relationship for determination of sludge yield parameters a and b. Notice again the presence of term $(S_o - S_e)/X_{v,a}t$ (substrate removal rate), which also occurred in Eqs. (5.18), (5.19), and (5.86) for determination of parameters k, a', and b'.

Note on units for Eq. (5.89) [or Eq. (5.90)]: From laboratory determinations, the value of $\Delta X_t/V$ is obtained in metric units, i.e., $\Delta X_t/V$ = mg total sludge yield/(day)(liter of liquor). From similar considerations as those for R_r (Section 6.3.2), it follows that this value is numerically equal to that expressed in English units, i.e., $\Delta X_t/V$ = lb sludge yield/(day)(Mlb liquor).

Therefore term $(\Delta X_t/V)/X_{v,a}$ in English units is

$$\frac{\Delta X_t/V}{X_{v,a}} = \frac{\text{lb total sludge yield/(day)(Mlb liquor)}}{\text{lb MLVSS/Mlb liquor}}$$

$$= \text{lb total sludge yield/(day)(lb MLVSS)}$$

Similarly,

$$\frac{\Delta X_t/V - (X_{NV,o} - X_{NV,e} + X_{v,o})/t}{X_{v,a}} = \text{lb MLVSS yield/(day)(lb MLVSS)}$$

According to Eq. (5.89) [or Eq. (5.90)] a plot of (accounting for presence of NVSS)

$$\left[\frac{\Delta X_t/V - (X_{NV,o} - X_{NV,e} + X_{v,o})/t}{X_{v,a}} = \frac{\Delta X_v/V}{X_{v,a}}\right] \text{vs. } (S_o - S_e)/X_{v,a}t$$

or simply (if NVSS is negligible)

$$\frac{\Delta X_v/V}{X_{v,a}} \text{ vs. } (S_o - S_e)/X_{v,a}t$$

yields a straight line from the slope and intercept of which design parameters a and b are obtained. A typical plot is shown in Fig. 5.17, and its construction from laboratory data is illustrated in Section 6.4, Example 5.5.

The abscissa intercept in Fig. 5.17 corresponds to a zero value for the ordinate. This occurs for a condition of net zero yield of MLVSS, i.e., $\Delta X_v = 0$. Referring to Eq. (5.68), for $\Delta X_v = 0$ it follows that production of MLVSS by synthesis, i.e., $a(S_o - S_e)Q$, is exactly balanced by loss of MLVSS oxidized by endogenous respiration, i.e., $bX_{v,a}V$. Therefore

$$a(S_o - S_e)Q = bX_{v,a}V$$

Thus, the length of abscissa intercept is $(S_o - S_e)/X_{v,a}t = b/a$, as indicated in Fig. 5.14.

In summary, the most important information derived from bench scale studies using this laboratory reactor is the organic removal capacity of an acclimated biological sludge receiving a predefined wastewater. Full scale plants operating on design criteria developed using this reactor produce an effluent which approximates the predicted quality. Moreover, oxygen utilization rates are scaled up with relative accuracy from bench scale reactors to full scale units. There is some difficulty, however, in scaling up and applying coefficients a and b developed from bench scale reactors to a full scale unit because of limitations due to low accuracy of the VSS test, and the difficulty of establishing a solids balance in small scale simulation studies. Using larger reactors of pilot-plant scale enhances the accuracy of these coefficients. Fortunately, the accuracy of coefficients a and b is less important for the designer than those for removal rate constant (k) and oxygen demand coefficients (a', b').

6.4. NUMERICAL EXAMPLES: DETERMINATION OF DESIGN PARAMETERS FOR AN ACTIVATED SLUDGE SYSTEM

Example 5.5

An industrial plant is considering an activated sludge system for treatment of their wastewaters. Preliminary tests are performed in laboratory scale continuous reactors (Fig. 5.10). The volume of the aeration chamber in laboratory reactors is 7 liters. Four reactors are operated in parallel until steady state conditions are obtained. Data taken are presented in Table 5.4.

The influent contains an average of 30 mg/liter of NVSS. In effluent, con-

TABLE 5.4
Laboratory Data

Reactor no.	Influent average concentration BOD_5 S_0 (mg/liter)	Effluent average concentration BOD_5 S_e (mg/liter)	Average MLVSS concentration $X_{v,a}$ (mg/liter)	Flow rate Q (liter/day)	Oxygen uptake rate R_r [mg O_2/(liter)(day)]	Sludge yield $\Delta X_t/V$ [mg sludge/(liter)(day)]	Sludge volume index SVI
1	880	100	3100	40.00	4025	2387	75.0
2	870	50	2800	14.90	1800	806	61.0
3	870	30	3000	8.75	1292	387	64.0
4	860	20	2900	3.50	780	295	69.0

TABLE 5.5
Removal Kinetics[a]

	Laboratory data				Calculated data					Lab. data
(1) Reactor no.	(2) S_o (mg/liter)	(3) S_e (mg/liter)	*(4)* $X_{v,a}$ (mg/liter MLVSS)	(5) Q (liter/day)	*(6)* Residence time $t = V/Q = 7.0/Q$ (day)	*(7)* $(7) = (2) - (3)$ $S_o - S_e$ (mg/liter)	*(8)* $(8) = (4) \times (6)$ $X_{v,a}t$ $\left[\frac{(\text{mg})(\text{day})}{\text{liter}}\right]$	*(9)* $(9) = (7) \div (8)$ $(S_o - S_e)/X_{v,a}t$ (day^{-1})	*(10)* $(10) = (2) \div (8)$ $F/M = S_o/X_{v,a}t$ $\left[\frac{\text{mg BOD}_5}{(\text{mg MLVSS})(\text{day})}\right]$	*(11)* SVI
1	880	100	3100	40.00	0.175	780	543	1.436	1.620	75.0
2	870	50	2800	14.90	0.470	820	1316	0.623	0.661	61.0
3	870	30	3000	8.75	0.800	840	2400	0.350	0.363	64.0
4	860	20	2900	3.50	2.000	840	5800	0.145	0.148	69.0

[a] Reactor volume is 7 liters.

centration of NVSS is approximately 20 mg/liter. The difference, 30 − 20 = 10 mg/liter, corresponds to NVSS settled in the secondary clarifier. Sludge underflow from the secondary clarifier consists of this NVSS settled *plus* net yield of VSS from reactor operation.

From data in Table 5.4 determine design parameters k, a, a', b, and b'. Also estimate nonbiodegradable matter concentration S_n (mg/liter). From Table 5.5 determine k and S_n. Also plot column (*9*) of the table vs. column (*3*). A graph of this plot is shown in Fig. 5.14.

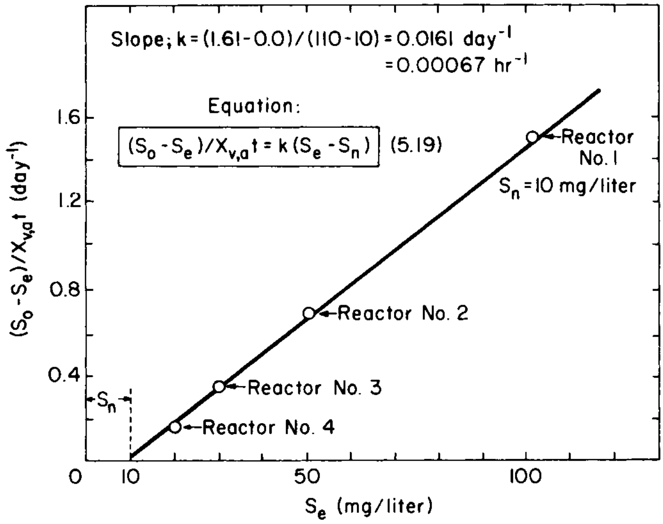


Fig. 5.14. Graphical determination of k and S_n (Example 5.5).

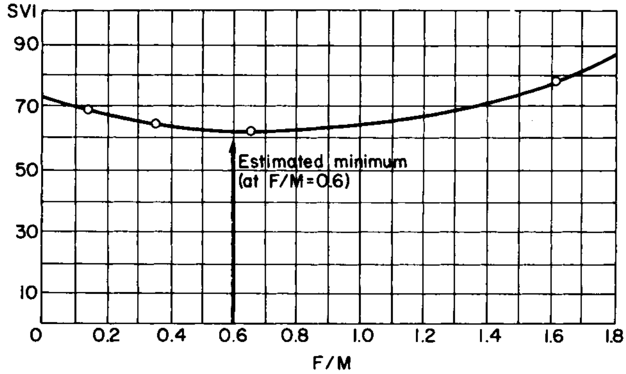


Fig. 5.15. Plot of SVI vs. F/M (Example 5.5).

Step 1. Determine the removal rate constant k (kinetics of BOD removal) [Eq. (5.19)].

$$k = 0.0161\ \text{day}^{-1} \qquad (0.00067\ \text{hr}^{-1})$$

$$S_n = 10\ \text{mg/liter}$$

A plot of SVI vs. F/M ratio is shown in Fig. 5.15. Estimated minimum SVI occurs at a value of F/M ratio ≈ 0.6.

Step 2. Determine oxygen utilization parameters a' and b' [Eq. (5.86)]. From Table 5.6 determine a' and b'.

Plot column (*4*) vs. column (*5*) (Table 5.6). The graph is shown in Fig. 5.16. Then

$$a' = 0.79\ \text{mg O}_2/\text{mg BOD}_r = 0.79\ \text{lb O}_2/\text{lb BOD}_r$$

$$b' = 0.15\ \text{day}^{-1}$$

TABLE 5.6
Oxygen Utilization Parameters

	Laboratory data		Calculated data	
(*1*) Reactor no.	(*2*) $X_{v,a}$ (mg/liter) (Table 5.4)	(*3*) R_r [mg O$_2$/(liter)(day)] (Table 5.4)	(*4*) (*4*) = (*3*) ÷ (*2*) $R_r/X_{v,a}$ (day^{-1})	(*5*) $(S_o - S_e)/X_{v,a}t$ (day^{-1}) (Table 5.5)
1	3100	4025	1.298	1.440
2	2800	1800	0.643	0.620
3	3000	1292	0.431	0.350
4	2900	780	0.269	0.145

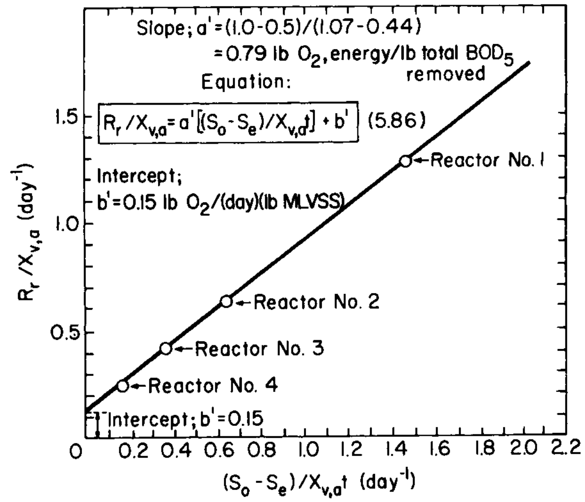


Fig. 5.16. ***Graphical determination of a′ and b′ (Example 5.5).***

TABLE 5.7
Sludge Yield Parameters

	Lab. data	Calculated data			Lab. data	Calculated data	
(*1*) Reactor no.	(*2*) $\Delta X_t/V$ [mg SS/(liter)(day)] (total SS)	(*3*) residence time t (days) (Table 5.5)	(*4*) $(X_{NV,o}-X_{NV,e})/t$ $=10/t$ [mg SS/(liter)(day)]	(*5*) $(\Delta X_t/V)-(X_{NV,o}-X_{NV,e})/t$ $=(2)-(4)$ [mg MLVSS/(liter)(day)]	(*6*) $X_{v,a}$ (mg/liter) (Table 5.4)	(*7*) $\dfrac{(\Delta X_t/V)-(X_{NV,o}-X_{NV,e})/t}{X_{v,a}}$ [mg MLVSS/(day)(mg MLVSS)]	(*8*) $(S_o-S_e)/X_{v,a}t$ (day^{-1}) (Table 5.5)
1	2387	0.175	57.0	2330.0	3100	0.75	1.440
2	806	0.470	21.3	784.7	2800	0.28	0.620
3	387	0.800	12.5	374.5	3000	0.125	0.350
4	295	2.000	5.0	290.0	2900	0.01	0.145

Step 3. Determine sludge yield parameters a and b [Eq. (5.89)].

$$(X_{NV,o}-X_{NV,e} = 30-20 = 10 \text{ mg SS/liter}) \qquad X_{v,o} = 0$$

From Table 5.7 determine a and b. Plot column (*7*) vs. column (*8*) (Table 5.7). The graph is shown in Fig. 5.17.

$$a = 0.575 \text{ lb MLVSS/lb total BOD}_5 \text{ removed}$$

$$b = 0.075 \text{ lb MLVSS/(day)(lb MLVSS)}$$

SUMMARY Design parameters (Example 5.5)

$k = 0.0161 \text{ day}^{-1} (0.00067 \text{ hr}^{-1})$

$S_n = 10$ mg/liter

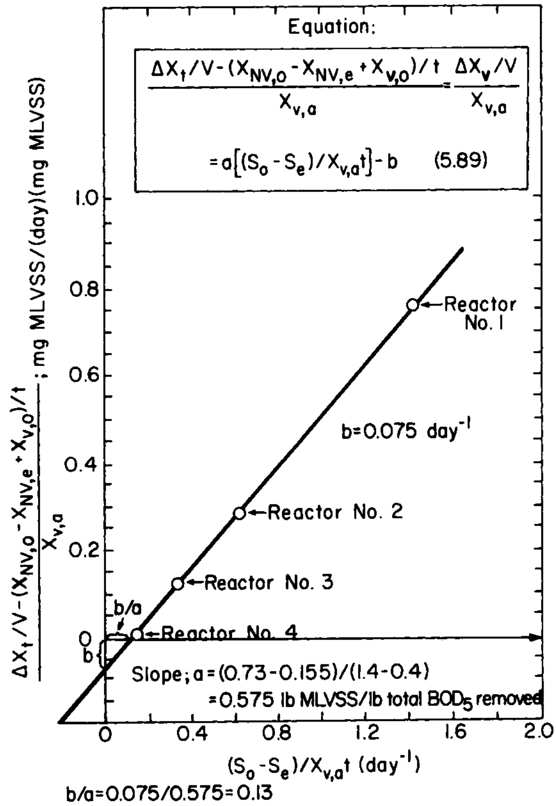


Fig. 5.17. Graphical determination of a and b (Example 5.5).

$a = 0.575$ lb MLVSS/lb total BOD_5 removed

$a' = 0.79$ lb O_2/lb total BOD_5 removed

$b = 0.075$ lb MLVSS/(day)(lb MLVSS)

$b' = 0.15$ lb O_2/(day)(lb MLVSS)

Example 5.6

If for a wastewater the lb O_2/day required for aerobic biological treatment is

$$\text{lb } O_2/\text{day} = 0.4(\text{lb BOD}_5 \text{ removed/day}) + 0.1(\text{lb MLVSS})$$

write an approximate equation for biological sludge yield in lb/day.

SOLUTION Here

$$a' = 0.4 \qquad (\text{basis BOD}_5\text{, i.e., } a'_{\text{BOD}_5})$$

$$b' = 0.1$$

The desired equation from Eq. (5.68) is

$$\Delta X_v(\text{lb MLVSS/day}) = a(\text{lb BOD}_5 \text{ removed/day}) - b(\text{lb MLVSS})$$

Approximate relationships for a [Section 4.1.6(c)] and b [from Eq. (5.60)] as functions of a' and b'

$$0.82a_{\text{BOD}_5} + 0.58a'_{\text{BOD}_5} = 1.0$$

$$b = b'/1.42$$

Therefore, the approximate value of a_{BOD_5} is

$$a_{\text{BOD}_5} = (1 - 0.58a'_{\text{BOD}_5})/0.82 = [1-(0.58)(0.4)]/0.82 = 0.94$$

The approximate value of b is

$$b = b'/1.42 = 0.1/1.42 = 0.07$$

The approximate equation for MLVSS yield is

$$\Delta X_v = 0.94(\text{lb BOD}_5 \text{ removed/day}) - 0.07(\text{lb MLVSS})$$

7. Design Procedure for an Activated Sludge Plant

7.1. INTRODUCTION

From knowledge of design parameters k, a, b, a', and b', design of the activated sludge plant is undertaken. For the laboratory reactor in Fig. 5.10 there is no recycle of sludge. Net sludge yield is withdrawn intermittently to maintain an average constant concentration ($X_{v,a}$) of MLVSS in the aeration chamber. For the actual plant, sludge is recycled as shown in Fig. 5.1.

A primary variable selected by the designer is concentration $X_{v,a}$ of MLVSS in the aerator. Rate of recycle sludge Q_R is calculated to provide this concentration. Usually $X_{v,a}$ is selected between 2000 and 4000 mg/liter of MLVSS. Another primary variable which is selected by the designer is the concentration $X_{v,u}$ of MLVSS in recycle sludge (stream 7 in Fig. 5.1), which is also equal to MLVSS concentration in underflow from the secondary clarifier [stream 5]. Concentration $X_{v,u}$ is also the same as that in stream 6 (wastage). Good settling sludge is expected to attain a concentration $X_{v,u}$ between 10,000 and 15,000 mg/liter of MLVSS.

At steady state conditions there is no accumulation of sludge. Thus, net yield of sludge in the aerator must be removed in wastage stream 6. For purposes of material balance calculations wastage is assumed to be continuous. In practice, it is usually performed intermittently by the arrangement shown in Fig. 5.18, since it is ordinarily too small to justify continuous withdrawal.

Return and wastage lines are valved as indicated. Valves are actuated by a time clock for intermittent sludge wastage (e.g., 5 min every hour).

7.2. MATERIAL BALANCE FOR DETERMINATION OF RECYCLE RATIO OF MLVSS

Write a material balance for MLVSS around the secondary clarifier in Fig. 5.1 (loop · · ·).

MLVSS, IN	MLVSS, OUT
1. MLVSS in reactor effluent [stream 3] $QX_{v,a}(8.34)$ (lb/day) or [from Eq. (5.5)] $Q_F(1+r)X_{v,a}(8.34)$ (lb/day)	1. MLVSS in net effluent [stream 4] zero (assuming complete sedimentation of MLVSS in secondary clarifier)
	2. MLVSS in wastage [stream 6] $\Delta X_v + Q_F X_{V,F}$ (8.34) (lb/day)
	3. MLVSS in recycled sludge [stream 7] $Q_R X_{v,u}(8.34) = rQ_F X_{v,u}(8.34)$ (lb/day)

Then

$$Q_F(1+r)X_{v,a}(8.34) = 0 + \Delta X_v + Q_F X_{V,F}(8.34) + rQ_F X_{v,u}(8.34)$$

Solving for the recycle ratio,

$$r = (8.34Q_F X_{v,a} - \Delta X_v - 8.34Q_F X_{V,F})/[8.34Q_F(X_{v,u} - X_{v,a})] \tag{5.91}$$

If net sludge yield (ΔX_v) and MLVSS concentration in fresh feed ($X_{V,F}$) are negligible by comparison with term $8.34Q_F X_{v,a}$, Eq. (5.91) simplifies to yield Eq. (5.92).

$$r = X_{v,a}/(X_{v,u} - X_{v,a}) \tag{5.92}$$

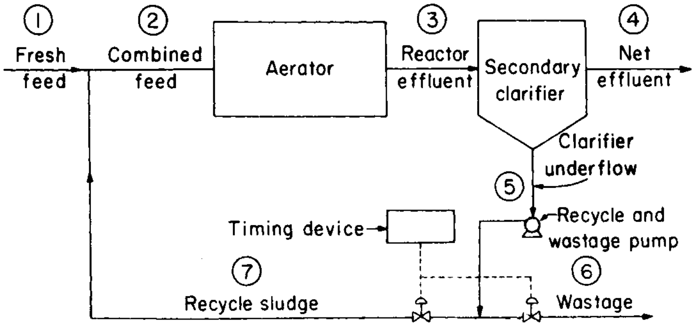


Fig. 5.18. Arrangement for sludge wastage.

Wastage flow Q'' is calculated by noting that it must contain the ΔX_v lb/day of net yield of MLVSS *plus* the MLVSS from fresh feed $(Q_F X_{V,F})$. Therefore, since concentration of MLVSS in stream Q'' is $X_{v,u}$,

$$\Delta X_v + Q_F X_{V,F}(8.34) = Q'' X_{v,u}(8.34) \tag{5.93}$$

$$Q'' = (\Delta X_v + 8.34 Q_F X_{V,F})/8.34 X_{v,u} \tag{5.94}$$

Q' is obtained by combining Eqs. (5.2) and (5.94):

$$Q' = Q_F - Q'' = Q_F - (\Delta X_v + 8.34 Q X_{V,F})/8.34 X_{v,u} \tag{5.95}$$

7.3. MATERIAL BALANCE FOR CALCULATION OF S_o

BOD of combined feed (S_o) is calculated by a BOD balance around the junction of fresh feed and recycle sludge to form combined feed, i.e., loop —·—·— in Fig. 5.1.

This material balance is as follows:

BOD IN: $Q_F S_F + Q_R S_e$ or $Q_F S_F + r Q_F S_e$

BOD OUT: QS_o or $Q_F(1+r)S_o$

Then

$$Q_F S_F + r Q_F S_e = Q_F(1+r)S_o$$

Therefore

$$S_o = (S_F + rS_e)/(1+r) \tag{5.96}$$

From Eq. (5.96) the difference $(S_o - S_e)$ between influent and effluent soluble BOD for the aerator is

$$S_o - S_e = [(S_F + rS_e)/(1+r)] - S_e$$

or

$$S_o - S_e = (S_F - S_e)/(1+r) \tag{5.97}$$

7.4. ALTERNATIVE EXPRESSIONS FOR NET YIELD OF BIOLOGICAL SLUDGE AND OXYGEN UTILIZATION IN THE AERATOR

1. *Net yield of MLVSS.* Substitution of Q and $(S_o - S_e)$ in Eq. (5.68) by their values given by Eqs. (5.5) and (5.97) yields after simplification

$$\Delta X_v = a(S_F - S_e)Q_F - bX_{v,a}V \tag{5.98}$$

Equation (5.98) is an alternative expression for ΔX_v. It is more convenient than Eq. (5.68), since it contains primary variables S_F and Q_F rather than S_o and Q. [S_o and Q are calculated from knowledge of S_F, Q_F, S_e, and r from Eqs. (5.96) and (5.5).]

The physical significance of the synthesis term $a(S_F - S_e)Q_F$ is clear. Combined feed Q (Fig. 5.13) is thought of as two hypothetical separate streams (Fig. 5.19). For stream Q_F soluble BOD is reduced from S_F to S_e, and biological sludge synthesized as a result of this BOD reduction is $a(S_F - S_e)Q_F$. The other stream (Q_R) enters and leaves the reactor with the same unchanged concentration of soluble BOD, i.e., S_e. Therefore it *does not* contribute to synthesis of biological sludge.

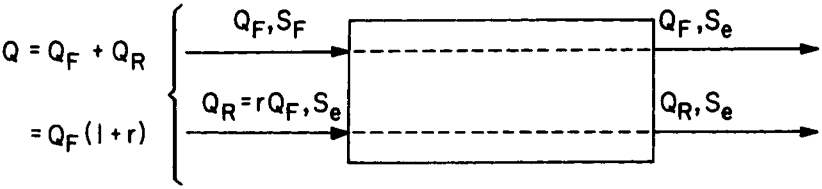


Fig. 5.19. Diagram corresponding to Eq. (5.98).

2. *Oxygen utilization in the aerator.* On substitution of $(S_o - S_e)$ and Q by their values given by Eqs. (5.97) and (5.5), respectively, Eq. (5.64) yields

$$\text{lb O}_2\text{/day} = a'(S_F - S_e)Q_F + b'X_{v,a}V \tag{5.99}$$

Significance of energy term $a'(S_F - S_e)Q_F$ is parallel to that of the synthesis term in Eq. (5.98). Only stream Q_F consumes oxygen since stream Q_R enters and leaves the reactor unchanged. Equation (5.99) is more convenient than Eq. (5.64), since it contains primary variables S_F and Q_F rather than S_o and Q.

7.5. CALCULATION OF RESIDENCE TIME IN REACTOR

Residence time in the reactor is calculated from two criteria in order to determine which one controls the design. These two criteria are

1. Effluent quality, which meets regulatory authority specifications.

Effluent quality depends on substrate removal rate given by Eq. (5.19), which solved for t yields

$$t = (S_o - S_e)/[kX_{v,a}(S_e - S_n)] \tag{5.100}$$

2. Organic loading, evaluated from F/M ratio for optimum flocculation and settling of sludge. This is given by Eq. (5.80), which solved for t yields

$$t = S_o/[X_{v,a}(F/M)] \tag{5.101}$$

Required residence time is calculated from Eqs. (5.100) and (5.101), the larger of the two values of t thus obtained being adopted for design. For wastes which are easily degradable (e.g., sugar refinery, dairy, brewery), the optimum flocculation condition is controlling for residence time calculations. For other wastes, e.g., in petroleum refineries and petrochemical plants, the effluent quality criterion controls residence time requirements since biological degradation is very slow.

7.6. EQUATIONS FOR SLUDGE RECYCLE RATIO r IN CASES WHEN EFFLUENT QUALITY AND ORGANIC LOADING CONTROL RESIDENCE TIME

Consider Eq. (5.91) for the sludge recycle ratio. ΔX_v is given by Eq. (5.68), which is rewritten including the factor 8.34 for use with Q in MGD; V in MG; and S_o, S_e, and $X_{v,a}$ in mg/liter as

$$\Delta X_v = 8.34a(S_o - S_e)Q - 8.34bX_{v,a}V \tag{5.102}$$

Utilizing Eq. (5.5),

$$\Delta X_v = 8.34a(S_o - S_e)Q_F(1+r) - 8.34bX_{v,a}V \tag{5.103}$$

Since reactor volume V is a value calculated by the designer, it is desirable to rewrite Eq. (5.103) as a function of residence time t, which is given by either Eq. (5.100) or (5.101). Term V in Eq. (5.103) is obtained by combining Eqs. (5.17) and (5.5):

$$V = Qt = Q_F(1+r)t \tag{5.104}$$

Substituting in Eq. (5.103) V by its value given by Eq. (5.104),

$$\Delta X_v = 8.34a(S_o - S_e)Q_F(1+r) - 8.34bX_{v,a}Q_F(1+r)t \tag{5.105}$$

Substitution of ΔX_v given by Eq. (5.105) in Eq. (5.91) yields after simplification

$$r = [X_{v,a} - a(S_o - S_e)(1+r) + bX_{v,a}(1+r)t - X_{V,F}]/(X_{v,u} - X_{v,a}) \tag{5.106}$$

Residence time t on the numerator of Eq. (5.106) is given by either Eq. (5.100) or (5.101), depending on whether residence time is governed by substrate removal rate or optimum flocculation conditions. Equations (5.100) and (5.101) are written in a generalized form as

$$t = (S_o - \alpha)/X_{v,a}\beta \tag{5.107}$$

Where effluent quality controls design (Case 1),

$$\alpha = S_e \tag{5.108}$$

$$\beta = k(S_e - S_n) \tag{5.109}$$

Where optimum flocculation conditions control design (Case 2),

$$\alpha = 0 \tag{5.110}$$

$$\beta = F/M \tag{5.111}$$

Substituting residence time t in the numerator of Eq. (5.106) by its value from Eq. (5.107) yields

$$r = \left[X_{v,a} - a(S_o - S_e)(1+r) + b\frac{S_o - \alpha}{\beta}(1+r) - X_{V,F}\right]\Big/(X_{v,u} - X_{v,a}) \tag{5.112}$$

Since S_o is *not* a primary variable it is desirable to eliminate it from the numerator of Eq. (5.112). Substituting $(S_o - S_e)$ by its value given in Eq. (5.97), and the value of S_o in term $(S_o - \alpha)$ by its value given in Eq. (5.96),

$$r = \left[X_{v,a} - a(S_F - S_e) + b\frac{S_F + rS_e - \alpha(1+r)}{\beta} - X_{V,F}\right]\Big/(X_{v,u} - X_{v,a}) \tag{5.113}$$

Now write Eq. (5.113) specifically for Cases (1) and (2).

For Case (1). Substituting in Eq. (5.113) α and β given by Eqs. (5.108) and (5.109) and simplifying,

$$r = \left[X_{v,a} - a(S_F - S_e) + b\frac{S_F - S_e}{k(S_e - S_n)} - X_{V,F}\right]\Big/(X_{v,u} - X_{v,a}) \tag{5.114}$$

Recycle ratio, Case (1): Effluent quality controls design

For Case (2). Substituting in Eq. (5.113) α and β given by Eqs. (5.110) and (5.111), solving the resulting expression for r, and simplifying [Eq. (5.115)],

$$r = \frac{[X_{v,a} - a(S_F - S_e)](F/M) + bS_F - X_{V,F}(F/M)}{(X_{v,u} - X_{v,a})(F/M) - bS_e} \tag{5.115}$$

Recycle ratio, Case (2): Optimum flocculation conditions control design

7.7. NEUTRALIZATION REQUIREMENTS

Optimum activity for bacteria occurs at pH values of 6–8. It should be checked if neutralization is needed preceding biological treatment. For alkaline wastes, it is taken as a rule of thumb that up to 0.5 lb of alkalinity (as

$CaCO_3$) is removed per lb of BOD removed. This happens because the CO_2 evolved from bacterial waste degradation reacts with alkalinity (OH^-) present in the waste to form bicarbonate ($HCO_3{}^-$), which buffers the system at a pH of about 8. Thus, neutralization preceding biological treatment may not be required for some alkaline wastewaters.

7.8. NUTRIENT REQUIREMENTS

The appropriate amount of certain nutrients is required for both synthesis and respiration phases of aerobic biological degradation of wastes. Required nutrients include nitrogen, phosphorus, calcium, magnesium, and vitamins. Most of these nutrients, which are required only in trace quantities, are usually present in wastewaters. However, many industrial wastewaters are deficient in nitrogen and phosphorus. Required amounts of nitrogen and phosphorus are estimated by the procedure discussed in this section. If deficiency exists, it is corrected by adding to the wastewater calculated weights of compounds containing nitrogen and phosphorus.

An estimate of requirements for nitrogen and phosphorus is based on the fact that wasted MLVSS (ΔX_v lb/day) contains approximately 2% of its dry weight as phosphorus and 12% as nitrogen. An estimate of weights of nitrogen and phosphorus to be added comprises

1. Weights of these nutrients which are lost by wastage of MLVSS, i.e.,

Nitrogen: $0.12\ \Delta X_v$ lb/day

Phosphorus: $0.02\ \Delta X_v$ lb/day

2. Weights of these nutrients which are lost in the effluent. (Total effluent $= Q' + Q'' = Q_F$.) Concentrations of soluble nitrogen and phosphorus present in effluent are usually estimated to be 1.0 and 0.5 mg/liter, respectively. Thus, the amounts of nitrogen and phosphorus lost in the effluent are

Nitrogen: $Q_F \times 1.0 \times 8.34$ lb/day

Phosphorus: $Q_F \times 0.5 \times 8.34$ lb/day

where Q_F is the effluent in MGD. Therefore, the total requirements of nitrogen and phosphorus are given by the sum of the estimates made under (1) and (2) [Eqs. (5.116) and (5.117)]:

$$\text{Nitrogen: } 0.12\ \Delta X_v + Q_F \times 1.0 \times 8.34 \text{ lb/day} \tag{5.116}$$

$$\text{Phosphorus: } 0.02\ \Delta X_v + Q_F \times 0.5 \times 8.34 \text{ lb/day} \tag{5.117}$$

In activated sludge plants, nitrogen and phosphorus requirements are provided by the addition of anhydrous or aqueous NH_3, H_3PO_4, or $(NH_4)_3PO_4$.

7.9. DESIGN PROCEDURE FOR ACTIVATED SLUDGE PLANTS

Step 1. Calculate the recycle ratio of MLVSS. Select values for $X_{v,a}$ and $X_{v,u}$, usually within the ranges of 2000–4000 and 10,000–15,000 mg/liter, respectively.

From sedimentation and SVI data (Fig. 5.7) select an appropriate value for the F/M ratio. Optimum F/M is usually in the range 0.3–0.7.

The recycle ratio is calculated (1) from Eq. (5.114), which assumes that effluent quality controls the design, and (2) from Eq. (5.115), which assumes that optimum flocculation conditions control the design. The decision on which condition controls design is arrived at in Step 3.

Step 2. Calculate BOD of the combined feed (S_o). S_o is calculated from Eq. (5.96) utilizing *both* values of r calculated in Step 1(1) and Step 1(2). These two parallel calculations are referred to hence as Steps 2(1) and 2(2), respectively.

Step 3. Calculate residence time in the reactor.

Case (*1*) Assuming substrate removal rate controls design [Eq. (5.100)], where S_o is the value calculated in Step 2(1).

Case (*2*) Assuming optimum flocculation conditions control design [Eq. (5.101)], where S_o is the value calculated in Step 2(2).

Possibly* the larger of these two calculated residence times is the one selected for design. Recycle ratio and BOD of combined feed for the specific case which controls design are then adopted. Calculated values for the other case are discarded.

Step 4. Calculate the reactor volume. Reactor volume is then calculated from Eq. (5.104) utilizing the value of the residence time selected in Step 3. At this stage, depth of tank is selected. Selection depends on type of aerator utilized (Chapter 4, Sections 14.4, 15.4, and 16.3). Cross-sectional area is then calculated.

Step 5. Calculate the net yield of MLVSS. Net yield of MLVSS ΔX_v is calculated from Eq. (5.105) or Eq. (5.98).

Note: A check on the material balance for MLVSS is made at this point. (See Example 5.7 for details.)

Step 6. Calculate Q'' and Q'. Q'' and Q' are then calculated from Eqs. (5.94) and (5.95).

Step 7. Calculate ΔX_{NV} and ΔX_t. ΔX_{NV} is calculated from Eq. (5.72) and total sludge yield is obtained from Eq. (5.1), where ΔX_v and ΔX_{NV} are the values calculated in Steps 5 and 7, respectively.

---

* The reason for the word *possibly* is that frequently a compromise is made in selection of residence time, so that not only reasonable BOD reduction is achieved (i.e., a low value of S_e), but also good flocculation conditions for the sludge (although not necessarily the optimum) are obtained.

Note: A check on the overall material balance for NVSS is made at this point. (See Example 5.7 for details.)

Step 8. Calculate oxygen requirements (lb/day) from Eq. (5.64) or Eq. (5.99).

Step 9. Specify aeration equipment. Aeration equipment is selected from oxygen requirements determined in Step 8 and manufacturer's specifications for aerators. This procedure is described in Chapter 4, Sections 14, 15, and 16. Calculated values needed are (1) total HP and number of aeration units; (2) power level, HP/1000 gal; and (3) spacing between aerators. A layout for aerators in the tank is selected. (Details are given in Example 4.5.)

Step 10. Check neutralization requirements. Verify if neutralization is required prior to biological treatment. For alkaline wastes, the rule of thumb stating that up to 0.5 lb of alkalinity (as $CaCO_3$) is removed per lb of BOD removed is frequently employed (Section 7.7).

Step 11. Evaluate nutrient requirements. Requirements for nitrogen and phosphorus (lb/day) are evaluated as described in Section 7.8 from Eqs. (5.116) and (5.117).

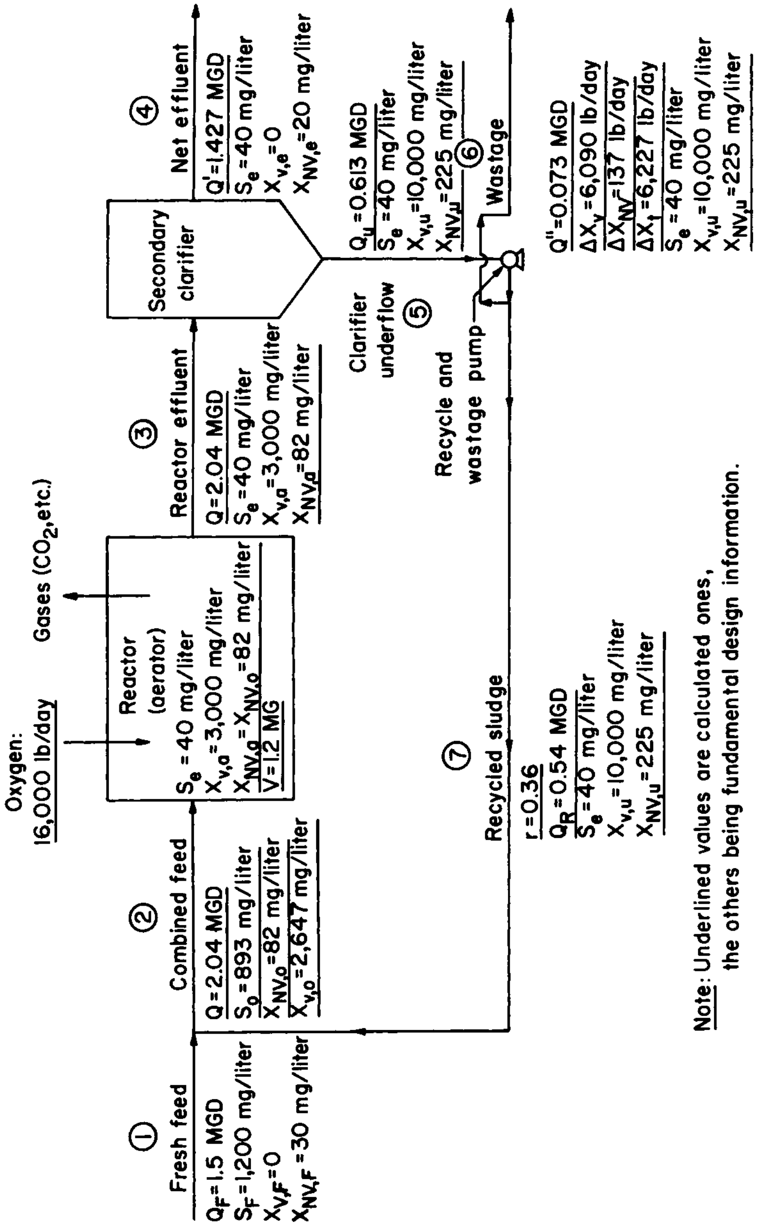
7.10. NUMERICAL EXAMPLE: DESIGN OF AN ACTIVATED SLUDGE PLANT

Example 5.7

An industrial plant (Example 5.5) considers an activated sludge system for treatment of their wastewaters. Base design on the following data in addition to information given in Example 5.5:

Flow: 1.5 MGD (1,500,000 gal/day)
Influent BOD_5: 1200 mg/liter
Effluent BOD_5: 40 mg/liter
OH^- alkalinity of raw wastewater: 90 mg/liter (as $CaCO_3$)
Total Kjeldahl nitrogen and phosphorus in fresh feed: 85 and 3 mg/liter, respectively
C_{SW}: Saturation DO of wastewater at the temperature and barometric pressure of the test
C_{SW} (summer conditions at 30°C): 7.4 mg/liter
C_{SW} (winter conditions at 18°C): 10.3 mg/liter
$\alpha = K_La(\text{wastewater})/[K_La(\text{water})] = 0.72$
Assume operating DO (level in aeration basin $\approx$ 1.0 mg/liter)
Characteristics of surface aerators: given by Fig. 4.17
Select for design purposes:
$X_{v,a}$ = 3000 mg/liter of MLVSS
$X_{v,u}$ = 10,000 mg/liter of MLVSS
Neglect VSS concentration in effluent from secondary clarifier and in fresh feed

SOLUTION A flowsheet for the proposed activated sludge plant is shown in Fig. 5.20. Follow design procedure summarized in Section 7.9.



Note: Underlined values are calculated ones, the others being fundamental design information.

Fig. 5.20. *Flow diagram for Example 5.7.*
]

Step 1. Calculate the recycle ratio of MLVSS.

(1) Assume substrate removal rate controls design [Eq. (5.114)]. Here

$X_{v,a} = 3000$ mg/liter

$a(S_F - S_e) = 0.575(1200-40) = 0.575 \times 1160 = 665$ mg/liter

$X_{v,u} - X_{v,a} = 10{,}000 - 3000 = 7000$ mg/liter

$X_{v,a} - a(S_F - S_e) = 3000 - 665 = 2335$ mg/liter

$X_{V,F} = 0$

Thus Eq. (5.114) yields

$$r = [2335 + b(S_F - S_e)/k(S_e - S_n)]/7000$$

Calculate separately term $b(S_F - S_e)/k(S_e - S_n)$, which is the only different term in equations for r in Step 1(1) and Step 1(2).

$$b(S_F - S_e)/k(S_e - S_n) = 0.075(1200-40)/0.0161(40-10) = 180$$

$$\therefore\ r = (2335+180)/7000 = 0.36$$

(2) Assume optimum flocculation conditions control design [Eq. (5.115)]. Here

$[X_{v,a} - a(S_F - S_e)] = 2335$ [calculated Step 1(1)]

$(X_{v,u} - X_{v,a}) = 7000$ [also calculated Step 1(1)]

$X_{V,F}(F/M) = 0$ since $X_{V,F} = 0$

and

$$(F/M) = 0.6$$

$$b = 0.075$$

$$S_e = 40$$

Thus Eq. (5.115) yields

$$r = (2335 \times 0.6 + 0.075 \times 1200)/(7000 \times 0.6 - 0.075 \times 40) = 0.353$$

Step 2. Calculate S_o from Eq. (5.96).

(1) Assume substrate removal rate controls design (i.e., $r = 0.36$).

$$S_o = (1200 + 0.36 \times 40)/(1 + 0.36) = 893 \text{ mg/liter}$$

(2) Assume optimum flocculation conditions control design (i.e., $r = 0.353$).

$$S_o = (1200 + 0.353 \times 40)/(1 + 0.353) = 900 \text{ mg/liter}$$

Step 3. Calculate residence time.

(1) Assume substrate removal rate controls design ($S_o = 893$ mg/liter). From Eq. (5.100)

$$t = (893-40)/[0.0161 \times 3000(40-10)] = 0.59 \text{ day}$$

or

$$24 \times 0.59 = 14.2 \text{ hr}$$

(2) Assume optimum flocculation conditions control design ($S_o =$ 900 mg/liter).
From Eq. (5.101)

$$t = 900/3000 \times 0.6 = 0.5 \text{ day}$$

or

$$24 \times 0.5 = 12 \text{ hr}$$

Thus substrate removal rate controls design, and calculations in part (2) for Steps 1–3 are discarded. F/M ratio for the reactor is [from Eq. (5.80)]

$$F/M = 893/3000 \times 0.59 = 0.504$$

From Fig. 5.15, this value of F/M is compatible with good flocculation conditions for the sludge. Therefore, no adjustment of selected residence time (14.2 hr) is required to achieve compatibility between BOD reduction and good flocculation conditions.

Step 4. Calculate reactor volume. Throughput rate [Eq. (5.5)]:

$$Q = 1.5(1+0.36) = 2.04 \text{ MGD}$$

Reactor volume [Eq. (5.104)]:

$$V = 2.04 \times 0.59 = 1.2 \text{ MG}$$

or

$$1{,}200{,}000 \text{ gal} \times \text{ft}^3/7.48 \text{ gal} = 161{,}000 \text{ ft}^3$$

For depths of 10 and 15 ft, for example, corresponding cross-sectional areas are

$$\text{For } d = 10 \text{ ft} \qquad A = 161{,}000/10 = 16{,}100 \text{ ft}^2$$

or

$$16{,}100 \text{ ft}^2 \times \text{acre}/43{,}500 \text{ ft}^2 = 0.37 \text{ acre}$$

$$\text{For } d = 15 \text{ ft} \qquad A = 161{,}000/15 = 10{,}700 \text{ ft}^2$$

or

$$10{,}700/43{,}500 = 0.246 \text{ acre}$$

Parallel basins might be recommended.

Step 5. Calculate net yield of MLVSS. From Eq. (5.105) (*a*, *b* from Example 5.5),

$$\Delta X_v = 0.575(893-40) \times 2.04 \times 8.34 - 0.075 \times 3000 \times 1.2 \times 8.34$$

$$\Delta X_v = 8342 - 2252 = 6090 \text{ lb/day}$$

Or from Eq. (5.98),

$$\Delta X_v = 0.575(1200-40) \times 1.5 \times 8.34 - 0.075 \times 3000 \times 1.2 \times 8.34$$

$$\Delta X_v = 8342 - 2252 = 6090 \text{ lb/day}$$

Note: At this point a check on material balance calculations is made. Calculate concentration of VSS in combined feed to the reactor ($X_{v,o}$) from Eq. (5.75), where $X_{V,F} = 0$.

$$X_{v,o} = (0+0.36\times 10{,}000)/(1+0.36) = 2647 \text{ mg/liter}$$

The difference between concentrations of MLVSS in reactor effluent (3000 mg/liter) and the value 2647 mg/liter in reactor influent must correspond to the net yield of MLVSS (i.e., $\Delta X_v = 6090$ lb/day). Therefore, $3000-2647 = 353$ mg/liter, i.e., 353 mg of MLVSS are produced per liter of liquor flowing through the reactor. Then based on flow $Q = 2.04$ MGD, net production of MLVSS is $353\times 2.04\times 8.34 \approx 6006$ lb/day, which agrees approximately with the value 6090 lb/day of ΔX_v calculated in Step 5 (within 1.4%).

Step 6. Calculate Q'' and Q' [Eqs. (5.94) and (5.95), respectively].

$$Q'' = 6090/8.34 \times 10{,}000 = 0.073 \text{ MGD} \qquad \text{for } X_{V,F} = 0$$

or

$$Q'' = 73{,}000 \text{ gal/day} \qquad (9.9 \text{ gal/min})$$

and

$$Q' = 1{,}500{,}000 - 73{,}000 = 1{,}427{,}000 \text{ gal/day}$$

Step 7. Calculate ΔX_{NV} and ΔX_t. From Eq. (5.72),

$$\Delta X_{NV} = 1.5(30-20) \times 8.34 + 0.073 \times 20 \times 8.34 = 125 + 12.2 \approx 137 \text{ lb/day}$$

The total sludge yield ΔX_t is [from Eq. (5.1), where $X_{VF} = 0$]

$$\Delta X_t = 6090 + 137 = 6227 \text{ lb/day}$$

Note: Check on material balance for NVSS in the influent.

$$\text{IN} = Q_F X_{NV,F} = 1.5 \times 30 \times 8.34 = 376 \text{ lb/day}$$

$$\text{OUT} = Q''X_{NV,u} = \Delta X_{NV} \qquad = \underline{137 \text{ lb/day}}$$

$$Q'X_{NV,e} = 1.427 \times 20 \times 8.34 \qquad = 239 \text{ lb/day (checks)}$$

$$X_{NV,u} = \Delta X_{NV}/8.34Q'' = 137/8.34 \times 0.073 = 225 \text{ mg/liter}$$

From Eq. (5.74),

$$X_{NV,o} = (30+0.36\times 225)/(1+0.36) = 82 \text{ mg/liter}$$

Step 8. Calculate oxygen requirements from either Eq. (5.64) or Eq. (5.99). From Eq. (5.64) (a', b' from Example 5.5)

$$\text{lb O}_2/\text{day} = 0.79(893-40) \times 2.04 \times 8.34 + 0.15 \times 3000 \times 1.2 \times 8.34$$
$$= 11{,}500 + 4500 = 16{,}000$$

From Eq. (5.99)

$$\text{lb O}_2/\text{day} = 0.79(1200-40) \times 1.5 \times 8.34 + 0.15 \times 3000 \times 1.2 \times 8.34$$
$$= 11{,}500 + 4500 = 16{,}000$$

or

$$16{,}000/24 = 665 \text{ lb O}_2/\text{hr}$$

Step 9. Specify aeration equipment. Specification of aeration equipment and layout for aerators in this application are presented in Chapter 4 (Example 4.5).

Step 10. Check neutralization requirements. Utilize rule of thumb: 0.5 lb of alkalinity (as $CaCO_3$) are removed per lb of BOD removed. Calculate lb of BOD removed per day.

$$(1200-40) \times 1.5 \times 8.34 = 14{,}512 \text{ lb/day}$$

Thus 14,512/2 = 7256 lb/day of alkalinity are removed. Calculate the lb/day of alkalinity in fresh feed.

$$90 \times 1.5 \times 8.34 = 1126 \text{ lb/day}$$

Since $1126 < 7256$, no neutralization is required prior to the biological process.

Step 11. Evaluate nutrient requirements.

Nitrogen

1. Nitrogen lost from system through wastage of sludge:

$$0.12\,\Delta X_v = 0.12 \times 6090 \qquad = 730 \text{ lb/day}$$

2. Nitrogen lost in effluent (1.0 mg/liter):

$$1 \times 1.5 \times 8.34 \qquad \text{(total nitrogen required)} \qquad \approx \frac{13 \text{ lb/day}}{743 \text{ lb/day}}$$

Nitrogen available is (85 mg/liter)

$$85 \times 1.5 \times 8.34 = 1070 \text{ lb/day}$$

Thus addition of nitrogen is *not* required.

Phosphorus

1. Phosphorus lost from system through wastage of sludge:

$$0.02\,\Delta X_v = 0.02 \times 6090 \qquad = 121.8 \text{ lb/day}$$

2. Phosphorus lost in the effluent (≈ 0.5 mg/liter):

$$0.5 \times 1.5 \times 8.34 \qquad \approx \quad \frac{6 \ \text{lb/day}}{128 \ \text{lb/day}}$$

Phosphorus available is (3 mg/liter)

$$3 \times 1.5 \times 8.34 = 37.6 \text{ lb/day}$$

Thus $128 - 37.6 \approx 91.0$ lb/day of phosphorus should be added.

8. The Michaelis–Menten Relationship

8.1. DERIVATION OF MICHAELIS–MENTEN RELATIONSHIP

Formulation of the Michaelis–Menten relationship is based on studies of pure cultures. However, it is used in determining kinetics of substrate degradation by a heterogeneous population of microorganisms, which is the case for the activated sludge process.

Degradation of wastes by microorganisms is accomplished through a complex series of chemical reactions. These reactions are catalyzed by organic catalysts (enzymes) present in the microorganisms. Enzymes are large molecular weight compounds. Usually enzymes are quite specific in their functions as catalysts, i.e., a given enzyme ordinarily catalyzes a specific chemical reaction. Bacteria contains a great variety of enzymes, each one being responsible for a minor step in the complex process of biological metabolism.

The action of enzymes is represented by the following chemical equation:

$$\text{Substrate} + \text{Enzyme} \rightleftharpoons \text{Enzyme–Substrate complex} \rightleftharpoons \text{Enzyme} + \text{Products}$$

or symbolically

$$[\mathrm{S}] + [\mathrm{E}] \underset{k_{-1}}{\overset{k_1}{\rightleftharpoons}} [\mathrm{E\text{–}S}] \underset{k_{-2}}{\overset{k_2}{\rightleftharpoons}} [\mathrm{E}] + [\mathrm{P}] \tag{5.118}$$

where k's stand for the reaction rate constants. As indicated by Eq. (5.118), substrate and enzyme unite to form an enzyme–substrate complex. This is followed by the breaking down of this complex, resulting in the end products. The enzyme remains unchanged and is ready to reenter the reaction, acting therefore as a catalyst.

The rate of substrate removal is obtained from Eq. (5.118) by making the assumption that the breaking down of the enzyme–substrate complex is irreversible. Then Eq. (5.118) is rewritten as

$$[\mathrm{S}] + [\mathrm{E}] \underset{k_{-1}}{\overset{k_1}{\rightleftharpoons}} [\mathrm{E\text{–}S}] \xrightarrow{k_2} [\mathrm{E}] + [\mathrm{P}] \tag{5.119}$$

This assumption is essentially correct if measurements are taken shortly after introduction of substrate, which means that very little product has been allowed to form. Under these circumstances, since the rate of the inverse reaction

$$[\mathrm{E}] + [\mathrm{P}] \xrightarrow{k_{-2}} [\mathrm{E{-}S}]$$

is given by*

$$k_{-2}[\mathrm{E}][\mathrm{P}]$$

and since $[\mathrm{P}] \approx 0$, it may be assumed that the breaking down of the enzyme–substrate complex is irreversible. Therefore Eq. (5.118) is rewritten as shown [Eq. (5.119)]. The rate of reaction measured under these conditions is that occurring immediately on contact of substrate and microorganism, and is referred to as the initial rate of reaction. To develop kinetic data it is necessary to measure a series of such initial rates, corresponding to different concentrations of substrate shortly after the substrate samples are brought into contact with the microorganism.

Substrate removal rate is denoted here by V. For a batch reactor, it corresponds to the slope of the BOD curve in Fig. 5.3 (Section 3.1) at any specified time t, corresponding to a concentration S of substrate. A specific substrate removal rate per mg/liter of MLVSS is utilized, i.e. [Eq. (5.120)],

$$V = -(1/X_{v,a})(dS/dt) \qquad (5.120)$$

(Minus sign is necessary since $dS/dt < 0$ and $V > 0$.)

For the continuous reactor as shown in Section 3.2, this corresponds to [in finite rather than differential form]

$$(S_o - S_e)/X_{v,a}t$$

where t is the residence time in the continuous reactor. The substrate removal rate is equal to the rate of formation of product P, and is given by Eq. (5.121).

$$V = k_2[\mathrm{E{-}S}] \qquad (5.121)$$

Similarly, the rate of formation of the enzyme–substrate complex (E–S) is

$$\text{Rate of formation of (E–S)} = k_1[\mathrm{S}][\mathrm{E}] \qquad (5.122)$$

The rate of conversion of enzyme–substrate complex to E and S is [Eq. (5.123)]

$$\text{Rate of conversion of (E–S)} = k_{-1}[\mathrm{E{-}S}] \qquad (5.123)$$

Therefore, the net change of concentration of enzyme–substrate complex is

$$d[\mathrm{E{-}S}]/dt = \underbrace{k_1[\mathrm{S}][\mathrm{E}]}_{\text{formation}} - \underbrace{k_{-1}[\mathrm{E{-}S}]}_{\text{destruction}} - \underbrace{k_2[\mathrm{E{-}S}]}_{\text{destruction}} \qquad (5.124)$$

---

* In the formulation to follow, symbols [S], [E], [E–S], and [P] are used to denote concentrations of substrate, enzyme, enzyme–substrate complex, and products, respectively.

Let the total concentration of enzyme in the reacting system be denoted as E_t. This includes not only free enzyme (E) but also enzyme in combined form as enzyme–substrate complex (E–S), i.e. [Eqs. (5.125) and (5.126)],

$$[E_t] = [E] + [E\text{–}S] \tag{5.125}$$

$$\therefore\ [E] = [E_t] - [E\text{–}S] \tag{5.126}$$

Substituting [E] in Eq. (5.124) by its value given in Eq. (5.126) yields

$$d[E\text{–}S]/dt = k_1([E_t]-[E\text{–}S])[S] - k_{-1}[E\text{–}S] - k_2[E\text{–}S] \tag{5.127}$$

At steady state conditions it is usually assumed that concentration of intermediate complexes (enzyme–substrate complex in this case) remains unchanged. This assumption is called the steady state approximation. Therefore

$$d[E\text{–}S]/dt = 0 \tag{5.128}$$

and Eq. (5.127) becomes Eq. (5.129).

$$k_1([E_t]-[E\text{–}S])[S] - k_{-1}[E\text{–}S] - k_2[E\text{–}S] = 0 \tag{5.129}$$

Solving for [E–S],

$$[E\text{–}S] = \frac{[E_t][S]}{[S] + (k_{-1}+k_2)/k_1} \tag{5.130}$$

Term $(k_{-1}+k_2)/k_1$ is called the Michaelis–Menten constant and is designated as K_s.

$$K_s = (k_{-1}+k_2)/k_1 \tag{5.131}$$

Then, Eq. (5.130) is rewritten as Eq. (5.132).

$$[E\text{–}S] = \frac{[E_t][S]}{[S]+K_s} \tag{5.132}$$

Substituting this value in Eq. (5.121), the following expression for the substrate removal rate V is obtained:

$$V = k_2\frac{[E_t][S]}{[S]+K_s} \quad \text{(Michaelis–Menten relationship)} \tag{5.133}$$

8.2. COROLLARIES OF MICHAELIS–MENTEN RELATIONSHIP

The two corollaries stated in Section 3.1 are obtained from Eq. (5.133).

Corollary 1: High substrate concentrations

At high substrate concentrations,

$$[S] \gg K_s \tag{5.134}$$

Neglecting K_s in the denominator of Eq. (5.133) as compared to [S] and simplifying,

$$V = k_2[\mathrm{E}_t] = V_{\mathrm{MAX}} \tag{5.135}$$

Equation (5.135) indicates that at high substrate concentrations, removal of substrate takes place at a maximum rate (V_{MAX}) independent of concentration. It is assumed that at these high substrate concentrations all active sites of the enzymes are saturated with substrate, and so reaction proceeds as fast as possible independent of substrate concentration (zero-order reaction). This corresponds to the section of the BOD curve in Fig. 5.3 (Section 3.1) from time zero up to time t', where the tangent to the BOD curve essentially coincides with the curve itself (constant slope).

