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Oxygen uptake in a completely mixed activated sludge system

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OXYGEN UPTAKE IN A COMPLETELY MIXED ACTIVATED SLUDGE SYSTEM

by

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I. INTRODUCTION

The progress of biological waste treatment, in common with all engineering processes, has been marked by condensation of time and space in which the desired treatment objectives are accomplished. Irrigation and sewage farming, first attempted in the United States in 1872, served only 30 to 250 persons per acre of treatment area (3, 8). Modern conventional activated sludge treatment will serve on the order of 25,000 to 50,000 persons per acre of site. Space requirements are reduced even further when so-called high rate processes are used. Greater understanding of the various mechanisms involved in biological waste treatment has become essential as this trend continues. This understanding is rapidly changing design and operation of waste treatment works from an empirical art to a rational process based upon scientific principles.

The activated sludge process of waste treatment was introduced in England in 1913-1914 by Ardern and Lockett, and has become the process of choice where high degree secondary treatment of wastes from large populations is required. Greater knowledge of the capabilities of the process and factors affecting its operation have led to its use on an ever increasing variety of industrial wastes and have reduced the size of the installation on which the process can be feasibly applied and successfully operated.

Conventionally the process consists of mixing the waste with a previously formed culture of microorganisms (return sludge) which is returned to process from a final settling tank, aerating the mixture (mixed liquor) for a period of from 4 to 6 hours in an aeration tank and separating the solids from the aeration tank effluent in the final settling tank. The metabolizable nutrients in the influent, which are the pollutants to be removed, are used by the microorganisms in the system to synthesize new cell material and furnish energy for synthesis. Sludge is usually wasted from the system at a rate corresponding to the synthesis rate to hold solids in the system at a predetermined concentration. Aeration is accomplished by diffusion of compressed air through various devices or by mechanical aerators.

Several modifications directed toward greater volumetric efficiency and/or operating cost reduction through lower air requirements have been introduced. All modifications, however, are governed by the same principles.

Since the process is biological in nature, understanding must be based upon knowledge of the microbiology of waste treatment and the mechanisms of biological degradation of metabolizable nutrients in the process feed. If this understanding can be achieved, rational predictions of performance become possible and full capabilities of the process can be utilized. Lacking understanding, designs and expected results

must be based on empirical observations of prior performance. Subtle differences in environment may have pronounced effects on the process. The impact of these differences is not reliably predicted by operating results unrelated to mechanisms of the process.

II. OBJECTIVES AND SCOPE OF THIS STUDY

In May, 1960, a continuous flow, activated sludge pilot plant was placed in operation at the P. F. Morgan Sanitary Engineering Laboratory of the State University of Iowa. Two years controlled operation of the plant showed that performance was unexplainable in terms of empirical predictions based on conventional activated sludge operating characteristics. A summary of pilot plant operating data for the first year of operation by Byrne (4) shows a low correlation between BOD removal and such conventional parameters as aeration time and loading per unit volume of aeration capacity. Byrnes recognized the importance of oxygen use and transfer but was unable to make quantitative measurements of oxygen use because of lack of suitable equipment and instrumentation during his studies.

The need for developing an oxygen use measuring system for use in this pilot plant was recognized by the late Prof. P. F. Morgan. Such a system was developed by the author (20) with the expectation of using it for studies of oxygen utilization.

It was originally expected that oxygen use studies would be made to determine the level of controllable operating parameters that would give maximum oxygen use from a given amount of air. Among these parameters are BOD-solids loading, BOD-aeration volume loading, mixed liquor suspended solids con-

centration and/or dissolved oxygen and aeration period. As the work progressed it became evident that little or no correlation was exhibited between oxygen use and these parameters. Moreover, no reliable correlation could be detected between such supposedly fixed factors as oxygen use and BOD removal.

Consequently, it became necessary to redefine the principle objective of this study as the derivation of a rational model of events occurring in the activated sludge system that would explain the level of oxygen use observed with the oxygen measuring system. If this model corresponded with those derived recently as a part of mathematical expressions for overall kinetics of activated sludge systems, this study would provide unique experimental proof of theoretically derived expressions for oxygen use. If the model derived from this study was at variance with those upon which current theoretical expressions were based, an experimentally proven model would be available from which modifications of theory could be made.

This pilot plant and its oxygen measuring system provide a unique opportunity for observing oxygen use in a continuously operating system in which normal operating environment is maintained at all times. This affords opportunity for measurement of operating variables under prototype conditions and flow schemes to a degree not possible with batch reactors. It offers considerably greater information than manometric measurements made on samples withdrawn from the system such as is

commonly done.

A secondary objective of the studies was to ascertain the true nature of the flow character and environment in the aeration tank in order to know the region of the biological growth curve in which the system was operating. This knowledge was necessary to determine the mathematical relations affecting biological growth and nutrient removal in the system so that performance results could be interpreted more reliably.

III. MECHANISMS OF BIOLOGICAL OXIDATION-WORK OF OTHER INVESTIGATORS

A. General Considerations

Some insight into the mechanisms and kinetics of the biological oxidation process is necessary for understanding the role of oxygen in the activated sludge process and for developing reliable models for explaining oxygen use in an operating system.

The activated sludge process accomplishes waste purification by removing organic materials from the waste and utilizing these materials as nutrients. These materials are biologically oxidized to produce new cell material and to yield the energy necessary both for production of new cell material and for carrying on the basic cell functions of movement and enzyme activation. The overall process, therefore, may be viewed from two standpoints:

1. From the standpoint of the waste (substrate), in which the removal of nutrients from the waste is the item of interest.

2. From the standpoint of the sludge mass (bacterial culture), in which the fate of the removed nutrients and their effect on the sludge mass is of interest.

Nutrients in the waste react with enzymes associated with the cells in the sludge mass to produce enzyme-nutrient complexes suitable for assimilation by the cells. These re-

actions are an intermediate process between removal from the substrate and assimilation by the cell.

Nutrient concentration is measured by its biochemical oxygen demand, commonly called BOD. The mass of cellular material entering into the process is the active fraction of the sludge in the system. The substrate is the waste being treated.

This section of this thesis reviews the basic concepts of nutrient removal, metabolism, cell synthesis and growth. Physical characteristics of aeration systems as they affect biological activity are then considered. Mathematical relationships pertaining to the type of system represented by the pilot plant as derived by current investigators are then summarized. Finally, the derived expressions relating to oxygen use are presented.

The present state of knowledge in this field is the result of work of many people over a long period. In the case of general biological phenomena, the references cited are typical works. In the case of recent work on mathematical expressions for process kinetics, the work of individuals or small groups is involved and the citations are correspondingly specific.

B. Mechanisms of Nutrient Removal from Substrate

Removal of nutrient (BOD) from a substrate (waste) occurs by one or more of the following mechanisms:

1. Physical adsorption of suspended nutrient by the sludge mass followed by enzymatic attack, solution and biological oxidation,
2. Direct enzymatic reaction and biological oxidation of dissolved nutrient,
3. Initial removal of both soluble and suspended nutrient by adsorption and enzymatic reaction and temporary storage for subsequent biological oxidation.

Nutrient is removed in the first two cases at the same rate that it is metabolized. In the third case, nutrient is removed on contact of the waste with the inoculuum (return sludge) at a higher rate than it is metabolized and the excess nutrient is stored in the sludge mass for subsequent oxidation.

The phenomenon of initial removal and storage was reported by Porges et al. (19) and is the basis for the biosorption process developed by Ullrich and Smith for Austin, Texas (22) and by Eckenfelder and Grich for cannery wastes (6). In this process the waste is aerated with the return sludge for a short period (15-30 minutes), the mixture is separated by settling and the settled sludge is aerated to complete the oxidation of the absorbed nutrient and prepare it for recycling. The degree of initial removal is a function of both sludge condition and nutrient character. McKinney points out

that the process works best on colloidal wastes (15).

The point at which assimilation of initially removed nutrient occurs in a self contained system is not clear but Eckenfelder presents evidence purporting to show that this happens when food becomes limiting (7). There is presumed to be no appreciable time lag between removal and assimilation in the first two cases. If this hypothesis of chronology of assimilation of stored material is accepted, nutrient removal from substrate by absorption and storage in the sludge mass would not be a significant removal mechanism in systems in which food is always limiting. Conversely, it would be an effective mechanism only when the food/organism ratio is so high that organism concentration rather than food concentration is limiting (log growth phase).

C. Nature of the Biological Oxidation Process

Nutrient removed from the substrate by the foregoing mechanisms is metabolized by bacteria for synthesis of new cells and energy. The substrate must contain sufficient nitrogen (NH_3), phosphorous and other trace elements to satisfy nutritional requirements. After the nutrient is exhausted, cells die and lyse releasing the nutrients in their own protoplasm for assimilation by still living cells in a process of auto-oxidation.

Concentration of nutrient or food is measured by its BOD. The mass of cellular material entering into the process is the active fraction of the sludge in the system.

Assimilation for synthesis and energy may be represented by:

$$BOD + O_2 + NH_3 \xrightarrow{cells} \text{new cells} + CO_2 + H_2O + \text{energy} \quad 1$$

Auto-oxidation may be represented by:

$$\text{Cells} + O_2 \xrightarrow{cells} CO_2 + H_2O + NH_3 \quad 2$$

Equation 1 shows that in the presence of food, free oxygen and available nitrogen, biological oxidation will proceed to produce new cells and energy. Thus, sludge mass will increase during assimilation. Equation 2 shows that in the presence of free oxygen and in the absence of food, auto-oxidation and consequent reduction of sludge mass will occur.

D. Basic Relationships of Biological Metabolism

As indicated by equation 1, nutrient is metabolized to synthesize new cells or protoplasm and to furnish the energy necessary for the synthesis, or

$$\text{Organic matter metabolized} = \text{Protoplasm Synthesized} + \text{Energy for Synthesis} \quad 3$$

Bacteria require a certain basal energy to maintain normal sustaining functions such as motivation and enzyme activation. This basal energy is called endogenous respiration.

It is a continuous process that results in metabolism of certain components of the protoplasm. It proceeds with breakdown of protoplasm even in the presence of excess nutrient. When nutrient is exhausted and the auto-oxidation phase depicted by equation 2 is entered, endogenous respiration becomes the only metabolic reaction. Thus,

Net Protoplasm Accumulated = Protoplasm Synthesized

- Protoplasm Metabolized by Endogenous Respiration 4

Equations 3 and 4 provide a basis upon which a complete description of biological metabolism can be constructed. They are the basis for McKinney's theory of biological oxidation (15) and his concepts of process kinetics (14).

Oxygen equivalent units may be used to convert equations 3 and 4 into dimensionally consistent form:

$$L_r = M_{so} + O_s \qquad 3a$$

$$M_{no} = M_{so} - O_e \qquad 4a$$

where:

L_r = total oxygen demand, BOD, removed

M_{so} = oxygen equivalent of protoplasm synthesized

M_{no} = oxygen equivalent of net protoplasm accumulated

O_s = Oxygen used for synthesis energy

O_e = Oxygen used for endogenous respiration.

All quantities are expressed in units of oxygen mass.

The oxygen equivalent of the protoplasm mass is usually

taken as 1.42 times the mass. This is the stoichiometric oxygen equivalent for oxidation of cellular material having a hypothetical formula, $C_5H_7NO_2$ as developed by Hoover et al. (11).

The 1.42 oxygen/mass ratio was approximately confirmed by chemical oxygen demand/volatile solids determinations during these studies (Table 3). The active cells constitute only part of the volatile solids in the mixed liquor or return sludge. However, the excess of volatile solids over active cell material is composed principally of organic material remaining after endogenous respiration. The most readily oxidizable materials are metabolized first leaving increasingly difficult to oxidize materials as time proceeds. In a continuous system fresh, readily oxidizable nutrient is constantly supplied in the feed. Thus only that part of the nutrient of cellular origin which is as easily oxidized as the feed nutrient will be metabolized. This means that the residual volatile solids will be only partially oxidized and will still possess substantial oxygen equivalent even though they are not active cell materials. The BOD/volatile solids values of Table 3 are slightly less than 1.42 and bear out this condition.

Using the factor of 1.42, equations 3a and 4a become

$$L_r = 1.42\ M_s + O_s \qquad 3b$$

$$1.42\ M_n = 1.42\ M_s - O_e \qquad 4b$$

where:

M_s = Mass of protoplasm synthesized

M_n = Net mass of protoplasm accumulated.

It has been shown that approximately 2/3 of the nutrient metabolized appears as new cell material (synthesis) and 1/3 is used for energy to carry out the synthesis. The magnitude of this relation has been confirmed by periodic solids - BOD removal balances in this pilot plant during 3 years of operation. With this relationship it is possible to eliminate the term O_s from equation 3b by expressing O_s in terms of M_s:

$$O_s = 1/2\ M_{so} = 1/2 \times 1.42\ M_s = 0.71\ M_s \qquad 5$$

from which

$$L_r = (1.42 + 0.71)M_s = 2.13\ M_s\ . \qquad 6$$

The foregoing equations may be used to show BOD removal or sludge growth rates by expressing them in their differential form:

$$\frac{dL_r}{dt} = 1.42\ \frac{dM_s}{dt} + \frac{dO_s}{dt} \qquad 3c$$

$$1.42\ \frac{dM_n}{dt} = 1.42\ \frac{dM_s}{dt} - \frac{dO_e}{dt} \qquad 4c$$

$$\frac{dL_r}{dt} = 2.13\ \frac{dM_s}{dt} \qquad 6a$$

where

t = reaction time.

Equations 4a, 4b and 4c show that synthesis is the sum of

net protoplasm accumulation and endogenous respiration. Since endogenous respiration is a first order process whose rate is proportional to the active protoplasm or active mass present, the rate of endogenous respiration may be expressed as:

$$\frac{dO_e}{dt} = k_e \; 1.42 \; M_a \qquad 7$$

where

k_e = endogenous respiration constant, mass of O_2 per unit time per unit mass of active mass present. Usually expressed as $\frac{mg}{1 hr}$ O_2 per $\frac{mg}{l}$ active mass.

M_a = active mass of protoplasm. Usually expressed as $\frac{mg}{l}$

1.42 = oxygen equivalent of active mass.

The active mass, M_a, is the total mass of live, metabolizing cell material present at any given time. It consists of the amount present at the beginning of a given time interval less that removed by the endogenous respiration of the original mass plus the net accumulation, M_n, during the interval. Thus, the rate of net accumulation, $\frac{dM_n}{dt}$, is the same as the rate of change of the active mass, $\frac{dM_a}{dt}$, and equation 4c may be written:

$$1.42 \frac{dM_s}{dt} = 1.42 \frac{dM_a}{dt} + 1.42 \; k_e \; M_a, \text{ or}$$

$$\frac{dM_s}{dt} = \frac{dM_a}{dt} + k_e \; M_a \; . \qquad 8$$

Substitution of equation 8 into equation 6a gives

$$\frac{dL_r}{dt} = 2.13 \left[\frac{dM_a}{dt} + k_e \; M_a\right] \qquad 9$$

or, integrating

$$L_r = 2.13\ M_a\ (1 + k_e t)\ . \qquad 9a$$

Equations 3 through 9 are simplified basic relations for organic matter metabolized, protoplasm growth and oxygen utilization. They are, however, useful only to the extent the value of k_e can be determined and only so long as the synthesis energy ratio is constant. They may be useful for describing events in a batch system in which no materials are added or removed during the reaction time. They are not useful for the more complex continuous flow systems normally encountered in practice.