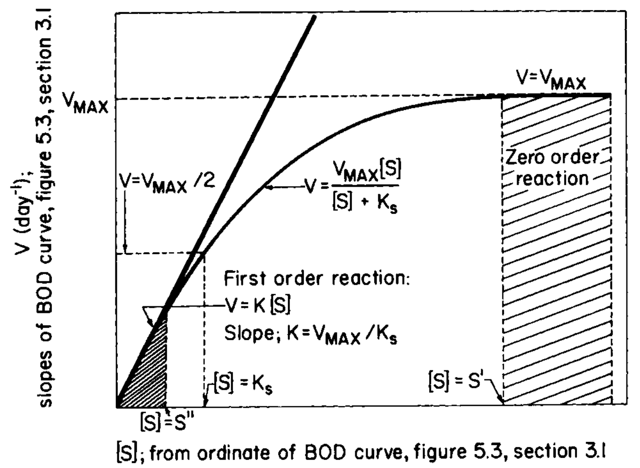


Fig. 5.21. Plot of V vs. **[S]**.

From Fig. 5.3, a plot of slopes of the BOD curve (V's or substrate removal rates) vs. the corresponding BOD values ([S]) can be constructed (Fig. 5.21). The region of high substrate concentration encompasses values from the right-hand extremity of the graph down to a concentration S' (corresponding to time t' in Fig. 5.3). This is the region where $V = V_{\mathrm{MAX}} =$ constant irrespective of substrate concentration.

From Eq. (5.135) Eq. (5.133) is rewritten as

$$V = V_{\mathrm{MAX}} \frac{[\mathrm{S}]}{[\mathrm{S}] + K_s} \tag{5.136}$$

Corollary 2: Low substrate concentrations

At low substrate concentrations,

$$[\mathrm{S}] \ll K_s \tag{5.137}$$

Neglecting [S] in the denominator of Eq. (5.136) as compared to K_s,

$$V = V_{MAX}[S]/K_s \tag{5.138}$$

Since V_{MAX} and K_s are both constant for a specific waste, Eq. (5.138) is rewritten as

$$V = K[S] \tag{5.139}$$

where

$$K = V_{MAX}/K_s \tag{5.140}$$

Equation (5.139) indicates that at low substrate concentrations, substrate removal follows first-order kinetics. In Fig. 5.21 this corresponds to the section of the curve from a value of the abscissa $S = 0$ up to a value S''. In this section, the curve is essentially replaced by a straight line passing through the origin (with slope = K). This situation corresponds to that encountered in continuous biological reactors operating at steady state conditions. In fact, Fig. 5.21 up to $[S] = S''$ is identical to Fig. 5.5 (Section 3.2), which was utilized for determination of the removal rate constant from a series of continuous laboratory reactors operating in parallel. Had these experiments been continued on higher concentrations of substrate, the straight line would have become a curve like the one in Fig. 5.21. However, operation of continuous reactors is always conducted at substrate concentrations much below 500 mg/liter (expressed as BOD_5). Under these conditions the straight line relationship applies.

8.3. SIGNIFICANCE OF MICHAELIS–MENTEN CONSTANT K_s

From Eq. (5.136) it is shown that K_s is equal to the substrate concentration when substrate removal rate V equals *half* the maximum, i.e., when $V = V_{MAX}/2$. This is shown by letting $V = V_{MAX}/2$ in Eq. (5.136) and solving for [S]. The final result is

$$[S] = K_s \qquad (\text{for } V = V_{MAX}/2)$$

This is indicated in Fig. 5.21.

8.4. DETERMINATION OF V_{MAX}: THE LINEWEAVER–BURK PLOT

The value of V_{MAX} estimated from Fig. 5.21 is inaccurate since it is an asymptotic value. A better way of determining V_{MAX} is as follows. Take the inverse of Eq. (5.136),

$$1/V = (K_s/V_{MAX})(1/[S]) + (1/V_{MAX}) \tag{5.141}$$

Based on Eq. (5.141) a linear plot of $1/V$ vs. $1/[S]$ is constructed, from which the characteristic constants V_{MAX} and K_s are determined from the slope and intercept of the straight line. This graph, which is shown in Fig. 5.22, is known as the Lineweaver–Burk plot [5]. As indicated the intercept at the abscissa corresponds to $(-1/K_s)$, since for $1/V = 0$ one obtains $1/[\mathrm{S}] = -1/K_s$ from Eq. (5.141).

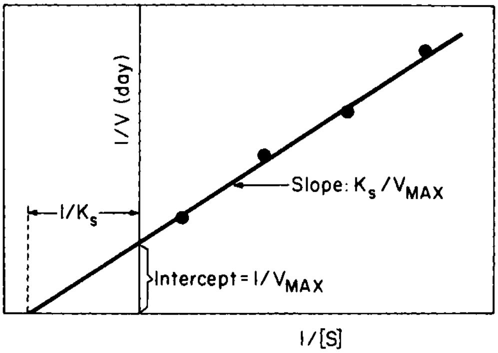


Fig. 5.22. Lineweaver–Burk plot.

8.5. MICHAELIS–MENTEN RELATIONSHIP WHEN NONBIODEGRADABLE MATTER IS PRESENT IN THE SYSTEM

If the concentration of nonbiodegradable matter is indicated as $[S_n]$, it is accounted for by substituting the value of [S] by $([\mathrm{S}]-[S_n])$ in Eq. (5.136). A similar substitution in Section 3.2 led to Eq. (5.19) from Eq. (5.18).

Therefore, modified Eq. (5.136) is

$$V = V_{\mathrm{MAX}} \frac{[\mathrm{S}]-[S_n]}{K_s+[\mathrm{S}]-[S_n]} \tag{5.142}$$

The two corollaries studied in Section 8.2 derived from Eq. (5.133) are also obtained from Eq. (5.142). Similarly, Fig. 5.21 is replotted when nonbiodegradable matter is present (Fig. 5.23).

From Eq. (5.142), it is shown that

$$K_s = [\mathrm{S}] - [S_n]$$

when $V = V_{\mathrm{MAX}}/2$

The Lineweaver–Burk equation when nonbiodegradable matter is present is written by replacing [S] in Eq. (5.141) by $([\mathrm{S}]-[S_n])$. The corresponding Lineweaver–Burk plot follows directly from the equation thus obtained.

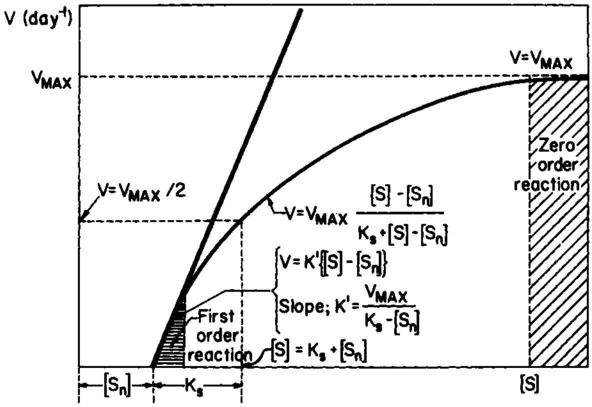


Fig. 5.23. Plot of V vs. [S] when nonbiodegradable matter is present.

8.6. MICHAELIS–MENTEN RELATIONSHIP IN TERMS OF SPECIFIC GROWTH RATE OF SLUDGE

The Michaelis–Menten relationship [Eq. (5.136)] is written as a function of the specific substrate removal rate [Eq. (5.120)]. An equivalent form is written as a function of the specific growth rate of sludge, defined as

$$\mu = (1/X_{v,a})(dX/dt) \tag{5.143}$$

Whereas values of dS/dt in Eq. (5.120) correspond to slopes of the BOD curve in Fig. 5.3, values of dX/dt in Eq. (5.143) correspond to slopes of the MLVSS curve.

Assuming that the specific growth rate of sludge is proportional to the specific substrate removal rate, i.e., that a constant fraction of the substrate removed is converted to cells ($\mu = aV$), Eq. (5.136) is rewritten as

$$\mu = \mu_{\mathrm{MAX}} \frac{[\mathrm{S}]}{[\mathrm{S}] + K_s} \tag{5.144}$$

From Eq. (5.144) it is shown, following the same procedure utilized in Section 8.3 for Eq. (5.136), that K_s is equal to the substrate concentration when the specific growth rate of the sludge is equal to half the maximum specific growth rate, i.e., $K_s = [\mathrm{S}]$, when $\mu = \mu_{\mathrm{MAX}}/2$. All corollaries, derivations, and graphical constructions studied in Sections 8.2 to 8.5 based on the specific substrate removal rate are also applicable in terms of the specific growth rate of the sludge.

9. The Concept of Sludge Age

Sludge age is defined as the mean residence time of MLVSS in the reactor. For the activated sludge plant shown in Fig. 5.1 this corresponds to

$$t_S = \text{lb MLVSS in reactor/net output of VSS from the system (lb/day)} \tag{5.145}$$

or

$$t_S = \frac{X_{v,a}V}{\begin{matrix}\text{total lb/day of VSS wasted, i.e.,} \\ \text{output of VSS} \quad \text{(lb/day)}\end{matrix} - \begin{matrix}\text{input of VSS in fresh} \\ \text{feed} \quad \text{(lb/day)}\end{matrix}} \tag{5.146}$$

or

$$t_S = \frac{X_{v,a}V}{(\Delta X_v + Q'X_{v,e}) - Q_F X_{V,F}} \quad \text{(days)} \tag{5.147}$$

In Eq. (5.147) numerator ($X_{v,a}V$) equals total lb of MLVSS in the reactor at any time (a constant value at steady state conditions). Terms between parentheses in the denominator represent total VSS wasted, including sludge wasted purposely (ΔX_v) and that lost in effluent from the secondary clarifier ($Q'X_{v,e}$). Term $Q_F X_{V,F}$ corresponds to input of sludge in fresh feed. The difference between the terms within parentheses and this value $Q_F X_{V,F}$ represents net output of VSS from this system.

Since in the formulation of the activated sludge process concentration of VSS in effluent from the secondary clarifier is neglected (i.e., $X_{v,e} \approx 0$), Eq. (5.147) reduces to Eq. (5.148).

$$t_S = X_{v,a}V/(\Delta X_v - Q_F X_{V,F}) \qquad (X_{v,e} = 0) \tag{5.148}$$

Finally, when concentration of VSS in fresh feed is also negligible (i.e., $X_{V,F} \approx 0$),

$$t_S = X_{v,a}V/\Delta X_v \qquad (X_{v,e} \approx 0,\ X_{V,F} \approx 0) \tag{5.149}$$

Sludge age is also referred to as mean cell residence time or solids retention time. The relationship between sludge age and hydraulic or liquid retention time ($t = V/Q$) is presented for two types of complete mix reactors: (1) complete mix—no recycle reactor; and (2) complete mix reactor with recycle (a) with wastage directly from reactor (or reactor effluent), and (b) with wastage from the sludge recycle line.

1. *Complete mix—no recycle reactor.* In this model, liquor in the reactor unit is completely mixed and there is no recycle. This *does not* correspond to the conventional activated sludge process, but rather to flow-through devices such as aerated lagoons (Chapter 6, Section 5), assuming complete mixing to occur in the lagoon. The situation is depicted by Fig. 5.24.

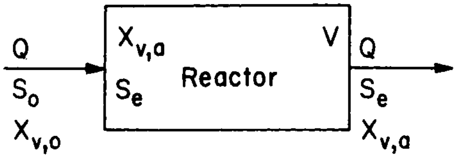


Fig. 5.24. Complete mix reactor without recycle.

Hydraulic or liquid retention time is $t = V/Q$, and the sludge age is [from Eq. (5.145)]

$$t_S = X_{v,a}V/Q(X_{v,a}-X_{v,o}) = [X_{v,a}/(X_{v,a}-X_{v,o})]\,t \tag{5.150}$$

Frequently, concentration of VSS in influent is negligible (i.e., $X_{v,o} \approx 0$). Then, Eq. (5.150) reduces to

$$t_S = X_{v,a}V/QX_{v,a} = V/Q = t \qquad (X_{v,o} \approx 0) \tag{5.151}$$

Thus for the complete mix reactor without recycle when concentration of VSS in influent is negligible, sludge age equals hydraulic (or liquid retention) time. Concentration of sludge in the reactor is kept at a constant value $X_{v,a}$. Since concentration of sludge in effluent also equals $X_{v,a}$, it follows that residence time is such that sludge is not washed out from the system faster than it can reproduce. In fact, since steady state is assumed (constant $X_{v,a}$ in reactor and effluent), residence time is such that sludge washed out in effluent is *exactly replaced* by an equal mass of net sludge yield for the same time interval.

2. *Complete mix reactor with recycle.* This model corresponds to the conventional activated sludge process (Fig. 5.1). Wastage of sludge is usually accomplished (Fig. 5.1) by drawing off from the sludge recycle line. However, the possibility of wasting sludge directly from the reactor (or reactor effluent) is also considered.

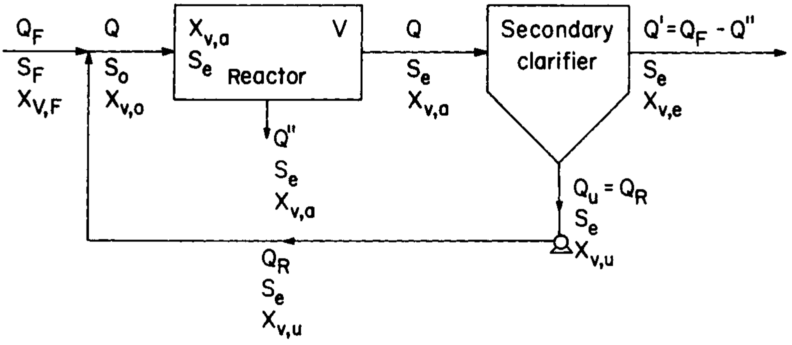


Fig. 5.25. Diagram of complete mix reactor with recycle, and wastage directly from the reactor.

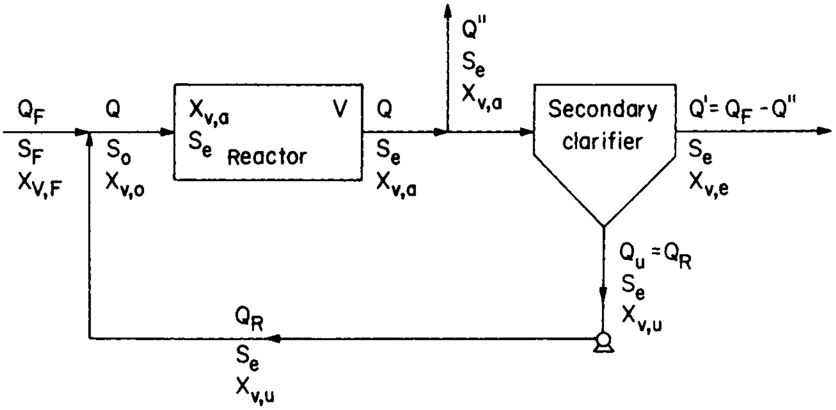


Fig. 5.26. Diagram of complete mix reactor with recycle, and wastage directly from the reactor effluent.

a. Complete mix reactor with recycle, and wastage directly from reactor (or reactor effluent). This corresponds to diagrams shown in Figs. 5.25 and 5.26, the former with wastage taken directly from reactor and the latter with wastage taken from reactor effluent. The hydraulic or liquid retention time for the models in these figures is $t = V/Q$, whereas sludge is [from Eq. (5.145)]

$$t_S = \frac{X_{v,a}V}{\begin{array}{c}\text{lb/day VSS wasted in } Q'' + \text{lb/day VSS} \\ \text{lost in effluent from secondary clarifier}\end{array} - \begin{array}{c}\text{input of VSS in fresh} \\ \text{feed} \quad \text{(lb/day)}\end{array}} \tag{5.152}$$

or

$$t_S = \frac{X_{v,a}V}{[Q''X_{v,a} + (Q_F - Q'')X_{v,e}] - Q_F X_{V,F}} \tag{5.153}$$

If the concentration of VSS in the effluent from the secondary clarifier is negligible (i.e., $X_{v,e} \approx 0$), Eq. (5.153) yields

$$t_S = X_{v,a}V/(Q''X_{v,a} - Q_F X_{V,F}) \qquad (X_{v,e} \approx 0) \tag{5.154}$$

Finally, when the concentration of VSS in fresh feed is also negligible (i.e., $X_{V,F} \approx 0$),

$$t_S = X_{v,a}V/Q''X_{v,a} = V/Q'' \qquad (X_{v,e} \approx 0;\ X_{V,F} \approx 0) \tag{5.155}$$

Comparing Eqs. (5.17) and (5.155), it follows that since $Q'' \ll Q$, then

$$t_S \gg t \tag{5.156}$$

b. Complete mix reactor with recycle, and wastage from recycle line. This corresponds to the flow diagram in Fig. 5.1. Since concentration of sludge in the wastage stream is equal to $X_{v,u}$, whereas it is $X_{v,a}(X_{v,a} < X_{v,u})$ when wastage is taken directly from the reactor or reactor effluent (Figs. 5.25 and 5.26), it follows that the volumetric wastage flow Q'' (which contains a total of $\Delta X_{v,a}$ lb/hr of sludge) is *less* for the case of Fig. 5.1. This is one advantage of taking wastage directly from the recycle line. Hydraulic or liquid retention time is $t = V/Q$, whereas sludge age is given from Eq. (5.152) as

$$t_S = \frac{X_{v,a}V}{[Q''X_{v,u} + (Q_F - Q'')X_{v,e}] - Q_F X_{V,F}} \tag{5.157}$$

If the concentration of VSS in the effluent from the secondary clarifier is negligible (i.e., $X_{v,e} \approx 0$), Eq. (5.157) yields

$$t_S = X_{v,a}V/(Q''X_{v,u} - Q_F X_{V,F}) \qquad (X_{v,e} \approx 0) \tag{5.158}$$

Finally, when concentration of VSS in fresh feed is also negligible (i.e., $X_{V,F} \approx 0$),

$$t_S = X_{v,a}V/Q''X_{v,u} \qquad (X_{v,e} \approx 0;\ X_{V,F} \approx 0) \tag{5.159}$$

Consequently, when wastage is taken from the recycle line, knowledge of both mixed liquor and recycled sludge microorganism concentrations are required for calculation of sludge age.

For the complete mix reactor with recycle (Figs. 5.25, 5.26, and 5.1), residence time is such that sludge is not wasted from the system faster than it reproduces. In fact, since a steady state condition is assumed, wastage (ΔX_v) equals exactly the net sludge yield for the same time interval if loss of VSS in the effluent from the secondary clarifier is negligible.

Example 5.8

For the activated sludge plant designed in Example 5.7 calculate the sludge age.

SOLUTION This is a case of a complete mix reactor with recycle, wastage being taken from the recycle line. Concentration of VSS in the secondary clarifier effluent is negligible (i.e., $X_{v,e} \approx 0$), and also $X_{V,F} = 0$. Equation (5.159) is then utilized to calculate the sludge age. Here

$X_{v,a} = 3000$ mg/liter

$X_{v,u} = 10{,}000$ mg/liter

$V = 1.2$ MG

$Q'' = 0.073$ MGD

Then from Eq. (5.159)

$$t_S = 3000 \times 1.2/0.073 \times 10{,}000 = 4.43 \text{ days}$$

Hydraulic residence time t is 14.2 hr (Example 5.7, Section 7.10).

A relationship between sludge age, substrate removal rate $[(S_o - S_e)/X_{v,a}t]$, and parameters a and b for sludge yield is written from Eq. (5.149) for the complete mix reactor with recycle. [In Eq. (5.149) it is assumed that concentrations of VSS in the effluent from the secondary clarifier and in fresh feed are negligible.]

If in Eq. (5.149) net sludge yield ΔX_v is replaced for the value given by Eq. (5.68),

$$t_S = X_{v,a}V/[a(S_o - S_e)Q - bX_{v,a}V]$$

and

$$1/t_S = [a(S_o - S_e)Q - bX_{v,a}V]/X_{v,a}V$$

Since $V/Q = t$, then

$$1/t_S = a[(S_o - S_e)/X_{v,a}t] - b \tag{5.160}$$

10. Kinetics of Continuous Treatment Systems: Plug Flow, Complete Mix, and Arbitrary Flow Reactors

In the formation of the activated sludge process, the model utilized for the continuous reactor was that of a complete mix vessel. The plug flow continuous reactor model was only briefly mentioned in Section 5 (Fig. 5.9). In this section three models for the continuous reactor (Fig. 5.27) are described: (1) plug flow reactor, (2) complete mix reactor, and (3) arbitrary flow reactor.

1. *Plug flow reactor.* In the plug flow reactor fluid particles travel through the vessel without mixing and therefore are discharged in the same sequence in which they enter. If a continuous tracer is introduced starting at time $t = 0$ (concentration of tracer in the influent being C_o), no tracer appears in effluent until a time t_r, equal to theoretical residence time of the fluid in the vessel, has elapsed. Then, the concentration of tracer in the effluent jumps from a zero value to the value C_0 and remains at that value as long as continuous injection of tracer is maintained.

If a first dose of slug tracer is introduced at time $t = 0$, none of it appears in the effluent until a time t_r has elapsed. At $t = t_r$, concentration of tracer in the effluent jumps from zero to C_0. At time $(t_r + dt)$ it is back again to zero. It jumps again to C_0 at time $t_r + \Delta t$, where Δt is the time interval between the first two discontinuous injections of tracer.

2. *Complete mix reactor.* In this reactor immediate dispersion of particles takes place as they enter the vessel. For a continuous tracer, its concentration

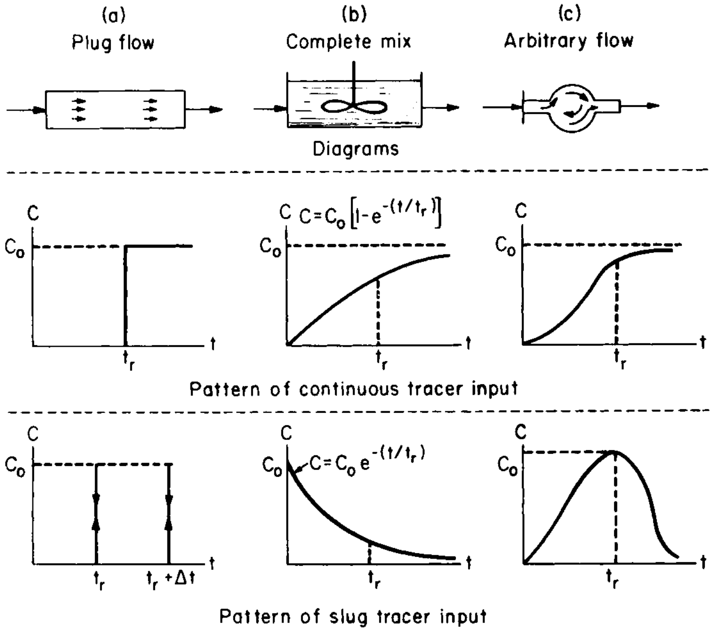


***Fig. 5.27. Continuous reactor models (adapted from Ref.* [7]).**

in the effluent as a function of time is determined by the following material balance for tracer around the reactor:

Rate of change in amount of tracer in reactor
= rate of input of tracer to reactor − rate of output of tracer from reactor

or

$$(dC/dt)V = QC_0 - QC \tag{5.161}$$

where C is the effluent concentration of tracer at any time t; V the volume of the reactor; Q the flow rate; and C_0 the concentration of tracer in the influent. From Eq. (5.161),

$$dC/dt = (Q/V)(C_0 - C) \tag{5.162}$$

Since $V/Q = t_r$ (hydraulic residence time, which is denoted here as t_r so as to distinguish it for time variable t), Eq. (5.162) yields

$$dC/dt = (1/t_r)(C_0 - C)$$

or

$$dC/(C_0 - C) = (1/t_r)\,dt \tag{5.163}$$

Integrating Eq. (5.163) and solving for C,

$$C = C_0[1 - e^{-(t/t_r)}] \tag{5.164}$$

This corresponds to the curve for the concentration of continuous tracer shown in Fig. 5.27(b). As steady state conditions are approached (theoretically at $t = \infty$), Eq. (5.164) yields

$$C = C_0$$

Thus, the curve approaches asymptotically the ordinate $C = C_0$. If addition of tracer stops when a steady state condition is reached, the corresponding value for tracer concentration in the effluent drops gradually following curve $C = C_0 e^{(-t/t_r)}$, also shown in Fig. 5.27(b). This corresponds to a reactor being purged of tracer. As $t \rightarrow \infty$ (steady state), the concentration of tracer in effluent approaches zero.

3. *Arbitrary flow reactor.* These reactors correspond to a partial mix condition between plug flow and complete mix types. Typical patterns for continuous and slug tracer input for arbitrary flow reactors are shown in Fig. 5.27(c). Mathematical analysis of this type of reactor is considerably more complicated than plug or complete mix types, and for this reason these two models are usually chosen to describe reactor performance.

It is interesting to compare efficiency of BOD removal for continuous reactors with recycle (typical activated sludge plant), adopting complete mix and plug flow models to describe the reactor in question. Comparison is made by computing for a given wastewater (i.e., k and S_n fixed) the effluent BOD (S_e) for fixed values of flow rate Q, influent BOD (S_F), recycle ratio (r), and MLVSS concentration ($X_{v,a}$) for various assumed residence times t. For the complete mix reactor, the kinetic model is given by Eq. (5.19). If in Eq. (5.19) S_o is eliminated utilizing Eq. (5.96), the result is

$$(S_F - S_e)/[(1+r)X_{v,a}t] = k(S_e - S_n) \tag{5.165}$$

Solving for S_e,

$$S_e = [S_F + kS_n(1+r)X_{v,a}t]/[1 + kX_{v,a}(1+r)t] \tag{5.166}$$

A typical plot of S_e vs. t obtained from Eq. (5.166) is shown by the curve labeled "complete mix model" indicated in Fig. 5.28. (For $t = 0$, $S_e = S_F$ and for $t = \infty$, $S_e = S_n$.)

A kinetic model for the continuous reactor with recycle under plug flow conditions is mathematically quite difficult to derive. A model has been obtained, however, by Lawrence and McCarthy [4]. This model predicts for a given residence time t a lower value of effluent BOD than that for the corresponding complete mix model. This is indicated by the dotted curve in Fig. 5.28. Thus, the plug flow recycle system is theoretically more efficient than the complete mix recycle system for stabilization of soluble wastes. In practice, however, the plug flow model is difficult to obtain because of longitudinal dispersion. Also, the complete mix systems handle sudden changes in influent

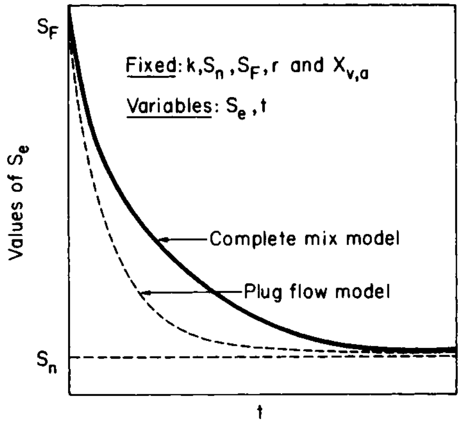


Fig. 5.28. Plot of S_e vs. t for continuous flow reactors with recycle (complete mix and plug flow models).

BOD (shock loads) much more satisfactorily than plug flow systems. In addition, there is the unfavorable situation of variable *F/M* ratios along plug flow reactors, and its possible undesirable effect on the settling characteristics of the sludge discussed in Section 5. All these factors tend to reduce differences in actual efficiency of BOD removal for the two models.

Figure 5.29 shows the progressive BOD reduction occurring in a plug flow reactor from value S_o at the inlet to the final value S_e. By dividing the aeration tank into a series of complete mix reactors (assume a uniform soluble BOD value for the liquor between any two dotted partitions in Fig. 5.29), an improvement in treatment performance is obtained without a major loss in ability of the system to handle shock loads. This is the idea behind the step aeration scheme (Chapter 6, Section 4.1, Fig. 6.6).

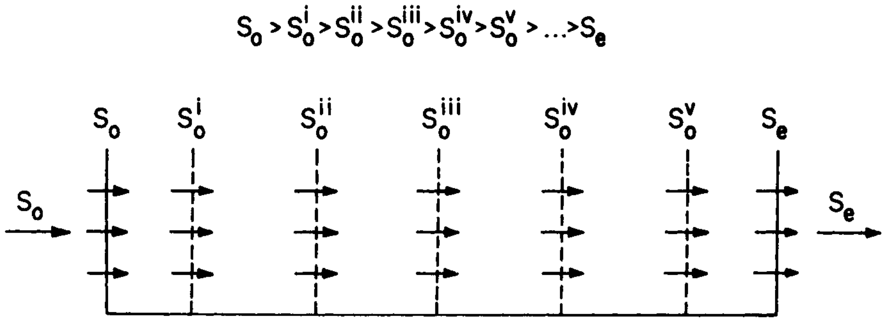


Fig. 5.29. BOD reduction in a plug flow reactor.

Problems

I. Determination of design parameters for an activated sludge project. An industrial plant considers an activated sludge system for disposal of wastewaters. Preliminary tests are performed in laboratory scale continuous reactors. Four reactors are operated in parallel until steady state conditions are obtained. Data taken are presented in the following tabulations.

TABLE 1
For Removal Kinetics

Reactor no.	Average BOD_5 of influent (mg/liter)	Average BOD_5 of effluent (mg/liter)	Average MLVSS concentration (mg/liter)	Residence time (hr)
1	850	100	2000	4.81
2	800	50	2500	7.32
3	750	25	3100	12.7
4	700	15	3100	18.4

TABLE 2
Oxygen Utilization and Sludge Production

Reactor no.	Oxygen uptake rate R_r [mg O_2/(liter)(day)]	Sludge yield $\Delta X_v/V$ [mg sludge/(liter)(day)]
1	3200	2500
2	2187	1450
3	1425	780
4	1008	403

From these data determine design parameters k (hr^{-1} and day^{-1}), S_n, a, a', b, and b'.

II. An organic chemical wastewater is to be treated by a proposed activated sludge plant to produce an effluent BOD of 50 mg/liter during summer conditions (20°C). Wastewater characteristics are

Flow = 2.0 MGD
Influent BOD = 1000 mg/liter

Treatment parameters are

$k = 0.0005\ \text{hr}^{-1}$ at 20°C
$a = 0.50$ lb MLVSS/lb BOD_r
$a' = 0.55$ lb O_2/lb BOD_r
$b = 0.1$ lb MLVSS/(day)(lb MLVSS)
$b' = 0.14$ lb O_2/(day)(lb MLVSS)
$F/M = 0.6$
$S_n = 0.0$ mg/liter

Take

$$X_a = 3000 \text{ mg/liter}$$
$$X_u = 12{,}000 \text{ mg/liter}$$

Neglect influent suspended solids.

Calculate

1. Reactor volume (Mgal) and sludge return rate (Mgal/day)
2. Oxygen required (lb O_2/hr)
3. Net sludge yield (lb MLVSS/day)
4. HP required for surface aeration. Characteristics of the aerator are given by Fig. 4.17.

Base calculation on 20°C operation and take

$$C_{SW} = 8.0 \text{ mg/liter}$$
$$C_L = 1.0 \text{ mg/liter}$$
$$\alpha = 0.8$$

Calculate required power level in HP/1000 gal

5. Nutrient requirements (lb/day) for nitrogen and phosphorus

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6

Secondary Treatment: Other Aerobic and Anaerobic Wastewater Treatment Processes

1. Introduction

The basic mechanism for aerobic treatment of wastewaters is described in Chapter 5, Section 4.1.1. The conventional activated sludge process is also studied in detail in Chapter 5.

In Sections 2–4 of this chapter, several modifications of the activated sludge process are described. Other types of wastewater treatment (aerated lagoons, stabilization ponds, and trickling filters) are discussed in Sections 5–7. These are mostly aerobic processes. Anaerobic treatment of wastewaters is studied in Section 8.

For both aerobic and anaerobic processes the approach utilized for the mathematical formulation, as well as the procedure followed to obtain design data from bench scale units, are those developed by Eckenfelder and associates.

2. Extended Aeration (or Total Oxidation Process)

2.1. INTRODUCTION

This process, also referred to as total oxidation, is a modification of the activated sludge process. The fundamental idea in extended aeration as compared to the conventional activated sludge process is to minimize the amount of excess sludge. This is achieved by increasing residence time; thus the reactor volume is comparatively larger than that required in the conventional activated sludge process. As a result, essentially all degradable sludge formed is consumed by endogenous respiration. Referring to Eq. (5.68), the condition for zero net yield of sludge is

$$\Delta X_v \approx 0 \tag{6.1}$$

or

$$aS_rQ = bX_{v,a}V \tag{6.2}$$

Theoretical residence time to achieve zero net yield of MLVSS is obtained from Eq. (6.2).

$$t = V/Q = aS_r/bX_{v,a} \tag{6.3}$$

The main advantage of the extended aeration process is that sludge handling facilities are minimal compared to those required for the activated sludge process.

2.2. COMPARISON OF EXTENDED AERATION AND ACTIVATED SLUDGE PROCESSES

There are four basic features which distinguish extended aeration from the conventional activated sludge process:

1. Longer detention time in aerator
2. Lower organic loadings. For the extended aeration process organic loading, expressed in terms of food to microorganism ratio (F/M) (Chapter 5, Section 5), is usually between 0.10 and 0.25, as compared to values of 0.3–0.7 for the conventional activated sludge process.
3. Higher concentration of biological solids in the aerator. These values range from 3500 to 5000 mg/liter for extended aeration, as compared to 2000–4000 mg/liter for the activated sludge process. Combination of features

considered under (2) and (3) (i.e., less food for greater microorganism population) results in starvation conditions for the microorganisms. Resulting "cannibalism" (endogenous respiration conditions) reduces concentration of MLVSS, and thus a minimization of sludge accumulation is achieved.

4. Higher consumption of oxygen in extended aeration process. For domestic wastewater treatment, Pasveer [11] reports for the extended aeration process an oxygen consumption approximately twice that for the conventional activated sludge process, namely 18 vs. 9 kwh/(capita)(year). With this comparatively small extra cost for energy, substantial savings in capital expenditure may be achieved. Table 6.1 presents a comparison of the main characteristics of conventional activated sludge and extended aeration processes.

TABLE 6.1
Comparison of Conventional Activated Sludge and Extended Aeration Processes

Characteristics	Activated sludge	Extended aeration
Food to microorganism ratio [lb BOD_5/(day)(lb MLVSS)]	0.3–0.7	0.10–0.25
MLVSS concentration in reactor (mg/liter)	2000–4000	3500–5000
Overall BOD_5 removal efficiency (% includes both soluble and suspended BOD_5)	90–95	85–98
Effluent characteristics		
Soluble BOD_5 (mg/liter)	10–20	10–20
Total BOD_5 (suspended + colloidal + soluble) (mg/liter)	15–25	20–40
Suspended solids (mg/liter)	<20	<70
Sludge yield (lb/lb BOD_5 removed)	≈ 0.03	≈ 0.01
O_2 requirement (as % of BOD_5 removed)	90–95	120

2.3. APPLICATION OF EXTENDED AERATION

The extended aeration process has been applied mostly in treatment of wastewaters when daily volume is less than 2000 gal/day. This includes treatment of domestic sewage for small communities, housing developments, recreational areas, and some industrial wastes. Extended aeration package units are commercially available. If well designed and operated, they should not present odor problems and thus can be located within populated areas.

2.4. EXTENDED AERATION UNITS

Figures 6.1 and 6.2 illustrate a conventional extended aeration unit and a variation known as the oxidation ditch. In the conventional aeration unit (Fig. 6.1), the influent passes first through a screen to remove large suspended

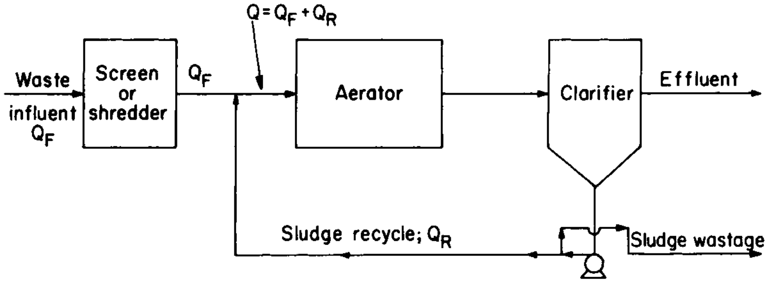


Fig. 6.1. Conventional extended aeration process.

solids, in order to protect the aerator unit from damage resulting from clogging. In some units a shredder is provided instead of a screen. The flow diagram of the conventional unit is essentially identical to that of the activated sludge process. Effluent from the clarifier may be chlorinated prior to discharge in the receiving water.

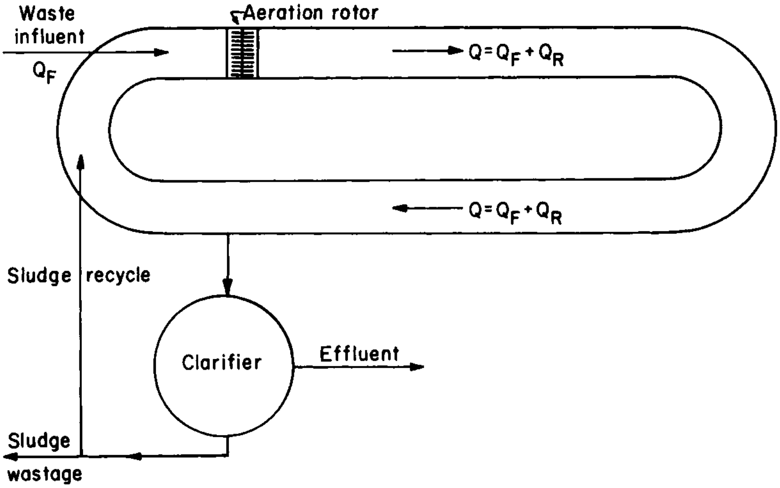


Fig. 6.2. Oxidation ditch.

Figure 6.2 shows a diagram of the oxidation ditch. An essential part of this system is an aeration ditch provided with an aeration rotor. This rotor has two functions: aeration and provision of a flow velocity to the mixed liquor in the ditch. Liquid flow velocity is of the order of 1 ft/sec. The mixture of sewage and activated sludge repeatedly passes over the aeration rotor at short intervals. A typical rotor has a diameter of approximately 30 in., revolves at about 75 rpm, has a depth of immersion of about 6 in., and an oxygenation capacity (OC) of the order of 6 lb/hr.

2.5. SETTLING OF SLUDGE FROM EXTENDED AERATION

Although theoretically sludge yield is nil for the extended aeration process, in practice this is not the case since part of the sludge is not biodegradable and therefore accumulates. The net sludge yield must be wasted.

Figure 5.7 (Chapter 5) reveals that for values of F/M in the range utilized for the extended aeration process (0.10–0.25), the microorganisms are driven to endogenous respiration and metabolize the cytoplasmic material of their "fellow" microorganisms. The remains of this "cannibalist feast" are nondegradable cellular shells which are relatively light compared to cytoplasmic material and settle with difficulty. Settling tanks for extended aeration systems should therefore provide longer retention time than for the conventional activated sludge process. Retention time is approximately 4 vs. 2 hr for the conventional activated sludge process.

2.6. NITRIFICATION IN EXTENDED AERATION

As organic loading (F/M ratio) is low in extended aeration and since a large excess of air is supplied, nitrification may occur to an appreciable extent involving the conversion of ammonia nitrogen to nitrite and nitrate. A problem related to nitrification is a drop in pH for the system due to formation of nitric acid. The pH may drop as low as 4.5, in which case the biological process may be adversely affected.

2.7. DESIGN CRITERIA FOR EXTENDED AERATION

2.7.1. Calculation of Residence Time for Extended Aeration Units

Since the detention period required for BOD removal is much shorter than that for autoxidation of sludge, aerator volume is controlled by the rate of sludge oxidation.

Formulation of design procedure for extended aeration is based on Eq. (6.2), which must be slightly modified prior to its application. The modifications are as follows: in Eq. (6.2), the left-hand member presupposes that all sludge formed (aS_rQ) is biodegradable. Experimental data indicate that approximately 77% of the sludge produced is biodegradable, the remaining 23% comprising nonbiodegradable cellular shells. Therefore, the left-hand member of Eq. (6.2) is rewritten as Eq. (6.4).

$$a_o S_r Q = faS_r Q \tag{6.4}$$

where a_o is the lb of biodegradable MLVSS produced/lb total BOD_5 removed or fa; f the lb biodegradable MLVSS produced/lb total MLVSS produced $\approx$ 0.77; and a the lb total MLVSS produced/lb total BOD_5 removed.

In the right-hand member of Eq. (6.2), MLVSS concentration corresponds only to biodegradable sludge, i.e., $X_{v,a}$ to be substituted by $fX_{v,a}$ (mg/liter of biodegradable sludge). As written in Eq. (5.68) parameter b represents the fraction of total MLVSS oxidized per day.

$$b = \text{lb MLVSS oxidized/(day)(lb total MLVSS in reactor)}$$

Define parameter b_o referred to lb of biodegradable sludge.

$$b_o = \text{lb MLVSS oxidized/(day)(lb biodegradable MLVSS in reactor)}$$

The relationship between b_o and b is

$$b_o = \text{lb MLVSS oxidized/(day)}[f\text{(lb total MLVSS in reactor)}] = b/f$$

If in the right-hand member of Eq. (6.2) b and $X_{v,a}$ are substituted by b_o and $fX_{v,a}$, respectively,

$$b_o(fX_{v,a})V = (b/f)(fX_{v,a})V = bX_{v,a}V \tag{6.5}$$

Therefore, the right-hand member of Eq. (6.2) is left unchanged.

Consequently, the modified Eq. (6.2) is obtained by equating Eq. (6.4) to Eq. (6.5).

$$faS_rQ = a_oS_rQ = bX_{v,a}V \tag{6.6}$$

Residence time is then shown in Eq. (6.7).

$$t = V/Q = a_oS_r/bX_{v,a} = faS_r/bX_{v,a} = fa(S_o - S_e)/bX_{v,a} \tag{6.7}$$

It is convenient to write the expression for residence time in terms of BOD for the fresh feed, S_F. If in Eq. (6.7) $(S_o - S_e)$ is substituted by the value given in Eq. (5.97), one obtains

$$t = (fa/bX_{v,a})[(S_F - S_e)/(1+r)] \qquad (f \approx 0.77) \tag{6.8}$$

2.7.2. Expression for Recycle Ratio *r*

Consider Eq. (5.91) for recycle ratio r (letting $X_{V,F} \approx 0$):

$$r = (8.34Q_FX_{v,a} - \Delta X_v)/[8.34Q_F(X_{v,u} - X_{v,a})] \tag{6.9}$$

For extended aeration, wastage ΔX_v corresponds to nonbiodegradable cells which are approximately 23% of the sludge formed.

$$\Delta X_v = 8.34(1-f)a(S_o - S_e)Q - \text{(effluent loss)} \tag{6.10}$$

where

$$1 - f \approx 1 - 0.77 \approx 0.23$$

Neglecting effluent loss in Eq. (6.10),* and substituting ΔX_v in Eq. (6.9) by its value from Eq. (6.10) yields after simplification

$$r = [Q_F X_{v,a} - (1-f)a(S_o - S_e)Q]/Q_F(X_{v,u} - X_{v,a}) \tag{6.11}$$

Substitution of Q and $(S_o - S_e)$ in Eq. (6.11) by their values from Eqs. (5.5) and (5.97) yields after simplification

$$r = [X_{v,a} - (1-f)a(S_F - S_e)]/(X_{v,u} - X_{v,a}) \tag{6.12}$$

2.7.3. Expression for Reactor Volume

Substitution of the residence time t in Eq. (5.104) by the value given in Eq. (6.8) leads to Eq. (6.13).

$$V = Q_F fa(S_F - S_e)/bX_{v,a} \tag{6.13}$$

The design procedure for an extended aeration unit is illustrated by Example 6.1.

Example 6.1

26,000 gal/day of an industrial wastewater are to be treated by extended aeration. Influent BOD_5 is $S_F = 1200$ mg/liter, and it is desired to reduce it to a value not over 50 mg/liter in the effluent (S_e). Take $X_{v,a}$ and $X_{v,u}$ as 4000 and 12,730 mg/liter, respectively. Values of design parameters a, b, a', and b' have been estimated as 0.7, 0.1, 0.5, and 0.142, respectively (units: BOD_5, day). Assume that 77% of the MLVSS formed is biodegradable and neglect $X_{V,F}$. Calculate (1) recycle ratio, (2) residence time in hr, (3) BOD_5 of combined feed, (4) combined feed in gal/day, (5) F/M ratio, (6) reactor volume in gal, and (7) oxygen requirements in lb/day.

SOLUTION

1. Recycle ratio from Eq. (6.12):

$$r = [4000 - (1-0.77)(0.7)(1200-40)]/(12{,}730-4000) = 0.437$$

2. Residence time from Eq. (6.8):

$$t = [(0.77 \times 0.7)/(0.1 \times 4000)][(1200-50)/(1+0.437)] = 1.078 \text{ days (26 hr)}$$

3. BOD_5 of combined feed from Eq. (5.96):

$$S_o = [1200 + (0.437)(50)]/(1+0.437) = 850.3 \text{ mg/liter}$$

4. Combined feed from Eq. (5.5):

$$Q = 26{,}000(1+0.437) = 37{,}360 \text{ gal/day}$$

---

* Effluent loss in extended aeration is more significant than in the conventional activated sludge process because as explained in Section 2.5, nonbiodegradable material is difficult to settle.

5. F/M ratio from Eq. (5.80):

$$F/M = 850.3/(4000 \times 1.078) = 0.197$$

6. Reactor volume from Eq. (5.104):

$$V = 37{,}360 \times 1.078 = 40{,}290 \text{ gal}$$

7. Oxygen requirements from Eq. (5.85):

$$R_r V = 0.5(850.3 - 50)(0.03736)(8.24) + (0.142)(4000)(0.04029)8.34$$

$$R_r V = 124.7 + 190.9 = 315.6 \text{ lb/day.}$$

3. Contact Stabilization

3.1. INTRODUCTION

Contact stabilization is another modification of the activated sludge process. A flow diagram for the system is shown in Fig. 6.3.

Influent wastewater is mixed with stabilized sludge, and this mixture is aerated in the initial contact tank for which detention time is only 20–40 min. During initial contact an appreciable fraction of suspended and dissolved BOD is removed by biosorption after contact with the well-aerated activated sludge. The mixed effluent from the initial contact tank flows into a clarifier. Clarified effluent is removed and underflow from the clarifier is taken to a stabilization tank, where it is aerated for a period of 1.5–5 hr.

During this stabilization period, biosorbed organics are broken down by aerobic degradation. Stabilized sludge leaving the stabilization tank is in a "starved" condition and ready to adsorb organic waste.

3.2. ADVANTAGE OF CONTACT STABILIZATION VS. CONVENTIONAL ACTIVATED SLUDGE PROCESS

Since only recycled sludge is subject to lengthy aeration, this system permits appreciable reduction in aeration basin volume. This is the main advantage of contact stabilization vs. the conventional activated sludge process. For a wastewater flow Q_F (ft^3/hr) and a sludge recycle of $0.3Q_F$, approximate tank volumes for the conventional activated sludge process and contact stabilization are $11Q_F$ and $4Q_F$, respectively. This corresponds to a nearly threefold tank reduction. Overall removal efficiencies are usually lower than in the conventional activated sludge process, but could easily reach 85–90% BOD_5 removal.

The contact stabilization process is suitable when the wastewater contains a high proportion of BOD in suspended and colloidal forms. Contact stabilization plants may operate without need of primary clarification.

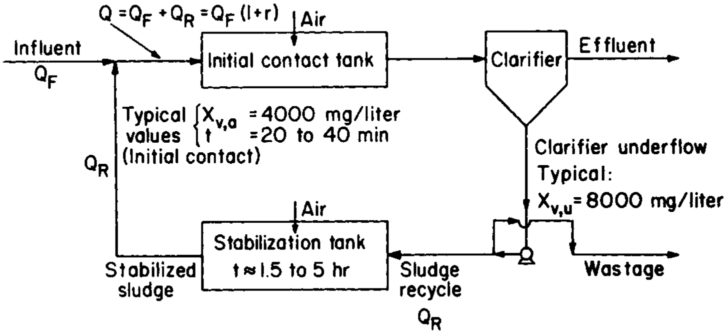


Fig. 6.3. Flow diagram of contact stabilization system.

3.3. SOLUBILITY INDEX (SI) AND OVERALL EFFICIENCY

The solubility index (SI) of a wastewater is defined as

$$\text{SI} = \text{soluble BOD/total BOD} \tag{6.14}$$

where $0 \leqslant \text{SI} \leqslant 1.0$.

As SI approaches zero total BOD tends to be of suspended or colloidal form, and the wastewater becomes suitable for treatment by contact stabilization since most BOD can be removed within a short initial contact period. As SI approaches one, total BOD tends to be of soluble form, and the conventional activated sludge process is more efficient.

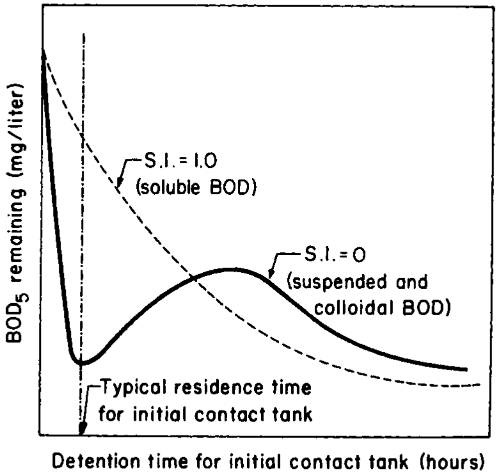


Fig. 6.4. Relationship of SI, BOD removal, and initial contact time.

The relationship of SI, BOD removal, and initial contact time is shown in Fig. 6.4. For typical residence times in the initial contact tank, considerably greater BOD reduction of effluent is achieved when the value of SI approaches unity. The rise in the curve for SI = 0 following the initial drop is due to overoxidation, a concept which is discussed in Section 3.4.

3.4. DESIGN OF CONTACT STABILIZATION SYSTEMS

3.4.1. Selection of Residence Times for Contact and Stabilization

The main objective in design of contact stabilization systems is the selection of residence times for initial contact and stabilization tanks. For a specific wastewater, laboratory tests are performed to determine the effect of a combination of various residence times for initial contact and stabilization tanks on the % BOD removal. A typical set of curves obtained for a specific wastewater is shown in Fig. 6.5. Each curve corresponds to a fixed initial contact detention time. The abscissa is stabilization time and the ordinate is the % BOD removal.

If the desired BOD removal cannot be obtained in the initial contact tank (i.e., at stabilization time = 0), stabilization time must be extended to reach the desired removal. For each contact time (that is, for each curve in Fig. 6.4) there is an optimum stabilization time as indicated, corresponding to a maximum % BOD removal. For stabilization times longer than those corre-

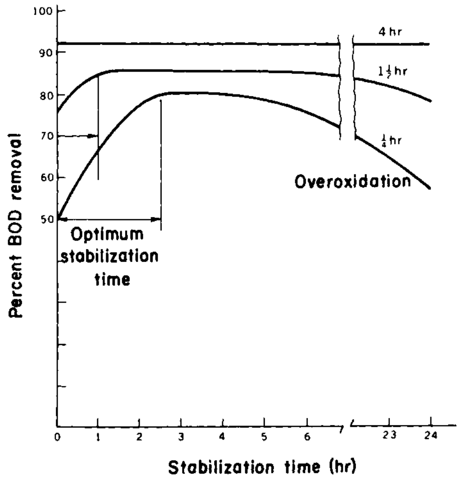


Fig. 6.5. Relationship of contact time, stabilization time, and % BOD removal [3].

sponding to these maxima, sludge degradation (endogenous phase) results in such a breaking down that sludge effluent from the stabilization tank is not adequate to perform biosorption in the initial contact. Therefore, % BOD removal drops. This is known as overoxidation.

3.4.2. Determination of Recycle Ratio

The recycle ratio is determined from Eq. (5.92), neglecting wastage. Typical values of $X_{v,a}$ and $X_{v,u}$ are 4000 and 8000 mg/liter, respectively, in which case $r = 0.5$. Once the recycle ratio and detention times have been determined, sizing of contact and stabilization tanks is straightforward.

3.4.3. Determination of Oxygen Requirements

Oxygen requirements are calculated from Eq. (5.64). Contact and stabilization tanks are considered separately and the corresponding values of a' and b' are used in the calculations.

4. Other Modifications of Conventional Activated Sludge Process: Step Aeration, Complete Mix Activated Sludge Process, and Tapered Aeration

4.1. STEP AERATION

Step aeration is a modification of the conventional activated sludge process in which fresh feed is introduced at several points along the aeration tank. This arrangement provides for an equalization of the F/M ratios along the tank.

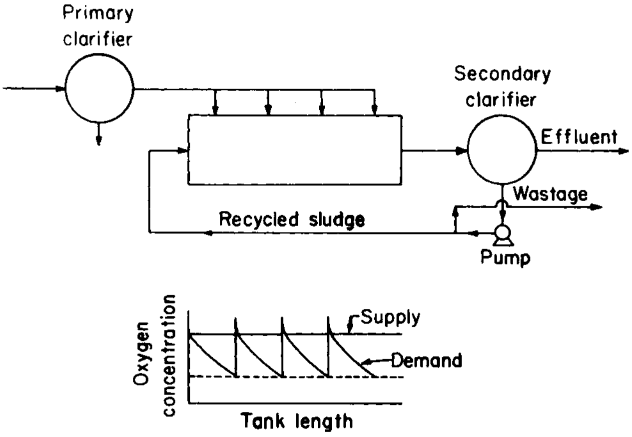


Fig. 6.6. Step aeration process [8].

The aeration tank is divided by baffles into several parallel channels. Each channel constitutes one step of the process, and the steps are linked together in series. This process, as well as oxygen supply and demand along the tank length, are illustrated by Fig. 6.6.

4.2. COMPLETE MIX ACTIVATED SLUDGE PROCESS

In this modification of the activated sludge process, fresh feed and recycled sludge are combined and then introduced at several points in the aeration tank from a central channel. Aerated liquor leaves the reactor from effluent channels on both sides of the aeration tank (Fig. 6.7).

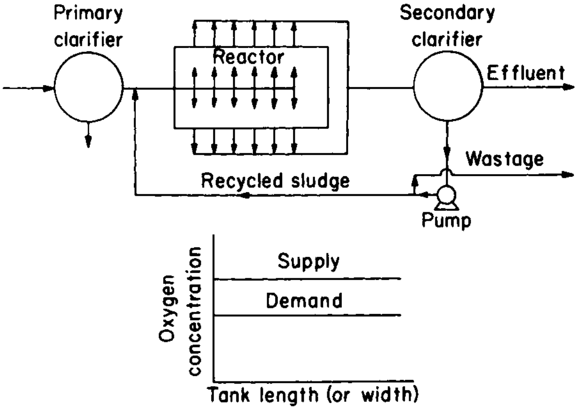


Fig. 6.7. Complete mix activated sludge process [8].

Oxygen supply and demand are uniform along the tank, as indicated by the graph accompanying Fig. 6.7. The mathematical model for the conventional activated sludge process developed in Chapter 5, Section 3.2 assumes complete mixing. If plug flow conditions are assumed, oxygen demand decreases along the length of the aeration tank, whereas the oxygen supply remains constant (Fig. 6.8).

4.3. TAPERED AERATION

The purpose of tapered aeration is to match the amount of air supplied with the oxygen demand along the aeration tank. Since at the inlet oxygen demand is the highest, aerators are spaced more closely to provide a higher oxygenation rate. Spacing between aerators is increased toward the outlet as oxygen demand decreases.

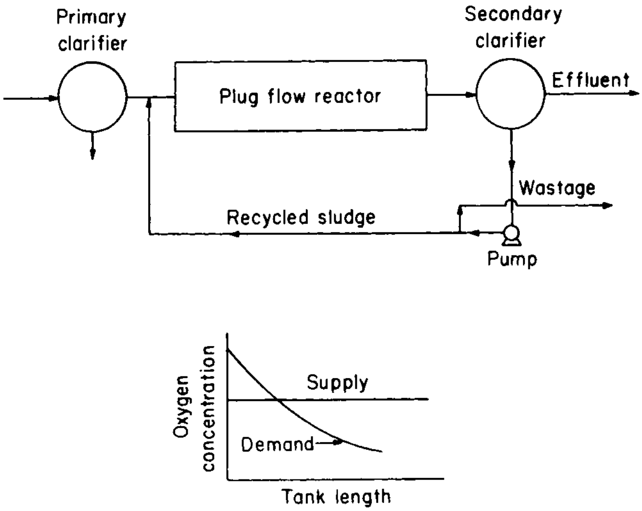


Fig. 6.8. Conventional activated sludge process with plug flow reactor [8].

5. Aerated Lagoons

5.1. INTRODUCTION

Aerated lagoons are basins having depths varying from 4 to 12 ft in which oxygenation of wastewaters is accomplished by aeration units. The fundamental difference between aerated lagoons and the activated sludge system is that recycling of the sludge is provided in the latter as a means of controlling the amount of biological sludge in the aerator. Aerated lagoons are flow-through devices, i.e., no recycle of sludge is provided. Solids concentration in the lagoon is a function of wastewater characteristics and detention time. It is usually between 80 and 200 mg/liter, i.e., much lower than that for activated sludge units (2000–4000 mg/liter).

5.2. MIXING REGIMES FOR AERATED LAGOONS

Turbulence level in lagoons is the basis for their classification into two types.

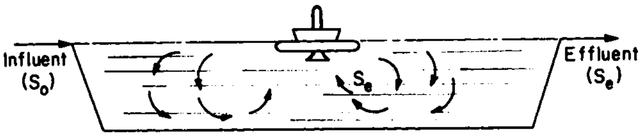
5.2.1. Completely Mixed Lagoons

The turbulence level is sufficient to maintain solids in suspension. Detention times are usually less than 3 days, and power levels are higher than 25 HP per million gallons of basin volume. Power levels for activated sludge units are in the vicinity of 0.25 HP/1000 gal (or 250 HP per million gallons), i.e., about ten times higher than for aerated lagoons. Assumption of complete mixing, which even for activated sludge units is an idealized approximation, is questionable for aerated lagoons. Nevertheless, this assumption is often made, leading to a simple mathematical model for the lagoon. Utilizing appropriate safety factors, this idealized approach is useful.

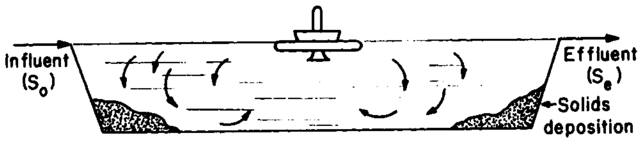
5.2.2. Facultative Lagoons

The turbulence level is insufficient to maintain all solids in suspension. Part of the solids settle to the bottom of the lagoon, where they undergo anaerobic decomposition. Detention times are usually over 6 days and power levels are 4–25 HP per million gallons of basin volume. There is a gradual buildup of residue which has to be desludged at periods of 1–10 years.

Selection of mixing regime is the result of an economic balance between power requirements (greater in completely mixed regime) and acreage cost (more acreage required for facultative lagoons). These two mixing regimes are illustrated in Fig. 6.9.



Completely mixed lagoon
(a)



Facultative lagoon
(b)

Fig. 6.9. Mixing regimes for aerated lagoons.

5.3. KINETICS OF BOD REMOVAL

Assuming appropriate environmental conditions (pH, presence of nutrients, etc.), the rate of BOD removal is a function of detention time, temperature, nature of wastewater, and concentration of suspended volatile solids.

Usually BOD removal rate is assumed to follow first-order kinetics, and the formulation for the continuous reactor assuming complete mixing (Chapter 5, Section 3.2) is utilized. Therefore, Eq. (5.18) is taken as the kinetic model for the lagoon. It is convenient to rewrite Eq. (5.18) in terms of the ratio S_e/S_o, i.e., the percentage of BOD remaining in the effluent. Letting $kX_{v,a} = K$, rearrangement of Eq. (5.18) yields

$$S_e/S_o = 1/(1+Kt) \tag{6.15}$$

where S_0 is the soluble BOD of the influent, S_e the soluble BOD of the effluent, K the removal rate constant, and

$$t \text{ (detention time)} = V/Q = AD/Q \qquad (6.16)$$

In Eq. (6.16), V is the lagoon volume, Q the flow rate, A the horizontal cross-sectional area of the lagoon, and D the depth of the lagoon.

It should be emphasized that this mathematical model is based on two fundamental assumptions, those of first-order kinetics and total mixing conditions. Neither assumption is valid in all cases, nevertheless, this formulation is useful for design purposes.

Values of K can be determined from bench scale data. A reactor like the one shown in Fig. 5.10 (Chapter 5, Section 6.1) is used to simulate an aerated lagoon. The sliding baffle is removed for this simulation. The value of K is determined from linear plots similar to those in Fig. 5.5 (Chapter 5, Section 3.2) based on Eq. (5.18). K values should be corrected for the lagoon temperature (summer and winter conditions) by the procedure described in Section 5.4. Values of K can also be determined from pilot-plant data or estimated from data on operating lagoons.

5.4. ESTIMATE OF LAGOON TEMPERATURE (T_w)

The rate of BOD removal nearly doubles for every 10°C of temperature rise (Chapter 2, Section 7.1). Thus it is necessary to estimate lagoon temperature under average summer and winter conditions. This is done by performing a heat balance. Consider the lagoon represented by Fig. 6.10. Temperature values shown are those for the summer conditions in Example 6.2.

Let

T_i = temperature of the influent

T_w = lagoon (and effluent) temperature (°F)

T_a = atmospheric temperature (°F)

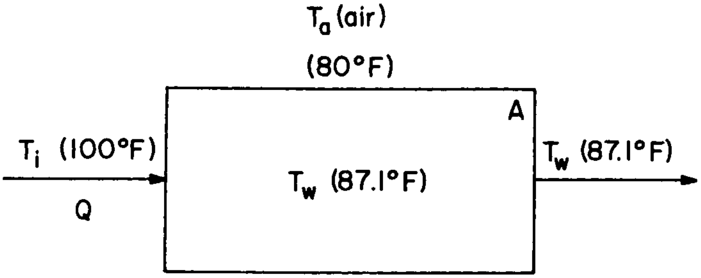


Fig. 6.10. Heat balance for aerated lagoon.

h = heat transfer coefficient [BTU/(day)(ft^2)(°F)] between lagoon and atmosphere (should take into account such factors as wind and humidity effects). In the absence of more accurate information a value of 100 BTU/(day)(ft^2)(°F) may be utilized for estimate purposes

Q = flow rate (Mgal/day)

A = lagoon surface (ft^2)

C = specific heat of wastewater [take as 1.0 BTU/(lb)(°F)]

The enthalpy change of the influent is

$$Q\frac{\text{Mgal}}{\text{day}} \times 8.34 \times 10^6 \frac{\text{lb liquor}}{\text{Mgal liquor}} \times C\frac{\text{BTU}}{(\text{lb liquor})(°\text{F})} \times (T_i - T_w)°\text{F}$$

$$= Q(T_i - T_w)\,8.34 \times 10^6 \qquad (\text{BTU/day}) \tag{6.17}$$

This should equal heat loss to surrounding air given by

$$h\,\text{BTU/(day)(ft}^2\text{)(°F)} \times A\,\text{ft}^2 \times (T_w - T_a)°\text{F} = hA(T_w - T_a) \qquad (\text{BTU/day}) \tag{6.18}$$

Equating Eqs. (6.17) and (6.18),

$$Q(T_i - T_w)\,8.34 \times 10^6 = hA(T_w - T_a) \tag{6.19}$$

Let

$$h \times 10^{-6}/8.34 = f$$

Note: If $h \approx 100$ BTU/(day)(ft^2)(°F), then $f \approx 12 \times 10^{-6}$.

One then writes

$$Q(T_i - T_w) = fA(T_w - T_a) \tag{6.20}$$

Solving for T_w,

$$T_w = (AfT_a + QT_i)/(Af + Q) \tag{6.21}$$

Equation (6.21) permits an estimate of lagoon temperature.

The effect of lagoon temperature (T_w) on BOD removal rate K is given by the empirical equation

$$K_{T_w} = K_{20}\,\theta^{T_w - 20} \tag{6.22}$$

where K_{T_w} is the BOD removal rate at temperature T_w, K_{20} the BOD removal rate at 20°C, and θ the temperature coefficient 1.056 (20–30°C) and 1.135 (4–20°C).

Consider Eqs. (6.15), (6.16), (6.21), and (6.22). Substituting in Eq. (6.15) values of t, K_{T_w}, and T_w given by Eqs. (6.16), (6.22), and (6.21), one obtains

$$S_e/S_o = 1/\{1 + (AD/Q)\,K_{20}\,\theta^{[(AfT_a + QT_i)/(Af + Q)] - 20}\} \tag{6.23}$$

Equation (6.23) permits evaluation of the effect of temperature on percentage of BOD remaining in the effluent.

5.5. OXYGEN REQUIREMENTS FOR AERATED LAGOONS

Oxygen requirements for aerobic oxidation processes are given by Eq. (5.64). Since the concentration of MLVSS ($X_{v,a}$) is low for aerated lagoons, the term $b'X_{v,a}V$ is usually neglected. Therefore

$$\text{lb O}_2/\text{day} \approx a'S_rQ \approx a'(\text{lb BOD removed/day}) \tag{6.24}$$

Values of a' for aerated lagoons vary from 0.9 to 1.4 depending on the nature of waste, mixing regime, and temperature.

5.6. SOLUBLE EFFLUENT BOD FOR AN AERATED LAGOON

The soluble effluent BOD for an aerated lagoon is calculated from Eq. (6.15). Solving for S_e,

$$S_e = S_o/(1+Kt) \tag{6.25}$$

This equation does not take into account BOD feedback to the lagoon due to anaerobic degradation of deposited solids. Usually a correction factor to account for this is introduced in Eq. (6.25). Since anaerobic BOD feedback is greater during the summer, two modified forms of Eq. (6.25) are recommended by Eckenfelder and Ford [5].

$$\text{Summer conditions: } S_e = 1.2S_o/(1+Kt) \tag{6.26}$$

$$\text{Winter conditions: } S_e = 1.05S_o/(1+Kt) \tag{6.27}$$

For activated sludge plants, design procedure consists of specifying a desired effluent quality (S_e) and then calculating residence time (t) necessary to achieve this specified quality. For aerated lagoons the design approach is normally the reverse of that for activated sludge plants. Uusually when an aerated lagoon is being considered, one has available a given acreage of land presumably at a reasonably low cost. The design procedure starts from the known value of the surface area A and an assumed reasonable depth. Therefore, the volume of the lagoon and consequently residence time are fixed. Effluent quality S_e achieved for this residence time is calculated from Eq. (6.25) [or Eqs. (6.26) and (6.27)]. For the activated sludge plant it is unimportant to write Eq. (6.25), since S_e is a *primary specification,* rather than a *calculated* value.

5.7. MLVSS CONCENTRATION IN AERATED LAGOONS

Consider an aerated lagoon for complete mix conditions indicated by Fig. 6.11. Let $X_{v,o}$ be the concentration of VSS in the influent (mg/liter) and $X_{v,a}$ the concentration of MLVSS for the lagoon (same as that in effluent, mg/liter).

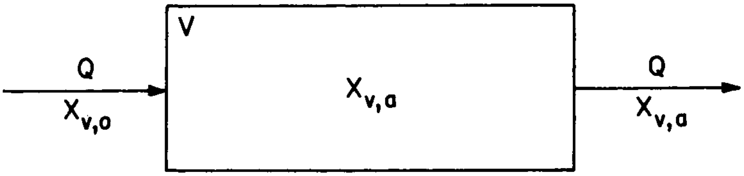


Fig. 6.11. Material balance for MLVSS.

A material balance for VSS yields

(Solids in) + (net synthesis in lagoon) = (solids out)

or

$$QX_{v,o} + (aS_rQ - bX_{v,a}V) = QX_{v,a}$$

Dividing through by Q, letting $V/Q = t$ = detention time, and solving for $X_{v,a}$,

$$X_{v,a} = (X_{v,o} + aS_r)/(1+bt) \tag{6.28}$$

For facultative lagoons, $X_{v,a}$ is lower than the value estimated from Eq. (6.28) depending on lagoon geometry, aerator spacing, power level, and the nature of influent solids.

Remarks similar to those made in Section 5.6 concerning the difference in design approach between activated sludge plants and aerated lagoons are applicable to Eq. (6.28). For activated sludge plants, MLVSS concentration in the aerator (i.e., $X_{v,a}$) is specified by the designer. For aerated lagoons, $X_{v,a}$ is a calculated quantity (from the specified residence time t from which S_e is calculated, and the value for $S_r = S_o - S_e$ is thus established). Equation (6.28) is then utilized to calculate $X_{v,a}$. For activated sludge plants it is unimportant to write Eq. (6.28), since $X_{v,a}$ is a primary specification rather than a calculated quantity.

5.8. RETENTION PERIOD REQUIRED FOR A SPECIFIED EFFLUENT SOLUBLE BOD

From the design approach in Section 5.6, retention period is usually a primary variable which is indirectly specified by the designer. Consequently, obtaining an equation for t for the aerated lagoon is less important than it was for the activated sludge plant. Nevertheless, an equation for residence time is derived which may be utilized to evaluate residence times for a specified value of S_e for several possible selected values of lagoon depth. Take Eq. (5.18) and substitute $X_{v,a}$ by its value given in Eq. (6.28). Simplifying and solving for residence time t, one obtains

$$t = S_r/[(X_{v,o} + aS_r)kS_e - bS_r] \tag{6.29}$$

where $S_r = S_o - S_e$. If the concentration of VSS in the influent ($X_{v,o}$) is negligible, Eq. (6.29) simplifies to yield

$$t = 1/(akS_e - b) \qquad (X_{v,o} \approx 0) \tag{6.30}$$

If in this equation the units to be utilized are t (days), a (lb MLVSS/lb BOD$_r$), k (hr^{-1}), S_e (mg/liter), and b [lb MLVSS/(day)(lb MLVSS)], a conversion factor of 24 hr/day is used. Therefore, Eq. (6.29) is rewritten as Eq. (6.31).

$$t = 1/(24akS_e - b) \qquad (X_{v,o} \approx 0) \tag{6.31}$$

5.9. TOTAL EFFLUENT BOD FOR AN AERATED LAGOON

So far, only soluble BOD of the effluent (S_e) has been considered. To this, one must add the BOD contribution corresponding to volatile suspended solids present in the effluent.* Assuming complete mixing, the concentration of VSS in the effluent is equal to that in the lagoon. BOD contribution due to VSS depends on sludge age, which for aerated lagoons is calculated from Eq. (5.150) [(or Eq. (5.151) if $X_{v,o} = 0$)]. As sludge stabilizes with age, its contribution to effluent BOD lowers. Correlation between sludge age and BOD contribution by VSS is shown in Fig. 6.12. The total BOD of the effluent is [Eq. (6.32)]

$$\text{Total BOD of effluent} = S_e + \psi X_{v,a} \qquad \text{(mg/liter)} \tag{6.32}$$

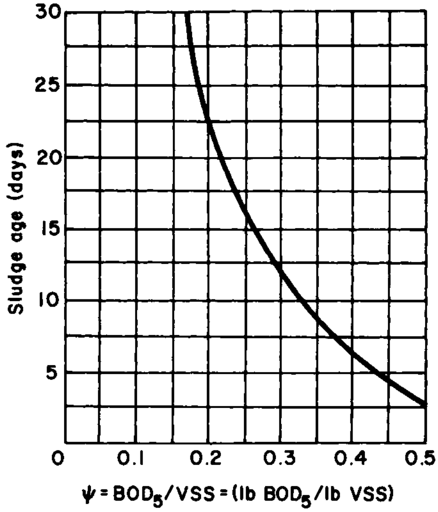


Fig. 6.12. Correlation for insoluble BOD [5].

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* For the activated sludge process, complete settling of MLVSS in the secondary clarifier is assumed. Therefore only soluble BOD is accounted for in the net effluent (refer to Fig. 5.1), where $X_{v,e} \approx 0$).

where S_e is calculated from Eq. (6.26) [or Eq. (6.27)], ψ is determined from Fig. 6.12, and $X_{v,a}$ is calculated from Eq. (6.28).

5.10. DESIGN PROCEDURE FOR AERATED LAGOONS

The procedure proposed by Eckenfelder and associates is illustrated by Example 6.2.

Example 6.2

An aerated lagoon is contemplated for treating an industrial wastewater. An area of 5 acres is available. The following information is taken as a basis for design:

Q = 1.5 MGD (average flow rate)

S_o = 600 mg/liter (BOD_5 of influent)

$X_{v,o}$ = 20 mg/liter (VSS in influent)

T_a = average air temperature, 80°F (summer); 35°F (winter)

T_i = 100°F (influent temperature)

Data obtained from bench scale equipment:

$K = 0.06\ \text{hr}^{-1} = 1.44\ \text{day}^{-1}$ (at 20°C)

a = 0.5 lb VSS/lb BOD_r

$b = 0.06\ \text{day}^{-1}$

a' = 1.1 lb O_2/lb BOD_r

h = 100 BTU/(day)(ft^2)(°F)

Calculate

1. Effluent soluble BOD_5 for summer and winter conditions. Base calculations on a lagoon depth of 8 ft.

2. MLVSS concentration at equilibrium for a completely mixed lagoon $X_{v,a}$ for summer and winter conditions.

3. Total BOD_5 in the effluent for summer and winter conditions.

4. Surface aeration requirements: lb O_2/day, required HP, and power level in HP/Mgal of basin volume.

SOLUTION: Part 1 BOD_5 of effluent

Step 1. Estimate T_w for summer and winter conditions [Eq. (6.21)]. Here

$$A = 5\ \text{acre} \times 43{,}560\ \text{ft}^2/\text{acre} = 217{,}800\ \text{ft}^2$$

and

$$f = 12 \times 10^{-6}$$

corresponding to

$$h = 100\ \text{BTU/(day)(ft}^2\text{)(°F)}$$

Summer: $T_w = (217{,}800 \times 12 \times 10^{-6} \times 80 + 1.5 \times 100)/(217{,}800 \times 12 \times 10^{-6} + 1.5)$

$= 87.1\text{°F} \quad (30.6\text{°C})$

Winter: $T_w = (217{,}800 \times 12 \times 10^{-6} \times 35 + 1.5 \times 100)/(217{,}800 \times 12 \times 10^{-6} + 1.5)$

$= 58.7\text{°F} \quad (14.8\text{°C})$

Step 2. Estimate BOD removal rate K for summer and winter conditions [Eq. (6.22)].

Summer: $K_{30.6\text{°C}} = 1.44 \times 1.056^{(30.6-20)} = 2.57\ \text{day}^{-1}$

Winter: $K_{14.8\text{°C}} = 1.44 \times 1.135^{(14.8-20)} = 0.745\ \text{day}^{-1}$

Step 3. Calculate detention time [Eq. (6.16)].

$$t = \frac{217{,}800\ \text{ft}^2 \times 8\ \text{ft} \times 7.48\ \text{gal/ft}^3}{1{,}500{,}000\ \text{gal/day}} = 8.7\ \text{days}$$

Step 4. Calculate S_e (soluble BOD_5 of effluent) for summer and winter conditions.

Summer: From Eq. (6.26) $S_e = 1.2 \times 600/(1 + 2.57 \times 8.7) = 30.8$ mg/liter

% soluble BOD removal: $[(600 - 30.8)/600] \times 100 = 95\%$

Winter: From Eq. (6.27) $S_e = 1.05 \times 600/(1 + 0.745 \times 8.7) = 84.2$ mg/liter

% soluble BOD removal: $[(600 - 84.2)/600] \times 100 = 86\%$

SOLUTION: Part 2 MLVSS concentration [Eq. (6.28)]

Summer: $X_{v,a} = [20 + 0.5(600 - 30.8)]/[1 + (0.06)(8.7)] = 200$ mg/liter

Winter: $X_{v,a} = [20 + 0.5(600 - 84.2)]/[1 + (0.06)(8.7)] = 182.6$ mg/liter

SOLUTION: Part 3 Total BOD_5 for effluent

Soluble BOD_5 of the effluent has been calculated in Solution, Part 1, Step 4.

Summer: $S_e = 30.8$ mg/liter

Winter: $S_e = 84.2$ mg/liter

Step 1. Estimate ψ from Fig. 6.12. First calculate sludge age [Eq. (5.150)].

Summer: $t_S = [200/(200 - 20)]\,8.7 = 9.67$ days

Winter: $t_S = [182.6/(182.6 - 20)]\,8.7 = 9.77$ days

Then from Fig. 6.12,

Summer: For $t_S = 9.67$ days, read $\psi = 0.332$

Winter: For $t_S = 9.77$ days, read $\psi = 0.330$

Step 2. Estimate VSS contribution to effluent BOD_5.

Summer: $\psi X_{v,a} = 0.332 \times 200 = 66.4$ mg/liter

Winter: $\psi X_{v,a} = 0.330 \times 182.6 = 60.3$ mg/liter

Step 3. Calculate total BOD_5 in effluent.

Summer:	Soluble BOD_5	30.8 mg/liter
	BOD_5 (VSS)	66.4 mg/liter
		97.2 mg/liter
Winter:	Soluble BOD_5	84.2 mg/liter
	BOD_5 (VSS)	60.3 mg/liter
		144.5 mg/liter

SOLUTION: Part 4 Surface aeration requirements

Step 1. Estimate oxygen requirements [Eq. (6.24)].

Summer: $BOD_r = 600 - 30.8 = 569.2$ mg/liter $= 569.2 \times 10^{-6}$ lb BOD_r/lb liquor

$$BOD_r/\text{day} = 569.2 \times 10^{-6} \text{ lb } BOD_r/\text{lb liquor} \times 1.5 \times 10^6 \text{ gal liquor/day} \times 8.34 \text{ lb liquor/gal liquor} = 7121 \text{ lb } BOD_r/\text{day}$$

$$\text{lb } O_2/\text{day} = 1.1 \text{ lb } O_2/\text{lb } BOD_r \times 7121 \text{ lb } BOD_r/\text{day} = 7833 \text{ lb } O_2/\text{day} = 326 \text{ lb } O_2/\text{hr}$$

Winter: $BOD_r = 600 - 84.2 = 515.8$ mg/liter $= 515.8 \times 10^{-6}$ lb BOD_r/lb liquor

$$BOD_r/\text{day} = 515.8 \times 10^{-6} \text{ lb } BOD_r/\text{lb liquor} \times 1.5 \times 10^6 \text{ gal liquor/day} \times 8.34 \text{ lb liquor/gal liquor} = 6453 \text{ lb } BOD_r/\text{day} = 269 \text{ lb } BOD_r/\text{hr}$$

$$\text{lb } O_2/\text{day} = 1.1 \text{ lb } O_2/\text{lb } BOD_r \times 6453 \text{ lb } BOD_r/\text{day} = 7098 \text{ lb } O_2/\text{day} = 296 \text{ lb } O_2/\text{hr}$$

Step 2. Estimate lb $O_2/(\text{HP} \times \text{hr})$ [Eq. (4.34)]. Base estimate on the following values:

$N_o = 2.5$ lb $O_2/\text{HP} \times \text{hr}$

$\alpha = 0.8$

$C_{sw} = 7.0$ mg/liter (summer); 9.5 mg/liter (winter)

$C_L = 1.0$ mg/liter

Then

Summer: For $T_w = 30.6°\text{C}$ (Part 1, Step 1)

$$N = 2.5[(7.0-1.0)/9.2]0.8 \times 1.024^{(30.6-20)} = 1.68 \text{ lb } O_2/(\text{HP} \times \text{hr})$$

Winter: For $T_w = 14.8°\text{C}$ (Part 1, Step 1)

$$N = 2.5[(9.5-1.0)/9.2]0.8 \times 1.024^{(14.8-20)} = 1.63 \text{ lb } O_2/(\text{HP} \times \text{hr})$$

Step 3. Calculate the required HP.

$$\text{HP} = \frac{\text{lb } O_2/\text{hr}}{\text{lb } O_2/\text{HP} \times \text{hr}}$$

where the numerator and denominator have been calculated in Part 4 (Steps 1 and 2, respectively).

Summer: HP = 326/1.68 = 194 HP

Winter: HP = 296/1.63 = 182 HP

Summer operation controls design.

Step 4. Estimate the power level based on 194 HP. The lagoon has a volume of $217{,}800 \text{ ft}^2 \times 8 \text{ ft} = 1{,}742{,}400 \text{ ft}^3$, or

$$1{,}742{,}400 \text{ ft}^3 \times 7.48 \text{ gal/ft}^3 = 13{,}033{,}152 \text{ gal}$$

$$\therefore \text{HP/Mgal} = 194/13.03 = 14.9 \text{ HP/Mgal} \qquad \text{(faculative lagoon level)}$$

6. Wastewater Stabilization Ponds

6.1. INTRODUCTION

The basic difference between the wastewater treatment process described in this section and those previously studied is that no aeration equipment is employed in stabilization ponds. Oxygen needs for ponds are provided by natural surface aeration and by algae, which produce oxygen by photosynthesis. Oxygen released by algae as a result of photosynthesis is utilized by bacteria for aerobic degradation of organic matter. Products of this degradation (carbon dioxide, ammonia, phosphates) are in turn utilized by algae. This cycle symbiotic relationship between algae and bacteria is shown diagrammatically in Fig. 6.13.

Wastewater stabilization ponds are feasible when large land areas are available at low cost and high quality effluent is not required. If BOD of the influent is high, oxygen demand is above that provided by photosynthesis and natural surface aeration. Under these circumstances DO concentration in the

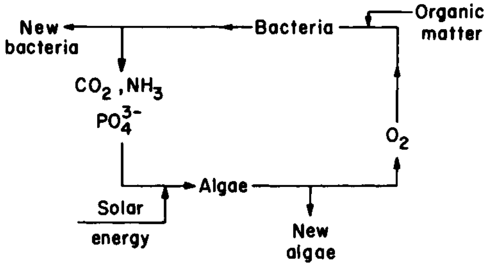

Fig. 6.13. Cyclic symbiotic relationship between algae and bacteria.

wastewater drops to a very low level and anaerobic decomposition prevails. Terminal products for anaerobic decomposition are $CH_4 + H_2O$, in contrast to $CO_2 + H_2O$ for aerobic decomposition. Chemical equations pertinent to anaerobic decomposition are studied in Section 8.1.

Ponds in which the upper layers are aerobic and the lower are anaerobic are referred to as facultative ponds. Most stabilization ponds fall in this category.

Whenever organic loading is very high, oxygen demand may be such that pond operation is anaerobic. When several ponds are operating in series, the first one receiving raw wastewater discharge is anaerobic and the second, which receives partially stabilized wastewater from the first, may be a facultative pond. The last one receiving relatively low BOD wastewater discharge from the preceding one might function as an aerobic pond.

Because of high detention time, frequently about 2 months, removal of refractory organic materials which cannot be accomplished by activated sludge or aerated lagoon processes may become possible in stabilization ponds. Thus, a convenient arrangement may be to provide stabilization ponds following an activated sludge (or aerated lagoon) unit to complete stabilization.

6.2. KINETICS OF BOD REMOVAL FOR STABILIZATION PONDS

An idealized approach similar to that utilized for aerated lagoons is often employed for stabilization ponds. Concentration of MLVSS is not a relevant parameter in stabilization ponds, and thus the term $X_{v,a}$ does not appear in the mathematical model. Equation 5.18 is rewritten as

$$(S_o - S_e)/t = KS_e \tag{6.33}$$

Equation (6.33) indicates that a plot of $(S_o - S_e)/t$ vs. S_e yields a straight line, and the value of K is determined from the slope. Typical graphs of this type are shown in Figs. 6.18 and 6.19.

6.3. LABORATORY SIMULATION OF STABILIZATION PONDS

Batch or continuous models of stabilization ponds have been used on bench or pilot scale. Two of these models, recommended by Eckenfelder and Ford [5], are shown in Fig. 6.14.

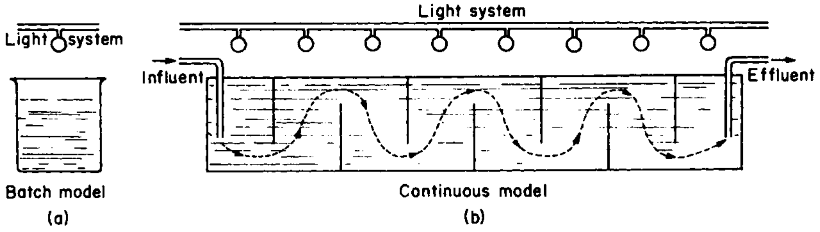

Fig. 6.14. Laboratory models for wastewater stabilization ponds.

From data obtained from these models, graphs such as Figs. 6.18 and 6.19 are constructed, and an estimate of K values is made.

6.4. MATHEMATICAL FORMULATION FOR SEVERAL STABILIZATION PONDS IN SERIES

Start from the modified form of the first-order kinetics removal equation for one pond. Solving Eq. (6.15) for t yields:

$$t = (1 - S_e/S_o)/K(S_e/S_o) \tag{6.34}$$

Ratio S_e/S_o represents the % BOD remaining in the effluent.

For two stabilization ponds in series (Fig. 6.15), one can write [Eq. (6.15)]

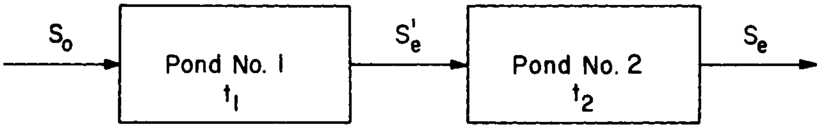

Fig. 6.15. Diagram for two stabilization ponds in series.

For pond No. 1: $S_e'/S_o = 1/(1+K_1 t_1)$ (6.35)

For pond No. 2: $S_e/S_e' = 1/(1+K_2 t_2)$ (6.36)

Combining Eqs. (6.35) and (6.36) by multiplication,

$$S_e/S_o = 1/(1+K_1 t_1)(1+K_2 t_2) \tag{6.37}$$

When retention period t and removal rate constant K are the same for both ponds, Eq. (6.37) yields

$$S_e/S_o = 1/(1+Kt)^2 \tag{6.38}$$

where $t_1 = t_2 = t$ and $K_1 = K_2 = K$. Solving Eq. (6.38) for t,

$$t = [1-(S_e/S_o)^{1/2}]/K(S_e/S_o)^{1/2} \tag{6.39}$$

For n ponds in series, Eqs. (6.38) and (6.39) yield

$$S_e/S_o = 1/(1+Kt)^n \tag{6.40}$$

$$t = [1-(S_e/S_o)^{1/n}]/K(S_e/S_o)^{1/n} \tag{6.41}$$

Further simplification is possible whenever the product Kt is much less than unity. Exponential e^{Kt} is given by the power series in Eq. (6.42).

$$e^{Kt} = 1 + Kt + (Kt)^2/2! + (Kt)^3/3! + \cdots \tag{6.42}$$

If $Kt \ll 1$, one can write as an approximation

$$e^{Kt} = 1 + Kt \tag{6.43}$$

Substituting in the denominator of Eq. (6.40) $(1+Kt)$ by the exponential e^{Kt}, gives Eq. (6.44),

$$S_e/S_o = 1/e^{nKt} = e^{-nKt} \tag{6.44}$$

from which

$$S_o/S_e = e^{nKt} \tag{6.45}$$

or

$$(S_o/S_e)^{1/n} = e^{Kt} \tag{6.46}$$

6.5. EFFECT OF TEMPERATURE ON REACTION RATE CONSTANT *K*

This effect is calculated from an empirical relationship recommended by Eckenfelder and Ford [5].

$$K_t = K_{25}\theta^{t-25} \tag{6.47}$$

where K_t is the reaction rate at t°C; K_{25} the reaction rate at 25°C; t the temperature (°C); and θ the temperature constant (1.06–1.09; take $\theta = 1.07$).

6.6. OXYGEN PRODUCTION IN AEROBIC PONDS

Aerobic stabilization ponds depend on algae to provide the oxygen necessary to satisfy BOD requirements. Since this oxygen is produced by photosynthesis, sunlight is required. This restricts the depth of aerobic ponds to a range of 6–18 in.

The amount of oxygen produced by algae is estimated from Oswald's equation [Eq. (6.48)] [9].

$$\mathrm{OP} = 0.25FI_L \tag{6.48}$$

where OP is the oxygen production [lb O_2/(acre)(day)]; F the light conversion efficiency (%); and I_L the light intensity [cal/(cm^2)(day)]. F is usually assumed to be 4%. Thus $F = 4$ and

$$\mathrm{OP} \approx I_L \tag{6.49}$$

I_L varies from about 100 to 300 cal/(cm^2)(day) during winter and summer, respectively, for a latitude of 30°. This means that the maximum BOD loading for aerobic operation of stabilization ponds to take place varies from 100 to 300 lb BOD$_5$/(acre)(day).

6.7. DEPTH OF OXYGEN PENETRATION IN STABILIZATION PONDS

Depth of oxygen penetration has been correlated by Oswald [10] to surface loading, expressed as lb BOD$_5$/(acre)(day) (Fig. 6.16). The greater the loading, the shallower the depth of oxygen penetration since oxygen demand is higher.

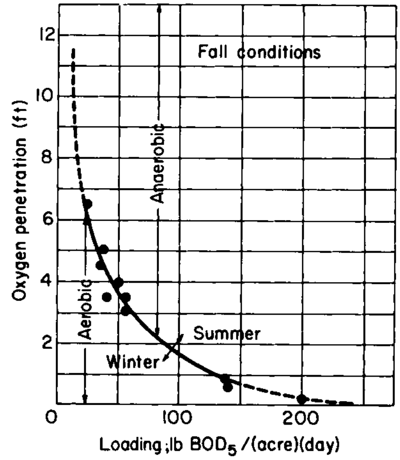

Fig. 6.16. ***Correlation for depth of oxygen penetration in stabilization ponds*** *(adapted from Oswald [10]). (Reprinted with permission, copyright by the University of Texas Press).*

6.8. FACULTATIVE PONDS: HERMANN AND GLOYNA'S EQUATION

The design formulation for stabilization ponds described in Sections 6.2 and 6.4 applies to facultative ponds. Facultative pond depths vary from 3 to 8 ft.

An empirical equation for facultative ponds has been developed by Hermann and Gloyna [6]. This formula is based on several assumptions and developed by analysis of results from bench scale, pilot-plant, and field ponds. It is applicable to domestic sewage for a 85–95% BOD reduction [Eq. (6.50)].

$$V = 10.7 \times 10^{-8} Q S_o (1.085^{35-t}) \tag{6.50}$$

where V is the pond volume (acre × ft); Q the wastewater flow (gal/day); S_o the BOD$_u$ of influent (mg/liter); and t the pond temperature (°C).

6.9. ANAEROBIC PONDS

Loading of anaerobic ponds is such that anaerobic conditions prevail throughout the liquid. Organic loadings range between 250 and 4000 lb BOD_5/(acre)(day). Removal efficiencies vary between 50 and 80%. Since this degree of BOD_5 removal is usually not adequate for discharging the effluent, anaerobic ponds are usually followed by faculative and aerobic ones. Depths from 8 to 15 ft are common, but greater depths are recommended to provide maximum heat retention, besides the resulting economy in terms of land cost.

6.10. SUMMARY OF DESIGN CRITERIA FOR WASTEWATER STABILIZATION PONDS

Design criteria are summarized in Table 6.2.

TABLE 6.2
Summary of Design Criteria for Wastewater Stabilization Ponds[a]

	Ponds		
Criteria	Aerobic	Facultative	Anaerobic
Depth (ft)	0.5–1.5	3–8	8–15
Detention time (days)	2–6	7–50	5–50
Loading			
lb BOD_5/(acre)(day)	100–200	200–500	250–4000
% BOD removal	80–95	70–95	50–80
Algae concentration (mg/liter)	100	10–50	—

[a] Adapted from Eckenfelder [4].

6.11. DESIGN CALCULATIONS FOR STABILIZATION PONDS

Design for stabilization ponds is illustrated by Example 6.3. The design procedure is that recommended by Eckenfelder and associates.

Example 6.3

Wastewater stabilization ponds are considered for treatment of organic chemicals waste. Total design flow is 1.0 MGD, and estimated pond temperatures are 15° and 30°C for winter and summer operations, respectively.

It is desired to reduce wastewater COD from 2000 to 400 mg/liter using two anaerobic ponds in series of equal detention time, and then to lower the COD of the effluent from the second anaerobic pond to 50 mg/liter by means of an aerobic pond. Bench scale tests are performed with wastewater for both anaerobic and aerobic pond conditions, and the laboratory data obtained at 25°C are tabulated below.

	S_o (mg/liter COD)	S_e (mg/liter COD)	Detention time (days)
Anaerobic pond			
Run 1	3000	1000	40
Run 2	2000	667	40
Run 3	1200	400	40
Aerobic pond			
Run 4	700	49	40
Run 5	400	28	40
Run 6	300	21	40

Design the treatment system.

SOLUTION See Figs. 6.17–19.

Step 1. Obtain constant K at 25°C for the anaerobic and aerobic ponds, from plots of laboratory data $[(S_o - S_e)/t$ vs. $S_e]$ shown in tabulation below.

	S_o (mg/liter COD)	S_e (mg/liter COD)	Detention time (days)	$(S_o - S_e)/t$
Anaerobic				
Run 1	3000	1000	40	50
Run 2	2000	667	40	33.3
Run 3	1200	400	40	20
Aerobic				
Run 4	700	49	40	16.3
Run 5	400	28	40	9.3
Run 6	300	21	40	7.0

In summary

$$K \text{ (anaerobic ponds)} = 0.05 \text{ day}^{-1}$$

$$K \text{ (aerobic pond)} = 0.335 \text{ day}^{-1}$$

Step 2. Obtain values of K at 15°C since winter conditions control the design [Eq. (6.47)].

For the anaerobic ponds

$$K_{15} = 0.05 \times 1.07^{(15-25)} = 0.0254 \text{ day}^{-1}$$

(was 0.05 day^{-1} at 25°C, laboratory conditions).

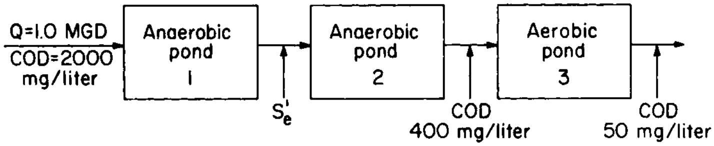

Fig. 6.17. Diagram for Example 6.3.

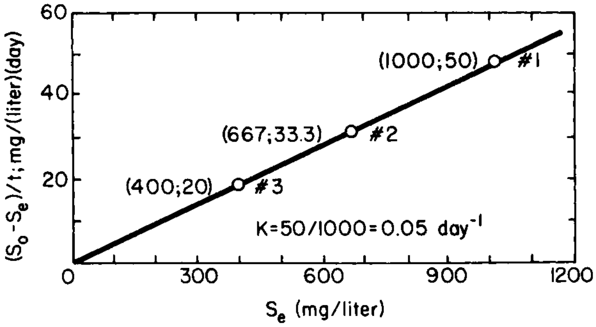

Fig. 6.18. Determination of K (anaerobic ponds) for Example 6.3.

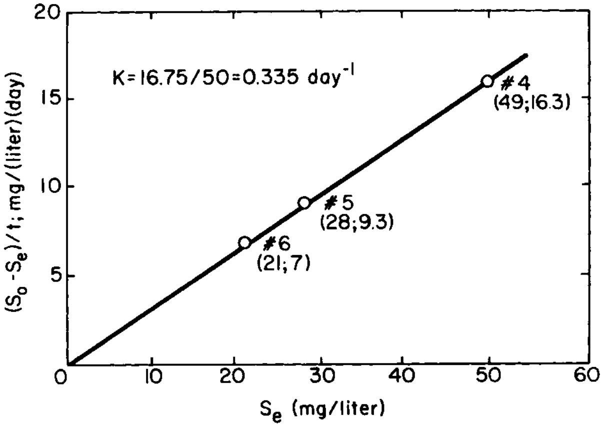

Fig. 6.19. Determination of K (aerobic pond) for Example 6.3.

For the aerobic pond

$$K_{15} = 0.335 \times 1.07^{(15-25)} = 0.17 \text{ day}^{-1}$$

(was 0.335 day^{-1} at 25°C, laboratory conditions).

Step 3. Calculate detention times.

For the two anaerobic ponds in series [Eq. (6.39)]

$$t = [1-(400/2000)^{1/2}]/0.0254(400/2000)^{1/2} = 48.7 \text{ days} \qquad \text{(each pond)}$$

Since for the two anaerobic ponds in Example 6.3 $K_1 = K_2 = K_{15} = 0.0254$ day^{-1} and $t_1 = t_2 = t = 48.7$ days, it follows that from Eqs. (6.35) and (6.36),

$$S_e'/S_o = S_e/S_e'$$

$$\therefore S_e' = (S_o S_e)^{1/2} = (2000 \times 400)^{1/2} = 894 \text{ mg/liter}$$

(COD of effluent from anaerobic pond 1)

For the aerobic pond [Eq. (6.34)]

$$t = (1 - 50/400)/0.17(50/400) = 41.2 \text{ days}$$

Step 4. Calculate pond area (acres). For the two anaerobic ponds assume a depth of 12 ft. Since $t = V/Q = Ah/Q$, where t is the residence time (day); V the volume of pond (ft^3); Q the flow rate (ft^3/day); A the area of pond (ft^2); and h the depth of pond (ft), then $A = tQ/h$ or

$$A = 48.7 \text{ days} \times (1 \times 10^6 \text{ gal/day} \times \text{ft}^3/7.48 \text{ gal} \times 1/12 \text{ ft} \times \text{acre}/43{,}560 \text{ ft}^2)$$

$$= 12.5 \text{ acres per pond}$$

Therefore, design two 12.5 acre ponds, each 12 ft deep. Surface loading [lb BOD$_5$/(acre)(day)] for each of the anaerobic ponds is shown below.

Anaerobic pond 1

Since 2000 mg/liter = 2000×10^{-6} lb COD/lb liquor

$$\text{lb COD/day} = 1 \times 10^6 \text{ gal liquor/day} \times 8.34 \text{ lb liquor/gal liquor}$$
$$\times 2000 \times 10^{-6} \text{ lb COD/lb liquor}$$
$$= 1 \times 8.34 \times 2000 = 16{,}680 \text{ lb COD/day}$$

Assume BOD$_5$/COD $\approx$ 0.7. Then lb BOD$_5$/day = (0.7)(16,680) = 11,676 and lb BOD$_5$/(acre)(day) = 11,676/12.5 = 934 (surface loading for anaerobic pond 1).

Anaerobic pond 2

lb COD/day = 1 × 8.34 × 894 = 7456

lb BOD$_5$/day = (0.7)(7456) = 5219

lb BOD$_5$/(acre)(day) = 5219/12.5 = 418 (surface loading for anaerobic pond 2)

For the aerobic pond, the design procedure is as follows:

1. Assume a depth of oxygen penetration, e.g., $h = 3$ ft.
2. From knowledge of residence time and flow rate, calculate a first approximation of the pond area (acres): $A = tQ/h$.
3. Calculate surface loading in lb BOD$_5$/(acre)(day) based on the assumed value of h.

4. From Fig. 6.16 determine the depth of oxygen penetration and compare it with the assumed value under (1). For aerobic operation, the depth of oxygen penetration should be *at least* equal to the assumed depth. If necessary, assume another value of h and iterate steps (1)–(4).

5. Oxygen production by algae is checked to assure that it is sufficient to satisfy the surface loading [Eq. (6.49)].

Calculations are as follows:

1. Assume $h = 3$ ft.
2. Calculate A in acres.

$$A = 41.2 \text{ days} \times (1 \times 10^6 \text{ gal/day} \times \text{ft}^3/7.48 \text{ gal} \times 1/3 \text{ ft} \times \text{acre}/43{,}560 \text{ ft}^2)$$

$$= 42.1 \text{ acres}$$

3. Surface loading. First calculate loading in terms of lb COD/(acre)(day). Since 400 mg/liter = 400×10^{-6} lb COD/lb liquor,

$$\begin{aligned}\text{lb COD/day} &= 1 \times 10^6 \text{ gal liquor/day} \times 8.34 \text{ lb liquor/gal liquor}\\ &\quad \times 400 \times 10^{-6} \text{ lb COD/lb liquor}\\ &= 3336 \text{ lb COD/day}\end{aligned}$$

Therefore,

$$\text{lb COD/(acre)(day)} = 3336/42.1 = 79.2 \text{ lb COD/(acre)(day)}$$

Assume $BOD_5/COD \approx 0.7$. Then the surface loading in terms of BOD_5 is

$$\text{lb BOD}_5\text{/(acre)(day)} = 0.7 \times 79.2 \approx 55 \text{ lb BOD}_5\text{/(acre)(day)}$$

From Fig. 6.16 for this loading, read $h \approx 3$ ft. Therefore, the assumed depth is appropriate and no further trial is necessary.

4. Check oxygen production by algae. From Eq. (6.49) it follows that oxygen production restricts maximum loadings from 100 to 300 lb BOD_5/(acre)(day). Since actual loading is only 55 lb BOD_5/(acre)(day), an excess of oxygen over BOD_5 requirements is available.

7. Trickling Filters

7.1. INTRODUCTION

The trickling filter is a packed media covered with biological slime through which wastewater is percolated. The slime layer, which usually has a total thickness between 0.1 and 2.0 mm, consists of one aerobic and one anaerobic sublayer, as shown diagrammatically in Fig. 6.20.

The biological aerobic process which takes place in the aerobic sublayer is typical (Chapter 5, Section 4.1.1, Fig. 5.6). The substrate is partially oxidized

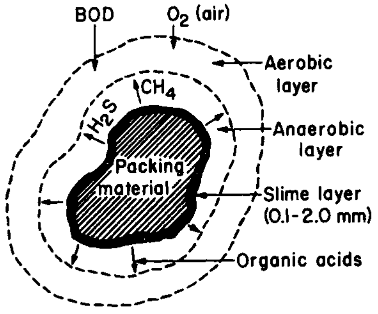

Fig. 6.20. Diagram of aerobic and anaerobic sublayers for a trickling filter.

to provide energy for the biological process. Another part of the substrate is utilized to synthesize new slime material.

In the anaerobic sublayer decomposition takes place with formation of organic acids, CH_4, and H_2S (Section 8.1). In the trickling filter, organic and colloidal matter are removed by aerobic oxidation, biosorption, coagulation, and anaerobic decomposition. Essentially there is no removal by mechanical filtration. The term "trickling filter" is misleading in this respect.

7.2. THICKNESS OF SLIME LAYER

Usually, the thickness of the slime layer is between 0.1 and 2.0 mm. It has an adverse effect on the operation of the trickling filter if it is thicker than 2.0 mm. Clogging of media may occur, thus impairing the wastewater flow and the transfer of oxygen to aerobic microorganisms.

Operational hydraulic loadings are low [0.4–4.0 gal/(min)(ft^2)] and are not sufficient to keep the slime layer scoured off. Thus, hydraulic loading cannot be used in controlling the thickness of the slime layer. This control is exerted mostly by larvae and worms, which thrive on the accumulated slime.

As the slime layer increases in thickness, organic matter in the wastewater is metabolized before it can reach the layer of microorganisms clinging to the surface of the media. These microorganisms are left without sufficient food and tend to enter the endogenous respiration phase. Thus, the slime layer loses its ability to cling to the media surface and is washed away. This phenomena, called sloughing, is a function of organic and hydraulic loading of the filter.

7.3. COMPARISON BETWEEN TRICKLING FILTERS AND ACTIVATED SLUDGE PROCESS

For BOD removal efficiencies of about 60%, it is usually found that trickling filters are more economical than the activated sludge process, in particular for small flow rates of wastewater. For higher BOD removal efficiencies (90% or above), the activated sludge process is more economical because packing material costs would be too high. These considerations suggest a possible two step operation: trickling filters followed by an activated sludge plant, a combination which in some cases may prove advantageous.

Some advantages of trickling filters over the activated sludge process are (1) no power requirements for aeration, (2) simple operation, and (3) slower response and quicker recovery to sudden changes of influent BOD.

7.4. PHYSICAL ARRANGEMENT OF TRICKLING FILTERS

Trickling filters are beds from 3 to 40 ft deep filled with packing such as broken rock, clinkers, or synthetic media (trade names Surfpac, Flocor, Actifil). These plastic material packings are available commercially (Dow Chemical Co., Ethyl Corporation, B.F. Goodrich, Norton Co.) in honeycomb and other shapes. Influent wastewater is usually distributed over the filter by a mechanical rotating arm mechanism and percolates through the packing, coming in contact with the biological slime layer.

Whereas beds filled with rocks, clinkers, or other materials are limited in depth from 3 to 8 ft, beds of synthetic materials are commonly 20–40 ft deep. The higher percentage of void space for synthetic packing allows an easier flow and reduces the risk of flooding.

For ordinary packing (rocks, clinkers, etc.) the following characteristics are typical: diameters: $1\frac{1}{2}$–2 in.; surface area: 24–34 $\mathrm{ft}^2/\mathrm{ft}^3$ of bulk volume; void %: 45–55%; and maximum hydraulic loadings: 0.5 gal/(min)(ft^2).

Advantages of synthetic packings are that they (1) allow packing depth up to 40 ft; (2) allow higher hydraulic loadings, up to 4 gal/(min)(ft^2); (3) have surface areas up to 70 $\mathrm{ft}^2/\mathrm{ft}^3$ of bulk volume; and (4) are less likely to be clogged by wastewaters carrying large amounts of suspended solids.

Disadvantages of synthetic packings are that they are (1) relatively expensive; and (2) inappropriate for wastewater treatment to arrive at a relatively high effluent quality as compared to ordinary packing.

7.5. TRICKLING FILTER SYSTEMS

Most common arrangements for trickling filters are shown in Fig. 6.21. (a) *Single filter system*—May be operated with or without recycle. Recycling is indicated for higher effluent quality. If influent BOD is greater than 500 mg/liter, recycling is usually recommended. (b) *Alternating double filtration*—The first filter is responsible for most BOD removal; the second one is an effluent polisher. Consequently, most slime growth occurs in the first filter. The cycle is reversed periodically (daily or weekly) as indicated by dotted lines in Fig. 6.21 (b). In this manner control of the slime layer thickness is achieved, maintaining a uniform slime thickness in both units. Higher BOD effluent

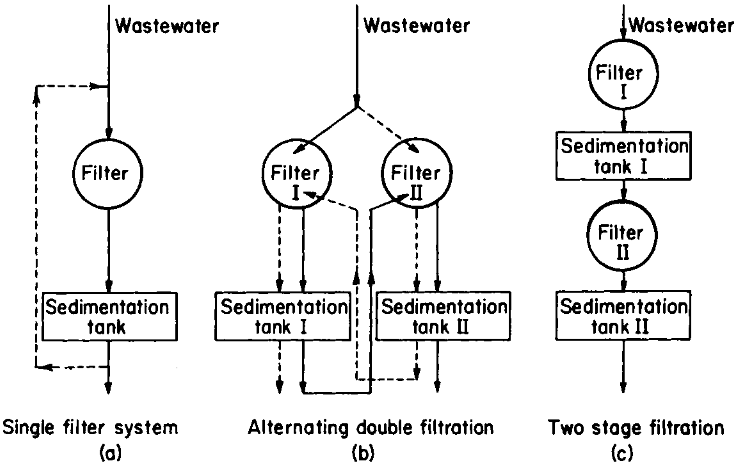

Fig. 6.21. Trickling filter systems.

quality is obtained by this system as compared to the single filters. (c) *Two stage filtration*—The first filter is a coarse one, usually filled with synthetic packing which removes 60–70% of the BOD. The second filter, where slime growth is considerably less, acts as the effluent polisher.

7.6. PRETREATMENT FOR TRICKLING FILTRATION

A pretreatment similar to that for the activated sludge process may be required for trickling filtration. It may be necessary to adjust the pH by neutralization to an optimum range from 7 to 9, because excess alkalinity or acidity disturbs the biological process.

7.7. DESIGN FORMULATION FOR TRICKLING FILTERS

The purpose of design formulation is to obtain a relationship among BOD removal, depth of the filter, and hydraulic loading. The following formulation is the one developed by Eckenfelder and associates. BOD removal is usually expressed as % BOD remaining in the effluent.

$$S_e/S_o = \text{effluent BOD} \quad (\text{mg/liter})/\text{influent BOD} \quad (\text{mg/liter})$$

The trickling filter depth is denoted by D (ft) and hydraulic loading by L [gal/(min)(ft^2)]. Assuming BOD removal to follow first-order kinetics [Eq. (6.51)],

$$dS/dt = -k'X_v S = -K'S \tag{6.51}$$

where $K' = k'X_v$.

Rearranging and integrating,

$$\ln(S_e/S_o) = -k'X_v t = -K't$$

or

$$S_e/S_o = e^{-k'X_v t} = e^{-K't} \tag{6.52}$$

where S_e is the effluent BOD (mg/liter); S_o the influent BOD (mg/liter); k' the removal rate constant (no volatile solids included); X_v the volatile solids; K' the removal rate constant (volatile solids included, $K' = k'X_v$); and t the residence time.

Residence time t is written as

$$t = CD/L^n \tag{6.53}$$

where D is the filter depth (ft); L the hydraulic loading [gal/(min)(ft^2)]; and C, n the constants which are functions of the filter media and specific surface. Specific surface is defined as ft^2 of packing surface per ft^3 of bulk volume. Substitution of t [Eq. (6.53)] in Eq. (6.52) yields

$$S_e/S_o = e^{-K'CD/L^n} = e^{-KD/L^n} \tag{6.54}$$

where $K = K'C$.

Equation (6.54) is the basic mathematical model [4] for trickling filters. It relates % BOD remaining (S_e/S_o) to depth of the filter (D) and hydraulic loading (L). Parameter K (rate constant) is a function of the ease of degradability of the substrate. Parameter n depends on the characteristics of the packing media.

7.8. APPLICATION OF BASIC MATHEMATICAL MODEL TO TRICKLING FILTERS WITHOUT AND WITH RECYCLE

Streams involved in operation of trickling filters without and with recycle are shown in the Fig. 6.22. Recycle improves efficiency of BOD removal and dilutes influent BOD to a level compatible with maintenance of aerobic conditions. When there is no recycle, Eq. (6.54) applies directly. For trickling filters with recycle, influent BOD (S_F) is diluted to a value S_o prior to entering the filter. The relationship among S_o, S_F, S_e, and the recycle ratio is obtained by a material balance for the BOD [loop (---) of Fig. 6.22(b)].

$$S_e/S_o = e^{-KD/L^n} \tag{6.55}$$

$$Q_F S_F + Q_R S_e = (Q_F + Q_R) S_o$$

$$\therefore\ S_o = (Q_F S_F + Q_R S_e)/(Q_F + Q_R)$$

Dividing both numerator and denominator by Q_F and letting $Q_R/Q_F = r =$ recycle ratio [Eq. (6.56)],

$$S_o = (S_F + rS_e)/(1 + r) \tag{6.56}$$

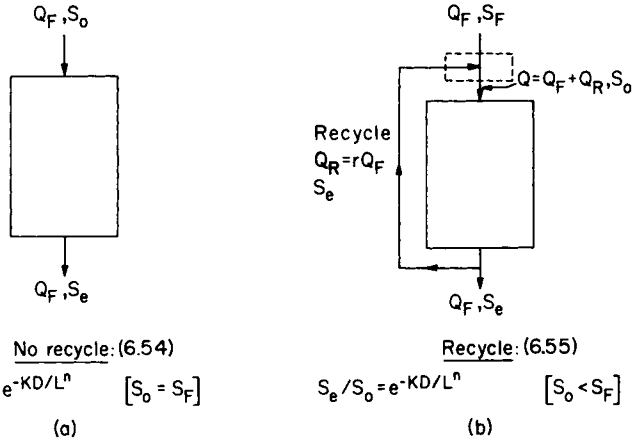

Fig. 6.22. Trickling filter without recycle (a) and with recycle (b).

Substituting this value in Eq. (6.55),

$$S_e/[(S_F + rS_e)/(1+r)] = e^{-KD/L^n} \tag{6.57}$$

For simplicity, let $KD/L^n = X$. Dividing both numerator and denominator by S_F and solving for S_e/S_F,

$$S_e/S_F = e^{-X}/(1 + r - re^{-X}) \tag{6.58}$$

For domestic sewage, an empirical relationship has been developed by Balakrishnan [2] based on Eq. (6.58):

$$S_e/S_o = \exp(-0.003 A_v^{0.644} S_e^{0.54} D/L^n) \tag{6.59}$$

7.9. PROCEDURE FOR DESIGN OF TRICKLING FILTERS WHEN BENCH SCALE OR PILOT-PLANT DATA ARE AVAILABLE

Three items will be discussed: (1) A model of bench scale trickling filter and procedure for obtaining design data from this model, (2) treatment of data obtained in order to determine constants n and K in Eq. (6.54); and (3) application of these results to design of a plant scale trickling filter. Both cases, without and with recycle, are studied. The approach followed is that proposed by Eckenfelder and Ford [5]. Example 6.4 presents a numerical illustration.

7.9.1. Model of Bench Scale Trickling Filter

A sketch of a bench scale model of a trickling filter developed by Eckenfelder and associates is shown in Fig. 6.23. Wastewater contained in the feed reservoir is pumped by a Sigmamotor pump to a perforated plate for flow distribution. There it is mixed with the recycle stream from the settling tank.