The term active mass, M_a, is used in the foregoing discussion. This is the living mass of microorganisms and, because of endogenous respiration, is not the complete mass of organic matter in the sludge. During the process of lysing and assimilation of protoplasm nutrients by living cells, there will be an accumulation of non-metabolized volatile organic matter derived from the lysed cells. Thus, only part of the total volatile organic content of the sludge is active mass which enters into the biological oxidation process. This distinction has not always been clear in prior work on process kinetics.

E. Growth

Synthesis will result in growth of the protoplasm mass. The total growth will be characterized by the phases described

below:

I. Lag Phase: A period of adjustment required to adapt the culture to the nutritional and physical environment.

II. Log Growth Phase: A period during which the concentration of nutrients is not limiting and growth rate is governed by the mean generation time of the organism: growth rate is proportional to the active mass present. This phase continues until sludge (protoplasm) mass increases and nutrient concentration decreases to such levels that nutrient concentration becomes the growth limiting factor.

III. Declining Growth Phase: A period during which food concentration is limiting and growth rate is proportional to food concentration rather than sludge mass. Growth continues in this phase at a decreasing rate until all oxidizable or metabolizable nutrient has been removed.

IV. Auto-Oxidation Phase: The period following exhaustion of substrate food when endogenous respiration becomes the only metabolic process; sludge mass decreases. This is, initially, a first order reaction whose rate is dependent upon the remaining sludge mass. In the latter stages the reaction be-

comes retarded due to the presence of a larger portion of less readily oxidizable material in the remaining sludge mass. The phase continues until the sludge mass reaches a final limit and the oxidation process ceases.

The progress of growth through these phases for a batch of nutrient inoculated with organisms at t = 0 is depicted on Figure 1. Detailed discussion of the growth phases is given by Stanier, Doudorff, and Adelberg (21), Eckenfelder and Weston (5), Eckenfelder and McCabe (7), McCabe and Eckenfelder (12) and McKinney (15).

BOD removal from the batch is related to sludge growth as shown on Figure 1 and described below:

II Log Growth Phase: the rate of BOD removal increases as the sludge mass increases

III Declining growth phase: the rate of BOD removal is proportional to the BOD remaining. BOD removal is complete at the end of this phase.

Figure 1 also shows how the rate of oxygen utilization changes during growth progression. Although Figure 1 depicts progression of growth and rate of oxygen utilization with time for a batch system it is useful for interpreting performance of a continuous flow system. The characteristics of the continuous flow system such as feed strength and treatability, solids concentration, aeration time and flow charac-

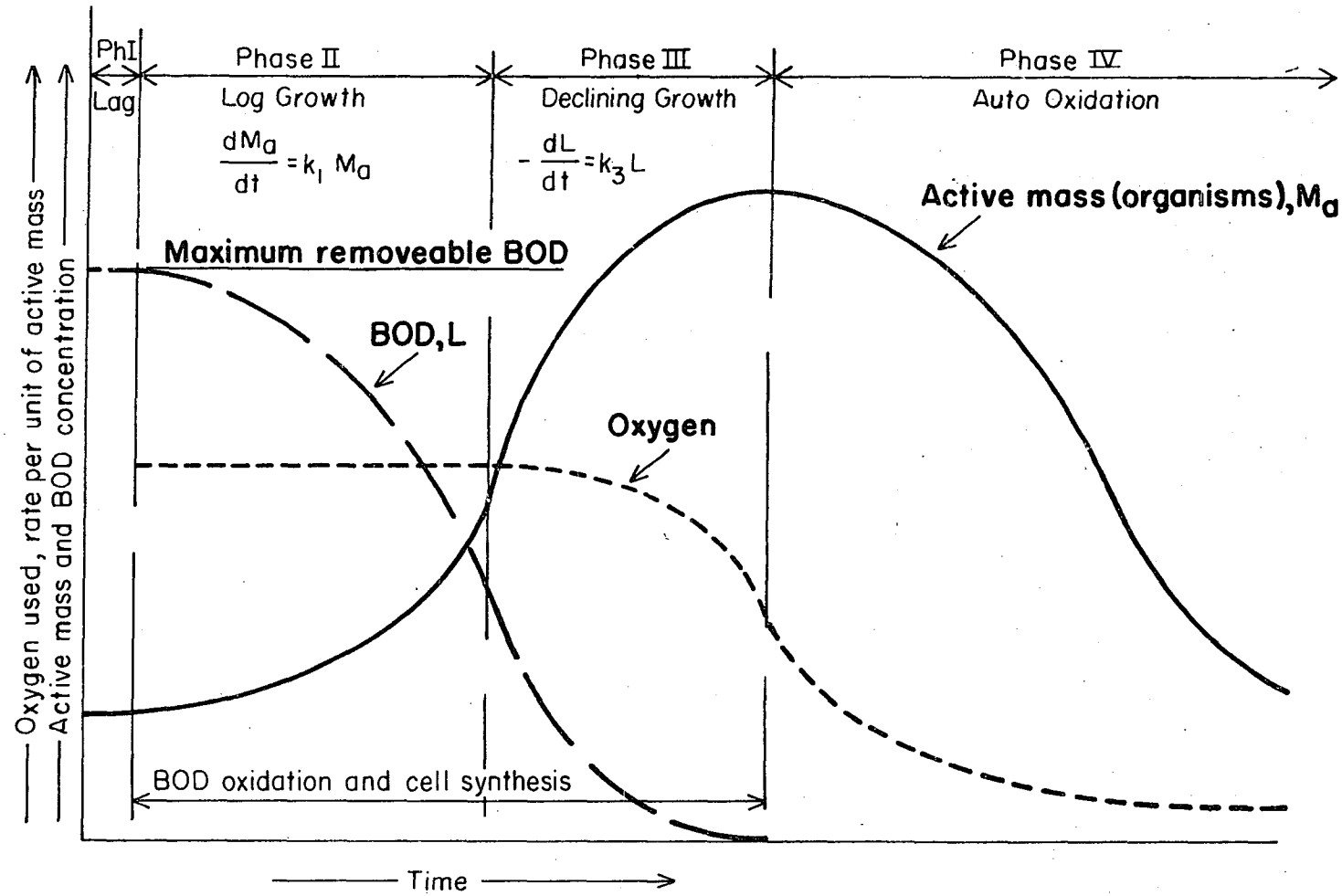


Figure I. Generalized growth, BOD removal and oxygen use rate

teristics ("plug" flow or mixed flow) will determine the ratio of food to microorganisms in any part of the system and hence the region of the growth curve in which that part of the system is operating. For example, to achieve high degree BOD removal the system must operate in the later stages of declining growth or in the autooxidation phase.

McKinney (15) states that log growth will prevail when the food/microorganism concentration ratio is 2.5 or more, declining growth between 2.5 and 0.006 and auto-oxidation below 0.006.

Growth rate in the log growth and declining growth phases may be expressed as differential equations according to the conditions of growth in each phase.

For log growth,

$$\frac{dM_a}{dt} = k_1 M_a \qquad \text{in mass units} \qquad 10$$

where

k_1 = log growth constant, $\frac{mg}{l\,hr}$ active mass per $\frac{mg}{l}$ active mass.

Equation 10 requires evaluation of active mass for practical use. As noted by equation 4, the net accumulation of active mass will be the result of synthesis less endogenous respiration. Likewise, if active mass was initially present, its net value after any time interval will be decreased due to endogenous respiration. Data indicate that less than 1 per

cent per hour of active mass is lost to endogenous metabolism whereas log growth rates of 20 per cent per hour are reported by Garrett and Sawyer and up to 1500 per cent by McKinney (14). Thus, it is seen that endogenous respiration is so low in relation to growth in the log growth phase as to have little practical significance. Active mass can, therefore, be considered as that initially present plus synthesis over a given time in considering equation 10.

The problem of determining active mass present at any given time still remains. This can be approximated from equation 7 by observing the oxygen uptake rate in some type of respirometer. After all nutrient has been stabilized the observed oxygen uptake rate will be that due to endogenous respiration and active mass will be the quotient of this uptake rate divided by the endogenous respiration constant times 1.42, the mass to oxygen conversion factor. Data indicate that uptake for endogenous respiration is about 0.5 per cent per hour. For example,

$$M_a = \frac{dO_2}{dt} \times \frac{1}{.005 \times 1.42} = \frac{dO_2}{dt} \times \frac{1}{0.007} \qquad 7a$$

where M_a is in mass units. This estimate of M_a can be used in equation 10.

For declining growth,

$$\frac{dM_s}{dt} = k_2 L \qquad 11$$

where

$\frac{dM_s}{dt}$ = rate of protoplasm synthesis, mass units, mg/hr

k_2 = declining growth constant, mg/1 hr of protoplasm synthesized per mg/1 ultimate BOD present

L = Ultimate BOD present, mg/1 .

Equation 11 in the form given is not usable because only a portion of the BOD removed is utilized for synthesis, the balance being used for energy. Unless this portion is known and expressed, nothing can be predicted as to the magnitude of either M_s or L. However, if it is assumed that the ratio of metabolism for synthesis to that for energy is constant throughout the declining growth phase, the relationship expressed in equation 11 can be modified to:

$$-\frac{dL}{dt} = k_3 L \qquad 12$$

where k_3 = the declining growth BOD removal rate constant, mg/1 hr ultimate BOD removed per mg/1 ultimate BOD present.

Thus, equations 10 and 12 provide simplified means of expressing the rate of active mass growth for systems in log growth or the rate of BOD removal for systems in declining growth. The equations are of practical significance only insofar as the various constants can be evaluated and are truly constant for a given waste or oxidation system.

The foregoing discussion is a condensation and interpreta-

tion of kinetics of growth and BOD removal developed by such investigators as Eckenfelder and Weston (5), Eckenfelder and McCabe (7, 12), McKinney (14, 15) and Garrett and Sawyer (10). It is noted that, for the most part, experimental work supporting these concepts has been conducted with small batch systems in the laboratory. Values of the constants expressed in the foregoing discussion together with such others as result from manipulation or solution of the various differential equations obtained from these laboratory batch studies have been applied to design of full scale continuous flow systems. There is a paucity of published data on comparison of laboratory batch results and those produced in the field. It is not clear to the author that such variable field conditions as mixed liquor solids concentration, return sludge rate and concentration, period of residence of return sludge in the final settling tank and dissolved oxygen concentration are or can be accounted for by application of the foregoing equations or their derivatives.

F. Batch Reaction vs. Complete Mixing

Two variations in process application are encountered which have pronounced effect on the reaction kinetics. These are

1. Batch reaction
2. Complete mixing.

1. Batch reaction

Batch reaction occurs when the nutrient and inoculuum are initially mixed and this mixture retains its identity during the reaction period without being mixed with additional nutrient and inoculuum which are added either prior to or after the given addition.

This is the condition encountered in a fill and draw system of waste treatment frequently used in laboratory work and occasionally encountered in the field. However, batch reaction conditions also exist in spiral flow aeration tanks which are commonplace in activated sludge practice. Even though there is violent transverse mixing in these tanks it has been shown that there may be only minor longitudinal mixing and the flow leaves the aeration tank in essentially the same order as it entered. This condition is called "plug flow."

In batch reactors, sludge growth and BOD removal will proceed in time along the courses depicted by Figure 1. The degree of BOD removal will depend on how far along the growth curve the reaction has progressed. This progress will be affected by the initial nutrient/organism ratio and the reaction time. The initial nutrient/organism ratio determines both the point on the growth curve at which the reaction starts and the amount of growth and BOD reduction that can be achieved within the time available for reaction. Conditions which allow the

reaction to be carried into the latter stages of declining growth or even into auto-oxidation are necessary to achieve high degree BOD removals. These are the conditions encountered in conventional activated sludge systems. On the other hand a lesser degree of treatment is expected if the initial nutrient organism ratio is higher and/or the reaction time is reduced, such as encountered in high rate or dispersed aeration systems.

Reduction of oxygen uptake rate per unit of sludge mass occurs along the length of a long spiral flow aeration tank. It follows the course of organism growth and BOD removal and is depicted by the variation in oxygen utilization rate shown on Figure 1.

2. Complete mixing

In a completely mixed system, the added nutrient and inoculuum are immediately dispersed throughout the aeration tank. The entire tank contents are homogeneous and the composition of the tank effluent is the same as that of the tank contents.

The effective nutrient/organism ratio is determined by the nutrient and organism concentration of the tank contents and is constant throughout the reaction period rather than variable from maximum at beginning to minimum at end as in a batch reactor. Thus, it is seen that a completely mixed system operates at a single point on the growth curve while a batch type system operates over a range of the growth curve.

3. Significance of differences

Differences between plug flow and completely mixed continuous flow biological reactors are significant for the following reasons:

1. The nature of the microorganism population is affected by the nutrient/organism ratio. In plug flow reactors, character of the inoculuum (return sludge) is that resulting from the lowest nutrient/organism ratio of the system whereas the nutrient/organism ratio when the innoculuum and feed are first mixed is the maximum of the system. Thus, the microorganism character must undergo adjustment before normal growth and BOD removal rates are established. In complete mixing the only change in nature of the population is that which may occur during residence in the final settling tank. Since the nutrient/organism ratio is low in the aeration tank, there will be little change during residence in the final tank and the microorganism population of the return sludge will be essentially the same character as that existing in the aeration tank. Thus, readjustment in a completely mixed system is not a factor.
2. In a completely mixed system the incoming nutrient is immediately dispersed throughout the tank. In the event of peak or shock loads their effect will be

minimized by the averaging effect of dispersion and dilution. No such leveling occurs in a plug flow system and the innoculuum must bear the full brunt of the shock thus extending the period of adjustment over that required when the leveling effect of dilution exists.

The differences in initial food to microorganism ratio require a clear understanding of the nature of flow in a given reactor if one is to predict and interpret results obtained with a given system. For example, high degree performance in a short period aeration system might be difficult to rationalize if considered from the standpoint of batch or plug flow. However, if the reactor is, in fact, a completely mixed unit it becomes obvious that it is operating at a low food to microorganism ratio throughout its volume and higher degree performance is to be expected.

G. Kinetics of Complete Mixing

1. General

The previously described relations governing growth of microorganisms and removal of BOD by biological oxidation together with the hydraulic or flow properties of a given reactor provide a point of departure for describing events and mechanisms occurring in an operating system. Such descriptions, if confirmed, are a means of predicting performance of a given

system and become valuable tools for the designing engineer. This study was principally concerned with the role of oxygen and mechanisms of its use in continuous flow biological oxidation processes. However, mathematical descriptions of events in various types of completely mixed systems were carefully studied and gave added insight into the process. Principal features of a mathematical description of completely mixed kinetics as developed by McKinney (13, 14) and discussed by Washington, Hetling and Rao (23) are described.

Two basic systems are recognized:

1. Aeration only. The wastes are continuously added to the aeration tank and a like rate of mixed liquor is displaced from the tank. There is no return of solids to the mixed liquor. Mixed liquor solids concentration is dependent upon the strength of the waste, synthesis rate and aeration time. Prolonged aeration periods (24 hours) are normally employed. Final settling may or may not be employed, but if used it has no function in the aerobic biological process. The aerated lagoon is a typical application of this system.
2. Aeration with sludge return. This system employs a final settling tank for gravity separation of solids from the mixed liquor discharged from the aeration tank. The settled solids are recycled as return sludge to the aeration tank to provide inoculuum for the raw waste. Solids must be wasted to offset accumulation

due to synthesis. Waste may be accomplished by allowing the system to reach an equilibrium condition such that carryover in the effluent maintains the desired mixed liquor solids concentration or by wasting a portion of the return sludge or mixed liquor. So called extended aeration "package plants" are examples of the first type of wasting. Short period aeration systems with separate sludge disposal as frequently used for industrial waste treatment are examples of the later type of wasting.