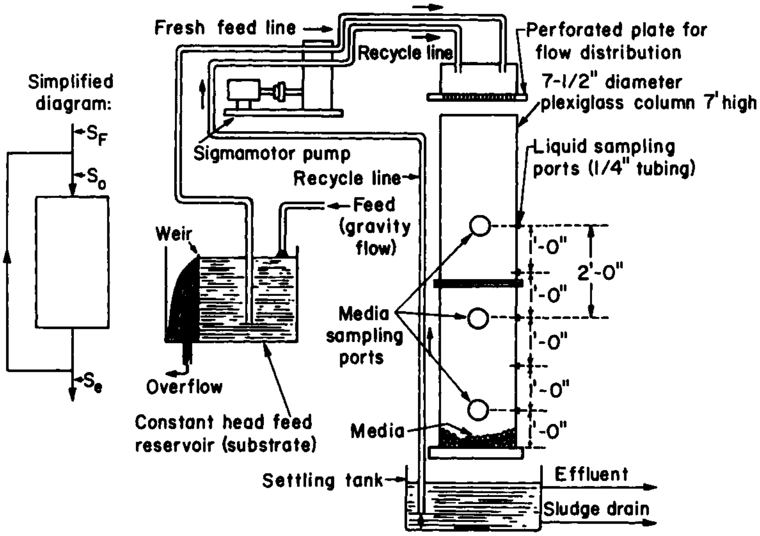

Fig. 6.23. Bench scale trickling filter.

As a preliminary step, it is necessary to generate an acclimated slime on the filter media. This may take from a few days to several weeks, depending on the nature of the wastewater. Samples are taken periodically at the media and liquid sampling ports, and BOD determinations are performed. Steady state BOD values are recorded. Pilot-plant units may also be utilized.

The procedure to obtain basic data is (1) select three or four hydraulic loadings [gal/(min)(ft^2)]. For deep filters ($D > 10$ ft) with plastic packing, flow rates of 0.5–4 gal/(min)(ft^2) are selected. For the pilot-plant unit (Example 6.4), hydraulic loadings of 1, 2, 3, and 4 gal/(min)(ft^2) are utilized (Table 6.3). (2) For each flow rate, sample at least three depths. For Example 6.4, four

TABLE 6.3
Data for Example 6.4[a]

D, depth (ft)	L, hydraulic loading [gal/(min)(ft^2)]			
	$L = 1$	$L = 2$	$L = 3$	$L = 4$
5	57.5	67.5	73.0	76.0
10	33.5	46.0	53.0	57.0
15	19.5	31.0	38.7	43.0
20	11.3	21.0	28.2	32.8

[a] Values in 4 × 4 matrix are those for $(S_e/S_o) \times 100$.

depths (5, 10, 15, and 20 ft) are sampled. Analyses of samples are usually expressed as % BOD remaining, i.e., $S_e/S_o \times 100$. Since in Example 6.4 four hydraulic loadings (L) and four sampling depths are selected, the S_e/S_o values form a 4×4 matrix (Table 6.3). Analyses can also be made in terms of COD or TOC. Determinations of pH, Kjeldahl nitrogen, and water temperature are also performed.

7.9.2. Treatment of Data Obtained in Order to Determine Constants *n* and *K*

Step 1. On semilog paper, plot the % BOD remaining $(S_e/S_o) \times 100$ vs. depth (D) for each value of hydraulic loadings (L). From Eq. (6.54), it follows that [Eq. (6.60)]

$$\ln(S_e/S_o) = -KD/L^n = -(K/L^n)D \tag{6.60}$$

Therefore a family of straight lines is obtained. The absolute value of their slopes corresponds to K/L^n. For Example 6.4, this plot is shown in Fig. 6.24, where four lines correspond to four runs [$L = 1, 2, 3$, and 4 gal/(min)(ft^2)].

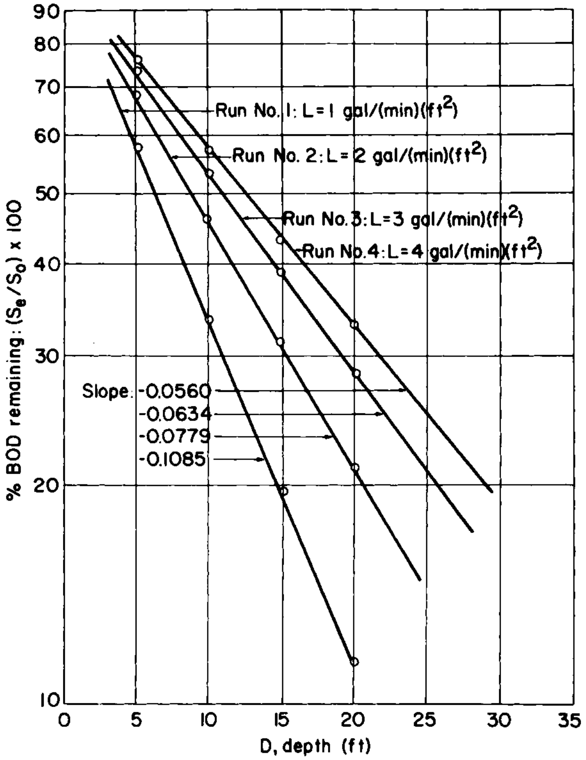

Fig. 6.24. Plot of % BOD remaining vs. depth.

Step 2. On log–log paper, plot the absolute values of the slopes for each of the lines from Step 1 vs. L (Table 6.4 and Fig. 6.25). Since

$$|\text{Slope}| = K/L^n \tag{6.61}$$

then

$$\log|\text{slope}| = \log K - n \log L \tag{6.62}$$

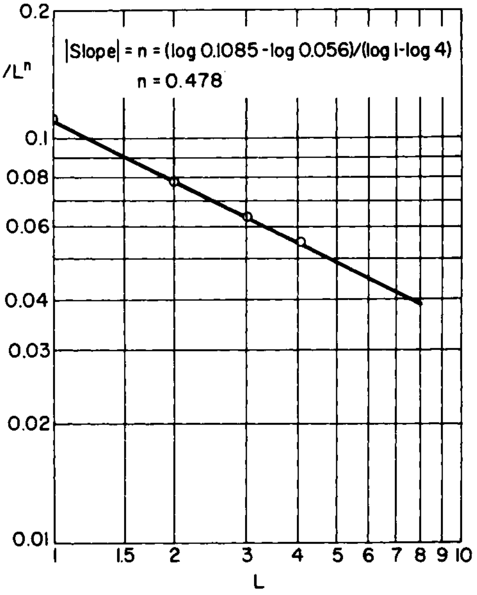

Fig. 6.25. Plot of K/L^n vs. L.

TABLE 6.4
K/L^n* vs. *L

L [gal/(min)(ft^2)]	[a]Absolute value of slopes, K/L^n
1	0.1085
2	0.0779
3	0.0634
4	0.0560

[a] From Fig. 6.24.

One straight line is obtained, each point of which corresponds to one experiment. Its slope yields the value of n in accordance with Eq. (6.62). For Example 6.4 this plot is shown in Fig. 6.25, where four points are utilized for the construction.*

---

* The value of K may be estimated from the ordinate intercept (for abscissa $L = 1$) of Fig. 6.25 as 0.11. It is preferable, however, to determine K from the slope of the straight line in Fig. 6.26, as described in Step 4.

Step 3. Perform the following calculations:

1. Determine the values of L^n for each experimental run utilizing the value of n determined in Step 2 (Table 6.5).

TABLE 6.5
Values of $L^n = L^{0.478}$ vs. L

Run no.	L (gal/(min)(ft^2)]	$L^n = L^{0.478}$
1	1	1.0
2	2	1.393
3	3	1.690
4	4	1.941

2. Calculate the values of D/L^n for each experimental run at each depth. For Example 6.4 this results in the 4 × 4 matrix (Table 6.6).

TABLE 6.6
Matrix for Values of $D/L^n = D/L^{0.478}$

D, depth (ft)	Values of L			
	$L=1$	$L=2$	$L=3$	$L=4$
5	5.0	3.59	2.96	2.57
10	10.0	7.17	5.91	5.15
15	15.0	10.75	8.87	7.71
20	20.0	14.35	11.83	10.30

3. Construct a table of D/L^n vs. $(S_e/S_o) \times 100$ from inspection of Tables 6.3 and 6.6. For Example 6.4, this is shown in Table 6.7. Values of D/L^n are read from Table 6.6 from left to right and from top to bottom, and corresponding values of $(S_e/S_o) \times 100$ are read from Table 6.3 in the same way. For Example 6.4, a summary of calculated values is shown as a convenient array in Table 6.8.

Step 4. On semilog paper plot the values of $(S_e/S_o) \times 100$ vs. D/L^n (Table 6.7). From Eq. (6.60) obtain

$$\ln(S_e/S_o) = -K(D/L^n) \tag{6.63}$$

The slope of the straight line thus obtained yields the value of K. For Example 6.4, this plot is shown in Fig. 6.26. Since in this example there is a 4 × 4 matrix, 16 points are utilized in the construction of this line. Therefore, the values of constants n and K for the basic mathematical model have now been obtained.

TABLE 6.7
$D/L^n = D/L^{0.478}$ vs. $(S_e/S_o)\times 100$

$D/L^n = D/L^{0.478}$	$(S_e/S_o)\times 100$
5	57.5
3.59	67.5
2.96	73.0
2.57	76.0
10.0	33.5
7.17	46.0
5.91	53.0
5.15	57.0
15.0	19.5
10.75	31.0
8.87	38.7
7.71	43.0
20.0	11.3
14.35	21.0
11.83	28.2
10.30	32.8

TABLE 6.8
Summary of Calculated Values for Example 6.4

D, depth (ft)	L [gal/(min)(ft^2)]	$L^{0.478}$	$D/L^{0.478}$	$(S_e/S_o)\times 100$
5	1.0	1.0	5.0	57.5
10	1.0	1.0	10.0	33.5
15	1.0	1.0	15.0	19.5
20	1.0	1.0	20.0	11.3
5	2.0	1.393	3.59	67.5
10	2.0	1.393	7.17	46.0
15	2.0	1.393	10.75	31.0
20	2.0	1.393	14.35	21.0
5	3.0	1.690	2.96	73.0
10	3.0	1.690	5.91	53.0
15	3.0	1.690	8.87	38.7
20	3.0	1.690	11.83	28.2
5	4.0	1.941	2.57	76.0
10	4.0	1.941	5.15	57.0
15	4.0	1.941	7.71	43.0
20	4.0	1.941	10.30	32.8

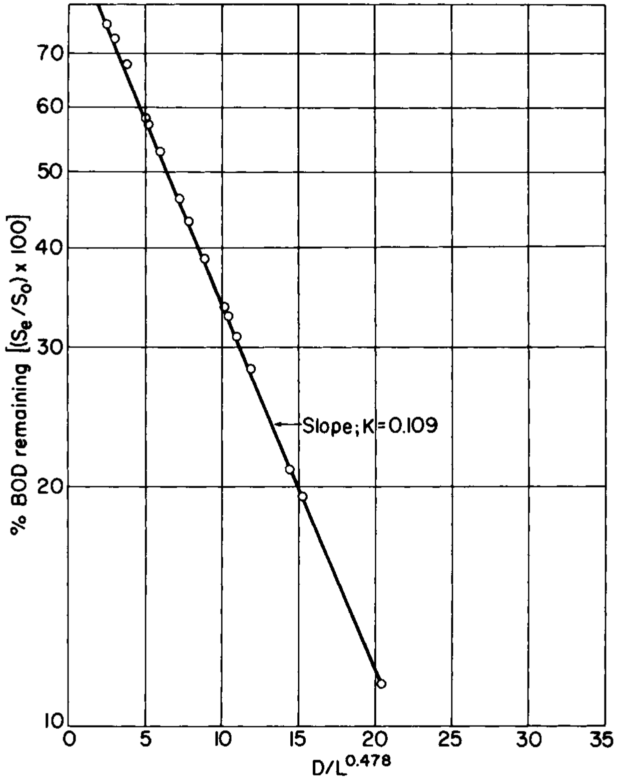

Fig. 6.26. Plot of $(S_e/S_o)\times 100$ *vs.* $D/L^{0.478}$.

7.9.3. Application of Results to the Design of a Plant Scale Trickling Filter

Case 1. Trickling filters without recycle [Fig. 6.22(a)]

Design is based on Eq. (6.54), for which values of K and n are determined as described in Section 7.9.2.

Step 1. Calculate required hydraulic loading L for specified BOD removal corresponding to required S_e/S_o. Assume a value for depth D and solve for hydraulic loading L [gal/(min)(ft^2)]. Solving Eq. (6.54) for L,

$$L = [-KD/\ln(S_e/S_o)]^{1/n} \tag{6.64}$$

Equation (6.64) yields the required hydraulic loading L in gal/(min)(ft^2).

Step 2. The required filter area A is

$$A = Q_F/L = \frac{\text{gal/day}}{\text{gal/(day)(ft}^2)} = \text{ft}^2 \tag{6.65}$$

Step 3. Filter diameter is given in Eq. (6.66).

$$d = (4A/\pi)^{1/2} = (A/0.785)^{1/2} \text{ ft} \tag{6.66}$$

Case 2. Trickling filters with recycle [Fig. 6.22(b)]

Step 1. Calculate S_o from Eq. (6.56) for a specified value of the recycle ratio r.

Step 2. Solving Eq. (6.55) for L, one obtains an equation similar to Eq. (6.64), from which the hydraulic loading is calculated. Steps 1 and 2 may be combined by solving Eq. (6.58) directly for L. Final result is

$$L = \left[\frac{-KD}{\ln\dfrac{(S_e/S_F)(1+r)}{1+r(S_e/S_F)}}\right]^{1/n} \tag{6.67}$$

Step 3. The required filter area A is shown in Eq. (6.68).

$$A = Q/L = Q_F(r+1)/L \qquad (\text{ft}^2) \tag{6.68}$$

Step 4. Filter diameter is calculated from Eq. (6.66).

Example 6.4

Data presented in Table 6.3 for % BOD remaining vs. depth are obtained from a pilot-plant trickling filter treating an industrial wastewater.

1. Determine values of parameters n and K and write the correlation of BOD remaining to depth and hydraulic loading.
2. Calculate trickling filter diameter necessary to obtain BOD reduction of 80% if wastewater flow is 2.0 MGD. A filter depth of 20 ft is provided with recycle ratio of 0.3. Base calculations on an influent BOD of 300 mg/liter.
3. If no recycle is used for the filter designed in Part 2, calculate the maximum wastewater flow permissible in Mgal/day to reach the desired BOD reduction of 80%.

SOLUTION: Part 1 Follow procedure described in Section 7.9.2.

Step 1. Plot of $(S_e/S_o) \times 100$ vs. D for a set of four values of L. This graph is shown in Fig. 6.24.

Step 2. Table 6.4 presents the absolute values of the slopes read from Fig. 6.24 vs. L. Fig. 6.25 is a plot of these values. From Fig. 6.25 obtain $n = 0.478$.

Step 3.

1. Values of $L^n = L^{0.478}$ for each experimental run are presented in Table 6.5.
2. Matrix for values of $D/L^n = D/L^{0.478}$ for each experimental run at each depth is shown in Table 6.6.
3. Values of $D/L^n = D/L^{0.478}$ vs. $(S_e/S_o) \times 100$ are presented in Table 6.7.

A summary of calculated values is given in Table 6.8.

Step 4. Plot $\log[(S_e/S_o)\times 100]$ vs. $D/L^{0.478}$ (Fig. 6.26). From the slope of the straight line in Fig. 6.26 one obtains $K = 0.109$. From Eq. (6.54) for $n = 0.478$ and $K = 0.109$,

$$S_e/S_o = e^{-0.109D/L^{0.478}}$$

which is the correlation of BOD remaining to depth and hydraulic loading.

SOLUTION: Part 2 Trickling filter diameter

Follow procedure described in Section 7.9.3 (Case 2).

Step 1. Calculate S_o [Eq. (6.56)].

Here,

$$S_e = (0.2)(300) = 60 \text{ mg/liter}$$

$$\therefore\ S_o = [300 + (0.3)(60)]/(1+0.3) = 244.6 \text{ mg/liter}$$

Step 2. Calculate L [Eq. (6.64)].

Here,

$$S_e/S_o = 60/244.6 = 0.245$$

$$\ln 0.245 = -1.405$$

and

$$n = 0.478$$

Therefore,

$$L = (-0.109\times 20/-1.405)^{1/0.478} = 2.506 \text{ gal/(min)(ft}^2\text{)}$$

or

$$2.506 \text{ gal/(min)(ft}^2\text{)} \times 60 \text{ min/hr} \times 24 \text{ hr/day} = 3609 \text{ gal/(day)(ft}^2\text{)}$$

Alternative calculation procedure [Eq. (6.67)]

Here, $S_e/S_F = 60/300 = 0.2$, $K = 0.109$, $n = 0.478$, $D = 20$ ft, and $r = 0.3$. Therefore,

$$\therefore\ L = \left[-\frac{0.109\times 20}{\ln[(0.2)(1+0.3)]/[1+(0.3\times 0.2)]}\right]^{1/0.478} = 2.506 \text{ gal/(min)(ft}^2\text{)}$$

Step 3. Determine the filter area [Eq. (6.68)].

$$A = 2{,}000{,}000(0.3+1)/3609 = 720 \text{ ft}^2$$

Step 4. Determine the filter diameter [Eq. (6.66)].

$$d = (720/0.785)^{1/2} = 30.3 \text{ ft}$$

SOLUTION: Part 3 No recycle

$$A = 720 \text{ ft}^2$$

$$S_e/S_o = 0.2$$

Step 1. Calculate the allowable hydraulic loading [Eq. (6.64)]. Here,

$$S_o = S_F = 300 \text{ mg/liter}$$

$$\therefore L = (-0.109 \times 20/\ln 0.2)^{1/0.478} = 1.89 \text{ gal/(min)(ft}^2\text{)}$$

or

$$1.89 \text{ gal/(min)(ft}^2\text{)} \times 60 \text{ min/hr} \times 24 \text{ hr/day} = 2722 \text{ gal/(day)(ft}^2\text{)}$$

Step 2. Determine the maximum wastewater flow [Eq. (6.65)].

$$Q_F = AL = 720 \text{ ft}^2 \times 2722 \text{ gal/(min)(ft}^2\text{)} = 1{,}960{,}000 \text{ gal/day}$$

or

$$Q_F = 1.960 \text{ MGD}$$

(was 2.0 MGD with $r = 0.3$).

7.10. DESIGN PROCEDURE WHEN EXPERIMENTAL DATA ARE NOT AVAILABLE

Experimental data necessary to determine parameters n and K by the procedure described in Section 7.9.2, i.e., a network of data as shown in Table 6.3, are often unavailable. It may not be feasible to conduct the tedious experimental work required to obtain this network of data. If this is the case parameter n is estimated from available data for different types of filter media (n depends *only* on the characteristics of packing). Parameter K, which is only a function of the nature of the wastewater, is also estimated from available data for wastewaters from several sources. Eckenfelder [4] presents a tabulation of K values for Surfpac filter media ($n = 0.5$) for several industrial wastewaters.

8. Anaerobic Treatment

8.1. INTRODUCTION

Anaerobic treatment is utilized for treatment of wastewaters as well as for digestion of sludges. Anaerobic treatment of wastewaters is described in this section. Anaerobic digestion of sludges is studied in Chapter 7 (Sections 2.6 and 2.7). The end products of anaerobic degradation are gases, mostly methane (CH_4), carbon dioxide (CO_2), and small quantities of hydrogen sulfide (H_2S) and hydrogen (H_2). The process comprises two stages: (1) acid fermentation and (2) methane fermentation.

In the acid fermentation stage, organic materials are broken down to organic acids, mainly acetic (CH_3COOH), propionic (CH_3CH_2COOH), and butyric ($CH_3CH_2CH_2COOH$). In the methane fermentation stage, "methane microorganisms" convert the longer chain acids to methane, carbon dioxide, and an acid having a shorter carbon chain. The acid molecule is repeatedly broken down in the same manner. The resulting acetic acid is directly converted to CO_2 and CH_4.

$$CH_3COOH \xrightarrow[\text{organisms}]{\text{methane}} CO_2 + CH_4$$

The methane fermentation stage controls the rate of anaerobic degradation. In Section 8.3.1 it is shown that a plot of $(S_o - S_e)/X_v t$ vs. S_e (Fig. 6.28) yields rate constant k. This is similar to the situation found in aerobic treatment [Eq. (5.18) and Fig. 5.5]. However, one straight line is obtained in aerobic treatment, whereas two result from anaerobic data. The rate constants can be evaluated from the slopes of these straight lines. From Fig. 6.28, the slope corresponding to the acid fermentation stage is much greater than that for the methane fermentation stage. If the two lines are plotted on the same scale as in Fig. 6.28, the one corresponding to the acid fermentation is nearly vertical by comparison with the other. This indicates that the methane fermentation stage controls the process rate. Therefore, for design purposes one should adopt the k value and other parameters evaluated from the methane fermentation stage. Since methane fermentation controls the process rate, it is important to maintain conditions of effective methane fermentation. Detention time for methane microorganisms must be adequate, or they are washed away from the system. Experimental data show that the required detention time varies from 2 to 20 days. Optimum pH range is 6.8–7.4.

Anaerobic treatment is relatively inexpensive because aeration equipment is not utilized. On the other hand, residence times required are much longer than for aerobic processes. Bad odors associated with anaerobic processes, due mainly to production of H_2S and mercaptans, may constitute a serious limitation, particularly in urban areas.

8.2. A QUANTITATIVE STUDY OF ANAEROBIC DEGRADATION OF AN ORGANIC WASTE

A quantitative study of anaerobic degradation of organic wastewaters has been made by Andrews [1], and results are summarized in Fig. 6.27.

The following observations may be made:

1. *pH.* At the beginning of anaerobic degradation (acid fermentation), the pH drops due to formation of organic acids. Since at a later stage (methane fermentation) these acids are broken down, pH increases. The rise starts after about 2 days, as depicted in Fig. 6.27.

2. *COD remaining.* This curve is relatively flat during the first 2 days of the acid fermentation stage, since at this time organic compounds are merely converted to soluble form; so there is no COD reduction. COD drops markedly after this initial stage.

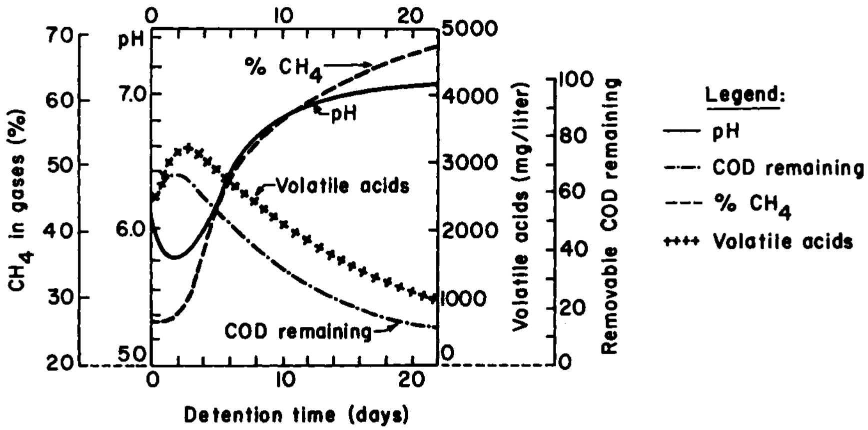

Fig. 6.27. Anaerobic degradation of an organic waste [1]. *(Data obtained with a batch reactor.)*

3. *Methane percentage and volatile acids.* There is a rapid increase in methane production with a corresponding decrease in volatile acids and an increase in pH after the first 2 days.

For long detention times, nearly all volatile acids are converted to CH_4 and CO_2. Since not all volatile solids are biodegradable, the curve for COD remaining approaches a limiting ordinate for large values of residence time.

8.3. MATHEMATICAL FORMULATION FOR ANAEROBIC DIGESTION PROCESS

The mathematical formulation described is the one proposed by Eckenfelder and associates [5].

8.3.1. BOD Removal Rate for Anaerobic Process

If BOD removal is assumed to follow first-order kinetics, one may utilize the formulation presented in Chapter 5, Section 3.2 for the aerobic process. Equation (6.69), which is similar to Eq. (5.18), is obtained.*

$$(S_o - S_e)/X_v t = kS_e \tag{6.69}$$

Owing to the presence of nonremovable BOD and following considerations similar to those in Chapter 5, Section 3.2, one may write a modified form of Eq. (6.69).

$$(S_o - S_e)/X_v t = k(S_e - S_n) \tag{6.70}$$

---

* Notation X_v is utilized in Eq. (6.69) instead of $X_{v,a}$ used in Eq. (5.18). Since subscript *a* refers to the aerator, it is omitted in the formulation of the anaerobic process.

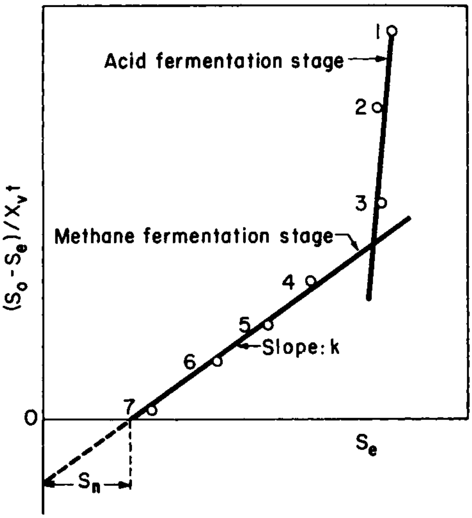

Fig. 6.28. Determination of BOD removal rate for anaerobic degradation.

Equation (6.70) is similar to Eq. (5.19). From Eq. (6.70) a plot of $(S_o - S_e)/X_v t$ vs. S_e yields two straight lines, as shown in Fig. 6.28. For design purposes, the value of k obtained from the slope of the line corresponding to the methane fermentation stage is the one adopted. In Fig. 6.28 as in Fig. 5.14, the abscissa at the origin of the ordinate axis corresponds to the nonbiodegradable substrate (S_n).

Figure 6.28 can be replotted by choosing as abscissa only the biodegradable portion of the effluent, S_e'.

$$S_e' = S_e - S_n \tag{6.71}$$

This corresponds to a translation of the axes indicated in Fig. 6.29, so that the straight line corresponding to methane fermentation passes through the origin.

The mathematical model corresponding to Fig. 6.29 is

$$(S_o - S_e)/X_v t = S_r/X_v t = kS_e' \tag{6.72}$$

If there is no nonbiodegradable matter, $S_n = 0$, and $S_e' = S_e$, then Eq. (6.72) reduces to Eq. (6.69).

8.3.2. Volatile Solids in the Anaerobic Reactor

Consider the anaerobic reactor depicted by Fig. 6.30. Let $X_{v,o}$ and X_v be the VSS concentrations in the influent and in the reactor, respectively. Assuming steady state operation and perfect mixing conditions, VSS concentration in the effluent is also equal to X_v. A material balance for VSS leads

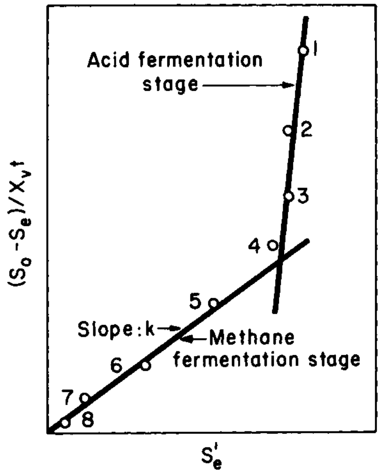

Fig. 6.29. Modified plot for determination of BOD removal rate for anaerobic degradation.

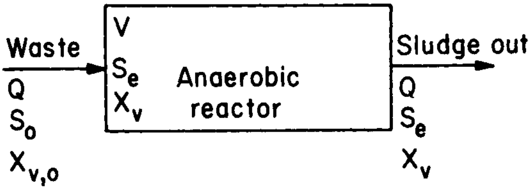

Fig. 6.30. Material balance for volatile solids in anaerobic reactor.

to an equation similar to Eq. (6.28) in Section 5.7, except for omission of subscript a in term X_v.

$$X_v = (X_{v,o} + aS_r)/(1 + bt) \tag{6.73}$$

Equation (6.73) is the expression for the concentration of VSS in the anaerobic reactor. If the concentration of VSS in the influent is negligible (i.e., $X_{v,o} \approx 0$), this equation is simplified to yield

$$X_v = aS_r/(1 + bt) \qquad (X_{v,o} \approx 0) \tag{6.74}$$

Parameters a and b in Eq. (6.73) are determined by writing it in linear form. By cross multiplying, dividing through by aX_v, and rearranging, Eq. (6.74) yields

$$(1/a)(1 - X_{v,o}/X_v) + (b/a)t = S_r/X_v \tag{6.75}$$

Equation (6.75) is the basis for determination of parameters a and b. A plot of $S_r/X_v = (S_o - S_e)/X_v$ vs. t yields a set of two straight lines: the first one, for lower t's, corresponding to the acid fermentation stage, and the second, for higher t's, corresponding to the methane fermentation stage. A typical plot is shown in Fig. 6.31. The determination of parameters a and b is indicated in this graph.

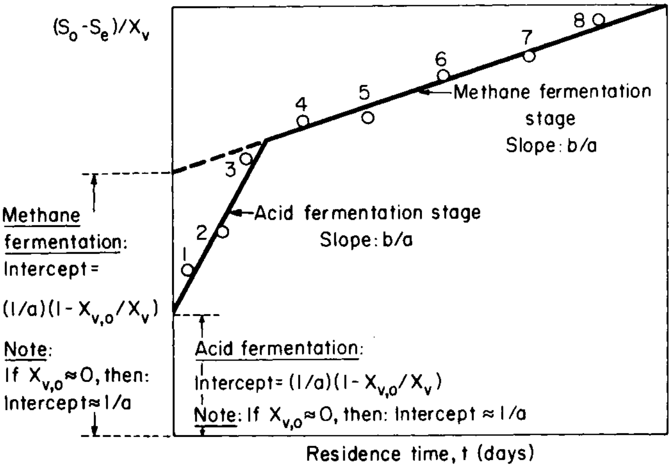

Fig. 6.31. Determination of parameters a and b.

When the concentration of VSS in the influent is negligible (i.e., $X_{v,o} \approx 0$) as shown in Fig. 6.35, Section 8.5, Eq. (6.75) yields Eq. (6.76).

$$1/a + (b/a)t = S_r/X_v \qquad (X_{v,o} \approx 0) \tag{6.76}$$

For calculation of detention time, the values of parameters a and b corresponding to the methane fermentation are the ones to be adopted since this is the controlling stage.

8.3.3. Calculation of Detention Time for Anaerobic Treatment

Consider Eq. (6.72) for BOD removal and Eq. (6.73) for VSS concentration. From Eq. (6.72),

$$t = S_r/kX_vS_e' \tag{6.77}$$

From Eq. (6.73),

$$S_r = [X_v(1+bt) - X_{v,o}]/a \tag{6.78}$$

Substituting S_r in Eq. (6.77) by its value from Eq. (6.78) and solving for t,

$$t = (X_v - X_{v,o})/(akX_vS_e' - bX_v) \tag{6.79}$$

In Eq. (6.79), S_e' is given by Eq. (6.71). When the concentration of VSS in the influent is negligible (i.e., $X_{v,o} \approx 0$), Eq. (6.79) is simplified to yield

$$t = 1/(akS_e' - b) \qquad (X_{v,o} \approx 0) \tag{6.80}$$

8.3.4. Gas Production in Anaerobic Treatment

Gas produced by anaerobic degradation is composed of CH_4, CO_2, and small quantities of H_2S and H_2. Lawrence and McCarthy [7] have shown that methane gas production can be estimated as 5.62 ft^3 (at STP)* per lb of COD removed, *excluding* the COD removed which is converted to cells. The proposed relationship to estimate gas production is then

$$G = 5.62(QS_r - 1.42QX_v) = 5.62Q(S_r - 1.42X_v) \qquad (6.81)$$

where G is the ft^3 of gas (at STP) produced per day; QS_r the lb COD removed/day; and $1.42QX_v$ the lb VSS in reactor effluent/day. [Ratio 1.42 corresponds to lb COD/lb VSS; see Chapter 5, Eq. (5.60).]

Attention should be given to the units in utilizing Eq. (6.81). If the following units are employed, where G is the ft^3 gas/day (at STP); Q the MGD; S_r the mg COD/liter (≈ lb COD/Mlb liquor); and X_v the mg VSS/liter (≈ lb VSS/Mlb liquor), then

$$G = 5.62\frac{\text{ft}^3}{\text{lb COD}}\left(Q\frac{\text{Mgal liquor}}{\text{day}} \times S_r\frac{\text{lb COD}}{\text{Mlb liquor}} \times 8.24\frac{\text{lb liquor}}{\text{gal liquor}}\right.$$
$$\left. - 1.42\frac{\text{lb COD}}{\text{lb VSS}} \times Q\frac{\text{Mgal liquor}}{\text{day}} \times X_v\frac{\text{lb VSS}}{\text{Mlb liquor}} \times 8.24\frac{\text{lb liquor}}{\text{gal liquor}}\right)$$

or finally [Eq. (6.82)],

$$G = 46.87Q(S_r - 1.42X_v)\ \text{ft}^3/\text{day} \quad (\text{STP}) \qquad (6.82)$$

8.4. LABORATORY ANAEROBIC REACTORS FOR OBTAINING BASIC DESIGN INFORMATION

Two models of a laboratory anaerobic reactor are shown in Figs. 6.32 and 6.33. The reactor in Fig. 6.32 is a batch type and the one in Fig. 6.33 is continuous.

The operational procedure followed to obtain basic design data is

1. To provide a seed of microorganisms, obtain a sample of settled sludge from a municipal wastewater treatment plant and place it in the laboratory reactor. Add the wastewater to be treated and maintain a temperature of 35°C throughout the test period.

2. If significant amounts of air are trapped in the reactor, purge the system with an inert gas.

3. The mixture is mixed continuously, either mechanically (Fig. 6.32) or by recirculation of the gas produced, by means of a gas pump (Fig. 6.33). Do not start feeding the wastewater or withdrawing the mixed liquor until gas production is ascertained.

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* STP stands for standard temperature and pressure, taken as $P = 1$ atm and $t = 60°\text{F}$.

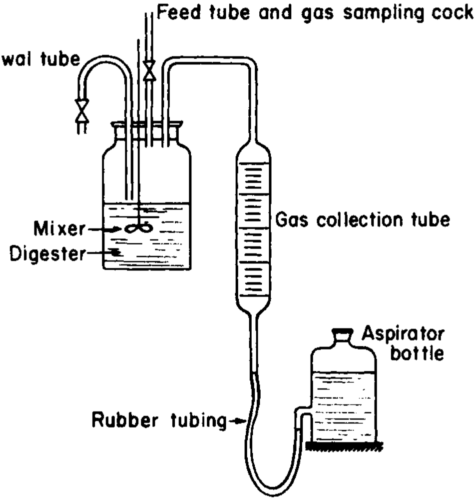

Fig. 6.32. Bench scale batch feed digester.

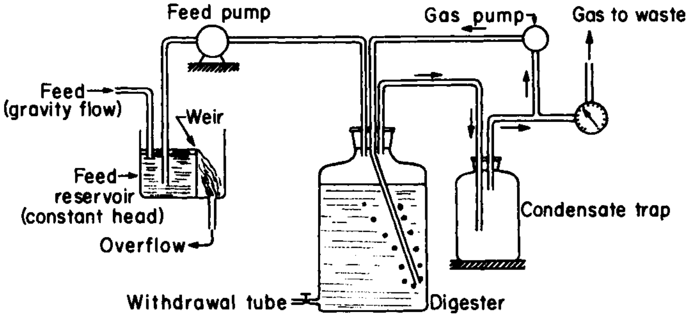

Fig. 6.33. Continuous feed digester.

4. Once gas production is noticed, feed portions of wastewater, withdrawing equal portions of mixed liquor.

5. During the entire start-up period pH is observed closely, and if it falls below 6.6, alkalinity is added to maintain the recommended range, i.e., pH 6.6–7.6.

6. Continue to feed and withdraw daily until effluent BOD (or COD) stabilizes. Analyze and record the data: raw wastewater: pH and COD (or BOD_5); effluent: pH, COD (or BOD_5), and VSS.

8.5. DESIGN PROCEDURE FOR ANAEROBIC DIGESTERS

Treatment of the data obtained in the laboratory to arrive at design parameters and utilization of this information in the design of an anaerobic digester are illustrated by Example 6.5.

Example 6.5

200,000 gal/day of a wastewater are to be treated by anaerobic degradation. The mean value for influent COD is 10,000 mg/liter. Effluent COD should be reduced to at the most 2000 mg/liter. Concentration of VSS in the influent is negligible.

A bench scale reactor is utilized to simulate the process, and MLVSS concentration is maintained at approximately 500 mg/liter in all experimental runs. Data obtained are presented in Table 6.9.

TABLE 6.9
Laboratory Data for Example 6.5

Run no.	Residence time (days)	Influent COD (mg/liter)	Effluent COD (mg/liter)
1	5	8,550	3,800
2	10	8,400	2,400
3	15	9,190	1,940
4	20	10,200	1,700
5	30	12,470	1,470

1. Obtain design parameters *k*, *a*, and *b*.
2. Calculate digester volume required (gal).
3. Calculate MLVSS concentration for the anaerobic digester (mg/liter).
4. Estimate gas production (ft^3/day at STP).

SOLUTION: Part 1

Step 1. Obtain values of $(S_o - S_e)/X_v$ and $(S_o - S_e)/X_v t$ necessary to plot Figs. 6.34 and 6.35. This is shown in Table 6.10.

Step 2. Prepare a plot of $(S_o - S_e)/X_v t$ vs. S_e and determine k from the slope of the line for the methane fermentation stage. This is shown in Fig. 6.34. Since all residence times considered are longer than 5 days, only the straight line corresponding to the methane fermentation stage is shown in Fig. 6.34. From the slope of this line one obtains $k = 0.0005$ day^{-1} and $S_n \approx 0$ (straight line passes through origin).

Step 3. Determine parameters a and b. Plot $(S_o - S_e)/X_v$ vs. t: This plot is shown in Fig. 6.35. As indicated in Step 2, only the straight line corresponding to the methane fermentation stage is obtained.

$$\text{Intercept} = 1/a = 7.0$$

$$\therefore a = 1/7 = 0.143$$

$$\text{Slope} = b/a = 0.5$$

$$\therefore b = a0.5 = 0.143 \times 0.5 = 0.0715$$

TABLE 6.10
Calculations for Example 6.5

Run no.	Residence time (days)	S_o (mg/liter) (COD)	S_e (mg/liter) (COD)	$S_o - S_e$ (mg/liter) (COD)	$(S_o - S_e)/X_v$ (lb COD/lb MLVSS)	$(S_o - S_e)/X_v t$ [lb COD/(day)(lb MLVSS)]
1	5	8,550	3,800	4,750	9.50	1.9
2	10	8,400	2,400	6,000	12.0	1.2
3	15	9,190	1,940	7,250	14.5	0.966
4	20	10,200	1,700	8,500	17.0	0.850
5	30	12,470	1,470	11,000	22.0	0.733

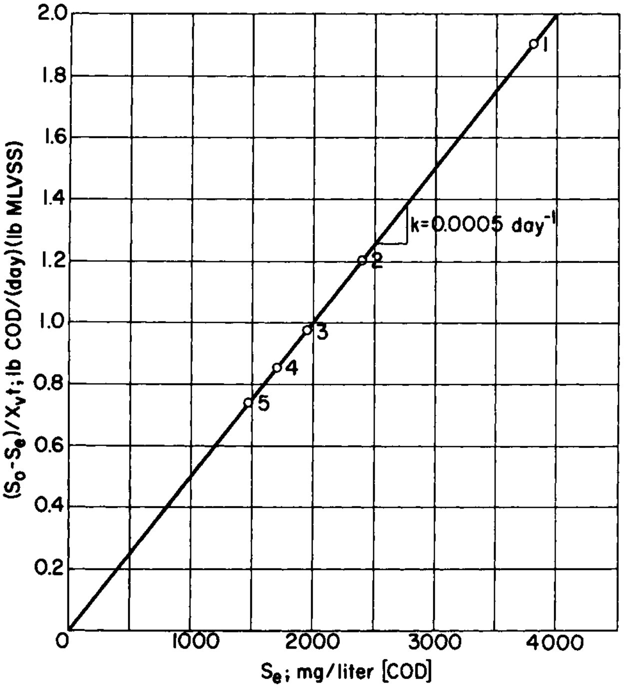

Fig. 6.34. Plot of $(S_0 - S_e)/X_v t$ vs. S_e for Example 6.5.

SOLUTION: Part 2 Digester volume (gal)

Step 1. Calculate the residence time from Eq. (6.80). Since $S_n \approx 0$, then $S_e' = S_e - S_n \approx S_e$.)

$$t = 1/(0.143 \times 0.0005 \times 2000 - 0.0715) = 14 \text{ days}$$

Step 2. Digester volume is then calculated.

$$V = Qt = 200{,}000 \text{ gal/day} \times 14 \text{ days} = 2{,}800{,}000 \text{ gal}$$

SOLUTION: Part 3 MLVSS concentration in anaerobic digester [Eq. (6 74)]

$$X_v = 0.143(10{,}000 - 2000)/(1 + 0.0715 \times 14) = 572 \text{ mg/liter}$$

SOLUTION: Part 4 Gas production from Eq. (6.82)
Since

$$Q = 200{,}000 \text{ gal/day} = 0.2 \text{ Mgal/day}$$

then

$$G = 46.87 \times 0.2[(10{,}000 - 2000) - 1.42 \times 572] = 67{,}380 \text{ ft}^3\text{/day at STP}$$

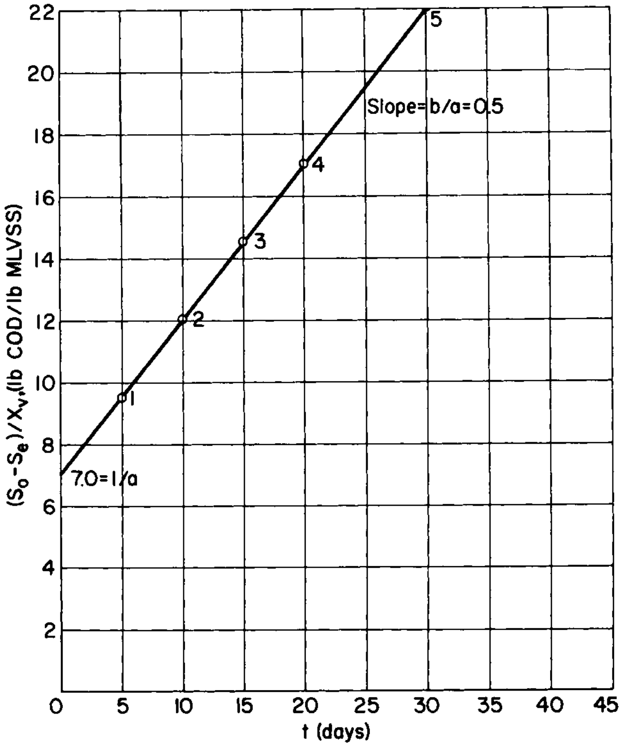

Fig. 6.35. Plot of $(S_o - S_e)/X_v$ vs. t for Example 6.5.

Problems

I. An aerated lagoon is being considered for treatment of an industrial wastewater flow of 1.0 MGD.

The following data are available:

Influent soluble BOD_5 = 200 mg/liter

Influent suspended solids concentration is negligible

Effluent soluble $BOD_5 \leqslant 20$ mg/liter

Wastewater temperature = 60°F

Since in most situations summer conditions control required aerator horsepower, base design upon an air temperature of 85°F

Take $h = 100$ BTU/(day)(ft^2)(°F)

Data obtained from laboratory simulation $k = 0.025$ day^{-1} (at 20°C), $a = 0.65$ lb VSS/lb BOD_r day, $b = 0.07$ day^{-1}, and $a' = 1.2$ lb O_2/lb BOD_5

Lagoon depth = 10 ft

Oxygen concentration to be maintained in liquid is 1.5 mg/liter

Determine

1. Lagoon surface in acres
2. Total effluent BOD (soluble + insoluble)
3. Oxygen requirements [lb O_2/(HP)(hr)] and total required HP and power level (HP/Mgal). Take $C_{SW} = 8$ mg/liter and $\alpha = 0.85$. Assume aerators are conservatively rated at $N_0 = 3.0$ lb O_2/(HP)(hr).

II. It is proposed to treat an industrial wastewater (flow, 1.0 MGD) utilizing stabilization ponds. Plan the operation in three stages. Influent BOD_5 (2000 mg/liter) should be reduced to a value not higher than 40 mg/liter (effluent from third stage).

Stage 1—Anaerobic pond. It should be designed for the purpose of reducing influent BOD_5(2000 mg/liter) to a value of 500 mg/liter in the effluent. Take $K = 0.1\ \text{day}^{-1}$. Plan on a 10-ft depth. Calculate the area of the pond in acres. Calculate surface loading in lb BOD_5/(acre)(day).

Stages 2 and 3—Facultative ponds. Design stages 2 and 3 for same residence time in each. Take $K = 0.2\ \text{day}^{-1}$ for both stages. For stage 2, the design is to be based on a surface loading of 500 lb BOD_5/(acre)(day). Calculate the area in acres and the depth in ft. Estimate depth of oxygen penetration in ft. For stage 3, the design is to be based on a surface loading of 250 lb BOD_5/(acre)(day). Calculate the area in acres and the depth in ft. Estimate depth of oxygen penetration in ft.

III. The data tabulated below for % BOD_5 remaining (i.e., $S_e/S_o \times 100$) were obtained in a pilot-plant trickling filter treating an industrial wastewater.

D, depth (ft)	Hydraulic loading [gal/(min)(ft^2)]			
	1	2	3	4
6	62	70	75	80
12	36	46	57	63
18	24	32	43	48.5
24	14	22	32	39

1. Correlate the data and develop a relationship between BOD removal, depth, and hydraulic loading.

2. Design a filter to obtain 80% BOD reduction from 5 MGD of settled wastewater with an initial BOD_5 of 250 mg/liter, using a depth of 30 ft (a) without recycle; (b) with recycle ratio of 1.5.

IV. A wastewater for which the mean influent COD may be taken as 10,000 mg/liter is to be treated by anaerobic degradation. Laboratory tests indicated that the nondegradable COD is approximately 2000 mg/liter. Average MLVSS influent concentration is 15 mg/liter. Bench scale determinations yield the following values for design parameters: $k = 0.0004\ \text{day}^{-1}$, $a = 0.14$ lb MLVSS yield/lb COD removed, and $b = 0.02$ lb MLVSS oxidized/(day)(lb MLVSS in reactor). Calculate % COD removal for residence times of 80, 60, 40, and 20 days, and plot % removal vs. residence time. On the same graph plot the corresponding values of % COD removal, neglecting the MLVSS in the influent.

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7

Sludge Treatment and Disposal

1. Introduction

Most primary treatment processes (Chapter 3) as well as secondary treatment sequences (Chapters 5 and 6) yield sludges which must be disposed of in some adequate way. Sludges resulting solely from solid–liquid separation processes (sedimentation, flotation) are referred to hence as primary sludges, and those resulting from biological processes are designated as secondary sludges.

One possibility in the treatment sequence is reduction of the total amount of organic and volatile content by submitting the sludge to digestion. Aerobic and anaerobic digestion of sludges are discussed in Section 2.

Another approach is to increase the percentage of solid contents in the sludge before final disposal by a sequence of processes which fall under the headings of thickening and dewatering, studied in Sections 3–7. Figure 7.1 illustrates the increase in solid content which might be expected in such treatment sequences.

Special treatment preceding dewatering becomes necessary for certain sludges which are difficult to dewater. These include chemical coagulation and heat-treatment processes, described in Section 8.

The end of the treatment sequence involves disposal of the remaining sludge, which is discussed in Section 9. Sludge disposal methods fall into two schemes involving either land disposal or incineration. All these alternatives are indicated in Fig. 7.1, which is the overall plan of study for this chapter.

2. Aerobic and Anaerobic Digestion of Sludges

2.1. INTRODUCTION TO AEROBIC DIGESTION OF SLUDGES

Aerobic digestion is a process in which a mixture of primary digestible sludge from primary clarification and activated sludge from aerobic biological treatment is aerated for an extended period of time. This results in cellular destruction with a decrease of volatile suspended solids (VSS).

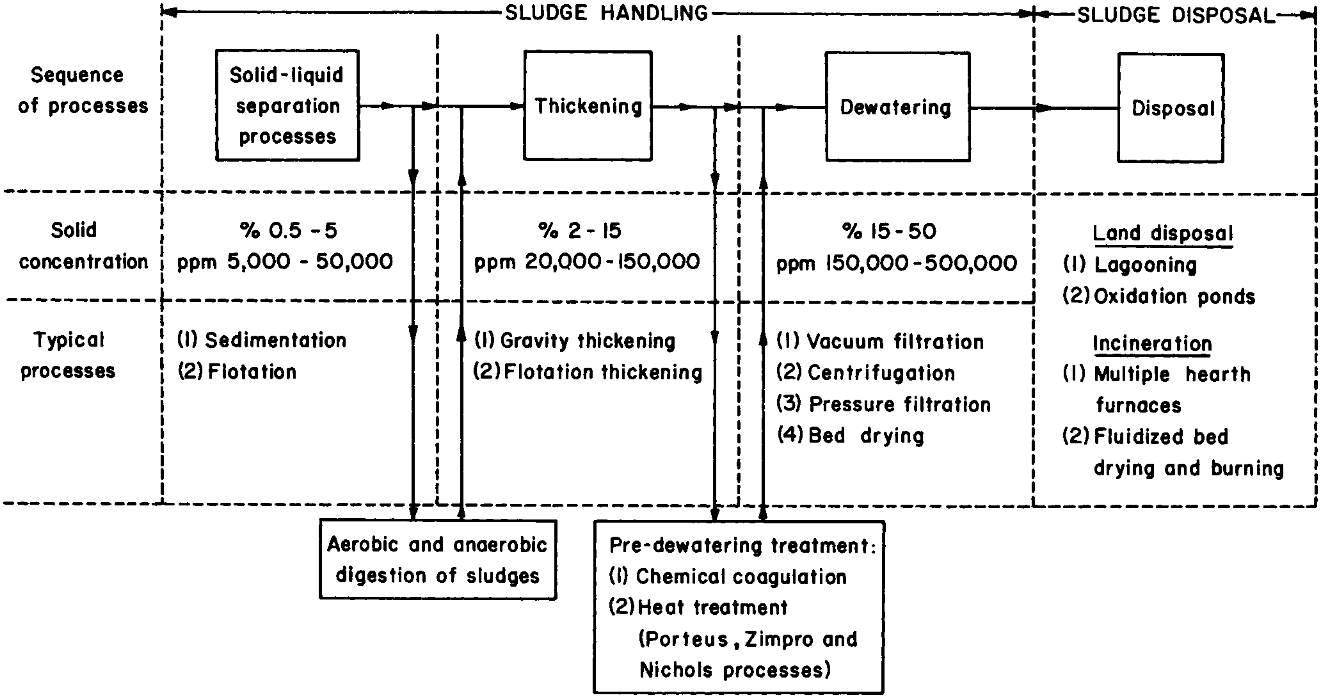

Fig. 7.1. Flow diagram (alternatives) for sludge treatment and disposal

The purpose of aerobic digestion is to reduce the amount of sludge which is to be disposed of subsequently. This reduction results from conversion by oxidation of a substantial part of the sludge into volatile products (CO_2, NH_3, H_2). If bacterial cells are represented by the formula $C_5H_7NO_2$, oxidation taking place in aerobic digestion is given by Eq. (5.51). This oxidation occurs when the substrate in an aerobic system is insufficient for energy and synthesis. It corresponds to the endogenous respiration phase (Fig. 5.3).

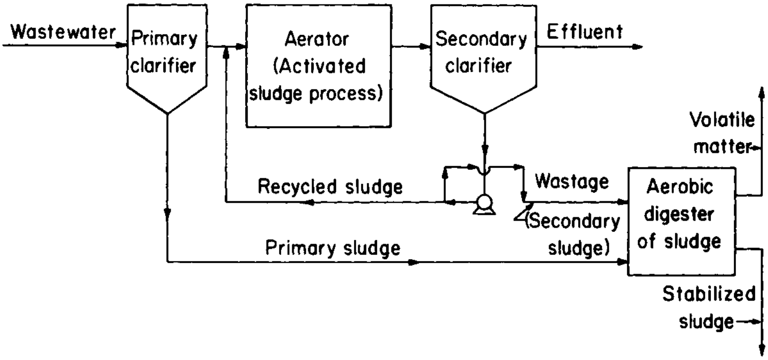

Fig. 7.2. Flow diagram showing aerobic digester of sludge.

Figure 7.2 shows a flow diagram of a continuous aerobic digester for the treatment sequence involving primary sedimentation, the activated sludge process, and aerobic digestion of sludge. When there is a small amount of sludge to be digested, batch operation is utilized with intermittent discharge of digested sludge. The aerobic digester shown in Fig. 7.2 handles a mixture of primary and secondary sludge.

Destruction rate of cells usually decreases when the food to microorganism ratio (F/M) increases. Therefore, the greater the proportion of primary sludge utilized in the process, the slower is the digestion because primary sludge has relatively high BOD (high F) and low MLVSS content (low M), thus leading to high F/M ratios.

2.2. SCHEMATIC REPRESENTATION OF AEROBIC BIOLOGICAL TREATMENT OF SLUDGES

Figure 5.3 shows two variables (mass of activated sludge and remaining soluble BOD) plotted vs. aeration time. The curve for remaining soluble BOD becomes nearly flat as the mass of MLVSS reaches its maximum. Since aerobic digestion of sludge occurs in the endogenous respiration phase region, there is essentially no soluble BOD removal. The fundamental objective of aerobic digestion is reduction of the mass of sludge for disposal, and not removal of soluble BOD.

2.3. CONCEPT OF SLUDGE AGE FOR THE CASE OF SLUDGE DIGESTERS

The concept of sludge age is discussed in Chapter 5, Section 9 and is given by Eq. (5.145), in which the denominator is the net output of VSS from the system. For all cases discussed so far, concentration of sludge in the reactor effluent was greater than that in the influent. As a result, there was a positive net output of VSS for the system. For sludge digesters, however, there is less sludge leaving than entering, the difference corresponding to sludge which is biodegraded. As a result, the denominator in Eq. (5.145) is negative, corresponding to a net input of VSS to the system. Therefore, for sludge digesters this equation is rewritten as

$$t_S = \frac{\text{lb MLVSS in the digester}}{\text{net input of VSS to the system} \quad (\text{lb/day})} > 0 \qquad (7.1)$$

From Eq. (7.1), sludge age corresponds to the average length of time the net input of sludge is subjected to digestion. A numerical example will clarify this concept (values from Example 7.1): 5000 lb/day of sludge are fed to an aerobic digester. Laboratory tests indicate that 55% of this sludge is nondegradable. This amounts to (0.55)(5000) = 2750 lb. Thus, the degradable portion is 5000 − 2750 = 2250 lb.

The aerobic digester is designed for the purpose of oxidizing 90% of this degradable VSS. The VSS which is not oxidized is 10% of 2250 lb or (0.10)(2250) = 225 lb. Assume that the digester is continuous, operating at steady state conditions, and that the concentration of VSS is maintained at 4000 mg/liter. Digester volume is 1,185,000 gal. These values are shown in Fig. 7.3(a).

Thus, the total output of sludge is

$$2750 + 225 = 2975 \text{ lb/day}$$

Since these 2975 lb/day enter and leave the digester without any change, the net sludge input (sludge which is actually oxidized in the digester) is

$$5000 - 2975 = 2025 \text{ lb/day}$$

Therefore, Fig. 7.3(a) is redrawn [Fig. 7.3(b)] showing the net input. Sludge age t_s from Eq. (7.1) is derived from Eq. (7.2).

$$t_S \ (\text{days}) = X_{v,a}V/(\text{sludge input} - \text{sludge output}) = X_{v,a}V/(\Delta X)_{\text{net}} \qquad (7.2)$$

Since $X_{v,a} = 4000$ mg/liter $= 4000 \times 10^{-6}$ lb sludge/lb liquor, then

$$t_S \ (\text{days}) = \frac{4000 \times 10^{-6} \text{ lb sludge/lb liquor} \times 8.34 \text{ lb liquor/gal liquor} \times 1{,}185{,}000 \text{ gal liquor}}{2025 \text{ lb sludge/day}}$$

$$= 19.5 \text{ days}$$

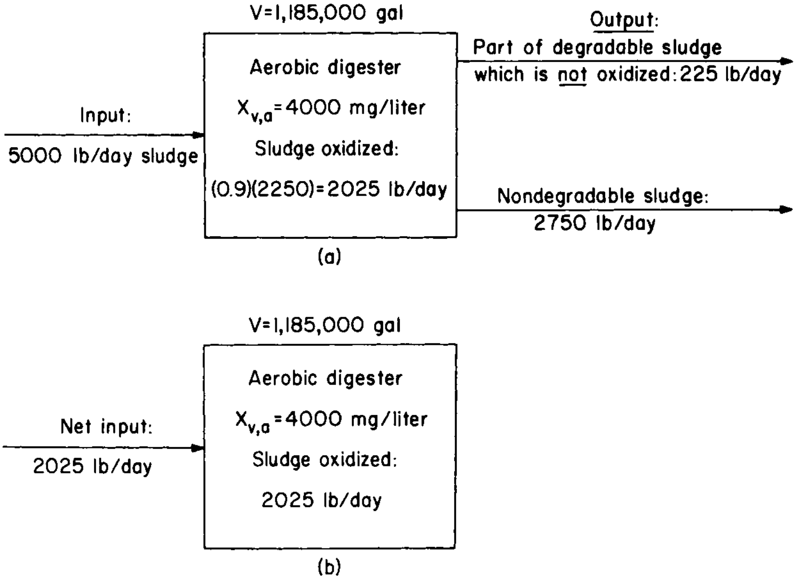

Fig. 7.3. Diagrams to illustrate concept of sludge age.