The systems are shown schematically on Figure 2.

2. Aeration only systems

A nutrient balance around an aeration tank in which declining growth conditions exist can be expressed as follows:

Net rate of change in tank nutrient = Inflow rate - outflow rate - rate of metabolism in tank 13

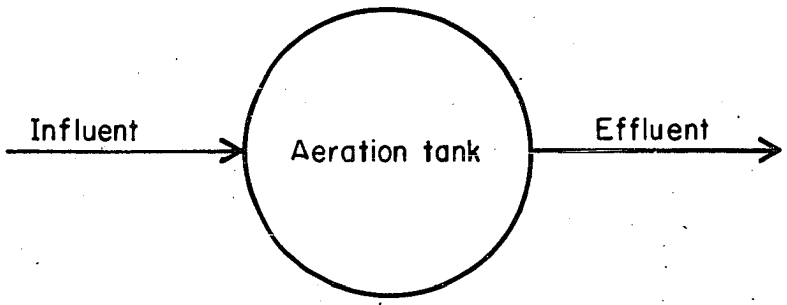
or

$$V_a \frac{dL_r}{dt} = QL_o - QL_r - k_3 V_a L_r \qquad 14$$

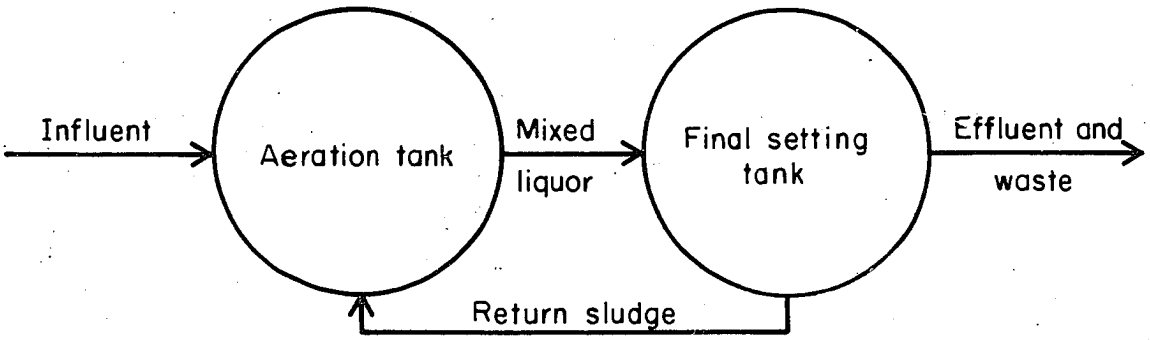
where symbols and customary units are:

V_a = aeration tank volume, liters

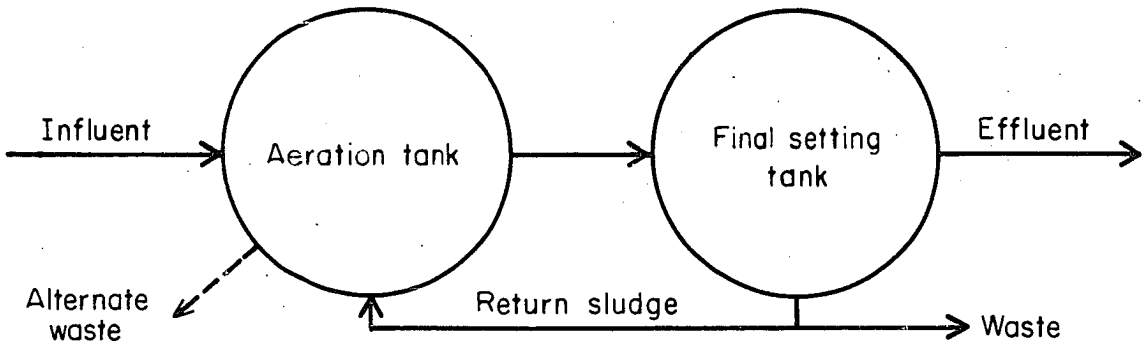
L_r = concentration of ultimate BOD remaining in tank, mg/l. Also concentration in tank effluent in a completely mixed tank



A. Aeration only; no final setting



B. Aeration with return sludge; excess solids wasted in effluent



C. Aeration with return sludge; separate solids waste system

Figure 2. Typical activated sludge flow diagrams

L_o = concentration of ultimate BOD in tank influent, mg/l

Q = Rate of flow, liters/hour

k_3 = declining growth BOD removal rate constant as previously defined for equation 12.

Since Q/V_a is the reciprocal of mean aeration time, $1/t$, equation 14 reduces to

$$\frac{dL_r}{dt} = 1/t\ L_o - 1/t\ L_r - k_3\ L_r \ . \qquad 14a$$

In a continuous flow system at equilibrium, $dL_r/dt = 0$ and

$$0 = 1/t\ L_o - 1/t\ L_r - k_3\ L_r \qquad 15$$

from which

$$L_r = \frac{L_o}{(1 + tk_3)} \ . \qquad 15a$$

Active mass may be balanced around the aeration tank as follows:

Net rate of change in tank contents = rate of synthesis in tank - rate of endogenous respiration in tank - outflow rate 16

or

$$V_a \frac{dM_a}{dt} = k_2\ V_a\ L_r - k_4\ V_a\ M_a - Q\ M_a \qquad 17$$

where new terms and customary units are:

k_2 = synthesis rate constant, mg/l hr mass increase per mg/l ultimate BOD remaining as used in equation 11. This is related to k_3 as follows:

k_3, oxygen units $\div 1.42 = k_3$, mass units

k_3, mass units x 2/3 (synthesis ratio) = $k_2 = 0.47k_3$ O_2 units

k_4 = endogenous respiration constant, mg/l hr mass decrease per mg/l active mass present. This is equal to $\frac{1}{1.42}$ times the k_e of equation 7 which is the oxygen rather than the mass rate constant,

or

$$V_a \frac{dM_a}{dt} = 0.47k_3 V_a L_r - k_4 V_a M_a - QM_a \qquad 17a$$

from which

$$dM_a/dt = 0.47k_3 L_r - k_4 M_a - M_a/t \; . \qquad 17b$$

At equilibrium,

$$\frac{dM_a}{dt} = 0, \text{ and}$$

$$M_a = \frac{0.47k_3 L_r}{1/t + k_4} \qquad 18$$

Further development of this line of analysis gives total mass of volatile solids in terms of active mass plus the inert organic mass (endogenous mass) resulting from endogenous respiration. Additional relationships can be developed to account for the effect of metabolizable volatile solids in the influent. A complete development is given by McKinney (14).

3. Aeration with return sludge

McKinney (14) develops relationships for systems with return sludge in which the solids level is maintained by wasting either through discharge of solids in the effluent or by sep-

arate wasting of a portion of the return sludge or mixed liquor. His approach is similar to that for aeration only but two important assumptions are presented without verification. They are

1. The effective time, or value of t, is based on the total volume of the system (aeration and final settling tanks) and is displacement time based on feed flow rate only, not on the combined feed plus return sludge flow rate.
2. Nutrient balance is made around the total system on the basis that metabolism and growth processes are occurring at equal rates in the aeration and final settling tank.

H. Oxygen Utilization

1. McKinney's description for complete mixing

As a part of McKinneys development of complete mixing kinetics, the following equation is presented for the rate of oxygen utilization in the aeration tank of an aeration only system:

$$dO/dt = k_5 L_r + k_e M_a \qquad 19$$

where

k_5 = BOD oxygenation constant, mg/1 hr oxygen used per mg/1 ultimate BOD remaining. This is related to k_3 in that it covers that part of the BOD used for energy, usually 1/3 of total BOD removed. Thus,

$k_5 = k_3/3$

k_e = Endogenous respiration constant previously defined for equation 7.

M_a = Oxygen equivalent of active mass or 1.42 M_a of equation 7.

Equation 19 simply states that rate of oxygen use is the sum of that required to produce energy from the BOD removed plus that required by endogenous respiration of the active mass present. The total oxygen requirement of a system is given by integrating equation 19 over the effective reaction time, or

$$O = (k_5 L_r + k_e M_a)t \qquad 20$$

since L_r and M_a are constant at equilibrium.

McKinney states that equations 19 and 20 apply equally to systems with aeration only and aeration with sludge return. As previously noted he assumes that t is the displacement time based on feed flow through the total volume of the system, including the final settling tank if present.

2. Eckenfelder's equation

Eckenfelder (5, 7) uses a similar concept to derive an equation for rate of oxygen use. His equation is

$$\frac{dO}{dt} = a' \frac{dL}{dt} + bS \qquad 21$$

where terms and units used in his work are

a' = fraction of 5-day BOD removed which is converted to energy

dL/dt = 5-day BOD removal rate, pounds/day

b = endogenous respiration rate, pounds oxygen/day per

pound of volatile suspended solids.

S = Volatile suspended solids in aeration tank, pounds.

It is to be noted that oxygen use for endogenous respiration in equation 21 is based on total volatile solids, not on the active mass. Thus, Eckenfelder's endogenous respiration constant b is not the same as the constant k_e of equations 7 and 19. If b is, in fact, constant then the ratio of active mass to total volatile solids must also be constant. This has not been demonstrated.

3. McKinney's simplified analysis

McKinney uses a similar but simplified approach to demonstrate the mechanics and magnitude of oxygen requirements (16). He reasons as follows:

1. One pound of 5-day BOD is approximately equivalent to 1.5 pounds of total or ultimate BOD for normal wastes.
2. On the basis of a synthesis-energy ratio of 2:1 this 1.5 pounds of ultimate BOD will
   a. Produce 0.7 pound of cells having an oxygen equivalent of 1 pound (1.42 pounds oxygen per pound of cell material),
   b. Use 0.5 pound of oxygen to provide the energy for production of the cells.
3. Approximately 75 per cent or 0.53 pound of the 0.7

pound of cells will undergo auto-oxidation (endogenous respiration) which will require 0.53 x 1.42 or 0.75 pound of oxygen. The remaining 25 per cent of the original 0.7 pound of cells or 0.17 pound of cells are inert, non-oxidizable organic materials.

This balance indicates that 0.5 pound of oxygen is required for energy and 0.75 pound of oxygen is required for endogenous respiration or a total of 1.25 pounds of oxygen are required for each pound of 5-day BOD removed. If McKinney's elemental balance is expressed as a rate of removal, it becomes functionally identical to Eckenfelder's more sophisticated expression, equation 21.

4. Verification of oxygen use relations

If equations 19, 20, and 21 and McKinney's elemental balance are valid descriptions of oxygen requirement in an activated sludge system, they can be used to predict the oxygen requirements. Furthermore, if measurement of the actual oxygen uptake can be made in an operating waste treatment system the results should correspond to the predictions of the derived equations. Complete validation of the equations requires confirmation of both time and total oxygen use.

The facilities available for this study did, in fact, permit direct measurement of oxygen utilization in a continuously operating system. The balance of this thesis describes this

system and the work leading to the conclusion that the foregoing representations of oxygen utilization, which are based solely on conditions within the aeration tank (and the effluent in a completely mixed tank) neglect the effect of the inflow mixture of raw waste and return sludge. Hence the foregoing developments represent only part of the elementary balance of oxygen demands around the aeration tank and are not valid representations of the total oxygen requirement in an activated sludge system with sludge return.

IV. EXPERIMENTAL WORK

A. Objectives

As previously discussed, the principle objective of this study was to test the validity of equations 19, 20 and 21 and of the direct relationship between BOD removal and oxygen use implied by McKinney (16) when these concepts were applied to a continuous flow aeration system.

A secondary objective was to ascertain the true flow character of the pilot plant in order to relate observed performance to actual conditions in the plant. An important part of this phase of the work involved making a series of observations to determine the amount of activity taking place in the aeration tank as compared with the final tank so that effective reaction volume and time could be defined.

B. Pilot Plant

The laboratory pilot plant was placed in operation in May 1960 and has been operated periodically for various studies since that date. The plant consists of the following units:

1. A 550 gallon steel, hopper bottom primary settling tank.
2. An identical holding tank wrapped in insulation and surrounded with cooling coils to hold waste feed temperature at approximately 20°C in warm weather.

3. A 6" x 10" x 8'-0" deep aeration column equipped with a 6" saran wrapped air diffuser in the bottom. Capacity of the tank is 25 gallons or 95 liters.
4. A 13.5" diameter by 7' - 11" average depth final settling tank equipped with a small rotary sludge scrapper in the bottom and with glass panels for viewing sludge character and level.
5. A centrifugal transfer and circulating pump for transferring settled sewage from the primary tank to the holding tank and for circulating contents of the holding tank to maintain a homogenous feed while that batch is being fed to the aerator.
6. A dual-3 cylinder Brosites pump, one side of which pumps feed from the holding tank to the aeration tank, the other side of which pumps return sludge from the bottom of the final settling tank to the aeration tank. All cylinders may be independently adjusted to control pumpage rate.
7. A time controlled sigma pump used for control of solids concentrations by wasting mixed liquor.

Figure 3 shows a schematic diagram of the pilot plant. Figure 4 is a photograph of the installation.

Iowa City screened sewage can be pumped to the primary tank by an air lift pump. This was the principle waste used in these studies although it was fortified with dry dog food