For higher values of sludge age corresponding to a larger volume for the aerobic digester, it is possible to oxidize over 90% of the degradable VSS. This means less sludge for disposal, though at the expense of a higher investment for the digester. An economic study can be made to determine the optimum fraction of degradable solids to be oxidized in the digester.

2.4. LABORATORY SCALE BATCH REACTOR TO OBTAIN BASIC DESIGN DATA FOR AEROBIC DIGESTERS

A typical batch laboratory scale reactor utilized for obtaining design data for aerobic digesters is shown in Fig. 5.2. Even when designing a continuous aerobic digester as shown in Fig. 7.2, the laboratory scale batch reactor is needed to obtain design information, because residence times required for aerobic digestion of sludges are high. Required flow rates for continuous operation are impractically small for the low volume laboratory reactor and cannot be measured with reasonable accuracy. Therefore, continuous laboratory scale reactors as in Fig. 5.10 are not recommended.

A continuous digester operates, assuming steady state conditions, with a constant concentration of suspended solids. For the batch laboratory scale unit utilized to simulate the plant scale continuous reactor, concentration of suspended solids decreases with time because of the gradual oxidation of VSS.

For extrapolation of batch laboratory scale data to a plant scale continuous reactor, one should work with an average value of the VSS concentration to simulate the prevailing constant VSS concentration in the continuous digester operating at steady state conditions. Similarly, the oxygen uptake rate for the batch reactor decreases with time. An average value is taken to simulate the prevailing constant oxygen uptake rate in the continuous digester operating at steady state conditions. The calculation technique for utilizing batch reactor data in designing a continuous digester is illustrated in Example 7.1.

Fundamental design information obtained from the batch laboratory scale reactor is (1) suspended solids vs. sludge age (days) and (2) oxygen uptake rate [mg/(liter)(hr)] vs. sludge age (days). Several units each with a capacity of about 2 liters are used simultaneously, and results obtained are averaged. The units are filled with sludge with an initial concentration which encompasses the range which might be expected in the proposed digester unit. Samples are withdrawn at selected time intervals, and VSS concentrations and oxygen uptake rates are determined.

Calculated design values are (1) required residence time (and volume of the digester) and (2) oxygen requirements (lb/hr) from which needed HP is calculated and aerators specified.

2.5. DESIGN PROCEDURE FOR AEROBIC DIGESTERS OF SLUDGE [1]

The procedure for utilizing laboratory data for this design is presented in Example 7.1. Design for a continuous digester is illustrated from data obtained in a laboratory batch reactor. For the design of a batch reactor the procedure is simpler, because the laboratory scale batch unit is an actual miniature of the plant scale digester.

The rate of degradation of sludge is temperature dependent. For a conservative design, laboratory data should be obtained at the lowest temperature anticipated in the field.

Example 7.1

Data in Table 7.1 were obtained from a laboratory batch reactor. 5000 lb/day of sludge are to be digested and it is desired to destroy 90% of degradable VSS. Assume an operating steady state concentration of 4000 mg/liter of VSS for the continuous digester to be designed. Calculate (1) digester volume (gal) and (2) oxygen requirement (lb/day).

SOLUTION: Part 1 Digester volume (gal)

Step 1. Plot VSS (mg/liter) [column (*2*), Table 7.1] vs. time of aeration [column (*1*), Table 7.1], as shown in Fig. 7.4. The curve is asymptotic to a VSS value estimated at 3050 mg/liter, which corresponds to nondegradable VSS in the sludge.

TABLE 7.1
Data for Example 7.1

(*1*) Time of aeration (days)	(*2*) VSS (mg/liter)	(*3*) R_r, oxygen uptake rate [mg/(liter)(hr)]
0	5550	—
1	5200	35.0
2	4950	28.0
5	4420	19.0
7	4170	16.0
10	3870	12.5
15	3500	8.8
20	3260	6.1
25	3200	4.2

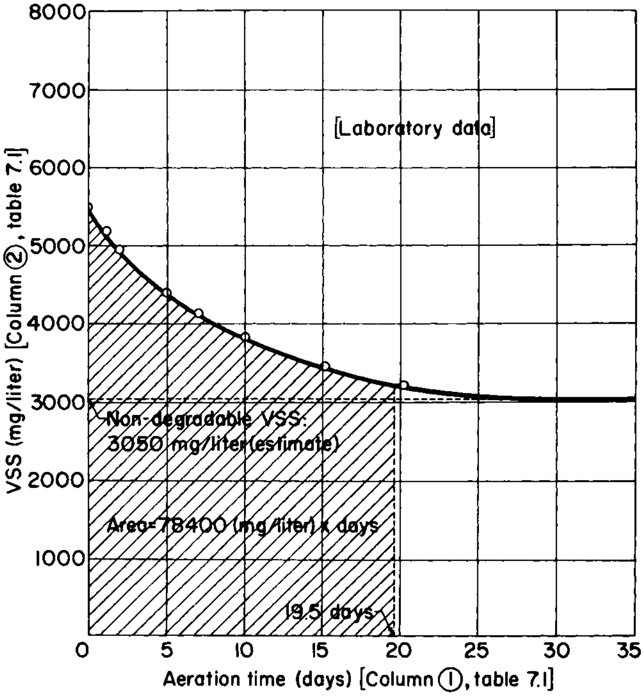

Fig. 7.4. VSS concentration vs. aeration time (Example 7.1).

Step 2. Obtain the oxidizable VSS remaining at any time by subtracting 3050 mg/liter from the values in column (*2*) of Table 7.1. The result is shown in Table 7.2 [column (*2*)].

TABLE 7.2
Degradable VSS Remaining vs. Time of Aeration (Example 7.1)

(*1*) Time of aeration (days)	(*2*) Degradable VSS remaining (mg/liter)
0	2500
1	2150
2	1900
5	1370
7	1120
10	820
15	450
20	210
25	150

Fraction of nondegradable sludge is 3050/5550 = 0.55 (or 55.0%), and that of degradable sludge is 1 − 0.55 = 0.45 (or 45.0%). At time zero, concentration of degradable sludge is 2500 mg/liter [first entry in column (*2*), Table 7.2]. It is desired to oxidize 90% of this sludge, which means that the degradable VSS remaining is 10% of 2500 mg/liter, or (0.1)(2500) = 250 mg/liter.

Step 3. Determine sludge age (days) for oxidizable VSS remaining to be 250 mg/liter. For convenient interpolation plot column (*2*) vs. column (*1*) (of Table 7.2) on semilog paper, as shown in Fig. 7.5. For an ordinate of 250 mg/liter, a sludge age of 19.5 days is read.

Step 4. From information obtained in the previous steps (utilizing laboratory data only), the material balance for the plant scale aerobic digester is written as follows:

Sludge IN: 5000 lb/day
Sludge OUT:
  Nondegradable VSS: (0.55)(5000) = 2750
  Fraction of degradable sludge which is *not* oxidized: degradable sludge = 5000 − 2750 = 2250 lb/day
  Fraction not oxidized: (0.1)(2250) = 225
  Total sludge output: 2975 lb/day

This material balance is shown in Fig. 7.3, which was used to illustrate the concept of sludge age.

Step 5. Calculate the digester volume from Eq. (7.2), where

$$\Delta X_{\text{net}} = (\text{sludge input} - \text{sludge output}) = 5000 - 2975 = 2025 \text{ lb/day}$$

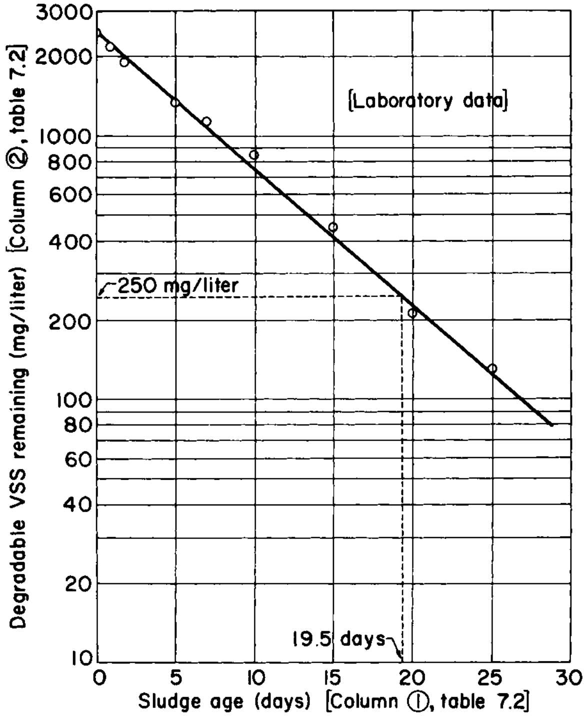

Fig. 7.5. Determination of sludge age (Example 7.1).

where $X_{v,a}$ = 4000 mg/liter and t = 19.5 days.

Therefore,

$$V = \frac{19.5 \text{ days} \times 2025 \text{ lb sludge/day}}{4000 \times 10^{-6} \text{ lb sludge/lb liquor} \times 8.34 \text{ lb liquor/gal liquor}} = 1{,}185{,}000 \text{ gal}$$

SOLUTION: Part 2 Oxygen requirement (lb/day)

Before making an estimate of oxygen requirements for the continuous digester, calculate average values for VSS concentration and oxygen uptake rate for the laboratory unit over a period of 19.5 days.

Step 1. Calculate the average VSS concentration for a laboratory unit. Average VSS concentration is calculated from Fig. 7.4. First determine by graphical integration the area bound by the curve from the abscissa zero to 19.5 days.

$$\text{Area} = \int_{t=0}^{t=19.5\text{days}} (\text{VSS concentration})\,dt = 78{,}400 \text{ (mg/liter)} \times \text{day}$$

Average VSS concentration is then

$$\text{Average VSS concentration} = [1/(t-0)]\int_{t=0}^{t=19.5\text{days}} (\text{VSS concentration})\,dt$$

$$= (1/t)\int_{t=0}^{t=19.5\text{days}} (\text{VSS concentration})\,dt$$

$$= (1/19.5)\,78{,}400 = 4020\ \text{mg/liter}$$

Step 2. Calculate the average oxygen uptake rate for a laboratory unit. Oxygen uptake rate [column (*3*) of Table 7.1] is plotted vs. time of aeration [column (*1*) of Table 7.1]. This is shown in Fig. 7.6. Determine by graphical integration the area bound by the curve from the abscissa zero to 19.5 days.

$$\text{Area} = \int_{t=0}^{t=19.5\text{days}} R_r\,dt = 303\ \text{mg O}_2/(\text{liter})(\text{hr}) \times \text{days}$$

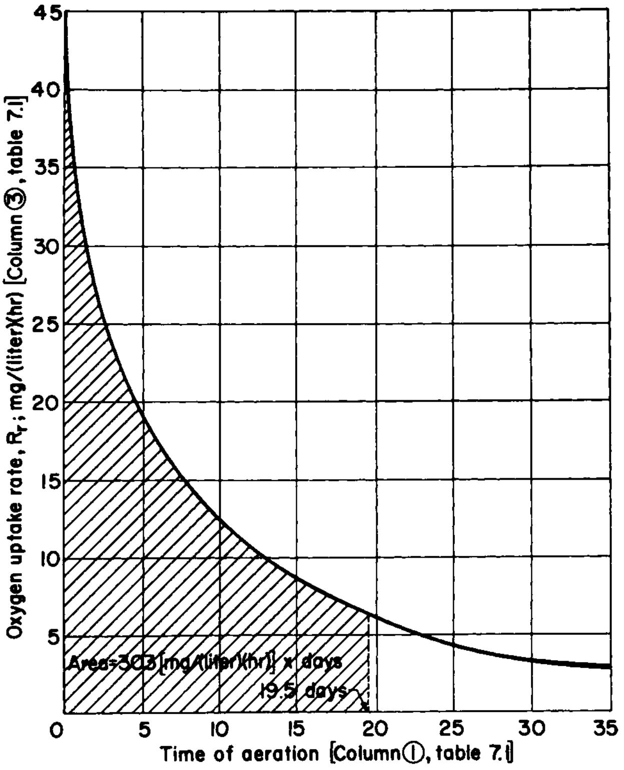

Fig. 7.6. Estimate of average oxygen uptake rate (Example 7.1).

or

$$303 \frac{\text{mg O}_2}{(\text{liter})(\text{hr})} \times \text{days} \times 24 \text{ hr/day} = 7272 \text{ mg O}_2\text{/liter}$$

Average daily oxygen uptake rate is then

$$\text{Average } R_r = [1/(t-0)] \int_{t=0}^{t=19.5\text{days}} R_r\,dt = (1/t) \int_{t=0}^{t=19.5\text{days}} R_r\,dt$$

$$= 1/19.5 \text{ days} \times 7272 \text{ mg O}_2\text{/liter}$$

$$= 373 \text{ mg O}_2\text{/(liter)(day)}$$

Step 3. Estimate oxygen requirements for a continuous aerobic digester. Average oxygen utilization rate for the laboratory batch unit is

$$373 \text{ mg O}_2\text{/(liter)(day)}$$

$$= 373 \times 10^{-6} \text{ lb O}_2\text{/(lb liquor)(day)}$$

$$= 373 \times 10^{-6} \text{ lb O}_2\text{/(lb liquor)(day)} \times 8.34 \text{ lb liquor/gal liquor}$$

$$= 3.11 \times 10^{-3} \text{ lb O}_2\text{/(gal liquor)(day)}$$

Calculated average VSS concentration for the laboratory batch unit is 4020 mg/liter (Solution, Part 2, Step 1), whereas VSS concentration at steady state conditions for the continuous aerobic digester is 4000 mg/liter. Oxygen requirements for the continuous aerobic reactor are estimated assuming proportionality between oxygen utilization and VSS concentration.

$$3.11 \times 10^{-3} \text{ lb O}_2\text{/(gal liquor)(day)} \times 1.185 \times 10^6 \text{ gal liquor} \times 4000/4020$$

$$= 3667 \text{ lb O}_2\text{/day}$$

where 4000/4020 is the correction factor required for translating data obtained with the laboratory batch reactor to actual operational conditions with the continuous reactor. Horsepower requirements and layout for aerators are obtained by the procedure discussed in Chapter 4, Sections 14–16.

2.6. INTRODUCTION TO ANAEROBIC SLUDGE DIGESTION

It has been known for at least a century that if settled sewage solids are kept in a closed tank for a period of time, they are converted to a liquid state and a combustible gas containing methane is generated. A patent was issued to Imhoff in 1904 for the design of anaerobic digestion vessels, which are known as Imhoff tanks. Most sludge digestion processes in use today are anaerobic, although aerobic digestion is increasingly more popular, especially for small units.

Anaerobic sludge digesters are usually of two types: (1) single-stage digesters and (2) two-stage digesters. A typical single-stage sludge digester is shown in Fig. 7.7.

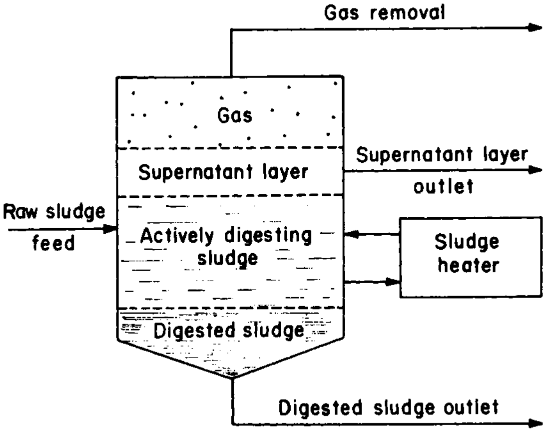

Fig. 7.7. Single-stage anaerobic sludge digester.

Raw sludge is fed to the zone where the sludge is actively digesting and gas is produced. As the gas rises, it lifts sludge particles and other materials (grease, oil, fats), forming a supernatant layer which is drawn off from the digester. Digested sludge is withdrawn from the bottom of the tank.

The digestion process is favored by high temperature (usually from 85°–105°F), so digesting sludge is heated either by steam coils within the vessel or by means of an external sludge heater (Fig. 7.7). Gas is removed from the top of the digester and often utilized as fuel, owing to its high content of methane. The purpose of the two-stage unit is fundamentally to provide a better volume

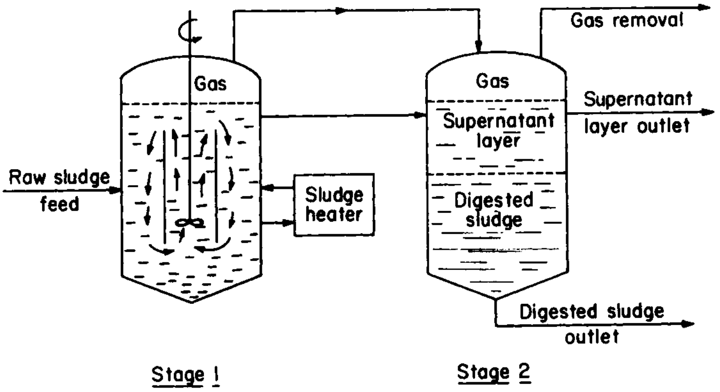

Fig. 7.8. Two-stage anaerobic sludge digester.

utilization. Volume utilization for the single-stage unit is poor, owing to stratification and poor mixing. Stage 1 is used for digestion. It is heated and mixing is provided either mechanically or by means of gas recirculation. Stage 2 is used for storage and separation of digested sludge and the supernatant layer.

2.7. SIZING OF ANAEROBIC SLUDGE DIGESTERS

Laboratory simulation for anaerobic digestion of sludges can be made by the same procedure described for aerobic digesters, except that aeration is not provided. Curves similar to the one shown in Fig. 7.4 are obtained from laboratory simulation, and the digester volume is calculated by a procedure similar to that for aerobic digesters (Example 7.1).

Empirical methods are usually employed for sizing anaerobic sludge digesters. Some of these methods are based on the population served by the sewage system (for domestic sludge) or on recommended values of sludge age. The latter values are temperature dependent, decreasing with increase of operating temperature. Table 7.3 presents suggested sludge age values as a function of temperature. Volume (SCFM) of gas generated is estimated from Eq. (6.82).

TABLE 7.3
Recommended Values of Sludge Age for Anaerobic Sludge Digester [3]

Temperature (°F)	Suggested sludge age t_S (days)
65	28
75	20
85	14
95	10
105	10

3. Thickening of Sludges

3.1. INTRODUCTION

Thickening is the usual first step in sludge disposal processing. It can be done (1) by gravity and (2) by dissolved air flotation.

3.2. ADVANTAGES OF THICKENING

1. It improves digester operation and reduces capital cost whenever sludge digestion is utilized.

2. It reduces sludge volume prior to land or sea disposal.
3. It increases economy of sludge dewatering systems (centrifuges, vacuum filters, pressure filters, etc.).

3.3. GRAVITY THICKENER

Thickeners are tanks of circular cross section provided with a rotating rake mechanism similar to the clarifiers discussed in Chapter 3, Section 3.7.

3.4. DESIGN PRINCIPLES FOR GRAVITY THICKENERS

The purpose of thickeners is to provide a concentrated sludge underflow. A fundamental parameter is thickener area required for a specific underflow concentration. This is expressed in terms of the unit area UA, defined as [Eq. (7.3)]

$$UA = \mathrm{ft}^2/(\mathrm{lb/day}) = (\mathrm{ft}^2)(\mathrm{day})/(\mathrm{lb}) \tag{7.3}$$

[i.e., the area (ft^2) required per lb/day of sludge in the influent]. The reciprocal of the unit area, which is termed mass loading (ML) is

$$ML = 1/UA = (\mathrm{lb/day})/\mathrm{ft}^2 = \mathrm{lb}/(\mathrm{day})(\mathrm{ft}^2) \tag{7.4}$$

The fundamental design problem consists of calculating the thickener area from the knowledge of (1) flow rate (Mgal/day) of sludge of initial concentration C_0 (mg/liter) and (2) desired final underflow concentration C_u (mg/liter). Once the value of UA (or ML) is determined, the area (ft^2) of the thickener is given in Eq. (7.5).

$$\text{Area} \quad (\mathrm{ft}^2) = (\text{lb/day of influent}) \times UA \qquad [\text{from Eq. (7.3)}] \tag{7.5}$$

or

$$\text{Area} \quad (\mathrm{ft}^2) = (\text{lb/day of influent}/ML) \qquad [\text{from Eq. (7.4)}] \tag{7.6}$$

The procedure discussed in Chapter 3, Section 3.6 may be utilized for design of thickeners. A specific method for thickener design based on an empirical equation proposed by Edde and Eckenfelder is described in Section 3.5.

3.5. EDDE AND ECKENFELDER'S EQUATION [2]

The degree to which sludges can be thickened depends on many variables. A correlation developed by Eckenfelder and Edde is

$$[(C_u/C_0)-1] = K/(ML)^n = K/(1/UA)^n \tag{7.7}$$

where C_u is the underflow concentration (mg/liter); C_0 the initial feed concentration (mg/liter); ML the mass loading [(lb/day)/ft^2]; UA the unit area [ft^2/(lb/day)]; and K and n constants. Constant n is only a function of the rheological properties of the sludge. Constant K is related not only to the initial concentration C_0 but also to the height of settling column.

Equation (7.7) is written in linear form by taking logarithms of both members.

$$\log[(C_u/C_0)-1] = \log K - n\log(ML) = \log K - n\log(1/UA) \qquad (7.8)$$

From Eq. (7.8) a plot on log–log paper of $[(C_u/C_0)-1]$ vs. $(1/UA)$ [or (ML)] yields a straight line of slope $= -n$ and intercept $= K$. A procedure for design of gravity thickeners starting from data obtained from bench scale units and based on Edde and Eckenfelder's equation is described by Eckenfelder and Ford [1].

3.6. FLOTATION THICKENING

Flotation thickening can be utilized for sludges and is specifically recommended for gelatinous ones such as those from the activated sludge process. Design of flotation equipment is discussed in Chapter 3, Section 4.

4. Dewatering of Sludges by Vacuum Filtration

4.1. INTRODUCTION

Vacuum filtration is the most widely used procedure for dewatering wastewater sludges. In vacuum filtration, water is removed under applied vacuum through a porous media which retains solids but allows liquids to pass through. Several types of media are used, such as nylon and dacron cloth, steel mesh, and tightly wound stainless steel coil springs.

The central unit is a rotary drum which revolves in a slurry tank (Fig. 7.9). Vacuum is applied to the submerged part of the drum and solids are retained

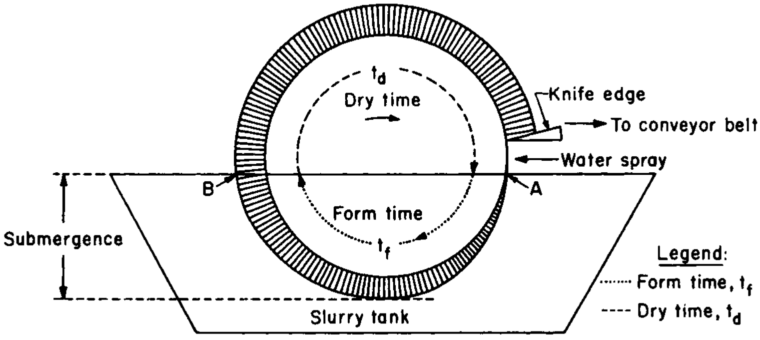

Fig. 7.9. Diagram of rotary filter.

on the drum surface. The cake starts building up at point A as the drum dives into the slurry tank, and reaches full thickness at point B as the drum emerges. Time elapsed from A to B (i.e., submergence time) is designated as form time (t_f). From B to A the cake is dewatered, and time elapsed is designated as dry time (t_d). The total cycle time (t_c) is

$$t_c = t_f + t_d \tag{7.9}$$

Since the drum revolves at a constant speed, the following relationship applies [Eq. (7.10)]:

$$t_f/t_c = t_f/(t_f + t_d) = (\%\ \text{submergence})/100 \tag{7.10}$$

Usually drum submergence between 10 and 60% is employed, i.e.,

$$0.6t_c > t_f > 0.1t_c \tag{7.11}$$

Therefore, form time is usually between 10 and 60% of total cycle time.

After being dried by liquid transfer to air drawn through the cake by the applied vacuum, the cake is removed by a knife edge onto a conveyor belt to disposal facilities. The filter media is then washed by a water spray prior to being immersed again in the slurry tank.

4.2. VARIABLES IN VACUUM FILTRATION

Variables to be considered fall into two groups: those related to sludge characteristics and those pertaining to filter operation. Variables related to sludge characteristics are as follows:

1. Solids concentration in the sludge. This is determined by the nature of the sludge (i.e., primary or secondary, domestic or industrial) and by thickening processes preceding the filtration step.
2. Viscosity of sludge and filtrate, the latter being approximately the same as that for water at similar temperatures.
3. Sludge compressibility, which is related to the nature of the sludge particles.
4. Chemical and physical composition, including particle size and shape, water content.

Filter operation variables are

1. *Operating vacuum.* Usually from 10 to 20 in. of mercury. Higher vacuums are more effective with incompressible cakes. For very compressible cakes, applying higher vacuums may result in clogging the filter media.
2. *Drum submergence.* This varies from 10 to 60%, high porosity sludges permitting higher submergences. Sludges of low porosity must be filtered with low submergence, otherwise the resulting compact and thick cake does not allow an adequate flow of filtrate.

3. *Sludge conditioning by chemical addition.* Many sludges require coagulants (e.g., $FeCl_3$, lime, polyelectrolytes) to coagulate smaller particles which might otherwise clog the filter media, resulting in reduction of filtration rate.

4. *Type and porosity of the filter media.* High porosity media results in higher filtration rates.

4.3. DEFINITION OF PARAMETER *c*

Filtration equations in Section 4.4 employ a parameter c, which is defined as the mass of solids deposited on the filter per unit volume of filtrate. An expression for c is derived as follows: Let c_i be the concentration of solids in the incoming slurry, expressed as grams of solids per ml of slurry. Slurry is filtered in a laboratory Büchner filter (Fig. 7.10). Furthermore, assume that for the filtrate the concentration of solids is reduced to a value indicated as c_f (also expressed as grams of solids per milliliter). Presumably $c_f \ll c_i$.

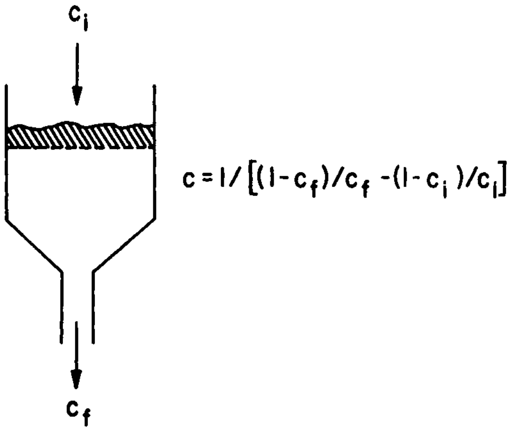

Fig. 7.10. Diagram to illustrate derivation of Eq. (7.12).

Since the slurry is relatively diluted, assume that the density is approximately that of water, i.e., 1 g per ml. Thus, 1 g of feed to the filter contains c_i gram of solid and $(1-c_i)$ gram of water. The ml of water per gram of solids in the incoming slurry are then $(1-c_i)/c_i$ (ml water/g solid). Similarly for the filtrate one has $(1-c_f)/c_f$ (ml water/g solid). The difference between these two values is

$$(1-c_f)/c_f - (1-c_i)/c_i \qquad \text{(ml water/g solid)}$$

The inverse of this quantity corresponds to parameter c, i.e., mass of solids deposited per unit volume of filtrate [Eq. (7.12)].

$$c = 1/[(1-c_f)/c_f - (1-c_i)/c_i] \tag{7.12}$$

4.4. FILTRATION EQUATIONS

The basic filtration equation derived from the Poiseuille and d'Arcy law is presented in Eq. (7.13).

$$dV/dt = \Delta P/(R_{\text{cake}} + R_{\text{media}}) \tag{7.13}$$

where V is the volume of filtrate; t the time; dV/dt the rate of filtration; ΔP the driving force = pressure difference. If expressed in psi ΔP is numerically equal to applied vacuum in gauge units, because $\Delta P = P_{\text{vacuum}} - P_{\text{atmospheric}}$. Hence, it is indicated as P = applied vacuum (psi); R_{cake} the resistance offered by cake to the flow of filtrate; and R_{media} the resistance offered by media to the flow of filtrate.

This equation is rewritten as [5]

$$dV/dt = PA^2/[\mu(rcV + R_m A)] \tag{7.14}$$

where V is the volume of filtrate; t the time; P the applied vacuum; A the filter area; c the parameter defined in Section 4.3, i.e., mass of solids deposited per unit volume of filtrate. Consequently product cV is the mass of cake (lb) corresponding to volume V of filtrate; r the specific resistance of cake to the flow of filtrate (i.e., resistance per lb of cake). Physical significance of r and its units is discussed in this section; R_m the initial resistance of the filter media. This resistance is usually negligible as compared to that developed by the filter cake; and μ the viscosity of the filtrate.

The physical significance of parameter r can be appreciated if in Eq. (7.14) media resistance R_m is neglected. Solving for r,

$$r = PA^2/[\mu cV(dV/dt)] \tag{7.15}$$

From Eq. (7.15) it follows that r is numerically equal to the pressure difference (applied vacuum P) required to produce a unit rate of filtrate flow (i.e., $dV/dt = 1.0$) through a unit mass of cake (i.e., $cV = 1.0$) and a unity filter area ($A = 1$), if filtrate viscosity is unity ($\mu = 1$, e.g., 1 cP) or $r = P$ if $dV/dt = 1.0$, $cV = 1.0$, $\mu = 1.0$, and $A = 1.0$. Thus, the specific resistance r measures the ability of the sludge to be filtered; the higher the value, the more difficult is the filtration.

Integration of Eq. (7.14) is usually performed assuming that specific resistance is constant throughout form time. From Eq. (7.14) if at $t = 0$, $V = 0$ and at $t = t$, $V = V$, integration of Eq. 7.14 yields

$$(\mu/A^2P)\int_0^V (rcV + R_m A)\, dV = \int_0^t dt$$

Assuming the specific resistance of cake to be constant,

$$(\mu/A^2P)\left(rc\int_0^V V\,dV + R_m A\int_0^V dV\right) = \int_0^t dt$$

or

$$(\mu/A^2P)[rc(V^2/2) + R_m AV] = t$$

Dividing both members by V and rearranging,

$$t/V = (\mu rc/2PA^2)V + \mu R_m/AP \tag{7.16}$$

From Eq. (7.16) it follows that a plot of t/V vs. V yields a straight line. Values of specific cake resistance r and media resistance R_m are evaluated from the slope and intercept of this line, respectively.

$$r = (2PA^2/\mu c)s \tag{7.17}$$

$$R_m = iAP/\mu \tag{7.18}$$

where s and i denote the slope and the intercept of the straight line. A typical plot of (t/V) vs. V is shown in Fig. 7.13. Specific resistance is primarily useful for comparing filtration characteristics of different sludges and determining optimum coagulant requirements to produce a cake offering a minimum resistance (Section 4.5).

4.5. LABORATORY DETERMINATION OF SPECIFIC RESISTANCE *r* AND OPTIMUM COAGULANT DOSAGE

Laboratory determination of specific resistance r is based on construction of a plot of t/V vs. V and calculation of r from Eq. (7.17). The laboratory equipment needed is an ordinary Büchner funnel apparatus (Fig. 7.11).

The procedure [1] is as follows:

1. Prepare the Büchner funnel and filter paper.
2. Wet filter paper and adjust vacuum to 15 or 20 in. of Hg.

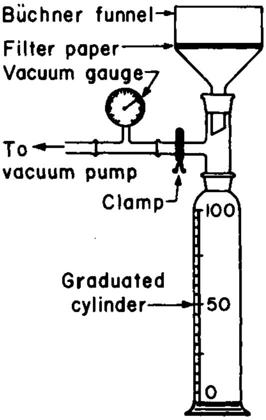

Fig. 7.11. Büchner funnel apparatus.

3. Record filtrate volumes at selected time intervals until the vacuum breaks. Vacuum exists as long as there is a pool of liquid over the cake. As soon as the liquid is drained off, air is sucked in and the vacuum breaks.

4. Measure solids content in incoming slurry and filtrate by evaporation and weighing. Let these be values c_i (initial concentration, g/ml) and c_f (concentration in filtrate, g/ml). Parameter c is then calculated from Eq. (7.12).

5. Calculate r from a plot of t/V vs. V utilizing Eq. (7.17).

6. Repeat Steps 1–5 using various concentrations of coagulant. Depending on the nature of the sludge, $FeCl_3$ and/or lime dosages are 2–10% by weight and polyelectrolyte dosages 0.1–1.5% by weight.

7. Compute specific resistance of all samples as indicated in Step 5. Determine optimum coagulant dosage from a plot of specific resistance vs. coagulant dosage. Optimum dosage corresponds to the minimum on the specific resistance curve (Fig. 7.12).

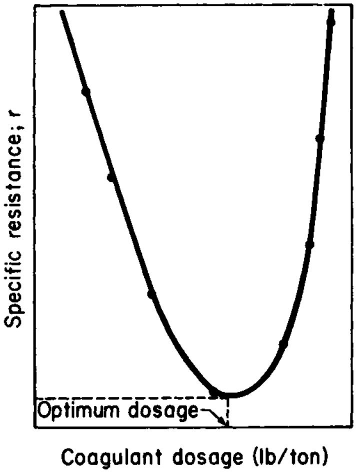

Fig. 7.12. Typical curve for the effect of coagulant dosage on sludge specific resistance.

4.6. UNITS FOR SPECIFIC RESISTANCE OF CAKE (*r*)

The following observations should be made:

1. Applied vacuum (in. Hg) is usually converted to g/cm^2 before utilization in Eq. (7.17). The conversion factor is

$$(\text{in. Hg})(34.5) = \text{g/cm}^2 \tag{7.19}$$

If applied vacuum is expressed in psi, the conversion factor is

$$(\text{psi})(70.1) = \text{g/cm}^2 \tag{7.20}$$

2. Specific resistance r is usually expressed in cm^2/g. From Fig. 7.13, units for slope s are (for $t = \text{sec}$, $V = \text{ml}$)

$$s = (t/V)/V = (\text{sec/ml})/\text{ml} \approx (\text{sec/cm}^3)/\text{cm}^3 = \text{sec/cm}^6$$

Units for c are

$$c = \text{g/ml} \approx \text{g/cm}^3$$

and those for μ are

$$\mu = \text{g/(cm)(sec)} = \text{Poise}$$

If P is expressed in g/cm^2 utilizing the conversion factors indicated in Eqs. (7.19) and (7.20), and the area is in cm^2, then from Eq. (7.17) units for r are

$$r = \frac{\text{g/cm}^2 \times \text{cm}^4}{\text{g/(cm)(sec)} \times \text{g/cm}^3} \times \text{sec/cm}^6 = \text{sec}^2/\text{g}$$

4.7. NUMERICAL EXAMPLE: DETERMINATION OF SPECIFIC CAKE RESISTANCE USING THE BÜCHNER FUNNEL

Example 7.2

The data tabulated below are obtained from a filtration laboratory test for an activated sludge utilizing a Büchner funnel apparatus.

Volume of filtrate (ml)	Time (sec)
25	48
50	150
75	308
100	520
125	788
150	1118

Operating vacuum is 20 in. of Hg and temperature is 25°C. Area of the filter is 500 cm^2. Assume that the filtrate has the properties of water at 25°C ($\mu = 0.8953$ cP). Take $c = 0.2\ \text{g/cm}^3$. Determine the specific cake resistance in sec^2/g.

Step 1. Set up the following tabulation from the data given above.

V (ml)	t (sec)	t/V (sec/ml)
25	48	1.92
50	150	3.00
75	308	4.11
100	520	5.20
125	788	6.30
150	1118	7.45

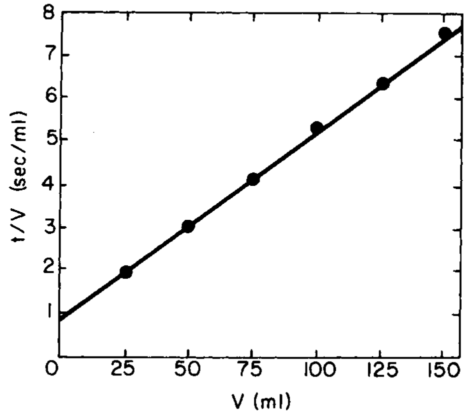

Fig. 7.13. Plot of t/V vs. V (Example 7.2).

Step 2. Prepare a plot of t/V vs. V. This is shown in Fig. 7.13.

Step 3. From Fig. 7.13,

$$\text{Slope} = s = (5.23-0.8)/(100-0.0) = 0.0443\ \text{sec/cm}^6$$

Since $P = (20)(34.5) = 690\ \text{g/cm}^2$, then r is calculated from Eq. (7.17).

$$r = \frac{2 \times 690\ \text{g/cm}^2 \times 500^2\ \text{cm}^4 \times 0.0443\ \text{sec/cm}^6}{0.008953\ \text{g/(cm)(sec)} \times 0.2\ \text{g/cm}^3} = 8.53 \times 10^9\ \text{sec}^2/\text{g}$$

4.8. SPECIFIC RESISTANCE FOR COMPRESSIBLE CAKES

Most industrial wastewater sludges form compressible cakes for which filtration rate and specific resistance are functions of the pressure difference across the cake. This effect is represented by Eq. (7.21).

$$r = r_o P^s \tag{7.21}$$

where s is the coefficient of compressibility. The larger is s, the more compressible is the sludge. When $s = 0$, the specific resistance is independent of pressure and the sludge is incompressible. Eq. (7.21) yields

$$r = r_o = \text{constant} \tag{7.22}$$

Some generalizations on filtration characteristics of sludges are as follows:

1. Ease of filterability of wastewater sludges decreases with degree of treatment.

$$\underset{\substack{\text{raw}\\\text{sewage}\\\text{sludge}}}{r} < \underset{\substack{\text{primary}\\\text{sludge}}}{r} < \underset{\substack{\text{secondary}\\\text{sludge}}}{r}$$

Raw sewage sludge is the easiest to filter (lower specific resistance), whereas secondary sludge is the most difficult.

2. Filterability is influenced by particle size, shape, and density, and by electrical charge on the particle. The larger the particle size, the higher the filtration rate (lower specific resistance), and thus the final cake moisture is lower. Addition of coagulants promotes agglomeration of particles, thus increasing filtration rate.

4.9. FILTRATION DESIGN EQUATION

For purpose of filter design it is convenient to modify Eq. (7.16). Neglecting resistance of media ($R_m \approx 0$) and recalling that t is the form time ($t = t_f$),

$$t_f/V = (\mu rc/2PA^2)V \tag{7.23}$$

It is convenient to write the equation in terms of filter loading (L_f), i.e., lb of deposited cake/(ft^2)(hr). Since product cV represents the weight of cake (Section 4.4), filter loading (L_f) based on form time (form loading) is

$$L_f = cV/At_f = \text{lb of deposited cake/(ft}^2\text{)(hr)} \tag{7.24}$$

Substituting in Eq. (7.23) r by its value given in Eq. (7.21) and rearranging,

$$V^2/A^2 = (2P^{1-s}t_f)/(\mu r_o c) \tag{7.25}$$

Substituting in the right-hand member of Eq. (7.25) the identities $t_f = t_f^2/t_f$ and $c = c^2/c$ and rearranging,

$$(cV/At_f)^2 = (2P^{1-s}c)/(\mu r_o t_f) \tag{7.26}$$

The left-hand member of Eq. (7.26) is the square of form loading [Eq. (7.24)]. Therefore,

$$L_f = [(2P^{1-s}c)/(\mu r_o t_f)]^{1/2} \tag{7.27}$$

TABLE 7.4
Units for Form Loading Equation

(*1*) Practical units	(*2*) Metric units	(*3*) Conversion factors
L_f = lb/(ft^2)(hr)	g/(cm^2)(sec)	L_f lb/(ft^2)(hr) × 454 g/lb × ft^2/30.5^2 cm^2 × hr/3600 sec = $1.356 \times 10^{-4} L_f$
P = psi	g/cm^2	P (psi) × 70.1 g/cm^2/psi = $P \times 70.1$ [Eq. (7.20)]
c = g/ml = g/cm^3	g/cm^3	—
μ = cP	g/(cm)(sec) = Poise	μ (cP) × Poise/10^2 cP = $\mu \times 10^{-2}$
r_o = (sec^2/g) × 10^{-7} = practical unit	sec^2/g	$r_o \times 10^7$
t_f (min)	sec	t_f (min) × 60 sec/min = $t_f \times 60$

which is the equation for form loading. If metric units are used (Example 7.2), the set of units for parameters in Eq. (7.27) is presented in column (*2*), Table 7.4.

For design purposes it is convenient to express form loading in terms of lb of cake/(ft^2)(hr) and other parameters in the practical units indicated in column (*1*), Table 7.4. Conversion factors from column (*1*) to column (*2*) are indicated in column (*3*). Substitution of these conversion factors in Eq. (7.27) leads to Eq. (7.28), where all parameters are in the practical units listed in column (*1*) of Table 7.4.

$$L_f = 35.7[(\mathrm{c}P^{1-s})/(\mu r_o t_f)]^{1/2} \tag{7.28}$$

Since most sludges have specific characteristics, Eq. (7.28) is modified for prediction of filtration performance.

$$L_f = 35.7(P^{1-s}/\mu r_o)^{1/2}(c^m/t_f^n) \tag{7.29}$$

[For units see Table 7.4, column (*1*).]

Equation 7.29 is referred to as the form loading equation. The usual ranges of values for parameters n and m are $1.0 > n > 0.4$, owing to variation in cake permeability while additional cake is being formed [Eq. (7.28) corresponds to a value of $n = 0.5$]; and $1.0 > m > 0.25$, owing to effect of variation in solids content fed to filter. [Eq. (7.28) corresponds to a value of $m = 0.5$.]

Determination of parameters n, s, m, and r_o is discussed in Section 4.10. Experimental technique utilized is described in Section 4.11. Example 7.3 (Section 4.12) illustrates the calculation procedure.

4.10. DETERMINATION OF PARAMETERS *n, s, m,* AND r_o IN Eq. (7.29)

1. *Determination of n.* If P and c are held constant (μ and r_o are constant for a specific experimental run), Eq. (7.29) becomes

$$L_f = K_1(1/t_f^n) \tag{7.30}$$

where

$$K_1 = 35.7(P^{1-s}/\mu r_o)^{1/2}c^m = \text{constant} \tag{7.31}$$

Writing Eq. (7.30) in logarithmic form,

$$\log L_f = -n \log t_f + \log K_1 \tag{7.32}$$

From Eq. (7.32), a logarithmic plot of L_f vs. t_f yields a straight line of slope $-n$. A typical plot is presented in Fig. 7.16.

2. *Determination of s.* If t_f and c are held constant (μ and r_o are constant for a specific experimental run), Eq. (7.29) becomes

$$L_f = K_2 P^{(1-s)/2} \tag{7.33}$$

where

$$K_2 = 35.7(1/\mu r_o)^{1/2}(c^m/t_f^n) = \text{constant} \tag{7.34}$$

Writing Eq. (7.34) in logarithmic form,

$$\log L_f = [(1-s)/2] \log P + \log K_2 \tag{7.35}$$

From Eq. (7.35), a logarithmic plot of L_f vs. P yields a straight line of slope $(1-s)/2$. A typical plot is presented in Fig. 7.17.

3. *Determination of m.* If t_f and P are held constant (μ and r_o are constant for a specific experimental run), Eq. (7.29) becomes

$$L_f = K_3 c^m \tag{7.36}$$

where

$$K_3 = 35.7(P^{1-s}/\mu r_o)^{1/2}(1/t_f) \tag{7.37}$$

Writing Eq. (7.37) in logarithmic form,

$$\log L_f = m \log c + \log K_3 \tag{7.38}$$

From Eq. (7.38), a logarithmic plot of L_f vs. c yields a straight line of slope m. A typical plot is presented in Fig. 7.18.

4. *Determination of r_o.* From Eq. (7.29) a plot of L_f vs. $35.7(P^{1-s}/\mu)^{1/2}$ $(c^m/t_f{}^n)$ yields a straight line of slope $(1/r_o)^{1/2}$. A typical plot is presented in Fig. 7.19. Since r_o is only a function of the nature of the sludge, it is a constant for all experiments performed with the same sludge.

4.11. LEAF TEST LABORATORY PROCEDURE FOR DETERMINATION OF THE PARAMETERS IN THE LOADING EQUATION

Eckenfelder and Ford [1] recommend a leaf test apparatus as shown in Fig. 7.14 for determination of the parameters in the loading equation. It is operated by alternatively submerging the test leaf in the sludge to simulate

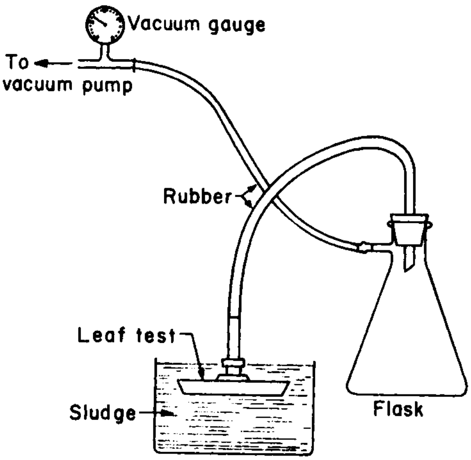

Fig. 7.14. Leaf test apparatus.

form time t_f, and taking it out to simulate dry time t_d. From Eq. (7.10) it follows that it is possible to simulate operation of a rotary filter of known submergence by proper choice of t_f and t_d. Usually, total cycle time (t_c) ranges from 1 to 6 min.

In practice, one selects an operating vacuum (10–20 in. Hg) and a submergence (10–60%). Then a dry time is selected to yield the desired % moisture for the cake. This is done by using the leaf test apparatus and preparing an experimental graph of % moisture for the cake vs. dry time. A sample graph is shown in Fig. 7.15. A value of t_d is then selected (Fig. 7.15) corresponding to the desired % moisture for the cake.

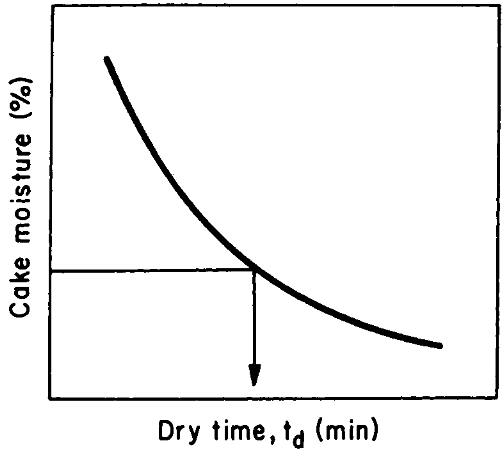

Fig. 7.15. Selection of t_d.

The procedure [1] for determination of parameters in the load equation is

1. Select t_d from Fig. 7.15.

2. Using the optimum coagulant dosage as determined in Section 4.5, run a series of leaf tests relating filter loading to form time (t_f), initial solids concentration (c), and vacuum (P). A series of 7–8 test runs should yield sufficient design data. Total cycle time is 1–6 min, with a range of drum submergence of 10–60%. This corresponds to values of t_f from 0.1 to 3.6 min (Table 7.5).

A possible series of runs is shown in Table 7.6, which corresponds to data in Example 7.3.

The following remarks are pertinent to Table 7.6: (a) for determination of n (P and c constant), runs 1–3; (b) for determination of s (t_f and c constant), runs 6–8; (c) for determination of m (t_f and P constant), runs 2, 4, and 5; and (d) for determination of r_o, runs 1–8.

TABLE 7.5
Usual Range of Values for t_f and t_d

t_c (min)	% submergence	$t_f = t_c$(% submergence/100) (min)	$t_d = t_c - t_f$ (min)
6	60%	3.6	2.4
6	10%	0.6	5.4
1	60%	0.6	0.4
1	10%	0.1	0.9

TABLE 7.6
Leaf Test Data

(*1*) Run no.	(*2*) Form time, t_f (min)	(*3*) Dry time, t_d (min)	(*4*) c	(*5*) Applied vacuum P (in. Hg)	P (psi)	(*6*) Loading, L_f [lb/(ft^2)(hr)]
1	0.25	1.5	0.03	20	9.80	73.0
2	0.50	1.0	0.03	20	9.80	47.0
3	1.00	0.5	0.03	20	9.80	29.2
4	0.50	1.5	0.04	20	9.80	70.0
5	0.50	1.5	0.05	20	9.80	90.0
6	1.50	1.0	0.03	10	4.90	21.0
7	1.50	1.0	0.03	15	7.35	22.5
8	1.50	0.5	0.03	20	9.80	23.5

3. Procedure for each run is as follows.

(a) Add the optimum coagulant dosage as determined in Section 4.5.

(b) Flocculate the mixture for 30 sec. In some cases a series of tests are made to determine optimum flocculation time.

(c) Submerge the leaf in flocculated sludge mixture for the specified form time (t_f) (Table 7.5). Maintain gentle mixing to avoid deposition of sludge.

(d) Remove leaf from sludge and hold it vertically for the specified dry time, keeping it under full vacuum.

(e) Transfer entire cake from the filter leaf to a tared dish. Compressed air may be gently applied to loosen the cake from the leaf.

(f) Weigh wet cake, dry at 103°C, and reweigh; measure and record cake thickness.

(g) The loading in lb/(ft^2)(hr) is

$$L_f = \text{dry weight of sludge in g} \times (\text{cycles/hr})/[454 \times \text{test leaf area (ft}^2)]$$

4. Prepare the following graphs:

(a) $\log L_f$ vs. $\log t_f$ (P and c are constants) (Fig. 7.16). Slope of the straight line yields parameter n in Eq. (7.29).

(b) $\log L_f$ vs. $\log P$ (t_f and c are constants) (Fig. 7.17). Slope of the straight line is $(1-s)/2$, from which parameter s is calculated.

(c) $\log L_f$ vs. $\log c$ (t_f and P are constants) (Fig. 7.18). Slope of the straight line is parameter m in Eq. (7.29).

(d) L_f vs. $35.7(P^{1-s}/\mu)^{1/2}(c^m/t_f{}^n)$ (Fig. 7.19). Slope of the straight line is $(1/r_o)^{1/2}$, from which parameter r_o is calculated.

4.12. ILLUSTRATION OF CALCULATION PROCEDURE FOR PARAMETERS *n, s, m,* AND *r_o*

Example 7.3

Laboratory tests on a leaf filter conducted for a sludge yield results presented in Table 7.6. Obtain the values for parameters n, m, s, and r_o in the form loading equation.

SOLUTION

(a) *Determination of n.* The logarithmic plot of L_f vs. t_f is shown in Fig. 7.16, from which slope $= -n = -0.653$, $n = 0.653$.

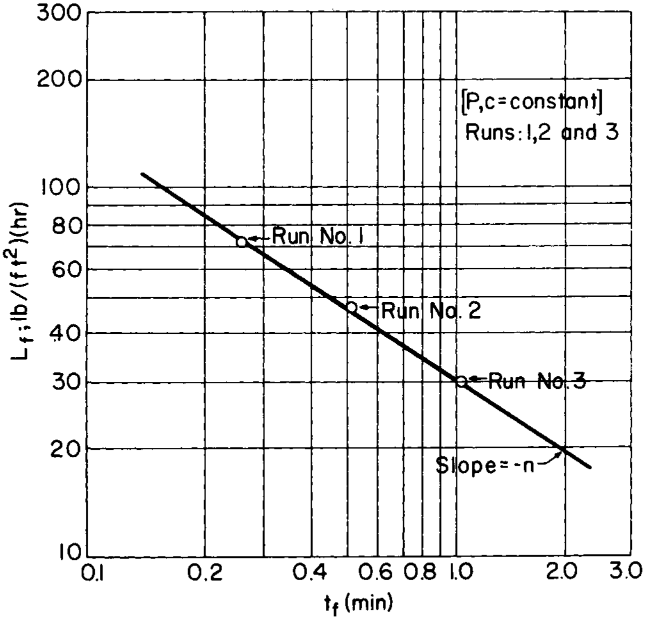

Fig. 7.16. Determination of n (Example 7.3).

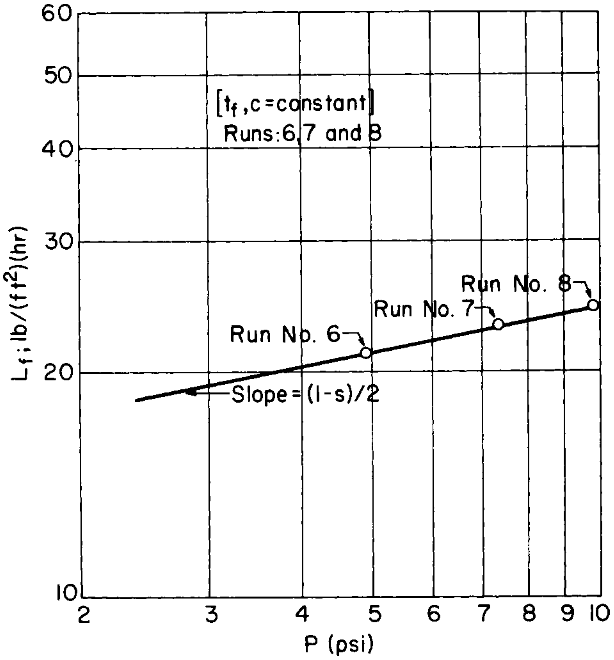

Fig. 7.17. Determination of s (Example 7.3).

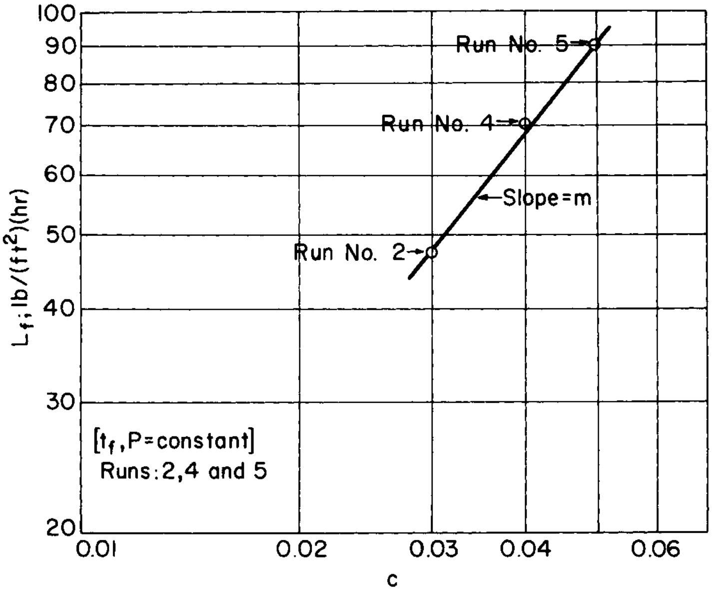

Fig. 7.18. Determination of m (Example 7.3).

(b) *Determination of s.* The logarithmic plot of L_f vs. P is shown in Fig. 7.17. From Fig. 7.17 read slope $= (1-s)/2 = 0.1635$. Therefore, $s = 1-2(0.1635) = 0.673$.

(c) *Determination of m.* The logarithmic plot of L_f vs. c is shown in Fig. 7.18. From Fig. 7.18 read slope $= m = 1.265$.

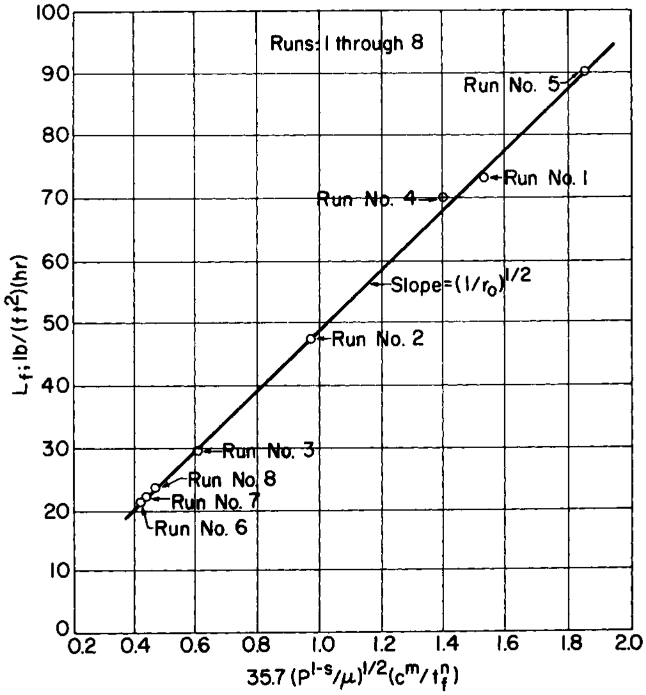

Fig. 7.19. Determination of r_0 (Example 7.3).

TABLE 7.7
Calculations for Determination of r_0 (Example 7.3)

Run no.	L_f	$P^{1-0.673} = P^{0.327}$	$c^{1.265}$	$t_f^{0.653}$	$35.7(P^{1-s}/\mu)^{1/2}(c^m/t_f^n)$
1	73.0	2.109	0.0119	0.404	1.525
2	47.0	2.109	0.0119	0.635	0.970
3	29.2	2.109	0.0119	1.000	0.615
4	70.0	2.109	0.0172	0.635	1.395
5	90.0	2.109	0.0227	0.635	1.850
6	21.0	1.683	0.0119	1.303	0.425
7	22.5	1.921	0.0119	1.303	0.451
8	23.5	2.109	0.0119	1.303	0.472

(d) *Determination of* r_o. The plot of L_f vs. $35.7(P^{1-s}/\mu)^{1/2}(c^m/t_f^n)$ is shown in Fig. 7.19; the necessary calculations are presented in Table 7.7. Take $\mu = 1$ cP.

From Fig. 7.19 slope $= (1/r_o)^{1/2} = 48.5$. Therefore, $r_o = (1/48.5)^2 = 0.000425$.

In summary, filtration parameters for the form loading equation for Example 7.3 are $n = 0.653$, $s = 0.673$, $m = 1.265$, and $r_o = 0.000425$.

4.13. PROCEDURE FOR ROTARY FILTER DESIGN [1]

Step 1. From Fig. 7.15 select an appropriate value of the dry time t_d for a desired cake moisture.

Step 2. For a selected % submergence and the chosen value of t_d, calculate form time t_f from Eq. (7.39), obtained by combining Eqs. (7.10) and (7.9). Solving for t_f,

$$t_f = t_d/[(100/\%\ \text{submergence}) - 1] \qquad (7.39)$$

If desired, adjust the values of t_f [calculated from Eq. (7.39)], t_c, and % submergence utilizing Eqs. (7.9) and (7.10). This is illustrated in Example 7.4.

Step 3. Calculate form loading L_f from Eq. (7.29). For calculation of L_f use (a) values of parameters n, m, s, and r_o determined from laboratory data (Example 7.3); (b) selected value of the operating vacuum; (c) form time t_f from Step 2; and (d) μ and c from data on characteristics of sludge. Ordinarily take $\mu = 1$ cP.

Step 4. Calculate cycle loading L_c from

$$L_c = L_f(\%\ \text{submergence}/100) \times 0.8 \qquad [\text{lb}/(\text{ft}^2)(\text{hr})] \qquad (7.40)$$

In Eq. (7.40), factor 0.8 compensates for the sector of the filter drum where cake is removed and media washed (Fig. 7.9).

Step 5. Calculate required filter area from

$$\text{Filter area} = \frac{\text{lb/hr of solids to be removed}}{L_c \quad [\text{lb}/(\text{ft}^2)(\text{hr})]} \qquad (\text{ft}^2) \qquad (7.41)$$

Example 7.4 illustrates this design procedure.

Example 7.4

Laboratory tests for the sludge in Example 7.3 indicated a dry time $t_d = 1.25$ min to be an appropriate choice.* It is desired to design a vacuum filter to dewater 30,000 lb/day of sludge (dry weight). Parameter c is 0.03. Filter operates 100 hr/week. Select a submergence of 25% and an operating vacuum of 15 in. of Hg. Determine the required filter area in ft^2.

---

* Details of these tests, including the graph (similar to Fig. 7.15) utilized for selection of t_d, are not shown in the text.

SOLUTION

Step 1. $t_d = 1.25$ min.

Step 2. Calculate t_f from Eq. (7.39).

$$t_f = 1.25/[(100/25)-1] = 0.417 \text{ min}$$

Take $t_f = 0.5$ min and adjust accordingly values of t_c and % submergence, i.e. [from Eq. (7.9)],

$$t_c = t_f + t_d = 0.5 + 1.25 = 1.75 \text{ min}$$

From Eq. (7.10), the adjusted % submergence is

$$\%\,\text{submergence} = 100(t_f/t_c) = 100(0.5/1.75) = 28.6\%$$

Step 3. Calculate form loading L_f from Eq. (7.29), where parameters $n = 0.653$, $s = 0.673$, $m = 1.265$, and $r_o = 0.000425$ (Example 7.3). Other data: $P = 15$ in. Hg ($= 7.35$ psi), $\mu = 1$ cP, $c = 0.03$, and $t_f = 0.5$ min.

$$L_f = 35.7[7.35^{1-0.673}/(1.0)(0.000425)]^{1/2}(0.03^{1.265}/0.5^{0.653}) = 44.7 \text{ lb/(ft}^2\text{)(hr)}$$

Step 4. Calculate cycle loading L_c [Eq. (7.40)].

$$L_c = 44.7(28.6/100) \times 0.8 = 10.23 \text{ lb/(ft}^2\text{)(hr)}$$

Step 5. Determine the required filter area [Eq. (7.41)].

$$\text{Filter area} = \frac{30{,}000 \text{ lb/day} \times 7 \text{ days/week} \times \text{week/100 hr}}{10.23 \text{ lb/(ft}^2\text{)(hr)}} = 205.3 \text{ ft}^2$$

5. Pressure Filtration

Filtration of sludges in filter presses is economically feasible whenever labor costs are relatively low, owing to the difficulty of full automation of the operation. Recently, systems with fully automatic operation have been available commercially with automatic opening of the press, cake discharge, and washing of the filter media between cycles. These new developments tend to make filter presses desirable for use more frequently in the future.

The main advantage of filter presses over rotary vacuum filters is that a drier cake can be obtained. This is especially advantageous if filtration is followed by incineration.

6. Centrifugation

Dewatering of sludges by centrifugation has been applied with increasing frequency in the last few years. A sketch of a typical centrifuge for this service (a continuous solid bowl type) is shown in Fig. 7.20.

The components of the centrifuge are (1) fixed casing, (2) rotating bowl, (3) rotating inner conveyor, (4) driving components (motor and gear system), (5) slurry inlet port, (6) solids discharge port, and (7) liquid discharge port.

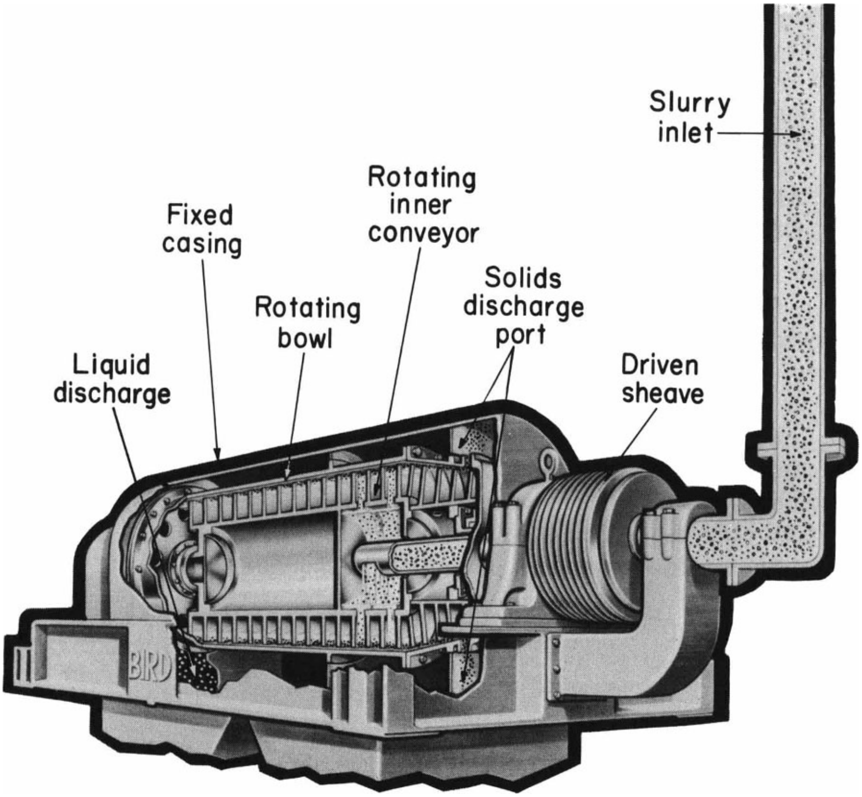

Fig. 7.20. Bird continuous solid bowl centrifuge. *(Courtesy of Bird Machine Company, Inc.)*

Sludge solids are compacted by centrifugal force against the inner walls of the rotating bowl, then picked up by the conveyor and taken to the solids discharge port. Liquid is discharged at the opposite end of the bowl. No established design procedure is available from laboratory data, but feasibility of centrifugation for a specific sludge may be evaluated from tests employing laboratory centrifuges.

7. Bed Drying of Sludges

7.1. INTRODUCTION

Air drying of sludges on sand beds is one of the most economical methods for dewatering. It is the most common method for small treatment plants, for both domestic and industrial wastewaters.

Economic feasibility of the process depends greatly on (1) availability of land at a reasonable cost, and (2) favorable climatic conditions (dry and warm climate) conducive to optimum evaporation conditions. The area needed is a function of (1) rainfall and evaporation rates, and (2) sludge characteristics (for example, gelatinous sludges require a larger area).

7.2. MECHANISMS OF DEWATERING SLUDGES ON SAND BEDS

Dewatering of sludges occurs by two mechanisms:

1. Percolation of water through the sand bed. The proportion of water removed by percolation varies from 20 to 55% depending on initial solids content of sludge and characteristics of solids. Percolation is generally complete in 1–3 days, resulting in solids concentration as high as 15–25%.

2. Evaporation of water. Evaporation occurs by mechanisms of radiation and convection. The rate of evaporation is slower than that of dewatering by percolation, and it is related to temperature, relative humidity, and air velocity. A typical evaporation rate curve is shown in Fig. 7.21 and exhibits two distinct sections corresponding to constant and falling rate periods.

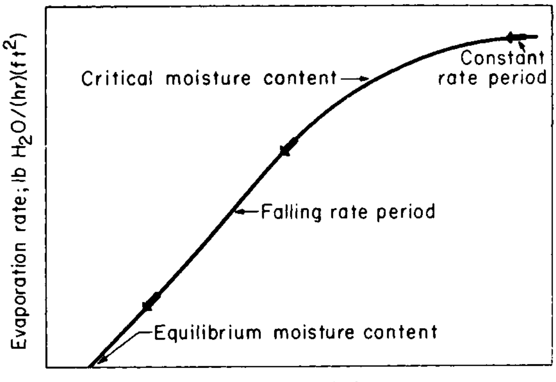

Fig. 7.21. Evaporation rate curve.

During the constant rate period, the sludge surface is wet, and rate of evaporation is relatively independent of the nature of the sludge. This rate is less than that which is observed from a free water surface (free water evaporation), owing to the fact that the plane of vaporization is below the surface of the solid. Evaporation proceeds at a constant rate until a critical moisture content is reached (Fig. 7.21).

When the critical moisture content is reached, water no longer migrates to the surface of the sludge as rapidly as it evaporates, and the falling rate period occurs. Rate of drying during this period is related to thickness of the sludge, its physical and chemical properties, and atmospheric conditions. Subsurface drying continues until an equilibrium moisture content is reached (Fig. 7.21).

7.3. CONSTRUCTION OF SAND DRYING BEDS

Typical sand bed construction is indicated in Fig. 7.22, which shows a vertical section of a drying bed. Underdrain piping may be of vitrified clay, with a minimum diameter of 4 in. and a minimum slope of 1%. The filtrate is returned to the treatment plant.

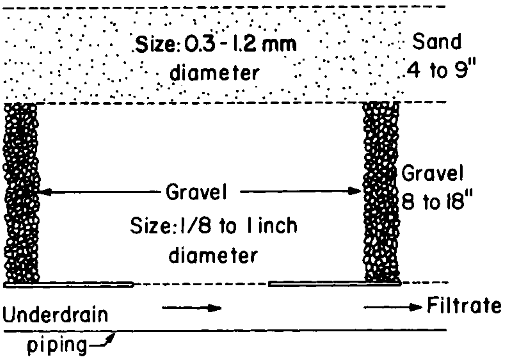

Fig. 7.22. Sand drying bed.

Sludge is usually applied to drying beds at depths of 8–12 in. It is left to dry until it reaches a solids content between 30 and 50%. It is removed when solids reach a liftable state, which varies with individual judgement as well as the final disposal means.

The period of time between application of sludge to sand bed and its removal in a liftable state is called bed turnover time. It varies between 20 and 75 days, depending on the nature of the sludge. It is possible to reduce substantially the bed turnover time by prior treatment with chemical coagulants, for example, alum and polyelectrolytes. By chemical pretreatment it is possible to reduce drying time by as much as 50%, and it is feasible to apply the sludge in a thicker layer. Bed yield is reported to vary linearly with coagulant dosage.

7.4. DRYING BED DESIGN

In the past, drying beds have been designed on an empirical basis of ft^2 of bed area/capita or lb of dry solids/(ft^2)(year). Values of these parameters employed in the United States are given in Ref. [7].

A rational method of design has been developed recently by Swanwick [6] and is recommended by Eckfelder and Ford [1]. The procedure is as follows, and Example 7.5 illustrates its applicating.

Step 1. Fill a glass cylinder (1–2 in. diameter) containing a sand base with test sludge to a depth of 8–12 in. (depth envisioned for the actual unit to be designed).

Step 2. Allow complete drainage of water from sludge. This requires a drainage period of 1–3 days, depending on sludge characteristics and initial moisture content. This corresponds to the percolation phase of the drying mechanism.

Step 3. Once drainage is completed, remove the sludge core from cylinder. Utilize a small fraction of it to determine moisture content (by oven drying and weighing before and after).

Step 4. Place the sludge core in an open dish to allow evaporation to occur. Check the sample periodically until desired moisture content is reached (moisture content when sludge cake is liftable).

Step 5. The difference between weights of water (moisture content) at the end of Steps 3 and 4 corresponds to the water to be evaporated (evaporation phase in the drying mechanism). Water to be evaporated is expressed in in. evaporated/ft^2 of bed area (details of calculation in Step 5, Example 7.5).

Step 6. Obtain local meteorological records for rainfall (in.) and evaporation (in.) tabulated on a monthly basis.

Step 7. Prepare in tabular form a record of inches of rainfall multiplied by a factor 0.57 vs. month. This is based on experimental evidence that 43% of the rainfall drains through the cake, leaving 57% to be evaporated. Rainfall fraction to be evaporated varies according to rainfall patterns and intensity. A value of less than 57%, for example, might be expected in regions where rainfall is intense and of short duration. Prepare also in tabular form a record of inches of evaporation multiplied by a factor 0.75 vs. month. This is based on experimental evidence that average evaporation of wet sludge is 75% of that for free water.

Step 8. Calculate average evaporation rate (days/in.) for each month.

Step 9. Based on values calculated in Step 8, determine time required to evaporate inches of water calculated in Step 5 for each month.

Step 10. Based on values calculated in Step 8 and inches of rainfall multiplied by factor 0.57, calculate time required to evaporate rainfall (days) for each month.

Step 11. The total time for evaporation of water calculated in Step 9 plus that to evaporate rainfall (Step 10) is then obtained for each month.

Step 12. Take the largest of the total times obtained in Step 11 as the design requirement. Calculate the ft^2 of bed area required as indicated in Example 7.5 (Step 12).

Example 7.5

Sludge drying beds are considered to dewater 4000 lb/day (dry weight) of a sludge produced in an industrial wastewater treatment plant. The sludge is applied to beds in 10-in. lifts.

A laboratory study shows that percolation increased solids concentration from its initial value of 5% to 20% in 25 hr. Sludge is considered liftable from drying beds at 25%solids. Meteorological records for the region are indicated in Table 7.8. Determine the required area of beds.

TABLE 7.8
Rainfall and Evaporation Record (Example 7.5)

	Jan.	Feb.	Mar.	Apr.	May	June	July	Aug.	Sept.	Oct.	Nov.	Dec.
Rainfall (in.)	4.0	3.0	3.1	4.1	4.0	3.5	2.1	3.0	3.2	3.0	2.7	2.8
Evaporation (in.)	5.8	6.5	7.5	8.7	11.2	11.0	13.2	11.3	9.1	6.5	4.6	3.1

$$10 \text{ in. lift} \rightarrow 10/12 = 0.83 \text{ ft}$$

$$(1 \text{ ft}^2) \times (0.83 \text{ ft})(62.4 \text{ lb/ft}^3) = 51.8 \text{ lb}$$

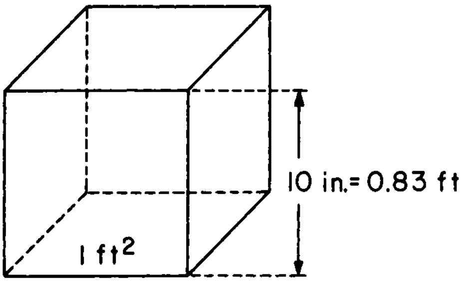

Fig. 7.23. Basis of calculations for Example 7.5.

SOLUTION Take sludge density as 62.4 lb/ft^3. Base calculations on total lb of wet sludge applied in a lift per ft^2 of bed, as indicated in Fig. 7.23.

Steps 1 and 2. Perform laboratory tests as described.

Step 3. Solids content at beginning of percolation equals 5% as stated. Based on a 10-in. lift per ft^2 of bed, this corresponds to (51.8 lb)(0.05) = 2.6 lb/ft^2 (dry solids). After drainage to 20% solids the total weight of wet cake per ft^2 of area (10-in. lift) is 2.6 lb/ft^2/0.2 = 13 lb/ft^2.

Step 4. Evaporation proceeds to a final cake containing 25% solids. This corresponds to a total weight of wet cake per ft^2 of area (10-in. lift) = 2.6 lb/0.25 = 10.4 lb/ft^2.

Step 5. Water evaporated per ft^2 of area (10-in. lift) is

$$\underset{\text{(Step 3)}}{13.0} - \underset{\text{(Step 4)}}{10.4} = 2.6 \text{ lb of water evaporated/ft}^2 \text{ of bed area}$$

Water evaporated is expressed in terms of inches evaporated per ft^2 of bed area. Volume of water evaporated (ft^3/ft^2 of bed area) is

$$2.6 \text{ lb/ft}^2 \times \text{ft}^3/62.4 \text{ lb} = 0.0417 \text{ ft}^3/\text{ft}^2$$

or

$$0.0417 \text{ ft}^3/\text{ft}^2 \times 12 \text{ in./ft} = 0.5 \text{ in. of water to be evaporated/ft}^2 \text{ of bed area}$$

Step 6. Maintain meteorological records (Table 7.8).

Step 7. Record inches of rainfall × 0.57; see columns (*2*) and (*3*) of Table 7.9. Record inches of evaporation × 0.75; see columns (*4*) and (*5*) of Table 7.9.

Step 8. Take the average evaporation rate (days/in.) for each month. This calculation is indicated in columns (*6*) and (*7*) of Table 7.9.

Step 9. Determine the time required to evaporate 0.5 in. of water (see Step 5). This calculation is indicated in column (*8*) of Table 7.9.

Step 10. Calculate the time required to evaporate rainfall (days). This calculation is indicated in column (*9*) of Table 7.9.

Step 11. Estimate the total time required to evaporate 0.5 in. of water *plus* rainfall. See column (*10*), Table 7.9.

Step 12. Take December as the control month [largest value of total time, i.e., 27.95 days, column (*10*) of Table 7.9]. Required area of sand bed is then

$$4000 \text{ lb dry sludge/day} \times \text{ft}^2/2.6 \text{ lb dry sludge} \times 27.95 \text{ days}$$

$$= 43{,}000 \text{ ft}^2 \text{ of bed area required}$$

8. Pre-dewatering Treatment of Sludges

Frequently, dewatering of sludges is difficult, especially when of gelatinous consistency. Dewatering these types of sludges by vacuum filtration, for example, is exceedingly difficult, and pre-dewatering treatment may be recommended. Two types of predewatering treatment are discussed in this section: (1) chemical coagulation and (2) heat treatment.

8.1. CHEMICAL COAGULATION

Addition of chemical coagulants promotes coalescence of sludge particles and thus improves their ability to be filtered. It may be preceded by washing of the sludge, an operation which is known as elutriation. Elutriation reduces alkalinity, and therefore minimizes coagulant requirements. $FeCl_3$, lime, and polyelectrolytes are the most common coagulants.

TABLE 7.9
Evaporation Time Requirement

(1) Month	*(2)* Rainfall (in.) (Table 7.8)	*(3)* Rainfall (in.) × 0.57 = *(2)* × 0.57 in.	*(4)* Evaporation (in.) (Table 7.8)	*(5)* Evap. (in.) × 0.75 = *(4)* × 0.75 in.	*(6)* No. of days per month	*(7)* Average evap. rate, *(6)*/*(5)* (days/in.)	*(8)* Time required to evaporate 0.5 in. (days) = 0.5 × *(7)*	*(9)* Time required to evaporate rainfall (days) = *(3)* × *(7)*	*(10)* Total days = *(8)* + *(9)*
Jan.	4.0	2.280	5.8	4.350	31	7.126	3.563	16.25	19.81
Feb.	3.0	1.710	6.5	4.875	28	5.744	2.872	9.82	12.69
Mar.	3.1	1.767	7.50	5.625	31	5.511	2.756	9.68	12.44
Apr.	4.1	2.337	8.7	6.525	30	4.598	2.299	10.75	13.05
May	4.0	2.280	11.2	8.400	31	3.690	1.845	8.41	10.26
June	3.5	1.995	11.0	8.250	30	3.636	1.818	7.25	9.14
July	2.1	1.197	13.2	9.900	31	3.131	1.566	3.75	5.32
Aug.	3.0	1.710	11.3	8.475	31	3.658	1.829	6.25	8.08
Sept.	3.2	1.824	9.1	6.825	30	4.396	2.198	4.81	7.01
Oct.	3.0	1.710	6.5	4.875	31	6.359	3.180	10.87	14.05
Nov.	2.7	1.539	4.6	3.450	30	8.696	4.348	13.38	17.73
Dec.	2.8	1.596	3.1	2.325	31	13.333	6.667	21.28	27.95

8.2. HEAT TREATMENT OF SLUDGES

Three processes of heat treatment are described in this section: (1) Porteus, (2) Zimpro, and (3) Nichols processes. These processes involve wet oxidation of sludge. This consists of chemical oxidation of organic solids in an aqueous phase by dissolved oxygen in reactors operating at high temperatures and pressures. Advantages of heat treatment are that sludge is sterilized, deodorized, and can be easily filtered either in vacuum or pressure filters. A combined flow diagram for the Porteus and Zimpro processes is shown in Fig. 7.24.

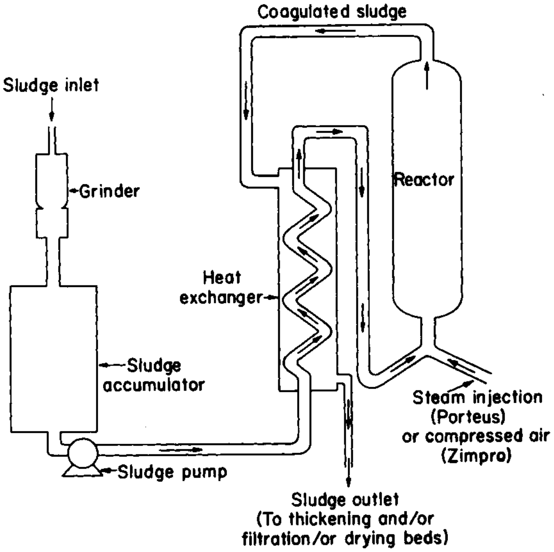

Fig. 7.24. Combined flow diagrams for the Porteus and Zimpro processes.

The Porteus process involves continuous operation under pressures of 180–210 psig and temperatures of approximately 400°F. After passing through a grinder, the sludge is pumped from the accumulator through a heat exchanger, where it is preheated by hot sludge effluent from the reactor. High pressure steam is injected into the reactor. Retention time in the reactor is approximately 30 min. An 80–90% reduction of organic matter can be accomplished; thus some organic matter and ammonia are found among the end products.

The Zimpro process differs from the Porteus primarily in the fact that air (instead of steam) is injected by an air compressor (Fig. 7.24). Maximum operating temperatures are 300°–600°F, and design operating pressures are 150–3000 psig. Combustion is 80–90% complete. Heat release per lb of air is 1200–1400 BTU.

The Nichols heat-treatment process [4] is a thermomechanical system that conditions all types of sewage sludge for effective dewatering. Sludge is subjected to a temperature of 395°F for a period of 30 min, breaking down gelatinous cell structures to release entrained water. After this thermal aging step, specific resistance and capillary suction time of the sludge are lowered to a point where it can easily be dewatered mechanically to a high solids concentration without addition of chemicals.

Heat treatment can be used to condition all types of combinations of organic wastewater sludge, including the difficult to handle waste activated sludge. The presence of industrial wastes has not been found to affect sludge conditioning.

As shown in Fig. 7.25, sludge first passes through grinders, then into high pressure feed pumps. Interlocks prevent improper flow levels and rates and assure continuous feed. Sludge travels directly to the heat exchanger section, where it is brought to process temperature in two stages; it then flows continuously through the reactor along serpentine tubes, where thermal aging takes place. The treated sludge then travels back through the heat exchanger, this time giving up its heat, and into a vessel that controls pressure and balances the system output against input. Heat-treated sludge is then discharged into a decanting tank. Thickened sludge is drawn continuously out of the decanting tank into a conditioned sludge tank equipped with mixers. Finally, conditioned, thickened, and homogenized sludge is pumped to the dewatering equipment.

The primary sludge heater employs water to transfer heat from outgoing to incoming sludge. This method permits all sludge to flow unhindered through the heat exchanger along piping designed to eliminate the high maintenance costs associated with plugging. The secondary heater also makes use of water to transfer heat to sludge. Since it is an indirect "closed loop" system, water is retained, and all heat that remains after the sludge has been heated to process temperature is recycled to the hot water generator for reuse. In addition, with indirect heating there is no increase in the volume of sludge to be dewatered. When required, the Nichols system can combine direct heating for quick start-up and indirect heating to provide maximum day-in, day-out efficiency. Because of automatic controls, safety interlocks, levels, temperature probes, and centralized valving, the entire Nichols heat-treatment system can be controlled by one operator. Nearly all system components can be located outside, effectively cutting down on indoor space requirements.

The automatic pressure control system prevents boiling of sludge and resultant corrosion of piping, in spite of flow control valve openings or equipment failures. This arrangement also allows the system to reach operating pressure automatically, as soon as the feed pumps are turned on.

Where a Nichols–Herreshoff multiple hearth furnace (Fig. 7.26) is used, the

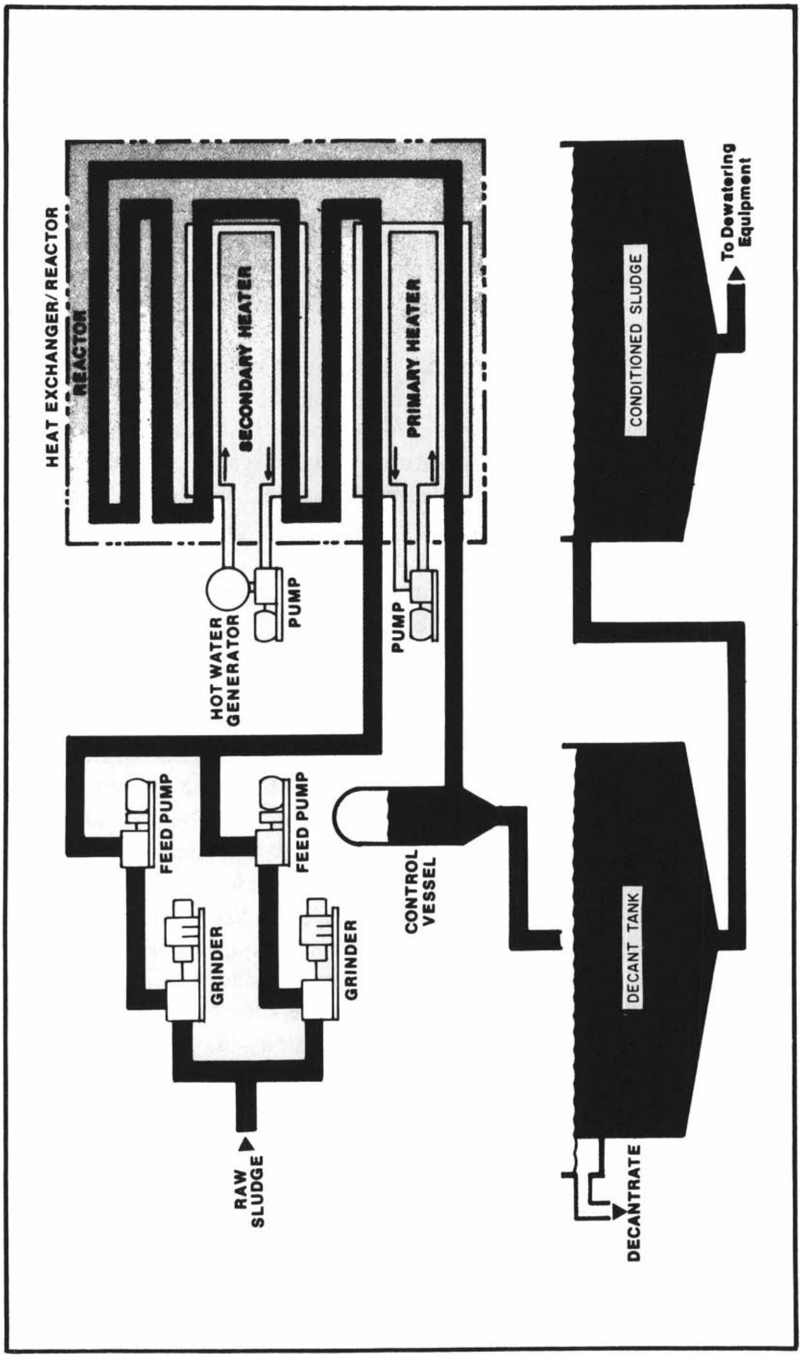

Fig. 7.25. Flow diagram for the Nichols heat-treatment process. *(Couresy of Nichols Engineering & Research Corporation.)*

heat-treatment system is operated, except for electrical power, from the waste heat recovered from the furnace. In addition, the conditioned sludge can be dewatered to such an extent that fuel for the furnace is required only at furnace warm-up.

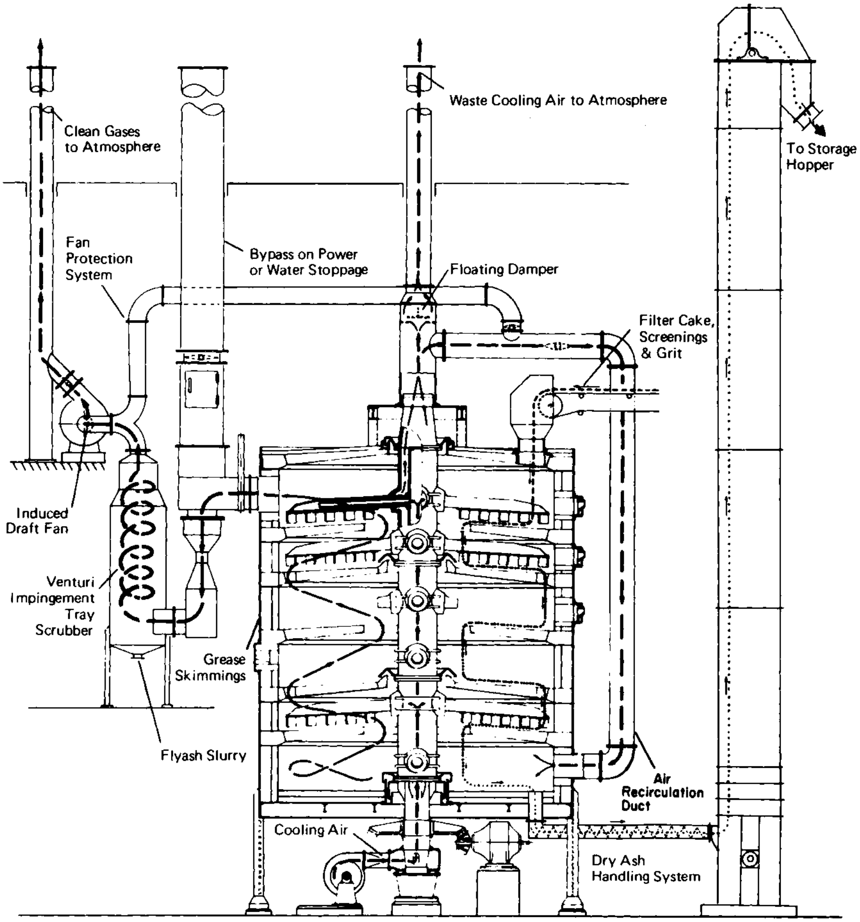

Fig. 7.26. Burning flow diagram of Nichols–Herreshoff multiple hearth furnace for sludge incineration. *(Courtesy of Nichols Engineering & Research Corporation.)*

9. Sludge Disposal

Two fundamental approaches are utilized in sludge disposal: land disposal or incineration.

9.1. LAND DISPOSAL OF SLUDGES

Two types of land disposal of sludges are mentioned here: (1) lagooning and (2) oxidation ponds. Lagooning may be an economical disposal method when large land areas are available at low cost. Supernatant liquor is removed continuously, and eventually the lagoon becomes filled with solids. In a 2- to 3-year period a 50–60% moisture content can be attained. Then the lagoon is abandoned and a new location is selected.

Oxidation ponds similar to those discussed for wastewaters in Chapter 6, Section 6 can be utilized. An aerobic condition is maintained on the surface, due mainly to the presence of algae. Anaerobic conditions prevail in deeper layers.

9.2. SLUDGE INCINERATION

Sludge incineration is a frequently employed disposal method. Self-sustaining combustion is sometimes possible after burning an auxiliary fuel which raises the temperature of the incinerator above ignition point. The combustion products are mainly carbon dioxide, sulfur dioxide, and ash.

Two types of incineration operations are (1) multiple hearth furnaces and (2) fluidized bed drying and burning. A diagram of a multiple hearth furnace for incineration of sludges (Nichols–Herreshoff furnace) is shown in Fig. 7.26.

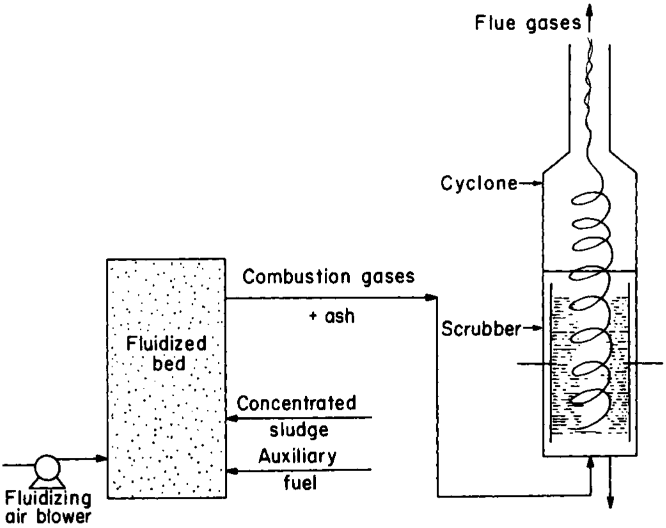

Fig. 7.27. Sketch of a fluidized bed disposal system.

Sludge passes downward through a series of hearths. Vaporization of moisture occurs in the upper hearths, followed by incineration in the lower ones. Ashes from the bottom of the furnace are collected by a dry ash handling system. An alternative water scrubbing system can also be provided.

Exhaust gases coming from the upper section of the furnace flow into a Venturi high energy type scrubber for removal of fly ash. Furnace temperature is 1000°–1800°F. The furnace is air cooled, air being blown by a fan. Cooling air is recirculated through the air recirculation duct. A fraction of the cooling air is wasted to the atmosphere.

A sketch of a fluidized bed disposal system is shown in Fig. 7.27. Sludge is fed into a bed of sand fluidized by air. The temperature of the fluidized bed is 1400°–1500°F. There is rapid drying and combustion of the sludge. Ash is carried up from the bed by the combustion gases and separated by water scrubbing and cycloning. Auxiliary fuel is utilized at least for starting the combustion process.

Problems

I. The following data are obtained from bench scale aeration of a sludge for which a continuous digester is to be designed.

Time of aeration (days)	Suspended solids (mg/liter)	Oxygen uptake rate [mg/(liter)(hr)]
0	6750	50
5	5650	21
10	4750	11
15	4200	7
20	3750	6
25	3600	5.5
30	3550	5.5

6000 lb/day of this sludge are to be treated in an aerobic digester which is designed to achieve 80% reduction of degradable volatile suspended solids. Assume a steady state concentration of 3500 mg/liter of suspended solids in the digester. Estimate basin volume (gal) and oxygen requirements (lb O_2/hr) for the aerobic digester.

II. Air drying is being considered to dewater 1500 lb/day (dry weight) of a sludge utilizing 8-in. lifts. A laboratory study has shown that percolation increased solids concentration from its initial value of 2% to 20% solids in 20 hr. Calculate the areas of drying beds (ft^2) required for liftable sludges with 30 and 35% solids, respectively. Base calculations on 3 in. of rainfall and 4.5 in. evaporation for a 30-day period.

References

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8

Tertiary Treatment of Wastewaters

1. Introduction

Tertiary treatment (also referred as "advanced wastewater treatment") consists of processes which are designed to achieve higher effluent quality than conventional secondary treatments described in Chapters 5 and 6. The following types of tertiary treatment are described in this chapter: (1) suspended solids removal, (2) carbon adsorption (organic removal), (3) ion exchange, (4) reverse osmosis, (5) electrodialysis, (6) chemical oxidation (chlorination and ozonation), (7) nutrient removal methods (nitrogen and phosphorus removal), and (8) sonozone wastewater purification process. These processes are not utilized extensively in wastewater treatment today, but their use on an increasingly larger scale is anticipated as effluent quality requirements become more stringent in the future.

2. Suspended Solids Removal

Suspended solids which have not been removed by conventional primary and secondary operations may consitute a major part of the BOD of effluents from wastewater treatment plants. The following removal processes for these suspended solids are available: (1) microscreening, (2) filtration, and (3) coagulation.

Microscreens are wound around rotating drums. Wastewater is fed continuously to the inside of the drum, flowing to a clear water storage chamber on the outside. Cleaning of the inner surface of the drum is performed by sprays of clear water, washing requirements usually being about 5% of feed volume. Microscreening results in 70–90% removal of suspended solids. Filtration is commonly used for suspended solids removal yielding removal efficiencies up to 99%. Sand, anthracite, and diatomaceous earth are the most commonly employed filter media. Coagulation is performed utilizing alum, polyelectrolytes, lime, and other chemical agents.

3. Carbon Adsorption

3.1. INTRODUCTION

Adsorption is the concentration of a solute at the surface of a solid. This phenomenon takes place when such a surface is placed in contact with a solution. A layer of molecules of solute accumulates at the surface of the solid due to imbalance of surface forces (Fig. 8.1).

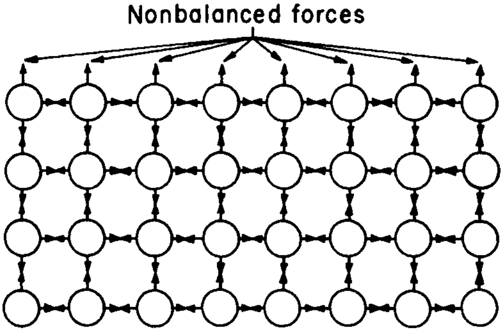


Fig. 8.1. Representation of forces in a solid.

In the interior of the solid, molecules are completely surrounded by similar molecules and therefore subjected to balanced forces, as indicated by the arrows in Fig. 8.1. Molecules at the surface are subjected to nonbalanced forces. Because these residual forces are sufficiently strong, they may imprison molecules of a solute with which the solid is in contact. This phenomenon is called physical (or van der Waals) adsorption. The solid (e.g., activated carbon) is termed the adsorbent and the solute being adsorbed is the adsorbate. Adsorption capacity is directly related to the total surface of adsorbent since the larger this surface is, the more residual (unbalanced) forces are available for adsorption.

3.2. ACTIVATED CARBONS AS ADSORBENTS

Activated carbons have been widely used as adsorbents in water treatment plants to remove taste and odor causing organics. It is expected that with the emphasis being placed on higher quality effluents, use of activated carbons in tertiary treatment of wastewaters will increase considerably in the future.

Preparation of activated carbons. Activated carbons are prepared from carbonaceous raw materials such as wood, lignite, coal, and nut shells by a process of thermal activation which yields a very porous structure with large surface areas (as high as 1000 m^2/g). Adsorption equilibrium is established when the concentration of contaminant remaining in solution is in dynamic balance with that at the surface of the solid.

Reactivation of activated carbons. The great advantage of activated carbon as an adsorbent lies in the possibility of reactivation (up to 30 or more times) without appreciable loss of adsorptive power. Usually, reactivation is done by heating spent carbon to about 1700°F in a steam–air atmosphere (thermal reactivation). This operation can be performed in multiple hearth furnaces or rotary kilns. Adsorbed organics are burned off, and activated carbon is restored basically to its initial adsorption capacity.

3.3. ADSORPTION ISOTHERMS

Equilibrium relationships between adsorbent and adsorbate are described by adsorption isotherms. In this section only the Langmuir and Freundlich isotherms are mentioned.

3.3.1. Langmuir Isotherm

In the development of the Langmuir isotherm it is assumed that the solute is adsorbed as a monomolecular layer at the surface of the adsorbent. This is the most often used adsorption isotherm, being given by the relationship

$$X/M = KbC/(1+KC) \tag{8.1}$$

where X is the weight of solute adsorbed (adsorbate) (mg); M the weight of adsorbent (g); K the equilibrium constant (cm^3 of adsorbent/mg of adsorbate); C the equilibrium concentration of solute (mg/liter); and b a constant which represents the monolayer coverage per unit weight of adsorbent (mg of adsorbate/g of adsorbent). A typical plot of X/M vs. C based on Eq. (8.1) is shown in Fig. 8.2.

Equation (8.1) is rewritten in linear form by taking the reciprocal of both members.

$$1/(X/M) = (1/Kb)(1/C) + (1/b) \tag{8.2}$$

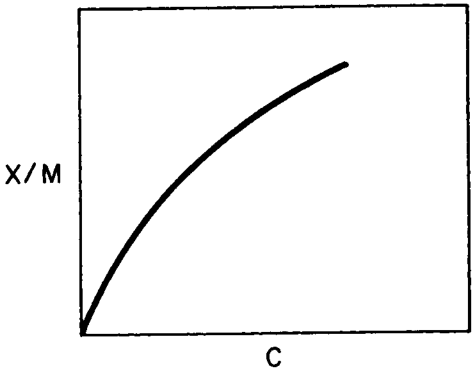


Fig. 8.2. Langmuir isotherm.

From Eq. (8.2) a plot of $1/(X/M)$ vs. $1/C$ yields a straight line, which permits determination of parameters K and b from its slope and intercept, respectively.

3.3.2. Freundlich Isotherm

The Freundlich isotherm is expressed by the equation

$$X/M = kC^{1/n} \tag{8.3}$$

X/M and C have the same meaning as in Langmuir's isotherm, and k and n are constants dependent on several environmental factors. Equation (8.3) is rewritten in linear form by taking logarithms of both members.

$$\log(X/M) = (1/n)\log C + \log k \tag{8.4}$$

Equation (8.4) reveals that a logarithmic plot of X/M vs. C yields a straight line which permits determination of parameters n and k from its slope and intercept.

3.4. ADSORPTION OPERATION

In practice, adsorption of organics in activated carbon is conducted either as a batch or continuous operation. In batch operation, powdered activated carbon is mixed with the wastewater and allowed to settle. Continuous operation is performed in columns containing granular carbon (40–80 mesh). It is more economical than batch operation and has found the widest application.

Removal of organics in activated carbon columns occurs by three mechanisms: (1) adsorption of organic molecules, (2) filtration of large particles, and (3) partial deposition of colloidal material. Percent removal depends primarily on contact time between wastewater and activated carbon.

As wastewater flows through the bed, carbon nearest to the feed point becomes saturated and must be replaced with fresh carbon. This is done by operating several suitably valved columns in series. The first column is replaced when exhausted, and the flow of wastewater is switched to make that column the last one in the series. In large installations regeneration of spent carbon is essential for economic feasibility.

3.5. DESIGN PROCEDURE FOR ACTIVATED CARBON ADSORPTION COLUMNS

The design procedure described is recommended by Eckenfelder and Ford [5] and is based on an equation derived by Bohart and Adams [1]. The following topics are discussed: (1) Bohart and Adams' equation for performance of activated carbon adsorption columns (Section 3.5.1); (2) laboratory tests with bench scale columns to obtain the necessary design parameters (Section 3.5.2); (3) "scale-up" of laboratory data and design of a plant scale unit (Section 3.5.3); and (4) derivation of Bohart and Adams' equation (Section 3.5.4).

3.5.1. Bohart and Adams' Equation

In the operation of an activated carbon adsorption column, wastewater with an influent solute concentration C_0 (mg/liter) is fed into the column. It is desired to reduce solute concentration in the effluent to a value not exceeding C_E (mg/liter), as determined by water quality requirements.

At the beginning of the operation, when activated carbon is fresh, effluent concentration is actually *lower* than allowable concentration C_E. As the operation proceeds and activated carbon approaches saturation, effluent concentration reaches value C_E. This condition is called the break point. Let t be the time elapsed to reach the break point (service time, hours). At time t, operation is discontinued and activated carbon is regenerated.

At time zero ($t = 0$), the theoretical depth of carbon which is sufficient to prevent effluent solute concentration from exceeding value C_E is called the critical bed depth D_0 (ft). Evidently, $D_0 < D$, where D is the actual bed depth (ft) (Fig. 8.3).

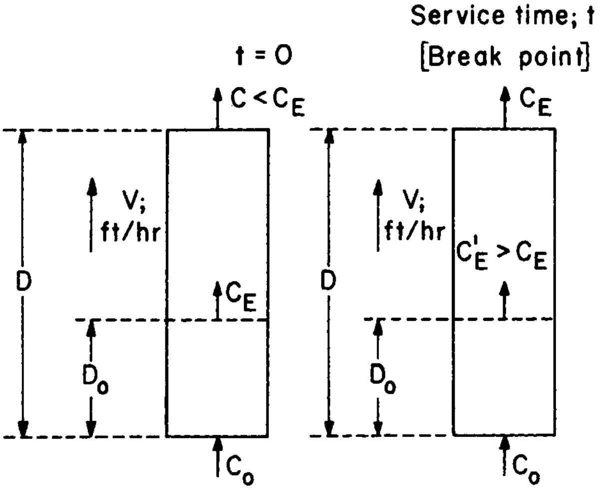


Fig. 8.3. Concept of critical bed depth (D_0).

Adsorptive capacity (N_0) is another important design variable. It is the maximum amount of contaminant solute that can be adsorbed by the carbon (lb solute/ft^3 of carbon) when saturation occurs. The break point is usually taken *before* saturation occurs. Another parameter which enters in Bohart and Adams' equation is the rate constant K. This assumes a first-order adsorption rate

$$r = Kc \tag{8.5}$$

$$\text{mg of solute/(g of carbon)(hr)} = K \text{ mg of solute/liter of solution}$$

$$\therefore K = \text{liter of solution/(g of carbon)(hr)}$$

or in English units

$$K = \text{ft}^3 \text{ of solution/(lb of carbon)(hr)}$$

Performance of continuous activated carbon columns may be evaluated by Eq. (8.6), which was developed by Bohart and Adams [1].

$$\ln[(C_0/C_E) - 1] = \ln(e^{KN_0D/V} - 1) - KC_0t \tag{8.6}$$

where C_0 is the influent solute concentration (mg/liter); C_E the allowable effluent solute concentration (mg/liter); K the rate constant [ft^3/(lb of carbon)(hr)]; N_0 the adsorptive capacity (lb of solute/ft^3 of carbon); D the depth of carbon bed (ft); V the linear flow rate (ft/hr); and t the service time (hr).

An equation for D_0 is written from Eq. (8.6). D_0 (critical depth of carbon bed, ft) is the theoretical depth of carbon sufficient to prevent effluent solute concentration from exceeding value C_E at $t = 0$. D_0 is obtained from Eq. (8.6) by letting $t = 0$ and solving for D (which equals D_0 in this case). Since the exponential term $e^{KN_0D/V}$ is usually much larger than unity, the unity term within brackets in the right-hand member of Eq. (8.6) is neglected. The final result is

$$D_0 = (V/KN_0)\ln[(C_0/C_E) - 1] \tag{8.7}$$

Expression for service time t. Solving Eq. (8.6) for t and neglecting the unity term within brackets in the right-hand member as compared to the exponential term.

$$t = (N_0/C_0V)D - \ln[(C_0/C_E) - 1]/KC_0 \tag{8.8}$$

Equation (8.8) is the basis for experimental determination of parameters N_0 and K from bench scale columns. The procedure followed is described in Section 3.5.2.

3.5.2. Determination of Parameters N_0, K, and D_0 from Laboratory Data

Laboratory equipment recommended by Eckenfelder and Ford [5] is shown in Fig. 8.4. Required data for removal of organics are obtained by passing wastewater containing a known concentration of organic material (C_0) through a series of columns (e.g., three columns in Fig. 8.4) and recording the times t (t_1, t_2, and t_3) at which concentrations of effluents from columns #1, #2, and #3 reach the allowable effluent solute concentration (C_E). Each set of experiments is performed at constant flow rate [gal/(min)(ft^2)], therefore linear velocity V (ft/sec) is held constant. Effluent from column #1 is the first to reach value C_E (record this time t_1); some time after, effluent from column #2 reaches value C_E, this time also being recorded (t_2). Finally,

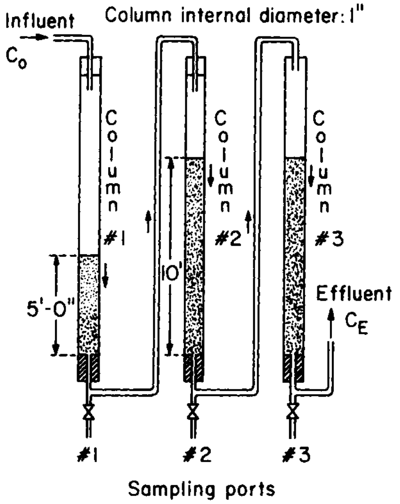


Fig. 8.4. Activated carbon columns (bench scale).

effluent from column #3 reaches concentration C_E. This time is also recorded (t_3) and the experiment discontinued. Consequently, the experiment is carried out until the break point of the last column is reached.

Table 8.1 shows a typical tabulation of t vs. D values at four different flow rates. For Experiment no. 1, three columns of 2.5 ft of carbon depth each are utilized, corresponding to total depths of 2.5, 5.0, and 7.5 ft at sampling ports. Flow rate is 2.0 gal/(min)(ft^2).

For Experiment no. 2, the first two columns contain a depth of 2.5 ft of carbon each, and the third 5 ft. Total depths corresponding to sampling ports are 2.5, 5.0, and 10 ft. Flow rate is 4.0 gal/(min)(ft^2).

For experiments no. 3 and 4, carbon depths are, respectively, 5-5-5 ft and 5-10-10 ft for the three columns, corresponding to total depths at sampling ports of 5-10-15 ft and 5-15-25 ft. Flow rates are 8 and 16 gal/(min)(ft^2), respectively. Equation (8.8) reveals that a plot of t vs. D yields a straight line for which slope (s) and intercept (i) are

$$s = N_0/C_0 V$$

$$\therefore N_0 = C_0 V s \tag{8.9}$$

$$i = \ln[(C_0/C_E) - 1]/KC_0$$

$$\therefore K = \ln[(C_0/C_E) - 1]/iC_0 \tag{8.10}$$

Values of parameters N_0 and K are determined from Eqs. (8.9) and (8.10), respectively. Figure 8.5 shows four straight lines corresponding to the four experiments for which data are presented in Table 8.1. Critical depth D_0 is calculated from Eq. (8.7).

TABLE 8.1
Data for Example 8.1

Experiment no.	(*1*) Flow rate [gal/(min)(ft²)]	(*2*) Bed depth (ft)	(*3*) Time, *t* (hr)	(*4*) Throughput volume (gal)[a]
1	2.0	$D_1 = 2.5$	740	484
		$D_2 = 5.0$	1780	1164
		$D_3 = 7.5$	2780	1818
2	4.0	$D_1 = 2.5$	180	235
		$D_2 = 5.0$	560	732
		$D_3 = 10.0$	1330	1740
3	8.0	$D_1 = 5.0$	170	445
		$D_2 = 10.0$	500	1308
		$D_3 = 15.0$	830	2171
4	16.0	$D_1 = 5.0$	60	314
		$D_2 = 15.0$	390	2040
		$D_3 = 25.0$	730	3819

[a] Calculation procedure for column (*4*) of Table 8.1. Cross-sectional area is

$$A = (\tfrac{1}{4})\pi(\tfrac{1}{12})^2 = 0.00545\ \text{ft}^2$$

$$\text{Throughput volume} = \text{gal/(min)(ft}^2) \times 0.00545\ \text{ft}^2 \times (t60)\text{min}$$

or

$$(4) = (1) \times (3) \times 0.327$$

Example 8.1

Data in Table 8.1 are obtained by using continuous bench scale activated carbon adsorption columns with 1-in. inside diameter. Four sets of experiments are performed. The wastewater contains 20 mg/liter of an organic solute removable by carbon adsorption. Experiments are carried out recording the time taken for effluents from the first, second, and third adsorption columns to reach a concentration $C_E = 1.0$ mg/liter of solute [column (*3*) of Table 8.1]. Prepare a plot of parameters N_0, K, and D_0 vs. flow rate [gal/(min (ft²)].

SOLUTION A plot of t vs. D [column (*3*) vs. column (*2*) of Table 8.1] is presented in Fig. 8.5. Adsorptive capacity N_0 is calculated from Eq. (8.9), where (20 mg/liter = 20×10^{-6} lb solute/lb liquor)

$$C_0 = 20 \times 10^{-6}\ \text{lb solute/lb liquor} \times 62.4\ \text{lb liquor/ft}^3\ \text{liquor}$$
$$= 1.248 \times 10^{-3}\ \text{lb solute/ft}^3\ \text{liquor}$$

Values of V (ft/hr) are

Exp. no. 1 $\quad 2.0\ \text{gal/(min)(ft}^2) \times \text{ft}^3/7.48\ \text{gal} \times 60\ \text{min/hr} = 16\ \text{ft/hr}$

Exp. no. 2 $4.0\ \text{gal/(min)(ft}^2) \times \text{ft}^3/7.48\ \text{gal} \times 60\ \text{min/hr} = 32\ \text{ft/hr}$

Exp. no. 3 $8.0\ \text{gal/(min)(ft}^2) \times \text{ft}^3/7.48\ \text{gal} \times 60\ \text{min/hr} = 64\ \text{ft/hr}$

Exp. no. 4 $16.0\ \text{gal/(min)(ft}^2) \times \text{ft}^3/7.48\ \text{gal} \times 60\ \text{min/hr} = 128\ \text{ft/hr}$

The calculation of N_0 is presented in Table 8.2.

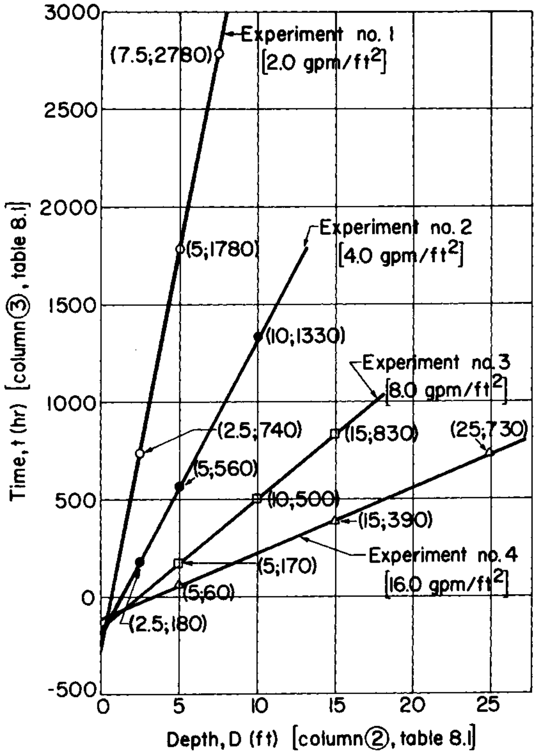


Fig. 8.5. Plots of t vs. D for Example 8.1.