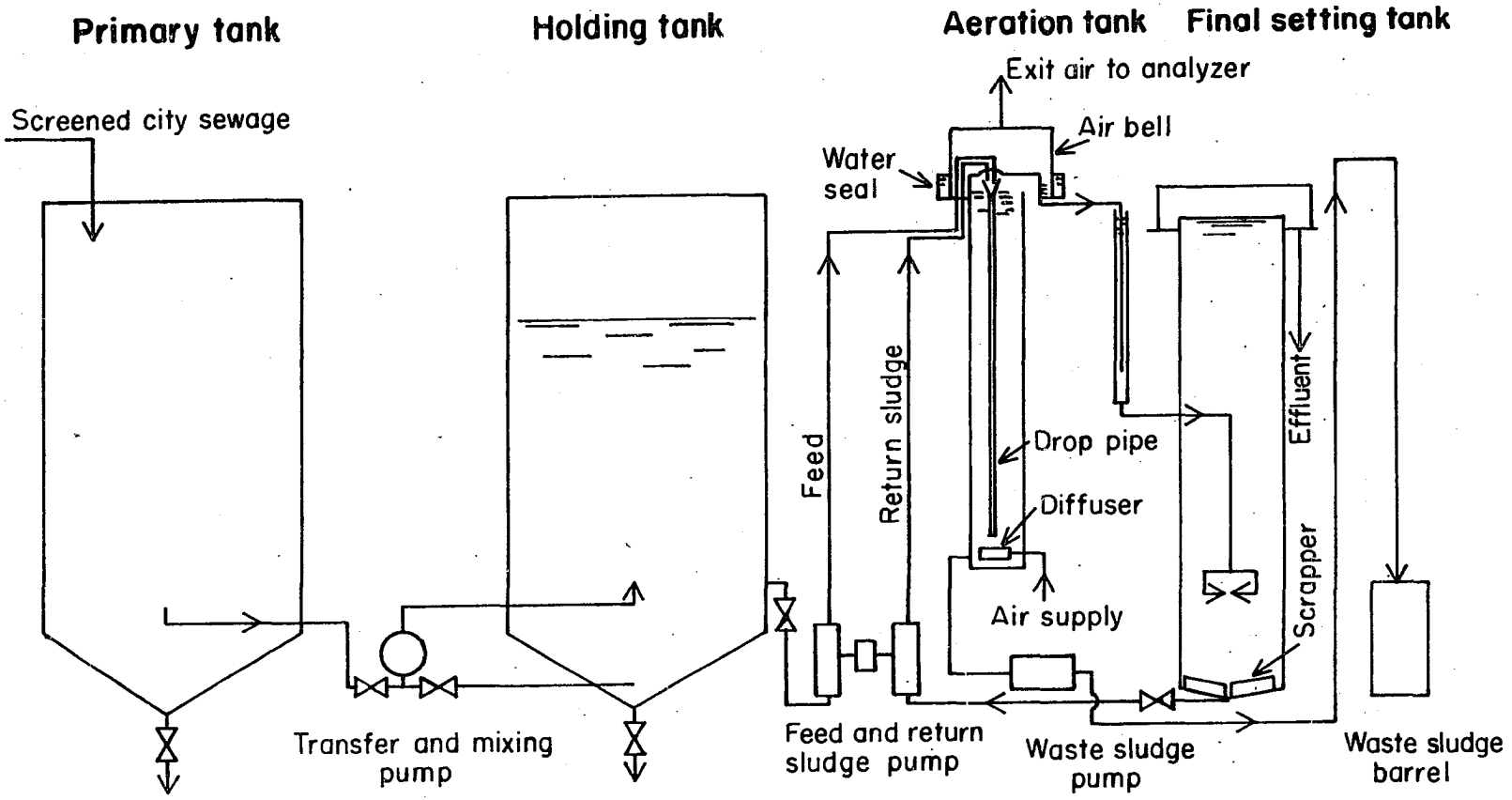


Figure 3. Schematic diagram of pilot plant

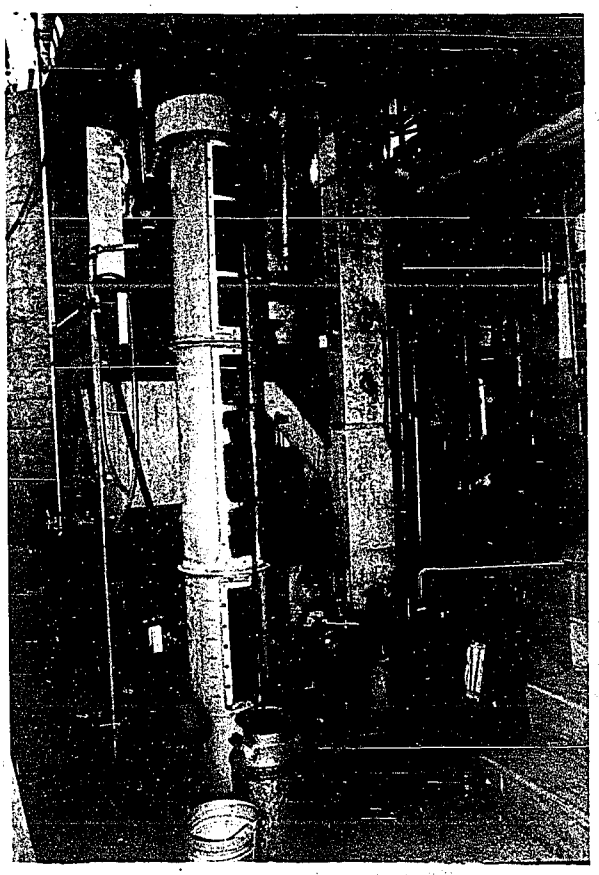


Figure 4. Pilot plant

or dried skim milk solids as discussed later. Air for aeration is supplied from the laboratory compressed air system. Inlet air flow is measured by an orifice system described later.

In operation, the primary tank is filled with screened city sewage. Raw sewage is settled for 30 to 60 minutes and sludge is then drawn from the hopper bottom. After settling and sludge withdrawal, supplemental nutrients, if any, are added to the primary tank and air mixed for about 10 minutes. After settling and mixing, if required, the settled waste is transferred to the holding tank from which it is fed to the aeration tank by the feed-return pump. The transfer pump is valved so that it may be used to circulate the holding tank contents to insure uniformity of feed quality during the batch feeding period. Previous studies have shown this system to maintain constant BOD over a 24 hour period so that a single BOD sample each day suffices for determination of daily feed quality (4). The holding tank provides sufficient volume for 24 to 48 hours of pilot plant feed depending on the rate of feed.

Both feed and return sludge are discharged at the top of the aeration tank to a 1" drop pipe which terminates one foot above the bottom of the aeration tank. This gives an intimate contact of feed and return sludge for from 3/4 to 2 minutes at the range of flows used in these studies before the mixture enters the aeration zone. The aeration tank exhibits complete

mixing characteristics as shown quantitatively by a chloride "washout" test and qualitatively by a dye test as described below. Aeration period, based on feed plus return sludge, varied from 0.6 to 2.2 hours during these studies.

Mixed liquor from the aeration tank is discharged to the final settling tank through 2 -3/4" opposed pipe nipples located in centrally in the final tank approximately 18" above the bottom. This arrangement dissipates the velocity energy of the influent without disturbing the sludge blanket as long as the blanket does not rise above the mixed liquor outlet. The area of the settling tank is 1 square foot and rise rates, based on through-put, varied from about 150 to 400 gallons per square foot per day in these studies.

Solids are automatically controlled by wasting mixed liquor from the aeration tank with the time controlled sigma pump. Wasted sludge is collected in a 50 gallon steel drum for measurement of waste sludge volume. The system affords good solids control as long as feed rate and strength remain constant and the final tank sludge blanket level remain below the mixed liquor outlets. Variations can be handled by adjusting the waste rate or by supplemental wasting from the bottom of the final settling tank.

Prior experience has indicated most satisfactory operation occurs when the return sludge rate is about 100 per cent of feed rate. This ratio was approximately maintained during these

studies. Lower return rates tend to cause increased sludge blanket depth in the final tank. When the top of the sludge blanket rises above the mixed liquor outlets, expansion of the sludge blanket by the up-flow makes it impossible to achieve the return sludge compaction necessary for high solids concentration required for low return/feed ratio. The sludge blanket did rise above the mixed liquor outlets on several occasions. When this occurred, the blanket expanded rapidly until almost the entire settling tank was filled with sludge. This condition, however, did not affect plant performance. The normal crystal clear effluent, substantially free of suspended solids, was maintained even with the expanded sludge blanket. In all such cases, a sharp zone of separation between sludge blanket and supernatant was present. This is not classical "bulking".

High sludge volume indices are the rule in this plant but there is no evident relation between sludge volume index and effluent quality. Neither was any relation between sludge blanket depth and sludge volume index observed. Radical changes in sludge depth seem to be more closely associated with hydraulic conditions within the final settling tank. These are merely qualitative observations since sludge conditions were not the primary purpose of this investigation as long as they did not affect effluent quality and the desired mixed liquor solids level could be maintained.

C. Aeration Tank Flow Characteristics

The differences between complete mixing and "plug" flow have been noted previously. It was assumed that the aeration tank had complete mixing flow characteristic but it was necessary to verify this assumption due to the important effect of flow characteristics on any interpretation of results.

If complete mixing exists, any tracer material added instantaneously will appear at once in the effluent and the material will be exponentially "washed out". The rate of change in tracer concentration in the tank and effluent will then be proportional to the amount remaining in the tank. This may be expressed as follows:

$$-\frac{dx}{dt} = Dx \qquad 22$$

where

x = tracer concentration in the tank and effluent

D = proportionality factor, also dilution factor or number of tank displacements per unit time

$\frac{dx}{dt}$ = rate of change of tracer concentration in the tank and effluent.