TABLE 8.2
Calculation of N_0 (Example 8.1)

Exp. no.	Flow rate [gal/(min)(ft²)]	V (ft/hr)	Slope (Fig. 8.5) (hr/ft)	$N_0 = C_0Vs$ Eq. (8.9) $\left(N_0 = \frac{\text{lb}}{\text{ft}^3}\cdot\frac{\text{ft}}{\text{hr}}\cdot\frac{\text{hr}}{\text{ft}} = \text{lb/ft}^3\right)$ $(C_0 = 1.248 \times 10^{-3}\ \text{lb/ft}^3)$
1	2.0	16	408	8.15
2	4.0	32	153	6.11
3	8.0	64	66	5.27
4	16.0	128	33	5.27

TABLE 8.3
Calculations of K and D_0 (Example 8.1)[a]

Exp. no.	Flow rate [gal/(min)(ft^2)]	V (ft/hr) (Table 8.2)	N_0 (lb/ft^3) (Table 8.2)	i (absolute value of intercept) (Fig. 8.5)	K, Eq. (8.10) [ft^3/(lb)(hr)]	D_0, Eq. (8.7) (ft)
1	2.0	16	8.15	283	8.3	0.695
2	4.0	32	6.11	203	11.6	1.33
3	8.0	64	5.27	151	15.6	2.29
4	16.0	128	5.27	112	21.1	3.31

[a] $C_0 = 1.248 \times 10^{-3}$ lb/ft^3.

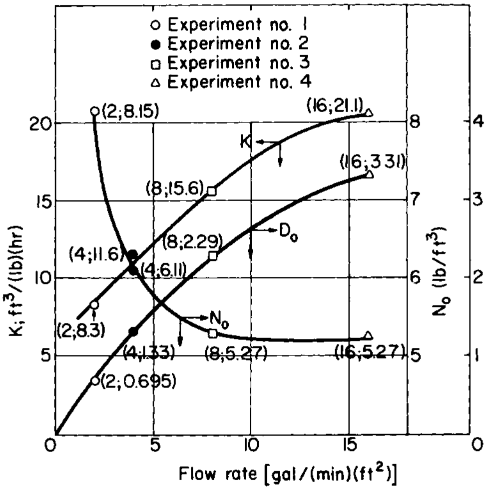


Fig. 8.6. Plots of N_0, K, and D_0 vs. flow rate (Example 8.1).

Values of K are calculated from Eq. (8.10), where $\ln[(C_0/C_E - 1] = \ln[(20/1) - 1] = 2.94$. Values of D_0 are calculated from Eq. (8.7). Calculations of K and D_0 are presented in Table 8.3. Values of N_0, K, and D_0 are plotted vs. flow rate in Fig. 8.6.

3.5.3. Scale-Up of Laboratory Data and Design of a Plant Scale Unit

Plant scale design is illustrated by Example 8.2.

Example 8.2

Wastewater utilized in laboratory experiments (Example 8.1) is treated for removal of solute (concentration = 20 mg/liter) to a residual value of 1 mg/liter in an activated carbon bed 2.5 ft in diameter and 6 ft deep. Flow is 25,000 gal/day.

1. Calculate service time (hours per cycle).
2. Calculate number of carbon changes required per year and annual carbon volume (ft^3).
3. Estimate solute removal (lb/year).
4. Calculate adsorption efficiency based on N_0 and D_0.
5. Estimate % error in neglecting unity factor in term $(e^{KN_0D/V}-1)$. This approximation is utilized in arriving at Eqs. (8.7)–(8.10).

SOLUTION Part 1. Calculation of service time

Step 1. Flow rate in gal/min is (25,000)/(24)(60) = 17.4 gal/min. Cross-sectional area of column is $A = (\frac{1}{4})\pi(2.5)^2 = 4.9$ ft². Flow rate in gal/(min)(ft²) is 17.4/4.9 = 3.6 gal/(min)(ft²), or

$$V = 3.6 \text{ gal/(min)(ft}^2) \times 60 \text{ min/hr} \times \text{ft}^3/7.48 \text{ gal} = 28.9 \text{ ft/hr}$$

Step 2. From Fig. 8.6, for flow rate 3.6 gal/(min)(ft²), one reads

$$K = 10.7 \text{ ft}^3/(\text{lb})(\text{hr})$$

and

$$N_0 = 6.35 \text{ lb/ft}^3$$

Step 3. Service time is calculated from Eq. (8.8).

$$t = [6.35/(1.248\times 10^{-3}\times 28.9)]\,6 - 2.94/(10.7\times 1.248\times 10^{-3}) = 836 \text{ hr/cycle}$$

SOLUTION Part 2. Number of carbon changes per year and annual carbon volume

$$\text{No. carbon changes/year} = (365\times 24)/836 = 10.5$$

$$\text{Annual carbon volume} = (6\times 4.9)\,10.5 = 309 \text{ ft}^3$$

SOLUTION Part 3. Estimation of solute removal (lb/year)

Solute removed per cycle is calculated from

lb solute removed per cycle

= lb solute in influent per cycle − lb solute in effluent per cycle

Since

Volume of wastewater per cycle

$= 25{,}000 \text{ gal/day} \times 836 \text{ hr/cycle} \times \text{day}/24 \text{ hr} \times \text{ft}^3/7.48 \text{ gal}$

$= 1.16 \times 10^5 \text{ ft}^3/\text{cycle}$

and

$$C_0 = 20 \text{ mg/liter} = 1.248 \times 10^{-3} \text{ lb/ft}^3$$

then

lb solute in influent per cycle $= 1.248 \times 10^{-3} \text{ lb/ft}^3 \times 1.16 \times 10^5 \text{ ft}^3/\text{cycle}$

$= 144.8 \text{ lb/cycle}$

Residual solute leaving with effluent per cycle is calculated from

$$\text{(lb solute in effluent per cycle)} = (1.16\times 10^5)\,\text{ft}^3/\text{cycle} \times C_{\text{ave}}\,\text{lb/ft}^3$$

$$= 1.16 \times 10^5\, C_{\text{ave}}\ \text{(lb/cycle)}$$

C_{ave} is given by

$$C_{\text{ave}} = (1/836) \int_{t=0}^{t=836\ \text{hr}} C dt$$

C stands for a series of effluent concentrations (increasing values), the highest one being that corresponding to $C_E = 1$ mg/liter (or 6.24×10^{-5} lb/ft^3), which occurs after 836 hr of operation. If intermediate values of effluent concentrations from time zero (start of cycle) to $t = 836$ hr (end of cycle) are recorded, one can evaluate the integral by graphical or numerical methods. In the absence of these values one can make a conservative (low) estimate of solute removed based on the C_E value of 1 mg/liter (6.24×10^{-5} lb/ft^3). The error in such an estimate is less than 5% (since in this case $C_0/C_E = 20$). Thus,

High estimate of lb solute in effluent per cycle

$$= 1.16 \times 10^5\ \text{ft}^3/\text{cycle} \times 6.24 \times 10^{-5}\ \text{lb/ft}^3 = 7.2\ \text{lb/cycle}$$

and

$$\text{Low estimate of lb solute removed per cycle} = 144.8 - 7.2 = 137.6\ \text{lb/cycle}$$

or

$$137.6\ \text{lb/cycle} \times 10.5\ \text{cycle/year} = 1445\ \text{lb/year}$$

SOLUTION Part 4. Adsorption efficiency

Based on N_0:

Total solute adsorbed: 1445 lb/year (Solution, Part 3)
Total adsorptive capacity: (N_0)(annual carbon volume) $= 6.35$ lb/ft$^3 \times 309$ ft$^3 =$ 1962 lb
Efficiency: $(1445/1962)\times 100 = 73.6\%$

Based on D_0: from Fig. 8.6 for flow rate 3.6 gal/(min)(ft^2), one reads $D_0 = 1.23$ ft.

Efficiency: $[(D-D_0)/D]\,100 = [(6-1.23)/6]\,100 = 79.5\%$

SOLUTION Part 5. Error in neglecting factor of unity

$$KN_0D/V = (10.7\times 6.35\times 6.0)/28.9 = 14.11$$

$$e^{14.11} = 1{,}342{,}441$$

$$e^{14.11} - 1 = 1{,}342{,}440$$

$$\text{Error}\,\% = (1/1{,}342{,}440)\,100 = 7.45 \times 10^{-5}\%$$

3.5.4. Derivation of the Bohart and Adams Equation [1]

Consider a mass of adsorbent. Its residual capacity (N) diminishes at a rate given by Eq. (8.11).

$$\partial N/\partial t = -KNC \tag{8.11}$$

where N is the residual adsorbing capacity [at $t = 0$, $N = N_0$ = adsorptive capacity (lb/ft^3)]; C the solute concentration (lb/ft^3); t the time (hr); and K the rate constant [ft^3/(lb)(hr)].

Consider now the solution from which solute is removed by adsorption. Solute concentration diminishes at a rate given by

$$\partial C/\partial D = -KNC/V \tag{8.12}$$

where D is the depth of adsorbent (total depth, $D = D_0$) (ft); and V the flow velocity of solution past the adsorbent (ft/hr).

The following boundary conditions (BC) apply: For BC-1 at $t = 0$, $N = N_0$ (initial solute capacity of adsorbent). For BC-2 at $D = 0$, $C = C_0$ (inlet concentration).

Perform the following changes of variable. Let

$$N' = N/N_0 \tag{8.13}$$

$$\therefore N = N_0 N' \tag{8.14}$$

$$C' = C/C_0 \tag{8.15}$$

$$\therefore C = C_0 C' \tag{8.16}$$

$$D' = KN_0 D/V \tag{8.17}$$

$$\therefore D = D'V/KN_0 \tag{8.18}$$

$$t' = KC_0 t \tag{8.19}$$

$$\therefore t = t'/KC_0 \tag{8.20}$$

With this change of variables, one writes two modified boundary conditions: For BC-1′ at $t' = 0$ (i.e., $t = 0$), since $N = N_0$, then $N' = N_0/N_0 = 1$. For BC-2′ at $D' = 0$ (i.e., $D = 0$), since $C = C_0$, then $C' = C_0/C_0 = 1$.

Equations (8.11) and (8.12) are written in terms of the new variables N', C', D', and t', substituting N, C, D, and t by their values given by Eqs. (8.14), (8.16), (8.18), and (8.20). The final result after simplification is

From Eq. (8.11)

$$\partial N'/\partial t' = -N'C' \tag{8.21}$$

or

$$\partial \ln N'/\partial t' = -C' \tag{8.22}$$

From Eq. (8.12)

$$\partial C'/\partial D' = -N'C' \tag{8.23}$$

or

$$\partial \ln C'/\partial D' = -N' \tag{8.24}$$

Differentiating Eq. (8.22) with respect to D' and Eq. (8.24) with respect to t' leads to

$$\partial^2 \ln N'/\partial D'\,\partial t' = -\partial C'/\partial D' = N'C' \tag{8.25}$$

$$\partial^2 \ln C'/\partial t'\,\partial D' = -\partial N'/\partial t' = N'C' \tag{8.26}$$

Subtracting Eq. (8.25) from Eq. (8.26),

$$\partial^2 \ln C'/\partial t'\,\partial D' - \partial^2 \ln N'/\partial D'\,\partial t' = 0$$

or

$$\partial^2 \ln(C'/N')/\partial t'\,\partial D' = 0 \tag{8.27}$$

Integration of Eq. (8.27) yields

$$\ln(C'/N') = f(D') + f(t') \tag{8.28}$$

where $f(D')$ and $f(t')$ are, respectively, functions of D' and t' *alone*.* Imposing boundary conditions BC-1′ and BC-2′, it follows that†

$$\ln(C'/N') = t' - D' \tag{8.29}$$

or

$$C'/N' = e^{t'-D'} \tag{8.30}$$

Equation (8.23) is rewritten as $[\partial C'/(-C')]/\partial D' = N'$. Dividing both members by C' and employing Eq. (8.30),

$$-(\partial C'/C'^2)/\partial D' = N'/C' = e^{D'-t'} \tag{8.31}$$

---

* Proof that Eq. (8.28) is a solution for differential equation (8.27): differentiating Eq. (8.28) with respect to D' and then with respect to t' yields Eq. (8.27), since

$$\partial f(t')/\partial D' = 0$$

and

$$\frac{\partial}{\partial t'}\frac{\partial f(D')}{\partial D'} = 0$$

[since $f(D')$ and $f(t')$ are, respectively, functions of D' and t' alone].

† Application of BC-1′ and BC-2′ to Eq. (8.29) leads to an identity

$$\ln 1/1 = 0 - 0 = 0.$$

Therefore $f(D')$ and $f(t')$ are satisfied by

$$f(D') = -D' \quad \text{and} \quad f(t') = t'$$

Equation (8.31) integrates to yield*

$$1/C' = e^{D'-t'} - \psi(t') \tag{8.32}$$

where $\psi(t')$ is a function of t' alone.

Function $\psi(t')$ is evaluated by imposing boundary condition BC-2'.†

$$\psi(t') = e^{-t'} - 1 \tag{8.33}$$

Thus Eq. (8.32) becomes

$$1/C' = e^{D'-t'} - e^{-t'} + 1 \tag{8.34}$$

From Eq. (8.34),

$$C' = 1/(e^{D'-t'} - e^{-t'} + 1) \tag{8.35}$$

Multiplying both numerator and denominator of Eq. (8.35) by $e^{t'}$,

$$C' = e^{t'}/(e^{D'} - 1 + e^{t'}) \tag{8.36}$$

Substituting in Eq. (8.36) C', D', and t' by their values given by Eqs. (8.15), (8.17), and (8.19), the result is

$$C/C_0 = e^{KC_0t}/(e^{KN_0D/V} - 1 + e^{KC_0t}) \tag{8.37}$$

Solving Eq. (8.37) for t, the final result is

$$t = (1/KC_0) \ln \frac{e^{KN_0D/V} - 1}{(C_0/C) - 1} \tag{8.38}$$

Equation (8.38) is rearranged to yield Eq. (8.6). (t is taken as the service time and therefore concentration C equals C_E.)

---

* Proof that Eq. (8.32) is a solution for Eq. (8.31) is obtained by differentiation of Eq. (8.32) with respect to D'. Notice that $\partial\psi(t')/\partial D' = 0$, and therefore

$$\partial(1/C')/\partial D' = \partial(e^{D'-t'})/\partial D'$$

or

$$-(\partial C'/\partial D')/C'^2 = e^{D'-t'}$$

and finally

$$-(\partial C'/C'^2)/\partial D' = e^{D'-t'}$$

which is Eq. (8.31).

† Application of BC-2' to Eq. (8.34) leads to an identity

$$1/1 = e^{-t'} - e^{-t'} + 1 = 1$$

Consequently, $\psi(t')$ is satisfied by Eq. (8.33).

4. Ion Exchange

4.1. INTRODUCTION

Ion exchange is a process where ions which are held to functional groups on the surface of a solid by electrostatic forces are exchanged for ions of a different species in solution. This treatment procedure has become increasingly important in the field of wastewater treatment.

Since complete demineralization is achieved by ion exchange, it is possible to use split stream treatment processes where part of the influent wastewater is demineralized and then combined with bypassed influent to produce an effluent of specified quality (e.g., a specified hardness).

Ion exchange resins. Until the 1940's natural zeolites were the only ion exchange resins available. Exchange capacity was relatively low, which limited their economic feasibility in the field of wastewater treatment. Since then, natural zeolites have been replaced by synthetic resins. These are insoluble polymers to which acidic or basic groups are added by chemical reaction procedures. These groups are capable of reversible exchange with ions present in a solution. The total number of functional groups per unit weight (or unit volume) of resin determines exchange capacity, whereas the type of functional group determines ion selectivity and position of exchange equilibrium. Resin particles have diameters of approximately 0.5 mm and are employed in packed columns with wastewater flows of 5–12 gal/(min) (ft^2).

4.2. BASIC MECHANISM OF ION EXCHANGE: CATION AND ANION EXCHANGERS

There are two basic types of ion exchangers: (1) cation and (2) anion exchangers.

Cation exchangers. Cation exchange resins remove cations from a solution, exchanging them for sodium ions (sodium cycle) or hydrogen ions (hydrogen cycle). Removal is represented by Eq. (8.39). [R denotes the resin and M^{2+} the cation (e.g., Cu^{2+}, Zn^{2+}, Ni^{2+}, Ca^{2+}, Mg^{2+}).]

Sodium cycle: $$Na_2R + M^{2+} \rightleftharpoons MR + 2Na^+ \qquad (8.39a)$$

or

Hydrogen cycle: $$H_2R + M^{2+} \rightleftharpoons MR + 2H^+ \qquad (8.39b)$$

Cu^{2+}, Zn^{2+}, Ni^{2+}, Ca^{2+}, and Mg^{2+} are retained on the resin and a soft effluent is produced. This soft effluent contains mainly sodium salts (if sodium cycle is employed) or acids (if hydrogen cycle is employed).

When the exchange capacity of the resin is exhausted, ionic concentration in effluent from the exchange column exceeds the specified value. This condition is called breakthrough. The resin must then be regenerated. Prior to regeneration, the column should be backwashed to remove solid deposits. Regeneration consists of passing through the column either a brine solution (NaCl for sodium cycle) or an acid solution, usually H_2SO_4 or HCl (for hydrogen cycle).

Regeneration reactions for the sodium and hydrogen cycles are shown below.

$$\begin{array}{rcccl} & \text{Regenerated resin} & & \text{Regenerant waste} & \\ MR + 2NaCl \rightleftharpoons & Na_2R & + & MCl_2 & (8.40a) \\ MR + 2HCl\,(\text{or } H_2SO_4) \rightleftharpoons & H_2R & + & MCl_2 & (8.40b) \\ & & & (\text{or } MSO_4) & \end{array}$$

Typical regenerant concentrations are 2–5% by weight with flow rates of 1–2 (gal)/(min)(ft^2). As indicated by Eq. (8.40), regenerant waste consists of cation salts. This waste stream amounts to 10–15% of influent volume treated before breakthrough. Following regeneration the exchanger bed is rinsed with water to remove residual regenerant.

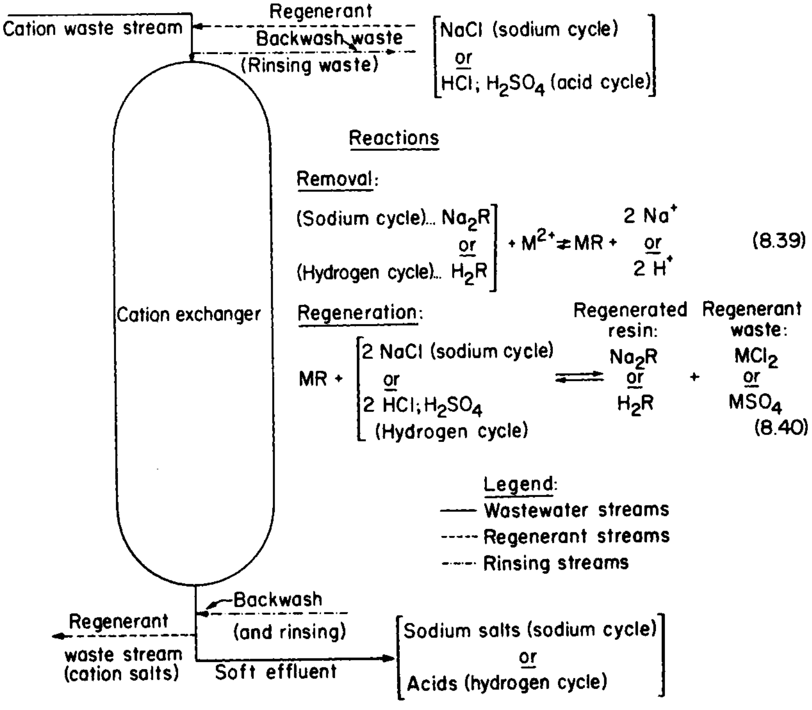


Fig. 8.7. Cation exchanger.

Cation exchanger reactions and directions of flow for different streams involved in the operation of a cation exchanger are indicated in Fig. 8.7. Hydrogen cycle cation exchange resins are weak or strong acids. Most acidic resins used in water pollution abatement are strong acids.

Anion exchangers. Anion exchange resins remove anions from a solution, exchanging them for hydroxyl ions. Removal is represented by Eq. (8.41) (where A^{2-} represents an anion).

$$R(OH)_2 + A^{2-} \rightleftharpoons RA + 2OH^- \tag{8.41}$$

Anions (e.g., SO_4^{2-}, CrO_4^{2-}) are thus removed from solution.

Regeneration is made after breakthrough, usually preceded by backwashing to remove solid deposits. Regenerants commonly used are sodium and ammonium hydroxides. The regeneration reaction is

$$\begin{array}{cccc} & \text{Regenerated} & & \text{Regenerant} \\ & \text{resin} & & \text{waste} \\ RA + 2NaOH \rightleftharpoons & R(OH)_2 & + & A^{2-} \\ \text{or} & & & \left[\begin{array}{c} Na_2A \\ \text{or} \\ (NH_4)_2A \end{array}\right] \\ 2NH_4OH & & & \end{array} \tag{8.42}$$

Typical regenerant concentrations are 5–10% by weight.

Usually, cation and anion exchangers are used in series. By adequate selection of ion exchangers, almost any wastewater problem of an inorganic nature can be handled. Basic exchange resins are either strong or weak bases. Anion exchanger reactions and direction of flow for the different streams involved in the operation of an anion exchanger are indicated in Fig. 8.8.

4.3. DESIGN OF ION EXCHANGE COLUMNS

The first step in designing an ion exchange system for a specific wastewater is to run a complete cation–anion analysis of the influent to be treated. In addition, data on total dissolved solids (TDS), dissolved CO_2, SiO_2, and pH are obtained.

Concentrations of individual ions present are expressed in either of two ways:

1. In meq/liter (milliequivalents per liter), e.g., for a solution containing 20 mg/liter of Cu^{2+}, the concentration in terms of meq/liter is

$$20/(63.5/2) = 0.63 \text{ meq/liter}$$

where 63.5 is the atomic weight of Cu and 2 is the valence.

2. In terms of calcium carbonate equivalents, e.g., for a solution containing 20 mg/liter of Cu^{2+} and the stoichiometric amount of Cl^-, the concentration of Cu^{2+} in terms of $CaCO_3$ is calculated as

$$20(134.5/100) = 27 \text{ mg/liter of } CaCO_3$$

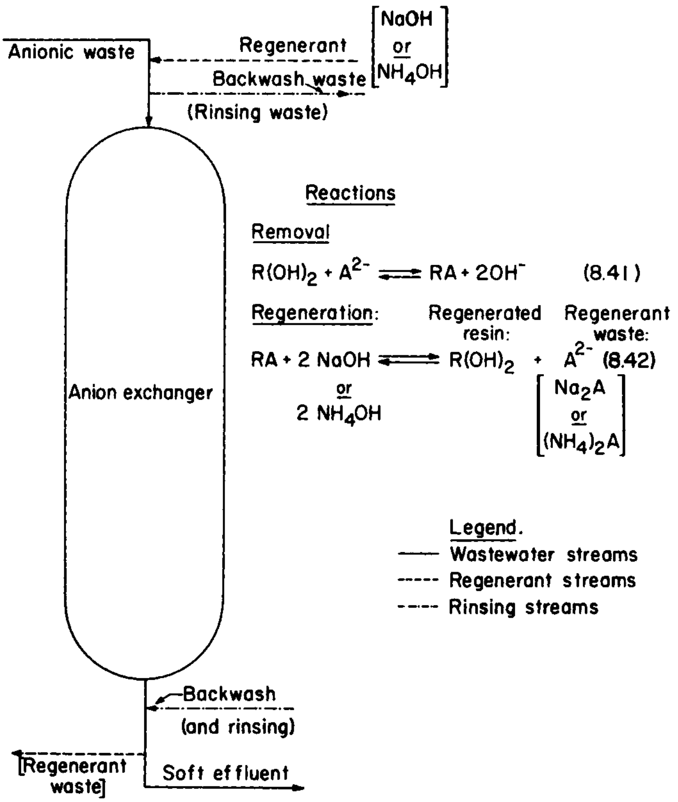


Fig. 8.8. Anion exchanger.

where 100 and 134.5 are the molecular weights of $CaCO_3$ and $CuCl_2$, respectively.

Design parameters determined by laboratory tests prior to design of an ion exchange column are

1. *Exchange capacity of resin.* Cation–anion resin capacities are usually expressed as equivalents of ion removed per unit bed volume (e.g., equivalents/liter of resin or equivalents/ft^3 of resin). They may be also expressed per unit weight of bed (e.g., equivalents/lb of resin). Exchange capacities are also expressed in terms of weight of $CaCO_3$ equivalent, either per unit volume or per unit weight of bed (e.g., lb $CaCO_3/ft^3$ of resin, lb $CaCO_3$/lb of resin).

2. *Regenerant requirements.* Regenerant requirements are expressed in terms of weight per unit volume of bed (e.g., lb H_2SO_4/ft^3). The degree of theoretical capacity attained (with respect to fresh resin) depends on the weight of regenerant employed. An economic balance between degree of theoretical capacity attained and weight of regenerant (lb regenerant/ft^3 of bed volume) is taken into consideration. Performance curves for regenerants

(exchange capacity of regenerated resin vs. weight of regenerant), determined from laboratory studies, are sometimes available from resin manufacturers. The exchange capacity of the column increases with weight of regenerant utilized.

3. *Rinsing water requirements.* Following regeneration, the exchanger bed is rinsed with water to remove residual regenerant. Rinsing requirements, also determined from laboratory studies, are sometimes available from resin manufacturers. They are expressed in terms of gallons of water per ft^3 of resin (range, 100–200 gal/ft^3). Characteristics of exchange resins are evaluated from bench scale units [5]. Plexiglass columns of 1 in. diameter are used at unit flow rates comparable to plant scale units. Table 8.4 shows typical data obtained from bench scale units.

TABLE 8.4
Characteristics of Cation and Anion Exchange Resins Utilized in Treatment of Plating Industry Wastewater

	Cation	Anion
Exchange capacity	70 eq/ft^3 of resin	3.5 lb CrO_3/ft^3 of resin
Regenerant	H_2SO_4	NaOH
Requirement (lb/ft^3 of resin)	11.0 (in lb H_2SO_4)	4.7 (in lb NaOH)
Concentration (%)	5.0	10.0
Flow rate [gal/(min)(ft^2)]	1.0	1.0
Rinsing water requirements (gal/ft^3 of resin)	130.0	100.0

Additional considerations concerning design of ion exchange systems are as follows:

1. Recovery of valuable wastewater constituents, an important factor in determining economic feasibility of ion exchange, is illustrated by Example 8.3. Chromates (CrO_4^{2-}) from a plating plant wastewater are held by an anion exchanger and subsequently recovered as chromic acid (H_2CrO_4) in a hydrogen cation exchanger. Nickel ions are salvaged from plating plant wastes.

2. Calculation of bed depth is illustrated in Example 8.3. Additional free height is provided to allow for expansion of the bed for backwashing and cleaning. As a rule of thumb a 50% allowance is taken.

4.4. DESIGN OF AN ION EXCHANGE SYSTEM [5]

Example 8.3

Design an ion exchange system to treat 120,000 gal of wastewater per day from a metal-plating industry. The main metal ions present in the wastewater are chromium, equivalent to 120 mg/liter of CrO_3 (present as chromate, CrO_4^{2-}); Cu^{2+}, 30 mg/liter; Zn^{2+}, 15 mg/liter; and Ni^{2+}, 20 mg/liter.

It is desired to remove Cu^{2+}, Zn^{2+}, and Ni^{2+} in a hydrogen cycle cation exchanger (cation exchanger no. 1) and CrO_4^{2-} in an anion exchanger downstream from the cation exchanger. The hydrogen cycle cation exchanger is regenerated by a 5% solution of H_2SO_4. The anion exchanger employed to remove CrO_4^{2-} is regenerated by a 10% solution of NaOH. Effluent from this regeneration contains sodium chromate (Na_2CrO_4). Valuable CrO_4^{2-} is recovered in another hydrogen cation exchanger (cation exchanger no. 2) as chromic acid (H_2CrO_4). This cation exchanger is also regenerated by a 5% solution of H_2SO_4. Acid effluents from regeneration of cation exchangers no. 1 and 2 are combined, neutralized, and then discarded to a sewer. Figure 8.9 shows a flowsheet of the proposed process.

Characteristics of the cation and anion exchangers to be employed are presented in Table 8.4. Both cation and anion units operate 6 days between regenerations. Design the three ion exchangers in the system and estimate regeneration and rinse water requirements.

SOLUTION The flow diagram is shown in Fig. 8.9, and chemical reactions involved are

Hydrogen cycle cation—exchanger no. 1

Removal

$$H_2R + M^{2+} \rightleftharpoons MR + 2H^+$$

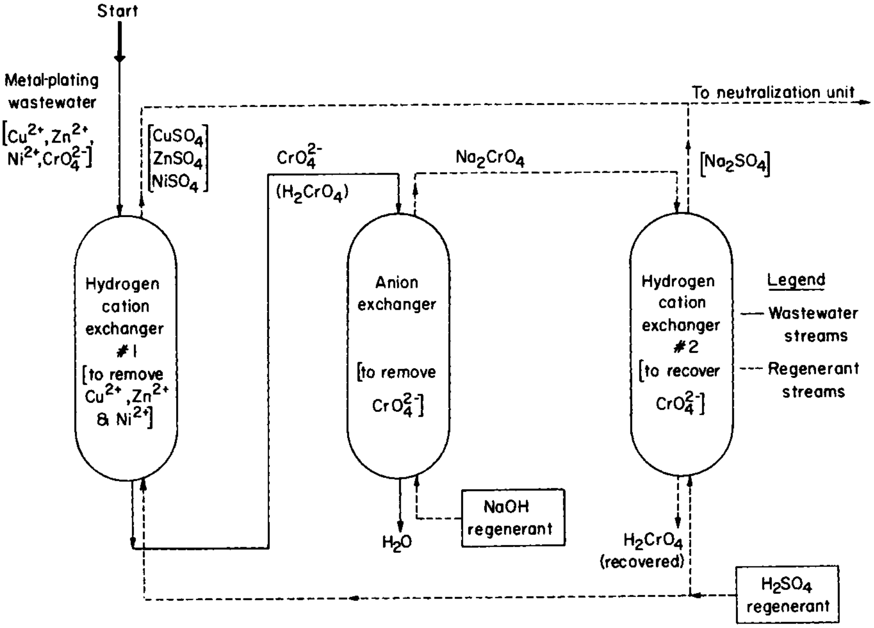


Fig. 8.9. Ion exchanger flow diagram (Example 8.3).

where $M^{2+} = Cu^{2+}$, Zn^{2+}, and Ni^{2+}.

Regeneration

$$MR + H_2SO_4 \rightarrow H_2R + MSO_4$$

Anion exchanger

Removal

$$R(OH)_2 + H_2CrO_4 \rightleftharpoons RCrO_4 + 2H_2O$$

Regeneration

$$RCrO_4 + 2NaOH \rightleftharpoons R(OH)_2 + Na_2CrO_4$$

Hydrogen cycle cation—exchanger no. 2

Removal

$$H_2R + Na_2CrO_4 \rightleftharpoons Na_2R + H_2CrO_4 \quad \text{(recovered)}$$

Regeneration

$$Na_2R + H_2SO_4 \rightleftharpoons H_2R + Na_2SO_4$$

Design of hydrogen cycle cation—exchanger no. 1

Step 1. Calculate equivalents of metal ions to be removed (see tabulation below).

Ion concentration	Equivalent wt	meq/liter
30 mg/liter Cu^{2+}	63.5/2 = 31.7	30/31.7 = 0.95
15 mg/liter Zn^{2+}	65.4/2 = 32.7	15/32.7 = 0.46
20 mg/liter Ni^{2+}	58.7/2 = 29.4	20/29.4 = 0.68
		Total = 2.09

Step 2. Determine the total equivalents/day to be removed.

$$2.09 \text{ meq/liter} \times 3.78 \text{ liter/gal} \times 120{,}000 \text{ gal/day} \times \text{eq}/1000 \text{ meq} = 948 \text{ eq/day}$$

Step 3. Calculate total resin requirements on the basis of 70 equivalents/ft^3 of resin (Table 8.4) and 6-day operation between regenerations.

$$\text{Resin requirement} = \frac{948 \text{ eq/day} \times 6 \text{ days/cycle}}{70 \text{ eq/ft}^3} = 81 \text{ ft}^3 \text{ of resin/cycle}$$

Step 4. Select a column diameter of 3 ft and calculate required depth of resin bed.

$$\text{Cross section} = (\tfrac{1}{4})\pi 3^2 = 7.07 \text{ ft}^2$$

$$\text{Depth} = 81/7.07 = 11.5 \text{ ft}$$

Allowing 50% free space for bed expansion for backwashing and cleaning, the height of the required column is (1.50)(11.5) = 17.3 ft. Utilize two columns in series, each 8.5 ft in height, each containing a 11.5/2 = 5.75-ft resin depth (free space, 2.75 ft for each column; ratio, 8.5/5.75 = 1.48).

Step 5. Calculate regenerant requirements. Regenerant is a 5% solution of H_2SO_4 as indicated in Table 8.4, and 11.0 lb H_2SO_4/ft^3 of resin are required. The lb of H_2SO_4 required are

$$11 \text{ lb/ft}^3 \times 81 \text{ ft}^3 = 891 \text{ lb of } H_2SO_4 \text{ per cycle}$$

or

$$891(100/5) = 17{,}820 \text{ lb of } 5\% \text{ solution}$$

Step 6. Calculate rinse water requirements. From Table 8.4, 130 gal of water are required for rinsing each cubic foot of resin. Required rinse water is

$$130 \text{ gal/ft}^3 \times 81 \text{ ft}^3\text{/cycle} = 10{,}530 \text{ gal per cycle}$$

Design of anion exchanger

Chromic acid (H_2CrO_4) passing through the cation unit is removed in the anion exchange unit, which is designed as follows:

Step 1. Total chrome removed per day (as CrO_3) is

$$120 \text{ mg/liter} \times 3.78 \text{ liter/gal} \times \text{g}/10^3 \text{ mg} \times \text{lb/454 g} \times 120{,}000 \text{ gal/day} = 120 \text{ lb/day}$$

Step 2. Calculate resin requirements. Basis: 3.5 lb CrO_3/ft^3 of resin (Table 8.4) and 6-day operation between regenerations.

$$\text{Resin requirement} = \frac{120 \text{ lb/day} \times 6 \text{ days/cycle}}{3.5 \text{ lb/ft}^3} = 206 \text{ ft}^3 \text{ of resin/cycle}$$

Step 3. Select a column diameter of 3 ft and calculate required depth of resin bed.

$$\text{Cross section} = (\tfrac{1}{4})\pi 3^2 = 7.07 \text{ ft}^2, \quad \text{depth} = 206/7.07 = 29 \text{ ft}$$

Allowing a 50% free space for bed expansion, the height of the required column is $(1.50)(29.0) = 43.5$ ft. Utilize four columns in series, each 11 ft in height ($11 \times 4 = 44$), and each containing $29/4 = 7.25$-ft resin depth (free space, 3.75 ft for each column; ratio, $11/7.25 = 1.5$).

Step 4. Calculate regenerant requirements. Regenerant is a 10% solution of NaOH as indicated in Table 8.4, and 4.7 lb of NaOH solution are required per ft^3 of resin. The lb of NaOH required are

$$4.7 \text{ lb/ft}^3 \times 206 \text{ ft}^3 = 968 \text{ lb of NaOH per cycle}$$

or

$$968(100/10) = 9680 \text{ lb of } 10\% \text{ solution}$$

Step 5. Calculate rinse water requirements. From Table 8.4, 100 gal of water are required for rinsing each cubic foot of resin. Required rinse water is

$$100 \text{ gal/ft}^3 \times 206 \text{ ft}^3 = 20{,}600 \text{ gal}$$

Design of hydrogen cycle cation—exchanger no. 2

Chromic acid (H_2CrO_4) is recovered from spent regenerant leaving the anion exchanger (which contains Na_2CrO_4) by passing it through a cation exchanger (Fig. 8.9). As calculated in Step 4 of the design for the anion exchanger, 968 lb of NaOH are required.

Step 1. Calculate sodium hydroxide equivalents to be removed by the cation exchanger.

$$968\ \text{lb} \times 454\ \text{g/lb} \times \text{eq}/40\ \text{g} = 10{,}987\ \text{eq}$$

since the equivalent weight of NaOH is 40 g.

Step 2. Calculate resin requirements on the basis of 70 equivalents/ft^3 of resin (Table 8.4).

$$\text{Resin requirement} = \frac{10{,}987\ \text{eq}}{70\ \text{eq/ft}^3} = 157\ \text{ft}^3\ \text{of resin}$$

Step 3. Select a column diameter of 3 ft, and calculate required depth of the resin bed.

$\text{Cross section} = (\tfrac{1}{4})\pi 3^2 = 7.07\ \text{ft}^2$

$\text{Depth} = 157/7.07 = 22.2\ \text{ft}$

Allowing 50% free space for bed expansion, the height of the required column is $(1.50)(22.2) = 33.3$ ft. Utilize three columns in series, each $33.3/3 \approx 11$ ft deep, each containing a $22.2/3 = 7.4$-ft resin depth (free space, 3.6 ft for each column; ratio, $11/7.4 = 1.48$).

Step 4. Calculate regenerant requirements. Regenerant is a 5% solution of H_2SO_4 as indicated in Table 8.4, and 11.0 lb of H_2SO_4/ft^3 of resin are required. The lb of H_2SO_4 required are

$$11\ \text{lb/ft}^3 \times 157\ \text{ft}^3 = 1727\ \text{lb}\ H_2SO_4$$

or

$$1727(100/5) = 34{,}540\ \text{lb of 5\% solution}$$

Step 5. Calculate rinse water requirements. From Table 8.4, 130 gal of water are required for rinsing each cubic foot of resin. Required rinse water is therefore

$$130\ \text{gal/ft}^3 \times 157\ \text{ft}^3 = 20{,}410\ \text{gal}$$

5. Reverse Osmosis

5.1. OSMOSIS AND OSMOTIC PRESSURE

Although osmotic phenomena have been known for over 200 years, the first precise experiments linking osmotic pressure to temperature and solute concentration were performed in the late 1800's by Pfeffer. A typical experiment with a sucrose solution is illustrated in Fig. 8.10.

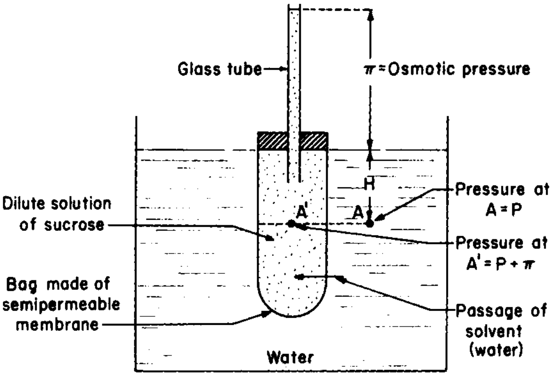


Fig. 8.10. Osmosis experiment.

The bag shown in Fig. 8.10 is made of a membrane permeable to the solvent (water in Fig. 8.10) but impermeable to the solute (sucrose). These are known as semipermeable membranes. Early semipermeable membranes utilized in osmosis were animal membranes (e.g., pig bladders). Synthetic membranes were developed later, cellulose acetate membranes now being the most widely used.

A dilute solution of sucrose (e.g., a 0.001 M solution) is placed inside the semipermeable bag, which is then dipped into a vat containing pure water. The water diffuses spontaneously from the vat to the interior of the semipermeable bag, as indicated by the arrow. As a result, a column of liquid rises through the glass tube connected to the dilute sucrose solution, reaching at equilibrium a height π above the level of water in the vat. At this moment, passage of solvent stops. Pressure exerted on points A' and A situated at the same elevation differs by the increment corresponding to height π of liquid. This value is called the osmotic pressure of the sucrose solution. Osmosis is defined as the spontaneous passage of a solvent from a dilute solution (pure water in the case of Fig. 8.10) to a more concentrated one through a semipermeable membrane.

Let P be the pressure at point A (atmospheric pressure plus pressure corresponding to a column of water of height H). The pressure at A' is $(P+\pi)$. Osmotic pressure π is a function of the concentration of the sucrose solution and temperature. The mathematical relationship for π as a function of concentration of solute (c) and absolute temperature (T) is given by the van't Hoff equation.

5.2. VAN'T HOFF EQUATION FOR OSMOTIC PRESSURE

Derivation of the van't Hoff equation is found in standard physical chemistry texts [3]. The equation is

$$\pi = nRT/V = cRT \tag{8.43}$$

where π is the osmotic pressure (atm); n the g mole of solute (e.g., sucrose); V the volume of the sucrose solution; $n/V = c$ the concentration of sucrose solution (g mole/liter); R the ideal gas constant [0.082 (atm)(liter)/(g mole)(°K)]; and T the absolute temperature (°K).

The van't Hoff equation shows a startling similarity to the ideal gas law, the solvent corresponding to the empty space between gas molecules and these latter corresponding to the molecules of solute, in the case of osmosis. Thus, one could consider osmotic pressure to be the result of bombardment exerted by molecules of solute on the membrane. Despite this analogy, it is deceptive to consider osmotic pressure as a sort of pressure which is exerted by the solute. Rather osmosis is the passage of solvent through the membrane due to momentary inequality of the chemical potential on the two sides of the membrane [3.] Osmotic pressure results from this passage of solvent.

Application of Eq. (8.43) to a 0.001 M solution of sucrose (i.e., $n/V = c = 0.001$ g mole/liter) at 20°C leads to a value of the osmotic pressure calculated as

$$\begin{aligned}\pi = cRT &= 0.001 \text{ g mole/liter} \times 0.082 \text{ (atm)(liter)/(g mole)(°K)} \times 293.2\text{°K} \\ &= 0.024 \text{ atm}\end{aligned}$$

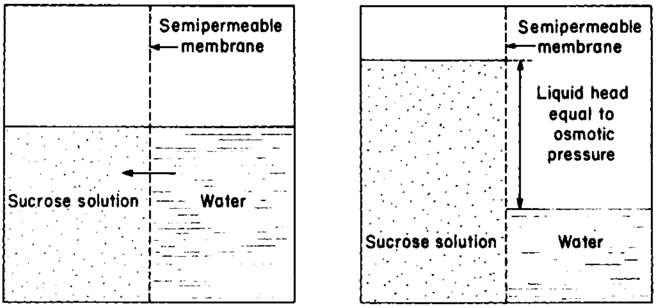
Assuming the specific gravity of the dilute solution to be that of pure water, this corresponds to a height π (Fig. 8.10) equal to

$$\pi = 0.024 \text{ atm} \times 34 \text{ ft water/atm} = 0.82 \text{ ft} \quad (\approx 10 \text{ in.})$$

If the van't Hoff equation could be applied to relatively concentrated sucrose solutions (e.g., a 1.0 M solution), height π would be 1000 times that just calculated, i.e., 820 ft. Inasmuch as the ideal gas law does not describe accurately gas behavior at higher pressures, the van't Hoff equation is not an adequate model for osmotic pressure at higher solute concentrations.

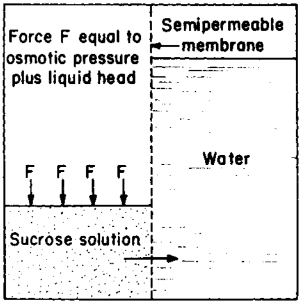
5.3. PRINCIPLE OF REVERSE OSMOSIS

The principle of reverse osmosis is illustrated by Fig. 8.11. Figure 8.11(a) depicts direct osmosis [e.g., condition existing at beginning of experiment with sucrose solution (Fig. 8.10)]. Solvent flows spontaneously through the semipermeable membrane. Figure 8.11(b) illustrates the equilibrium condition. Here the liquid head which has developed as a result of solvent flow through the membrane is equal to the osmotic pressure. Solvent flow stops. Figure 8.11(c) illustrates what happens when a force F in excess of the value of osmotic pressure is applied to the sucrose solution. Solvent flow is reversed, i.e., from the compartment containing the sucrose solution to the water compartment. This phenomenon is called reverse osmosis (hence abbreviated as RO).



(a)
Direct osmosis:
solvent flows spontaneously through the semipermeable membrane

(b)
Equilibrium condition:
solvent flow stops



(c)
Reverse osmosis:
requires applied force in excess of the osmotic pressure

Fig. 8.11. Illustration of the principle of reverse osmosis.

In the treatment of wastewaters by reverse osmosis, contaminated influent is placed in contact with a suitable membrane at a pressure in excess of the osmotic pressure of the solution [same situation as in Fig. 8.11(c), except that the left compartment contains wastewater instead of sucrose solution]. Under these conditions, water with a very small amount of contaminants permeates the membrane. Dissolved contaminants are concentrated in the wastewater compartment. This concentrate, which hopefully is a small fraction of the total volume of wastewater to be treated, is discarded. Purified water is obtained from the other compartment.

The compartments indicated in Fig. 8.11 are a schematic representation of RO process. In practice, the RO process is conducted in a tubular configuration system (Fig. 8.12). Raw wastewater flows under high pressure (in excess of the value of its osmotic pressure) through an inner tube made of a semipermeable membrane material and designed for high pressure operation. Purified water is removed from the outer tube, which is at atmospheric pressure and is made of ordinary tubular material. Typical values of operating pressures, water fluxes (yield of purified water per unit area of membrane), and product quality are discussed in the following sections.

5.4. PREPARATION OF RO MEMBRANES

Reverse osmosis found its earlier applications in desalination of ocean water. Considerable research and pilot-plant work are being done for utilization of RO in removal of contaminants from wastewaters. Some of these (e.g., nitrogen and phosphorus compounds, chromates, and some organic compounds) are not adequately removed by other processes. Consequently, usual processes may be complemented by RO, provided economic considerations are favorable. Research indicates that in principle, RO can be used to obtain an effluent of virtually any desired degree of purity while still maintaining reasonable flow rates.

Many natural materials have semipermeable characteristics. Animal and plant membranes are well-known examples. Collodion, cellophanes, porous glass frits, finely cracked glass, and inorganic precipitates such as copper ferrocyanide, and zinc and uranyl phosphates have been used. All these, however, have the shortcomings of developing leaks and exhibiting short-lived selectivity and poor reproducibility.

Cellulose acetate membranes (hence denoted as CA membranes) are the most successful semipermeable membranes developed. Sourirajan and Loeb [9] developed a technique for preparation of CA membranes yielding both high permeabilities and high degrees of solute separation from aqueous solutions of sodium chloride. Their technique is as follows:

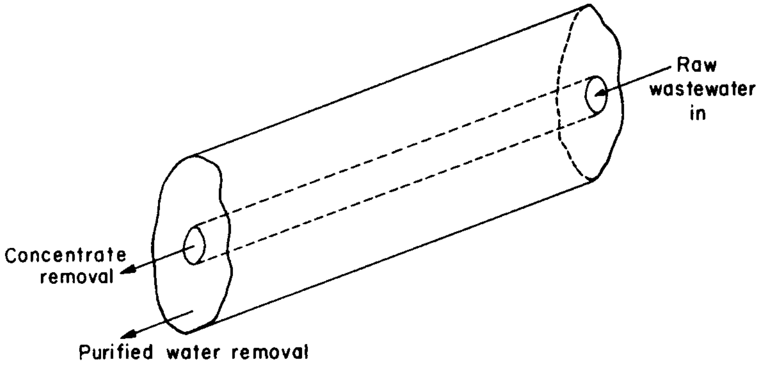


Fig. 8.12. Diagram of a tubular configuration system for wastewater treatment by reverse osmosis.

1. *Casting step*. The film-casting solution contains cellulose acetate dissolved in acetone, to which is added an additive soluble in water and not affecting the solubility of cellulose acetate in acetone (e.g., magnesium perchlorate). With this solution, membranes are cast on flat or tubular surfaces (e.g., glass plates or tubular surfaces), either at laboratory or lower temperatures ($\approx -10°\text{C}$). One of the most significant developments in the field of CA membrane technology is casting in tubular form. Tubular-shaped membranes are entirely lined within a porous fiberglass reinforced tube.

2. *Evaporation step*. After casting, part of the solvent (acetone) is allowed to evaporate from the surface of the membrane at casting temperature.

3. *Gelation step*. The membrane is immersed in ice-cold water for at least 1 hr. The film sets to a gel, from which the additive (e.g., magnesium perchlorate) and the solvent (acetone) are leached out, leaving a tough solid porous film on the flat or tubular surface.

4. *Shrinkage step*. Membranes from the gelation step exhibit pores which are too large to permit efficient operation (diameters ≈ 4000 Å). These large pores are a result of the leaching process. Consequently, the membrane receives a thermal treatment by shrinking it in hot water for about 10 min. Adjusting the hot water temperature, it is possible to obtain variable porosities, which results in different degrees of wastewater separation. Higher hot water temperatures (usual range is 70°–98°C) yield smaller pores.

5.5. PREFERENTIAL SORPTION–CAPILLARY FLOW MECHANISM FOR REVERSE OSMOSIS

Several mechanisms have been proposed by different investigators to explain reverse osmosis. Of these, only the preferential sorption–capillary flow mechanism is described here. This mechanism, proposed by Sourirajan [9], is summarized as follows: RO separation is the combined result of an interfacial phenomenon and fluid transport under pressure through capillary pores. Figure 8.13 is a conceptual model of this mechanism for recovery of fresh water from aqueous salt solutions (ocean water).

The solution is in contact with a porous membrane, the surface of which has a preferential sorption for water and/or preferential repulsion for the solute. A continuous removal of preferentially sorbed interfacial water is effected by flow under pressure through the membrane capillaries. The preferentially sorbed water layer at the interface is of a monomolecular nature (indicated in Fig. 8.13 below the single dotted line), and results from interaction between interfacial surface tension and adsorption of solute.

For a maximum separation and permeability, this model gives rise to the concept of critical pore diameter, which is equal to twice the thickness of the preferentially sorbed interfacial water layer (Fig. 8.14). From an industrial standpoint, application of the reverse osmosis technique for a given separation

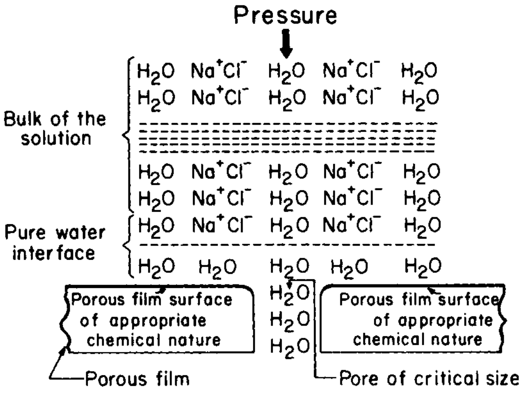


Fig. 8.13. Schematic representation of preferential sorption–capillary flow mechanism **[9]**. (*Reprinted with permission from Ind. Eng. Chem. Copyright by the American Chemical Society.*)

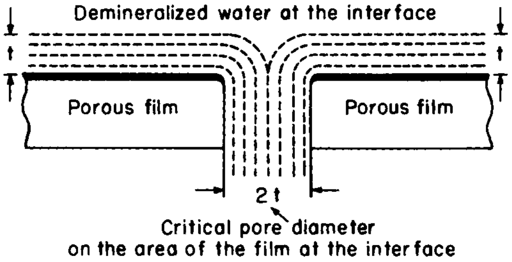


Fig. 8.14. Critical pore diameter for maximum separation and permeability **[9]**. (*Reprinted with permission from Ind. Eng. Chem. Copyright by the American Chemical Society.*)

problem involves the choice of the appropriate chemical nature of the film surface and development of methods for preparing films containing the largest number of pores of the required size. This approach is the basis of the successful development of the Sourirajan–Loeb type of porous CA membranes for saline water conversion and other applications.

5.6. CHARACTERIZATION OF MEMBRANE PERFORMANCE

The two basic parameters for characterizing RO systems are (1) production of purified water per unit area of membrane (water flux) and (2) product quality, i.e., purity of purified water (rejection factor). These parameters are discussed in Sections 5.7 and 5.8.

5.7. WATER FLUX

Production of purified water is measured by the water flux, defined as quantity of product recovered per day per unit area of membrane. English units are used for water flux in field work [gal/(day)(ft^2)], whereas metric units [g/(sec)(cm^2)] are used in laboratory tests.

Flux through a specific membrane is determined by (1) physical characteristics of the membrane, e.g., thickness, chemical composition, and porosity, and (2) system conditions, e.g., temperature, differential pressure across the membrane, salt concentration of solutions touching the membrane, and velocity of feed moving across the membrane.

In practice, physical characteristics of the membrane as well as temperature and concentrations of solute in feed and product streams are fixed for a given process. Therefore, water flux is a function of the differential pressure across the membrane, being given approximately by

$$F_{H_2O} \approx A(\Delta P - \Delta\pi) \tag{8.44}$$

where

$$\Delta P = P_F - P_P \tag{8.45}$$

$$\Delta\pi = \pi_F - \pi_P \tag{8.46}$$

and F_{H_2O} is the water flux [gal/(day)(ft^2)]; A the permeation coefficient for a unit area of membrane [gal/(day)(ft^2)(atm)]. This term includes physical variables of the membrane and is relatively constant; $\Delta P = (P_F - P_P)$ the pressure exerted on feed solution (P_F) minus pressure on product (P_P) (atm); and $\Delta\pi = (\pi_F - \pi_P)$ the osmotic pressure of feed solution (π_F) minus osmotic pressure of product (π_P) (atm).

For a laboratory experiment with a feed of pure water $\Delta\pi = 0$, Eq. (8.44) reduces to a classical flux equation:

$$F_{H_2O} = A\,\Delta P \tag{8.47}$$

i.e., flux = resistance × driving force. When wastewater feed is relatively concentrated in solute and product is a very dilute solution (nearly pure water), correction factor $\Delta\pi$ is subtracted from differential pressure ΔP. Actually, $\Delta\pi$ should be equal to the difference in osmotic pressure between solutions touching the membrane on each side (i.e., feed and product sides). This is *not* exactly equal to $\pi_F - \pi_P$ because of concentration polarization, a condition discussed in Section 5.15. This is why Eq. (8.44) is approximate. An accurate version is written in Section 5.15.

5.8. REJECTION FACTOR

Improvement of quality between feed and product streams is expressed quantitatively by the rejection factor, defined as

$$f = (C_F - C_P)/C_F \tag{8.48}$$

where f is the rejection factor (dimensionless); C_F the solute concentration in feed solution; and C_P the solute concentration in product.

Thus, $f = 0.9$ means that $C_P = 0.1C_F$, i.e., the product contains one-tenth the concentration of solute in the feed; therefore, 90% of the solute is rejected by the membrane. Rejection factor is therefore a measure of membrane selectivity. Solute concentrations C_F and C_P are obtained by determination of total dissolved solids (evaporation to dryness). When the solute is an electrolyte (e.g., NaCl), analysis is conveniently performed by conductivity measurements.

The ability of a membrane to reject solutes is a complicated problem depending on a combination of physicochemical characteristics of solute, membrane, and water. Properties of the solute which have the most influence on rejection of individual species are (1) valence charge—rejection increases with value of charge of ion; (2) molecular size—rejection increases with molecular size of solute; and (3) hydrogen bonding tendency—permeation increases with strength of hydrogen bonding. Ability of a membrane to reject salts decreases with operating time. Variation of rejection with time is illustrated by Fig. 8.15.

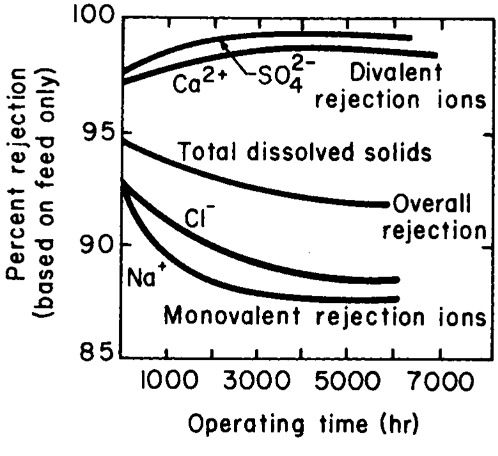


Fig. 8.15. Variation of salt rejection with operating time **[6].** *(Reprinted by special permission from Chemical Engineering, April 1973. Copyright by McGraw-Hill, Inc., New York, N.Y., 10020.)*

At first, decrease is more pronounced for small, univalent ions such as sodium (Na^+) and chloride (Cl^-). These are normally among the most permeable, showing the lowest initial rejection and have the highest rate of decline. Divalent ions such as calcium (Ca^{2+}) and magnesium (Mg^{2+}) and anions such as sulfate ($SO_4{}^{2-}$) are initially rejected very well and show a very slow rate of decline. Rejection of such ions may actually increase for a while (Fig. 8.15), becoming nearly constant after that. Overall salt rejection tends to follow the monovalent curve.

Progressive decrease in salt rejection may be caused by hydrolysis of the membrane, with subsequent loss of bonding sites. Another cause may be increase in pore size due to membrane swelling. Most probably, both effects play a part in the result.

5.9. EFFECT OF SHRINKAGE TEMPERATURE ON PERFORMANCE OF CA MEMBRANES

Adjusting hot water temperature in the shrinkage step for preparation of CA membranes (Section 5.4) permits control of membrane porosity, thus leading to various degrees of wastewater separation. At higher shrinkage temperatures, pore sizes obtained are smaller, leading to greater rejection. Water flux, however, decreases at higher shrinkage temperatures, as expected. For specific RO applications one seeks an economic balance between water flux and rejection. Figure 8.16 shows rejection and flux data obtained by Kopecek and Sourirajan for three types of CA membranes.

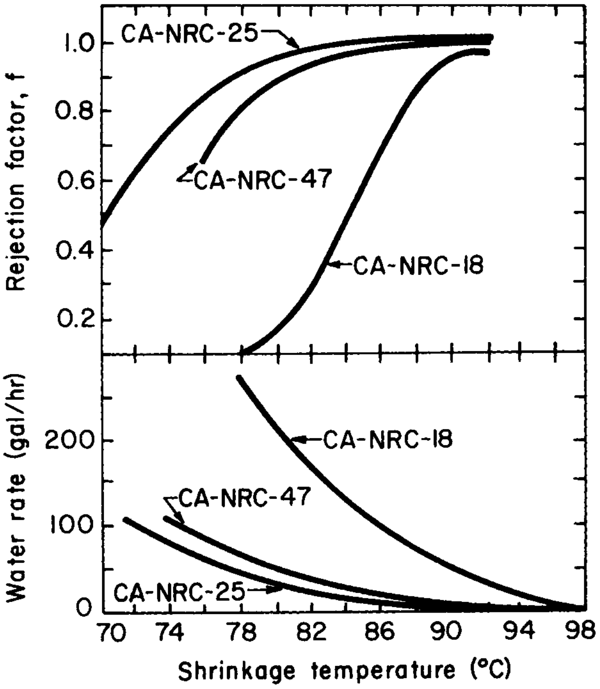


Fig. 8.16. ***Effect of shrinkage temperature on the performance of different CA membranes*** **[9].** (*Reprinted with permission from Ind. Eng. Chem. Copyright by the American Chemical Society.*)

5.10. EFFECT OF FEED TEMPERATURE ON FLUX [6]

Flux is also affected by feed temperature. Water permeability for the membrane increases about 1.5% per °F. Flux for a membrane is usually specified at 75°–77°F, and a correction factor is applied at other temperatures. This correction factor can be derived by theoretical considerations from diffusivity and viscosity values, but experimentally determined corrections are more reliable.

A correction factor which is a multiplier of the required membrane area is presented in Fig. 8.17. This curve was developed by Gulf Environmental Systems Co. for modified cellulose acetate membranes. Application of the correction factor is illustrated in Example 8.4.

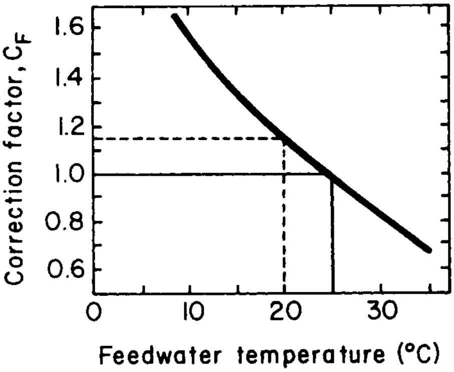


Fig. 8.17. Flux/temperature correction curve **[6].** (*Reprinted by special permission from Chemical Engineering, April 2, 1973. Copyright by McGraw-Hill, Inc., New York, N.Y., 10020.*)

Example 8.4 [6]

It is desired to specify the membrane area for a 100,000 gal/day RO system to treat brackish water. Records indicate that the lowest water temperature expected for any prolonged period is 68°F (20°C). The cellulose acetate membrane chosen is expected to have an average flux of 15 gal/(day)(ft^2) during its 2-year life, operating at 600 psig. Average flux was determined at a base temperature of 77°F. Determine required membrane area for operation at 68°F.

SOLUTION Membrane area is shown in Eq. (8.49).

$$\text{Membrane area (ft}^2) = \frac{\text{water production in gal/day}}{\text{water flux in gal/(day)(ft}^2)} \tag{8.49}$$

At 77°F the required membrane area is

$$\text{Area (77°F)} = \frac{100{,}000 \text{ gal/day}}{15 \text{ gal/(day)(ft}^2\text{)}} = 6667 \text{ ft}^2$$

For operation at 68°F (20°C), the correction factor C_F from Fig. 8.16 is $C_F = 1.15$. Therefore,

$$\text{Area (68°F)} = 1.15 \times 6667 = 7667 \text{ ft}^2$$

If additional product is not required when temperature is above 68°F, operating pressure is reduced.

5.11. FLUX DECLINE [6]

Application of pressure to the membrane results in compaction and consequently in a decline of flux. Compaction results from densification of the thin membrane layer and corresponds to narrowing of the pores through which water must pass. As the channels narrow, flow decreases.

Another cause of flux decline is hydrolysis of acetyl groups which takes place during the life of the membrane. Therefore, RO membranes are limited to a pH operating range of 3–7, outside of which rapid hydrolysis and membrane degradation occur. The optimum range is believed to be pH 5–6. Hydrolysis of acetyl groups results in loss of hydrogen bonding sites, which reduces the water transport. This hydrolysis is also a source of salt leakage because there are fewer water bridges blocking passage of foreign materials through the pores.

Loss in productivity happens slowly to every membrane and is permanent. Chemical rejuvenation and low pressure operation to relax and swell the matrix have been tried without success; the membrane simply ages and flux decreases until economics dictate replacement.

5.12. FOULING: CAUSES AND CURES [6]

Fouling, resulting in temporary flux reduction, is caused by foreign materials coating the membrane surface, as only hydrogen bonding substances (water, ammonia) pass through the discriminating pores of the membrane. Nonbonding materials are left in the quiescent film known as the liquid boundary layer.

The composition of deposits in boundary layers reflects the composition of feedwater. As expected, the most common constituents are calcium carbonate, sulfate scales, hydrates of iron and aluminum oxides, silicates, miscellaneous particulates, and biological growths.

Fouling is minimized by taking the following precautions: (1) pretreating feed to remove iron and control pH, (2) limiting the process to nonscaling concentrations of wastewater, (3) filtration of wastewater feed, and (4) injection of small amounts of biocides (e.g., chlorine). Unfortunately, no matter how thorough the protection, fouling always occurs and the membranes should be periodically cleaned. The usual cleaning procedure is as follows:

1. Flush the membrane with feedwater at reduced pressure of two or three times normal velocity; the turbulent action of the fluid loosens fouling deposits and carries them away. Water flushing is the preliminary step in every cleaning operation.

2. Hardness scales (carbonates and sulfates salts of calcium and magnesium) are sometimes removed by soaking the membrane in distilled water for long periods of time. Since this is time consuming, more often a warm solution of 1–2% citric acid is vigorously circulated through the unit, dissolving large amounts of metallic ions and keeping them in solution by chelation.

3. Microbiological growths occur in most natural waters, and are a particular problem when treating effluents from biological processes. These are often removed by recirculating washes of enzyme detergents.

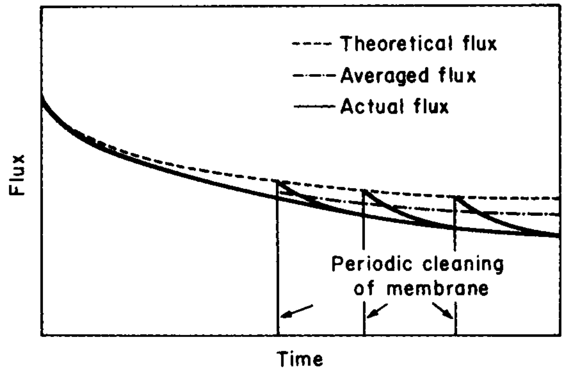


Fig. 8.18. ***Effect of membrane cleaning on flux*** [6]. (*Reprinted by special permission from Chemical Engineering, April 2, 1973. Copyright by McGraw-Hill, Inc., New York, N.Y., 10020.*)

Flux degradation by fouling is an additional loss superimposed on the permanent losses discussed in Section 5.11. This is shown in Fig. 8.18, which illustrates the effect of membrane cleaning on flux. As shown in Fig. 8.18, the actual flux curve follows a declining, saw-toothed pattern when the membrane is cleaned periodically. Without cleaning, flux would follow the lowest curve, a projection of the initial smooth decline. Flux from a hypothetical membrane that is never fouled (theoretical flux) is shown by the upper line, the theoretical flux curve which touches only the peaks of the saw-toothed curve.

5.13. PREDICTION OF FLUX [6]

Water output from a membrane begins to decline as soon as pressure is applied, and continues to degrade slowly thereafter. The loss is irreversible, and if more flux is required feed pressure must be increased. This alternative is self-defeating since additional pressure, while producing more water, also compresses the membrane further and hastens flux decline. Normal practice is to overspecify the membrane area slightly and to keep the operating pressure constant as long as possible, resorting to additional pressure late in the life of the membrane.

The output of a membrane is predictable because the decline per unit area of the membrane is quite uniform and can be projected. A plot of flux vs. operating time at a specified pressure (like the lower curve in Fig. 8.18) yields a curve with an initial steep descent followed by a prolonged and moderate decline. A logarithmic plot yields a straight line which is adequate for prediction purposes for 1 and probably 2 years. Thus, flux can be predicted from such plots once the initial flux and slope are known.

Manufacturers provide initial flux values and estimates of slopes at various operating pressures. Initial flux is the production for the first 24 hr divided by the membrane area in the test unit. Decline slope is computed or determined graphically from flux values taken at time intervals such as 10, 100, and 1000 hr. Time intervals in multiples of 10 are conveniently selected because computation is simplified. Decline rate is given by Eq. (8.50).

$$m = (\log F_i - \log F_x)/(\log t_i - \log t_x) = \log(F_i/F_x)/\log(t_i/t_x) \qquad (8.50)$$

where F_i is the initial flux [gal/(hr)(ft^2)]; F_x the flux at time x [gal/(hr)(ft^2)]; t_i the operating time (hr) for initial flux (obtained by computing initial flux and determining from a logarithmic plot of F_x vs. t the time t_i corresponding to this calculated value); and t_x the operating time (hr).

Determination of flux over the life of the membrane is basic to the design of RO systems, because these values are used to estimate membrane area required for a desired plant capacity by utilizing Eq. (8.49). Three possible approaches are

1. Take an average of the initial and final flux as a compromise to obtain an average area. When this choice is made, it is expected that during the later life of the membrane, flux will be brought up to its initial value by increasing the operating pressure.

2. Specify membrane area from final or smallest flux value. Although fixed costs are higher, lower system pressures decrease operating costs.

3. Select initial flux as the design basis. This minimizes membrane area

and capital costs, but raises operating costs as pressures are increased to maintain production. This is a logical choice for intermittent and short-term projects in which initial costs must be minimized.

5.14. MEMBRANE LEAKAGE [6]

Calculated solute rejection is always higher than experimental values, even those that are determined under the most careful conditions. For example, theoretical rejection of sodium chloride from a modified CA membrane is calculated as 99.7%, but experimental results show no better than 97–99% rejection. The difference is caused by minute imperfections in the membrane through which pressurized brine can flow and contaminate the product water.

All membranes have imperfections; these are probably not manufacturing faults, but a property of the membrane that must be adjusted or optimized to suit a specific service. Hence, very porous CA membranes are used to screen out large molecules (20–500 Å diameter) and very small particles for ultrafiltration applications. Less porous membranes are selected for high water flow and moderate salt rejection service in saltless softeners, and membranes heat treated to low porosity are used for applications requiring high rejection.

Fortunately, most imperfections are small and easily plugged. Distribution runs from numerous holes with diameters 100 Å and smaller to a few of 1-μm in diameter. The major source of product contamination results from solute passage through larger holes, since from Poiseuille's law for viscous flow, salt leakage increases proportionately to the fourth power of pore diameter.

A number of techniques have been tried to reduce membrane leakage:

1. *Heat treating and modifying dope formula.* This is the only permanent way to reduce leakage. Membranes for sea water must be heat treated at high temperatures to yield a film that can reject 99.5% of the salt in the feed.

2. *Addition of certain chemicals to the feed.* Chemicals of large molecular size added to the feed are utilized to plug leaking pores. Loeb [7] discovered that trace amounts of aluminum salts occurring in Los Angeles tap water plugged leaking pores of test membranes and improved salt rejection. Zephiran (tetraalkylaluminum chloride) was used at University of California at Los Angeles (UCLA) to gain a similar and more reproducible effect. Other materials that improve rejection include polyvinyl methyl ether and Dowfax.

Unfortunately, leak-stopping additives have serious drawbacks that limit their use: (1) they are more effective on low flux than on high flux standard CA membranes, now almost exclusively used; (2) they dissipate quickly and must be regularly replenished; (3) most of them reduce water flux as they reduce solute leakage; and (4) they are expensive and therefore useful only in special situations where cost is not a factor.

5.15. SOLUTE PERMEABILITY AND CONCENTRATION POLARIZATION

Theoretically, solute flux is a function of membrane permeability and the difference between solute concentrations in the high and low pressure sides of the membrane, i.e.,

$$F_{\text{solute}} = \beta(C_H - C_L) = \beta\,\Delta C \approx \beta C_H \tag{8.51}$$

where F_{solute} is the solute flux [g/(cm^2)(sec)]; β the solute permeability coefficient (cm/sec); C_H the concentration of solute on high pressure side of membrane, i.e., concentrate side (g/cm^3); and C_L the concentration of solute on low pressure side of membrane, i.e., product side (g/cm^3).

Unlike water flux [Eq. (8.44)], normal solute flux is independent of pressure [Eq. (8.51)]. Theoretically, if pressure in the RO system is increased, solute diffuses at a constant rate while water flow increases. The result is greater production of pure water.

Since C_L is usually small as compared to C_H, Eq. (8.51) is written approximately as indicated, i.e., salt flux is essentially governed by concentration of solute in the boundary layer next to the membrane on the concentrate side. Solute concentration C_H can be substantially higher than the concentration in the feed C_F owing to an effect called concentration polarization. In membrane processes solute accumulates in a relatively stable layer (boundary layer) next to the membrane. Concentration polarization is the ratio of solute concentration at this boundary layer to that in the bulk of the solution. Initially, solute concentration at the boundary layer is the same as in the bulk of the solution. However, since the membrane is permeable to solvent and impermeable to solute, the boundary layer becomes heavily populated with solute as solvent leaves through the channels of the membrane. The boundary layer grows thicker and more concentrated, because the rate of solute diffusion away from the membrane cannot keep pace with solvent flow through the membrane.

The result of concentration polarization follows from Eq. (8.51). Write $C_H = C_{\text{BL}}$, where C_{BL} stands for the progressively increasing solute concentration at the boundary layer. Therefore

$$F_{\text{solute}} \approx \beta C_{\text{BL}} \tag{8.52}$$

Consequently, concentration polarization results in an increased solute flux or a lower product quality. This is undesirable, so one strives to reduce the concentration polarization effect. This is accomplished in two ways:

1. *Higher feed velocity.* This reduces thickness and concentration of the boundary layer as it is scoured away by the feed stream at high velocity. The extra flow through the unit results in an overall lower product recovery, i.e., smaller ratio of product to feed. This increases power consumption and amount of concentrate (waste) produced.

2. *Addition of turbulence promoters.* This is a more efficient way to reduce the boundary layer. Tubular membrane units are provided with plastic balls on the high pressure side to break up the smooth flow of feed solution. Turbulence promoters may require prefiltration of the feed so that particulates or precipitates do not get lodged in narrow flow spaces and plug the membrane. The usual concentration polarization ratio is 1.2–2.0, which means that solute concentration at the boundary layer is 1.2–2.0 times that in the bulk of the feed.

The approximate equation for water flux, Eq. (8.44), is rewritten taking into account the concentration polarization effect, in which case $\Delta\pi$ is

$$\Delta\pi = \pi_{\mathrm{BL}} - \pi_P \qquad (8.53)$$

where π_{BL} is the osmotic pressure of the concentrated solution at the boundary layer ($\pi_{\mathrm{BL}} > \pi_F$). Since $\Delta\pi$, given by Eq. (8.53), is higher than the value calculated from Eq. (8.46), this implies from Eq. (8.44) a decreased water flux owing to the concentration polarization effect. Substitution of Eqs. (8.53) and (8.45) in Eq. (8.44) yields

$$F_{\mathrm{H_2O}} = A[(P_F - P_P) - (\pi_{\mathrm{BL}} - \pi_P)] \qquad (8.54)$$

There is no simple experimental technique available to determine solute concentration at the boundary layer (and therefore π_{BL}). An indirect approach consists of determining experimentally water flux, $F_{\mathrm{H_2O}}$, and permeation coefficient A (Section 5.16). Then π_{BL} is the only unknown in Eq. (8.54), and therefore

$$\pi_{\mathrm{BL}} = \pi_P + (P_F - P_P) - (F_{\mathrm{H_2O}}/A) \qquad (8.55)$$

The concentration of solute at the boundary layer is then obtained from a previously prepared plot of osmotic pressure vs. solute concentration [$\pi = f(c)$].

5.16. EXPERIMENTAL TECHNIQUE FOR LABORATORY PREDICTION OF MEMBRANE PERFORMANCE

A laboratory apparatus for the prediction of membrane performance in reverse osmosis is described by Sourirajan and Agrawal [9], and a schematic flow diagram is shown in Fig. 8.19. Feed solution is pumped through a surge tank into the cell containing the membrane. Pressure is controlled by a pressure regulator operating on a nitrogen back pressure system.

Experimentally determined variables at a specific pressure are (1) rejection factor [Eq. (8.48)]; (2) product recovery, i.e., ratio of product to feed; and (3) permeation coefficient A. Permeation coefficient A, which includes physical variables of the membrane, is relatively constant (Section 5.7). Therefore, it

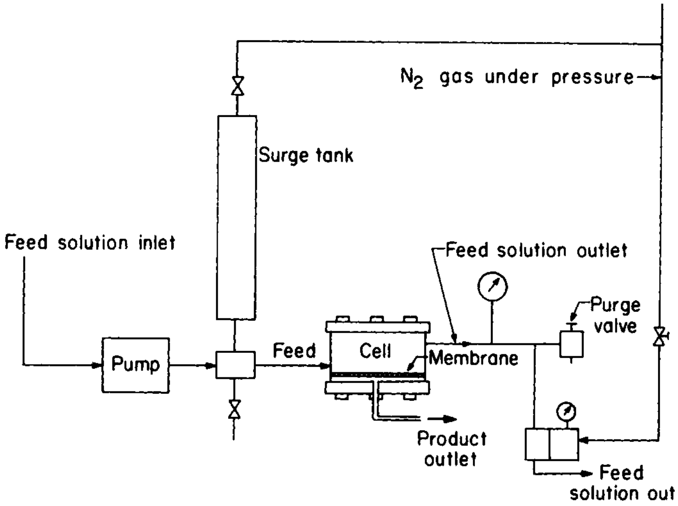


Fig. 8.19. Schematic flow diagram of RO laboratory unit [9]. (*Reprinted with permission from Ind. Eng. Chem. Copyright by the American Chemical Society.*)

can be determined once and for all by operating the system at a specific pressure with pure water and measuring the rate at which it permeates through the membrane. Equation (8.53) for pure water operation yields

$$\Delta\pi = \pi_{BL} - \pi_P = 0$$

From Eq. (8.54),

$$A = F_{H_2O}/\Delta P \tag{8.56}$$

where

$$\Delta P = P_F - P_P$$

and F_{H_2O} is the pure water flux [g/(hr) (cm^2)]. Product flow (g/hr) in this case is called pure water permeability (PWP). Therefore from Eq. (8.56),

$$A = PWP/(M \times S \times 3600\Delta P) \tag{8.57}$$

where A is the permeation coefficient [g mole of water/(sec)(cm^2)(atm)]; PWP the pure water permeability (g/hr); M the molecular weight of water (18); S the effective membrane area (cm^2) (7.6 cm^2 for Sourirajan's apparatus in Fig. 8.19); and ΔP the differential pressure (atm).

5.17. FINAL REMARKS ON REVERSE OSMOSIS

Reverse osmosis is still too expensive for widespread utilization in wastewater treatment. It is also limited to treating soluble wastes since suspended solids clog the membranes. Consequently, pretreatment of the feed is required whenever there are suspended solids, thus raising costs.

Operating pressures employed vary from 600 up to 1500 psig. One of the main goals in current research is to develop better membranes that can operate at lower pressures (around 250 psig) and still yield relatively high product rates, of the order of 40 gal/(day)(ft^2). Once these difficulties are overcome, RO could become a process of widespread application in the field of wastewater treatment.

6. Electrodialysis

6.1. INTRODUCTION

Electrodialysis was originally developed for desalination of ocean water. It is a promising method for removal of inorganic nutrients (phosphorus and nitrogen) from wastewaters, and thus it is a possible final stage in wastewater treatment processes.

A diagram of an electrodialysis cell is shown in Fig. 8.20. The basic components of the cell are a series of membranes made of ion exchange resins. These membranes are permeable only to ionic species and are selective to a specific type of ion. There are two types of membranes utilized in an electrodialysis cell: (1) cation membranes, which possess a fixed negative charge, allowing cations (positive ions) to pass through them but repelling anions (negative ions); and (2) anion membranes, which possess a fixed positive charge, allowing anions (negative ions) to pass through them but repelling cations (positive ions).

Passage of ions through the membranes is accelerated by application of a constant voltage across a series of cation- and anion-permeable membranes, as indicated in Fig. 8.20. The cathode and anode are located at two extremes of the cell, so that the membrane closest to the cathode is cation permeable and that closest to the anode is anion permeable. Raw wastewater is fed continuously into the concentrating compartments, and treated wastewater is withdrawn continuously from the diluting compartments.

6.2. VOLTAGE REQUIRED FOR ELECTRODIALYSIS

Voltage required is calculated from Ohm's law [Eq. (8.58)].

$$E = IR \tag{8.58}$$

where E is the applied voltage (V), I the current (A), and R the total electrical resistance of membranes and solutions in cells (ohms).

6.3. CURRENT REQUIRED FOR ELECTRODIALYSIS

The current required is proportional to the ionic strength of solution (expressed in terms of normality), and the number of cells. It is calculated

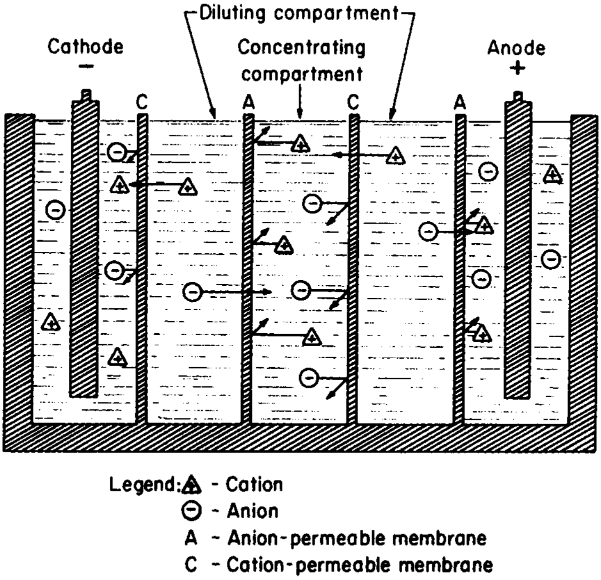


Fig. 8.20. Diagram of an electrodialysis cell.

from Faraday's law: 96,500 A × sec of electricity transfer one gram-equivalent weight of an electrolyte from one electrode to another. Thus, current required for electrodialysis for a fixed applied voltage E is given in Eq. (8.59).

$$I = FqNe/n\varepsilon \tag{8.59}$$

where I is the current (A), F the Faraday constant (96,500 A × sec/g equivalent), q the flow rate (liters/sec), N the normality of solution (g equivalents/liter), e the removal efficiency ($0 < e < 1.0$), n the number of cells between electrodes, and ε the current efficiency ($0 < \varepsilon < 1.0$).

Increasing the number of cells for a fixed applied voltage E results in an increase of the total electrical resistance R. Consequently, from Ohm's law the current I decreases.

6.4. PRETREATMENT OF WASTEWATERS IN ELECTRODIALYSIS

For proper operation of the electrodialysis cell, particulate matter, large organic ions, and colloidal matter must be removed prior to the process. If this is not done, these materials cause fouling of membranes, which results in an increase of total electrical resistance. For a constant applied voltage, the current passing through the cell is thus lowered. Therefore, demineralizing capacity of the equipment is lowered.

Fouling of membranes is the greatest problem to be overcome in order to achieve economic operation of electrodialysis in treatment of wastewaters. Fouling is minimized by

1. Pretreatment of raw wastewater in order to remove particulate and colloidal matter and large organic ions. This is done by addition of coagulants, filtration through microscreens, and/or adsorption in activated carbon columns. Cost of pretreatment may render the process uneconomical.
2. Fouling is minimized by periodic plant shutdown for cleaning.
3. Frequent current reversals tend to minimize the effects of fouling.

7. Chemical Oxidation Processes (Chlorination and Ozonation)

7.1. CHLORINATION OF WASTEWATERS

7.1.1. Utilization and Purposes of Chlorination

Chlorination is a widely used process in the treatment of domestic and industrial wastewaters. Some industrial effluents which are commonly chlorinated prior to discharge into receiving waters are those from beet sugar, cannery, dairy, pulp and paper, textile, tanning, petrochemical, pharmaceutical, and metal processing (chromium, electroplating) plants.

Purposes of chlorination are summarized as follows:

1. *Disinfection.* Primarily a disinfectant owing to its strong oxidizing capacity, chlorine destroys or inhibits growth of bacteria and algae.
2. *BOD reduction.* Chlorine accomplishes BOD reduction by oxidation of organic compounds present in wastewaters.
3. *Elimination or reduction of colors and odors.* Color and odor-producing substances present in wastewaters are oxidized by chlorine. The oxidizing ability of chlorine is employed for odor control and color removal in treatment of many industrial effluents (beet sugar, cannery, dairy, pulp and paper, textiles).
4. *Oxidation of metal ions.* Metal ions which are in a reduced state are oxidized by chlorine (e.g., ferrous to ferric ion and manganous to manganic ions).
5. *Oxidation of cyanides to innocuous products.* This application is described in Section 7.1.4.

7.1.2. Reactions of Chlorine in Water

Added to water as either a gas or solution, chlorine reacts to form hypochlorous acid (HOCl), which subsequently dissociates according to the following chemical equations:

$$Cl_2 + H_2O \rightleftharpoons HOCl + H^+ + Cl^-$$

$$HOCl \rightleftharpoons H^+ + OCl^-$$

In the presence of ammonia, hypochlorous acid reacts to form monochloramine, dichloramine, and nitrogen trichloride. Relative proportions of these products depend on pH and concentration of ammonia present. Corresponding chemical equations are

$$\mathrm{NH_3 + HOCl \rightleftharpoons NH_2Cl + H_2O}$$
(monochloramine)

$$\mathrm{NH_2Cl + HOCl \rightleftharpoons NHCl_2 + H_2O}$$
(dichloramine)

$$\mathrm{NHCl_2 + HOCl \rightleftharpoons NCl_3 + H_2O}$$
(nitrogen trichloride)

Reactions of chlorine in water are illustrated by Fig. 8.21, which shows the relationship between chlorine added and chlorine residual.

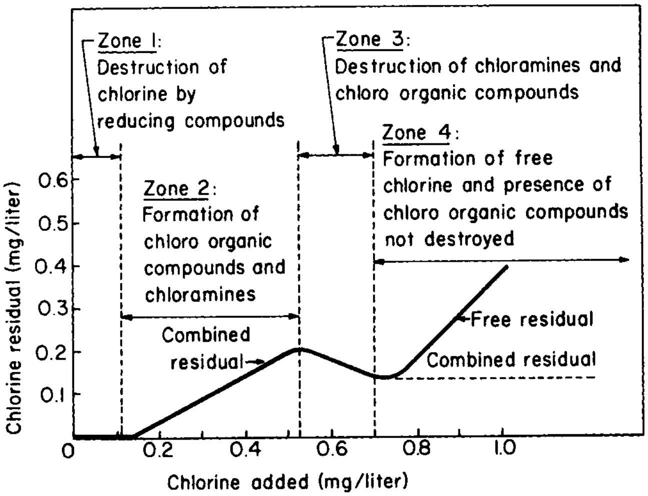


Fig. 8.21. Reactions of chlorine in water [8].

The initial amount of chlorine added is reduced by compounds which react rapidly with chlorine (e.g., $\mathrm{Fe^{2+}}$ and $\mathrm{Mn^{2+}}$). This corresponds to Zone 1 for which residual chlorine is nearly zero. Continued addition of chlorine results in residual chlorine in the form of chloro organic compounds or chloramines (combined residual). Chlorine residual is always less than the chlorine added. This corresponds to Zone 2 of the curve in Fig. 8.21. By adding more chlorine, chloro organic compounds are frequently oxidized: the molecule is broken down and chlorine is liberated. This results in a *decrease* of residual chlorine (Zone 3 in Fig. 8.21). Finally, when all reducing compounds have been oxidized, the additional quantity of chlorine added to the water results in an equivalent residual chlorine. This corresponds to Zone 4 in Fig. 8.21, which displays a 45° straight line for the free residual, and a straight line parallel to the abscissa for the constant combined residual.

7.1.3. Chlorine as a Disinfectant

Typical chlorine dosages required for disinfection are shown in Table 8.5. The effectiveness of chlorine for killing bacteria is given by Chick's law [4], which is written in differential form as Eq. (8.60).

$$dN/dt = -kN \tag{8.60}$$

TABLE 8.5
Typical Chlorine Dosages for Disinfection [8]

Effluent from	Dosage range (mg/liter)
Untreated wastewater (prechlorination)	6–25
Primary sedimentation	5–20
Chemical precipitation plant	2–6
Trickling filter plant	3–15
Activated sludge plant	2–8
Multimedia filter following activated sludge plant	1–5

where N is the bacteria count, t the time, dN/dt the rate of bacteria kill, and k the rate of kill constant.

Separating the variables and integrating from time $t = 0$ to any time t,

$$N/N_0 = e^{-kt} \tag{8.61}$$

where N_0 is the bacteria count at $t = 0$ and N the bacteria count at time t. The rate of kill constant k is a function of pH, temperature, and applied concentration of chlorine. It is estimated from the slope of a straight line plot of $\ln N/N_0$ vs. t based on Eq. (8.61), which written in logarithmic form is

$$\ln(N/N_0) = -kt \tag{8.62}$$

The effectiveness of chlorination for destruction of various organisms corresponds to values of k from 0.24 to 6.3 for 99% kill (i.e., $N/N_0 = 1/100 = 0.01$) at 0°–6°C.

Chick's law is an idealized portrayal of the situation. Usually, ideal conditions do not exist owing to variations in cell resistance, decrease in chlorine concentration, etc. Rates of kill sometimes increase or decrease with time. Consequently, a modified form of Eq. (8.61) containing an extra constant m to be determined from experimental data is written as

$$N/N_0 = e^{-kt^m} \tag{8.63}$$

If m is less than 1, the rate of kill decreases with time, and if m is greater than 1, it increases. Constants in Eq. (8.63) are determined by plotting $-\ln(N/N_0)$ vs. contact time t on logarithmic paper. The linear relationship is

$$\log[-\ln(N/N_0)] = \log k + m \log t \tag{8.64}$$

The effect of chlorine concentration is defined by the relationship

$$C^n t_p = \text{constant} = K \tag{8.65}$$

where C is the chlorine concentration (mg/liter); t_p the time required for a given percentage kill; and n the constant to be evaluated from the experimental data.

Constants in Eq. (8.65) are determined by plotting on logarithmic paper the concentration of chlorine vs. time for a given percentage kill. The linear form of Eq. (8.65) is

$$\log C = -(1/n) \log t_p + (1/n) \log K \tag{8.66}$$

The slope of this line corresponds to value of $-(1/n)$.

7.1.4. Utilization of Chlorine for Destruction of Cyanides

Chlorine is utilized to oxidize cyanide to innocuous products. This is done in an alkaline media at values of pH greater than 8.5. Oxidation takes place in two stages according to Eqs. (8.67) and (8.68).

First stage: $$CN^- + 2OH^- + Cl_2 \rightarrow CNO^- + 2Cl^- + H_2O \tag{8.67}$$

Second stage: $$2CNO^- + 4OH^- + 3Cl_2 \rightarrow 2CO_2 + N_2 + 6Cl^- + 2H_2O \tag{8.68}$$

From Eqs. (8.67) and (8.68) the theoretical stoichiometric ratio is $5Cl_2/2CN^-$ or $2.5Cl_2/1CN^-$. In practice, a large excess of chlorine is utilized, of the order of 7.5 parts chlorine per one part CN^-.

7.1.5. Economics of Chlorination of Wastewaters

Although the use of chlorination is widespread, it should be pointed out that chlorine is a relatively expensive chemical. If economics is a consideration for a given application, other methods should be evaluated. Chlorine oxides (Cl_2O, Cl_2O_5, Cl_2O_7) have been used in the disinfection of waters. A mixture of Cl_2O_5/Cl_2O_7 is available commercially.

7.2. OZONATION OF WASTEWATERS

Chemical oxidation with ozone is an effective method for treating wastewaters, based on the following factors [5]:

1. Ozone reacts readily with unsaturated organics in wastewaters.

2. Foaming characteristics of wastewaters are reduced following ozone treatment.

3. Ring opening and partial oxidation of aromatics renders the wastewater more susceptible to conventional biological treatment.

4. Ozone in the effluent quickly reverts to oxygen once it has served its purpose. This dissolved oxygen is beneficial to the receiving stream and helps support aquatic life. In contrast, chlorine (which is the most widely used bacteria kill agent) lingers in the effluent and becomes a pollutant itself). Laboratory scale ozonation equipment for evaluating amenability of wastewater to ozone oxidation is described by Eckenfelder and Ford [5].

Ozone may replace chlorine in treatment of cyanide wastewaters. Oxidation takes place in two stages according to Eqs. (8.69) and (8.70).

$$\text{First stage:} \quad CN^- + O_3 \rightarrow CNO^- + O_2 \tag{8.69}$$

$$\text{Second stage:} \quad 2CNO^- + 3O_3 + H_2O \rightarrow 2HCO_3^- + N_2 + 3O_2 \tag{8.70}$$

8. Nutrient Removal

8.1. INTRODUCTION

Removal of nutrients (phosphorus and nitrogen compounds) from wastewaters is an important operation, because these compounds play a critical role in lake eutrophication (Chapter 1, Section 7). Emphasis has been given to phosphorus removal for two reasons: (1) phosphorus is the most critical nutrient, and (2) nitrogen removal processes are less efficient and more expensive. Most nutrient removal treatment processes in operation today are designed for phosphorus removal alone.

8.2. PHOSPHORUS REMOVAL

8.2.1. Processes for Phosphorus Removal

Processes for phosphorus removal include (1) chemical precipitation, (2) activated sludge process (Chapter 5), (3) stabilization ponds (Chapter 6, Section 6), (4) reverse osmosis (Chapter 8, Section 5), and (5) electrodialysis (Chapter 8, Section 6).

The activated sludge process, although primarily intended for removal of organic contaminants, removes both phosphorus and nitrogen, since biological cells contain approximately 2.0% phosphorus and 12% nitrogen by weight. For domestic sewage this amounts to a phosphorus removal rate of 20–40% (or 1–2 mg/liter).

Stabilization ponds yield relatively high phosphorus and nitrogen removal, provided light and temperature conditions are favorable to the growth of algae. During the summer, removal of about 80% of the nutrients may be obtained, whereas during the winter it may drop to 20% or less. Reverse osmosis and electrodialysis are still too costly for generalized use. The discussion of phosphorus removal in this section is exclusively concerned with chemical precipitation processes.

8.2.2. Chemical Precipitation Processes for Phosphorus Removal

Precipitants which have been mainly employed in phosphorus removal are Fe^{3+} (as $FeCl_3$), Ca^{2+} (as lime), Al^{3+} [as alum, $Al_2(SO_4)_3 . 16H_2O$], and combinations of Fe^{3+} and lime. The mechanism of phosphorus removal is mostly precipitation in the form of phosphates of Ca^{2+}, Fe^{3+}, and Al^{3+}.

$$\begin{aligned} 3Ca^{2+} + 2PO_4{}^{3-} &\rightarrow Ca_3(PO_4)_2\downarrow \\ Fe^{3+} + PO_4{}^{3-} &\rightarrow FePO_4\downarrow \qquad (8.71) \\ Al^{3+} + PO_4{}^{3-} &\rightarrow AlPO_4\downarrow \end{aligned}$$

Adsorption also plays a role in the removal of some phosphates which are adsorbed on the precipitating floc. Considerations on utilization of different precipitants are as follows:

1. Fe^{3+} (as $FeCl_3$). $FeCl_3$ (in dosages of 10 mg/liter) is the most commonly employed precipitating agent for phosphorus, yielding around 90% removal.

2. Ca^{2+} (as lime). Lime is less efficient than $FeCl_3$. Utilized in dosages of 500–700 mg/liter yields removal of at most 80% at pH values of 10.5–11. Another drawback of lime utilization resides in the large volumes of sludge produced, which causes a disposal problem. Recovery of lime by calcination of this sludge may be performed. Because of the presence of organic materials, the sludge may be capable of supporting its own combustion.

3. Al^{3+} [as alum, $Al_2(SO_4) . 16H_2O$]. Although alum precipitation yields approximately 95% phosphorus removal at dosages of 200–250 mg/liter, it is less frequently applied owing to the high cost of alum. Removal of 50–60% of organic materials (carbonaceous and nitrogenous) is obtained simultaneously with phosphorus removal.

4. Combination of Fe^{3+} and lime. Combinations of $FeCl_3$ solutions and lime in respective proportions of 100–150 mg/liter and 2–5 mg/liter have been used, yielding phosphorus removal of approximately 95%.

The precipitation operation is usually carried out within a treatment process. Figure 8.22 summarizes three alternative precipitation operations for phosphorus removal within an activated sludge plant.

Alternative No. 1—precipitation in primary clarifier. The precipitant, usually a combination of $FeCl_3$ and lime with typical dosages indicated previously, is added to raw sewage, phosphates being precipitated and removed in

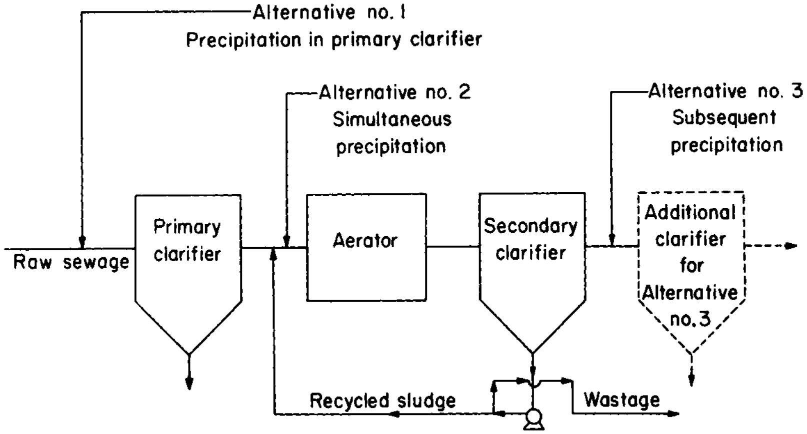


Fig. 8.22. Alternative precipitation operations for phosphorus removal.

the primary clarifier. Phosphorus removal is 90–95%, and a considerable amount of sludge is obtained. The addition of lime in this process raises the pH of the effluent from the primary clarifier to nearly 10.0. This is not an obstacle to the proper operation of the biological process in the aerator, which requires a pH near neutrality. The natural buffering capacity of the activated sludge process is sufficient to provide the required neutralization, owing to production of carbon dioxide.

Alternative No. 2—simultaneous precipitation. Precipitant is added directly to the aeration tank. The precipitate settles in the secondary clarifier together with the activated sludge. It seems that the chemicals do long-term damage to the biological cells, and that their organic removal efficiency is impaired to some extent. On the other hand, addition of chemicals aids settling and compaction of the activated sludge in the secondary clarifier.

Alternative No. 3—subsequent precipitation. Precipitating chemicals are added to the effluent from the secondary clarifier. An additional clarifier is required to remove precipitated phosphates, thereby increasing capital costs. On the other hand, the phosphorus removal efficiencies obtained are somewhat higher than those for the two previous alternatives.

Estimating chemical requirements for phosphorus removal cannot be done from simple stoichiometric relationships, because the actual mechanism of phosphorus removal is not known. Empirical relationships have been developed for estimating chemical requirements, and two of these are mentioned next.

1. For estimating Fe^{3+} or Al^{3+} requirements,

$$\varepsilon = C_1 + 0.5 \log(m/P_i) \tag{8.72}$$

where ε is the efficiency of phosphorus removal [$\varepsilon = (P_i - P_f)/P_i$, where P_i

is the initial phosphorus concentration in wastewater (moles of P/liter); P_f the final phosphorus concentration after precipitation operation (moles of P/liter)]; C_1 a constant with a value of 0.614 for $FeCl_3$ and 0.662 for alum; and m the required molality of precipitant (moles Fe^{3+} or Al^{3+}/liter).

Solving for m,

$$m = P_t 10^{(\varepsilon - C1)/0.5} \tag{8.73}$$

Equations (8.72) and (8.73) apply for a range $0.45 < \varepsilon < 0.95$.

2. For estimating lime requirements, a rule of thumb consists in utilizing a quantity of lime (moles/liter of Ca^{2+}) equal to 1.5 times the carbonate hardness of the wastewater.

Laboratory evaluation of phosphorus removal can be performed. These tests not only permit plotting curves of phosphorus removal efficiency vs. dosage of precipitant, but also evaluation of effects of pH and temperature. Curves of phosphorus removal efficiency vs. dosage of precipitant are exponential. They reach a plateau beyond a certain concentration of chemical added. It is usually uneconomical, therefore, to attempt reducing phosphorus concentration below 0.10 mg/liter owing to extremely high chemical requirements.

8.3. NITROGEN REMOVAL

8.3.1. Introduction

Nitrogen together with phosphorus contributes to the process of lake eutrophication. Also, nitrogen in the form of NH_4^+ or nitrites (NO_2^-) exerts an oxygen demand because of its oxidation to nitrates (NO_3^-).

$$NH_4^+ \rightarrow NO_2^- \rightarrow NO_3^- \tag{8.74}$$

These facts justify the desirability of nitrogen removal from wastewaters prior to discharge into receiving waters.

For domestic wastewaters the split among the various forms of nitrogen is relatively constant: NH_4^+ nitrogen, 50–60%; organic nitrogen, 40–59%; and nitrites and nitrates, 0–5%. In the effluent from activated sludge treatment, most nitrogen is present either as nitrites or nitrates, owing to the occurrence of nitrification. Nitrification invariably occurs during activated sludge treatment at values of F/M ratio less than 1.0. (Optimum range of F/M ratios for the activated sludge process is from 0.3 to 0.7; see Chapter 5, Section 5.)

8.3.2. Processes for Nitrogen Removal

Processes for nitrogen removal include (1) nitrification–denitrification process; (2) ammonia stripping; (3) ion exchange; (4) biological processes such as activated sludge and stabilization ponds; and (5) precipitation processes.

1. *Nitrification–denitrification.* The main process geared specifically for nitrogen removal is nitrification–denitrification, which is a modification of the activated sludge process and takes place in two steps:

(a) Nitrification consists in aeration employing F/M ratios less than 1.0, usually $F/M \approx 0.3$, and a longer aeration period than for the conventional activated sludge process. Nitrogen compounds are converted to nitrites (NO_2^-) and nitrates (NO_3^-) in the presence of *Nitrosomonas* and *Nitrobacter* microorganisms.

$$\begin{gathered} NH_4^+ \xrightarrow{Nitrosomonas} NO_2^- \\ NO_2^- \xrightarrow{Nitrobacter} NO_3^- \end{gathered} \tag{8.75}$$

(b) Denitrification is an anaerobic step which takes place in the secondary clarifier by extending residence time and adding an organic carbon source, usually methanol. Nitrites and nitrates are converted to nitrogen gas and nitrogen oxide, which are vented out.

$$NO_2^- \text{ or } NO_3^- \xrightarrow{\text{denitrification}} N_2 + N_2O \tag{8.76}$$

Sludge age in secondary clarifier is at least 2–3 days, higher values being required at lower operating temperatures. Denitrification is performed even without addition of the organic carbon source to help meet denitrifying bacteria requirements. Endogenous respiration provides these requirements, but addition of the carbon source is helpful. A two-step nitrification–denitrification process is shown in Fig. 8.23.

A more sophisticated *three-step nitrification–denitrification process* is shown in Fig. 8.24. The first step is a high rate activated sludge process where 75–85% of the carbonaceous material is removed. Residence time in the aerator is approximately 2 hr. The second step is nitrification in the presence of an

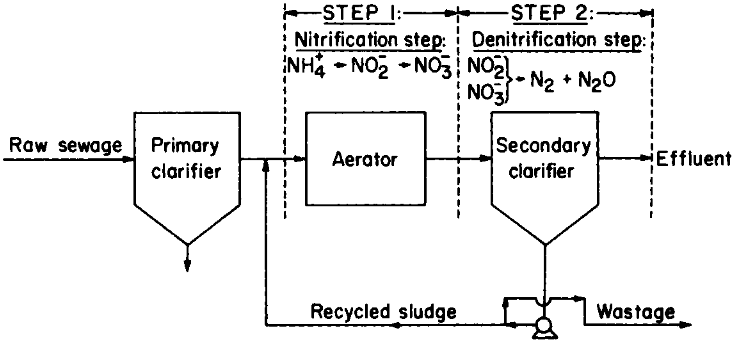


Fig. 8.23. Two-step nitrification–denitrification process.

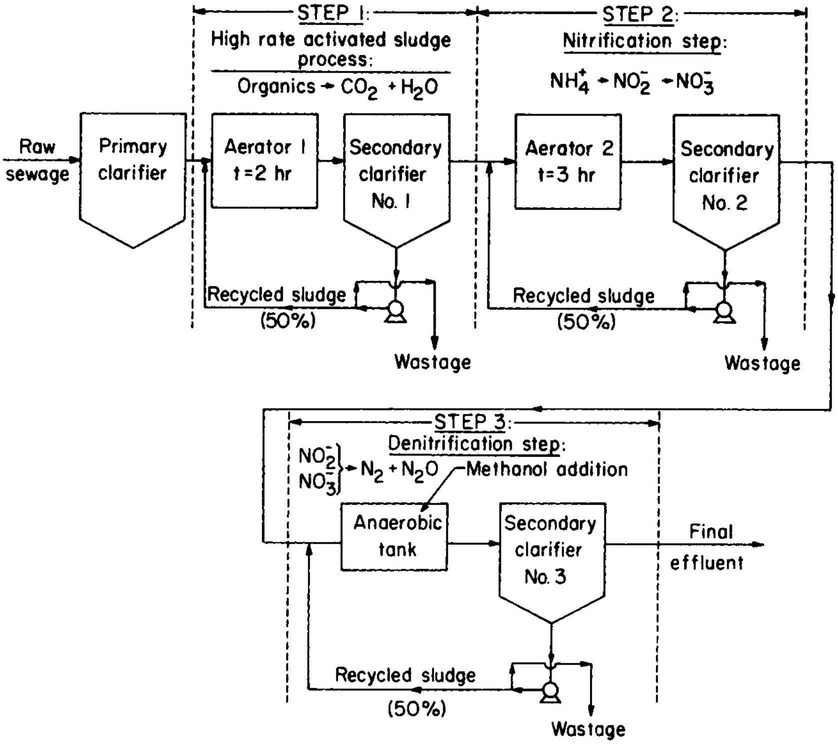


Fig. 8.24. Three-step nitrification–denitrification process.

enriched culture of nitrifying bacteria. Residence time in the aerator is approximately 3 hr. The third step is denitrification, which is an anaerobic operation, methanol being added to provide an adequate C/N ratio.

2. *Ammonia stripping.* This process consists of adjusting pH of the wastewater to a value above 10 (lime is utilized for this purpose), and then air-stripping the ammonia (at $\mathrm{pH} > 10$ nitrogen is present as NH_3) in a stripping tower. Nitrogen removal efficiencies of approximately 90% are obtained at air/liquid ratios of 350 ft^3/gal and liquid rates of about 3.0 gal/(min)(ft^2). Removal efficiency is increased close to 98% with an air/liquid ratio of 800 ft^3/gal. However, removal efficiency drops considerably in cold weather.

Ammonia stripping is employed either before or after secondary treatment. If the stripping operation is followed by biological treatment, it is necessary to lower the pH to a value near the neutral point. This is usually done by recarbonation, i.e., bubbling through flue gas containing carbon dioxide. It is important to leave enough nitrogen unremoved to satisfy nutritional requirements for the biological process.

Simultaneously with nitrogen removal, air-stripping accomplishes phosphorus removal, BOD reduction, and removal of suspended solids. The primary variables involved in the ammonia-stripping process are pH, air/liquid ratio, hydraulic loading [gal/(min)(ft^2)], packed height, and characteristics of the packing elements.

3. *Ion exchange.* A cation exchange resin (clinoptilolile) which is selective for NH_4^+ is utilized in treatment of activated sludge effluents from the secondary clarifier. This resin is regenerated with lime, and the regenerant is reused after being air-stripped of ammonia in a stripping tower. The cost of operation is high, and usually, this process is not economically feasible. Anion exchange resins are used for removal of nitrates (NO_3^-). Phosphates and other anions are simultaneously removed. The resin is regenerated with brine and restored by treatment with acid and methanol. Pretreatment by filtration may be required in ion exchange processes, thereby increasing costs.

4. *Biological processes,* i.e., activated sludge and stabilization ponds, have already been mentioned in Section 8.2.1 in connection with phosphorus removal.

5. *Precipitation processes.* Section 8.2.2 describes these processes in connection with phosphorus removal. As far as nitrogen removal is concerned, precipitation methods are rather inefficient, accounting for less than 30% removal of total nitrogen.

8.4. ADDED BENEFITS IN NUTRIENT REMOVAL

Although initial costs in the processes studied are high, there are added benefits in nutrient removal processes which in some cases justify the costs. Among these are (1) BOD reduction, (2) suspended solids removal, (3) bacterial and viral removal, and (4) removal of volatile organics.

9. Sonozone Wastewater Purification Process

This tertiary treatment process has been developed at the University of Notre Dame (Lobund Laboratory), and sonozone plants are commercially available from the Ecology Division of Telecommunications Industries, Inc. (Lindenhurst, New York). A pilot plant at the University of Indiana with a capacity of 20,000 gal/day is being used as a research model, treating on-campus sewage. A full-size plant (capacity, 570,000 gal/day) is in operation at Indiantown, Florida.

Telecommunications Industries, Inc. claims that the sonozone process provides tertiary treatment at costs comparable to secondary treatment. The sonozone process combines ultrahigh frequency sound and ozone treat-

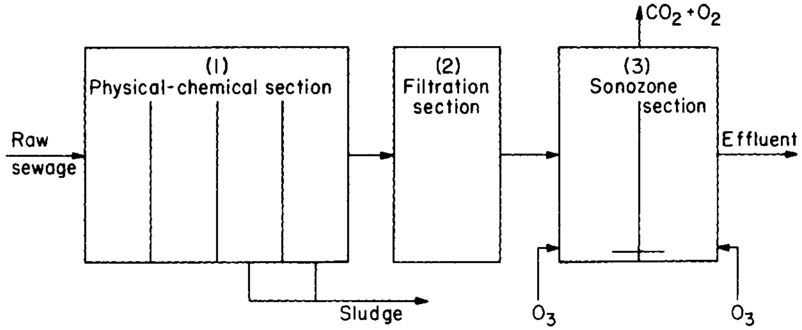


Fig. 8.25. Flow diagram of sonozone process **[10]**. (*Courtesy of Telecommunications Industries, Inc.*)

ment. A simplified flow diagram is shown in Fig. 8.25. Treatment involves three stages, of which the two initial ones are pretreatment units.

1. *Physical-chemical section.* Sludge is removed by a series of primary treatment contacts utilizing coagulation followed by clarification.

2. *Filtration section.* The filtration system is designed to remove micro-sized solids and organics from clarified wastewater.

3. *Sonozone section.* The central unit is the ozone and sonics unit. It consists of a small vibrating metal disk at the bottom of a tank through which wastewater flows. A steady stream of ultrasonic waves is sent out by the vibrating disk, and simultaneously ozone is bubbled into the tank from a nearby generator, which produces ozone by shooting electric arcs through the air.

The exact mechanism for the phenomena taking place when wastewater is subjected to a combination of ozone and ultrasonic waves is still under investigation. High frequency sound waves rattle bacteria and dissolved particles, breaking them into submicron size. In this form, they become highly susceptible to the strong oxidizing effect of ozone, so less of it is required. Carbonaceous materials are oxidized, yielding CO_2 and O_2. This corresponds to the simplified equation [Eq. (8.77)]

$$2O_3 + C \rightarrow CO_2 + 2O_2 \tag{8.77}$$

In tests conducted at the 20,000-gal/day pilot unit at the University of Notre Dame, less than 60 sec of sonozone treatment destroyed 100% of fecal bacteria and viruses, 93% of phosphates, and 72% of nitrogen compounds. Effluent purity was just short of that obtained by distillation.

In the effluent, ozone quickly reverts to oxygen. This dissolved oxygen is beneficial to the receiving stream and helps support aquatic life. In contrast, chlorine, the most largely used bacteria kill, lingers in the effluent and becomes a pollutant itself.

Sludge from the physical–chemical section of the sonozone system is processed by any of the standard methods such as sand bed drying, vacuum filtration, centrifugation, or incineration. The Indiantown plant uses sand bed drying, with liquid waste from the sludge reprocessed through the sonozone system. Sludge is collected and periodically transported to a sludge disposal area.

One important characteristic of the process is its compactness. It is estimated that the sonozone system requires approximately 20% of the land area of conventional systems; thus real estate acquisition costs are considerably reduced. In addition, the smaller housing of the plant is more attractive than the large horizontal aeration and sedimentation tanks of conventional systems.

Problems

I. For the activated carbon column in Example 8.2, calculate the residual concentration (C_E) if depth is increased to 10 ft, keeping the same service time. Prepare a plot of C_E vs. depth utilizing depths of 5, 10, 15, and 20 ft.

II. Verify if the carbon adsorption data in the tabulation below are fit by the Langmuir isotherm. If so, determine constants K and b.

C (mg/liter)	X/M (g/g)
10	0.20
20	0.28
30	0.33

III. The wastewater stream from a plating industry has the following characteristics: flow rate, 50 gal/min for 12 hr a day; chemical composition: copper, 30 mg/liter as Cu^{2+}; zinc, 12 mg/liter as Zn^{2+}; nickel, 20 mg/liter as Ni^{2+}; and chromate, 125 mg/liter as CrO_3. Characteristics of cation and anion exchange resins used are as given in the tabulation below.

	Cation	Anion
Regenerant	H_2SO_4	NaOH
Dosage (lb/ft^3)	11.0	4.7
Concentration (%)	5	10.0
Flow rate [gal/(min)(ft^2)]	0.5	0.5
Exchange capacity	1.5 eq/liter	3.8 lb CrO_3/ft^3
Rinsing water (gal/ft^3 of resin)	130	100

1. Design an anion exchange system to remove chromate. Calculate volume of resin required, NaOH for daily regeneration, and rinsing water requirements.
2. Design a cation exchanger to remove Cu^{2+}, Zn^{2+}, and Ni^{2+}. Calculate volume of resin required, H_2SO_4 for daily regeneration, and rinsing water requirements.

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Appendix

Conversion Factors from English to Metric Units

LENGTH

in. × 2.54 = cm
ft × 30.48 = cm
ft × 0.3048 = cm

AREA

in.^{2} × 6.4516 = cm^2
ft^2 × 929.03 = cm^2
ft^2 × 0.092903 = m^2
acre × 4046.8 = m^2 (1 acre = 43,560 ft^2)

VOLUME (U.S. GAL UTILIZED)

in.^{3} × 0.01638 = liter
in.^{3} × 16.386 = cm^3
ft^3 × 28.316 = liter
ft^3 × 0.02831 = m^3
ft^3 × 2.8316×10^4 = cm^3
gal × 3.7854 = liter
gal × 3785.4 = cm^3

FLOW RATE (U.S. GAL UTILIZED)

gal/min × 0.0631 = liter/sec
gal/hr × 1.052×10^{-3} = liter/sec
gal/day × 4.381×10^{-5} = liter/sec
MGD × 43.81 = liter/sec (*Note*: MGD = million gallons/day)

FLOW RATE PER UNIT AREA (U.S. GAL UTILIZED)

gal/(min)(ft^2) × 40.75 = liter/(min)(m^2)
gal/(min)(ft^2) × 0.679 = liter/(sec)(m^2)
gal/(min)(ft^2) × 0.004075 = liter/(min)(cm^2)
gal/(min)(ft^2) × 0.2445 = liter/(hr)(cm^2)
gal/(min)(ft^2) × 2445 = liter/(hr)(m^2)
gal/(hr)(ft^2) × 0.679 = liter/(min)(m^2)
gal/(hr)(ft^2) × 0.0113 = liter/(sec)(m^2)
gal/(hr)(ft^2) × 6.79×10^{-5} = liter/(min)(cm^2)
gal/(hr)(ft^2) × 0.004075 = liter/(hr)(cm^2)
gal/(hr)(ft^2) × 40.75 = liter/(hr)(m^2)

FLOW RATE PER UNIT VOLUME (U.S. GAL UTILIZED)

gal/(hr)(ft^3) × 2.2277 = liter/(min)(m^3)
gal/(hr)(ft^3) × 0.0371 = liter/(sec)(m^3)

MASS

lb × 453.59 = g
lb × 0.45359 = kg

QUANTITY OF HEAT

BTU × 0.25198 = kg calorie
BTU × 251.98 = g calorie

TEMPERATURE

$°C = (1/1.8)(°F - 32)$

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