Equation 22 may be solved by integration between limits t_o and t. The solution becomes

$$\ln \frac{x}{x_o} = -Dt \qquad 23$$

where x_o is the tracer concentration at t_o. This is a typical exponential decay equation. The value of D can be determined by the time required for the tracer concentration to

decrease to some given fraction of the original concentration from a plot of log x/x_o, or relative concentration, against time. For convenience, the time for the tracer concentration to reduce to 1/2 the original value is used. Equation 23 then becomes

$$\ln 1/2 = - Dt_{1/2} = -\ln 2$$

or

$$D = 0.693/t_{1/2} \qquad 23a$$

D has the dimensions of t^{-1} and if the reactor is completely mixed its value will be the same as the number of displacements per unit time.

The aeration tank flow characteristic was first checked qualitatively by adding 200 ml of flourescein dye solution to the mixture of feed and return sludge at the top of the drop pipe. After approximately 1 minute, which was the time for flow through the drop pipe, dye appeared initially and in maximum intensity in the tank effluent. This indicated probability of complete mixing in the aeration tank.

The flow pattern was then checked quantitatively by adding 10 grams of sodium chloride dissolved in about 250 ml water to the aeration tank through the drop pipe. Prior to addition, the chloride level in the mixed liquor filtrate was determined by titration with $AgNO_3$. Tank effluent samples were collected immediately upon addition of the Nacl solution and at frequent intervals thereafter. A chloride determination was made on

the filtrate from each sample. Figure 5 is a semi-log plot of the excess chloride concentration against time compared with the line resulting from a displacement rate of 0.97×10^{-2}/ minute resulting from the measured flow of 348 gallons per day. The plot shows excellent correlation between observed values and the theoretical line for complete mixing. The dispersion and appearance of chloride in the tank effluent at near maximum value immediately after residence time in the drop pipe is also clearly indicated. This test justified the use of completely mixed flow conditions through the aeration tank in developing concepts of mechanisms in this particular plant.

D. Growth Condition

The region of the growth curve, Figure 1, in which a system is operating must be known to depict growth, BOD removal and/or oxygen use mechanisms. The pilot plant has been shown to have a completely mixed aeration tank and, as noted in subsequent discussion, essentially all BOD removal and oxidation processes are shown to occur in the aeration tank. Therefore, the effective food/microorganism ratio is the ratio of the BOD remaining in the mixed liquor filtrate to the active mass of the mixed liquor solids. Volatile solids in the mixed liquor will exceed actual active mass due to the presence of biologically inert volatile material. Volatile solids may, however, be used as an indicator of the active mass present.

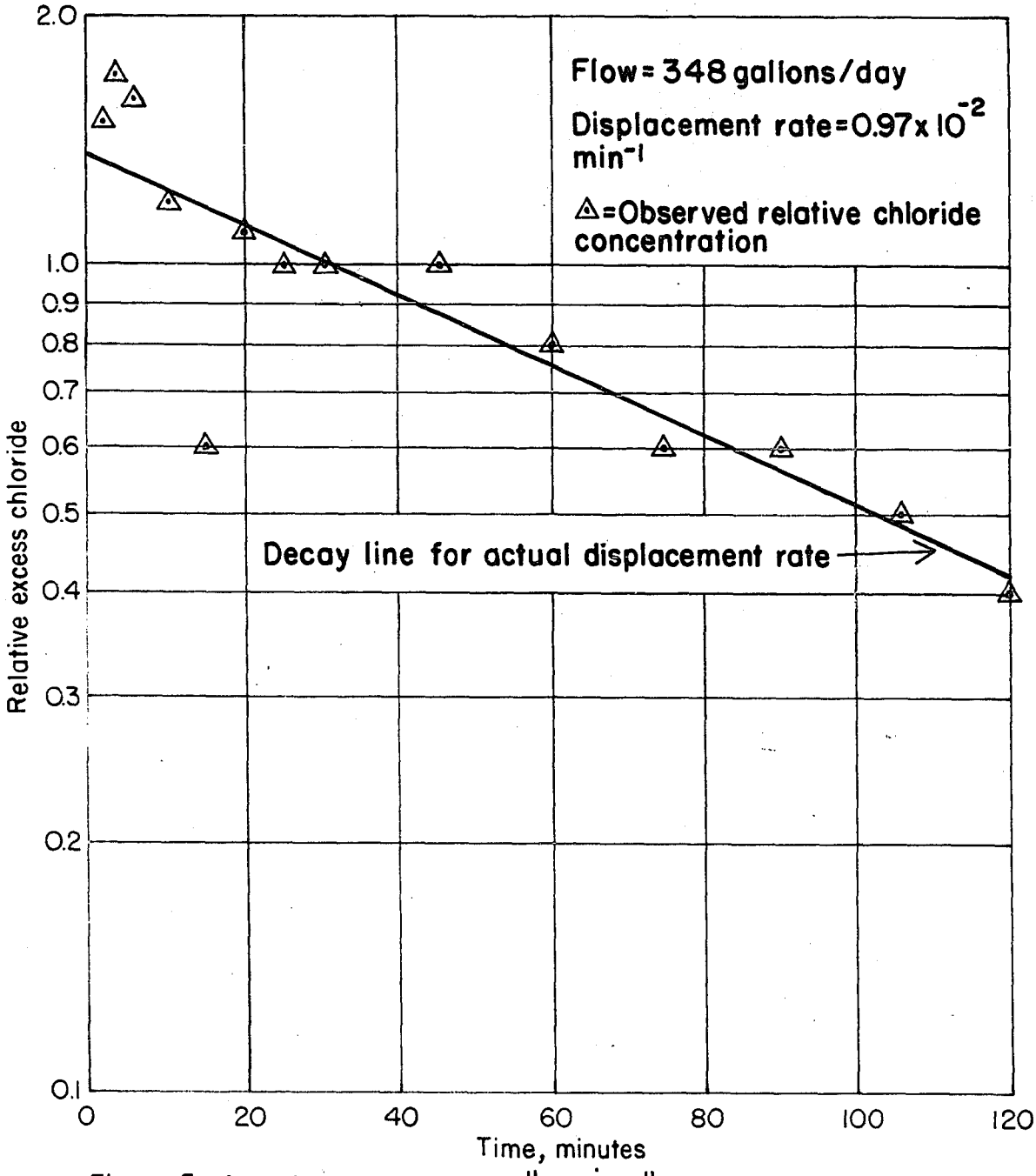


Figure 5. Aeration tank chloride "washout"

The BOD/mixed liquor volatile solids ratio ranged from about 0.01 to 0.003 during these studies. As previously noted, log growth conditions exist when the BOD/microorganism ratio exceeds 2.5 and declining growth conditions exist when the ratio is between 0.006 and 2.5. Auto-oxidation occurs when the ratio is less than 0.006. The actual food/microorganism ratio will be higher than the BOD/mixed liquor volatile solids level but the actual ratio will not be more than two to three times the BOD/volatile solids ratio. Therefore, it is clear that this system operates in the late stages of declining growth (or an early stage of auto-oxidation) and BOD removal is proportional to BOD remaining, assuming a constant synthesis/energy ratio. It is also clear that a high degree of BOD removal is to be expected.

Moreover, if Eckenfelder's hypothesis that absorbed nutrient removed initially from the feed is metabolized at the end of the log growth phase is true, initial removal by adsorption need not be considered since the system is operating so far along the growth curve that any adsorbed material will have been metabolized and no unmetabolized material will be carried to the final tank in the sludge mass. This is substantiated by the small change in COD through the final tank as noted in subsequent discussion.

E. Air Stream Oxygen Balance

Quantitative measurement of oxygen difference between the inlet and exit air streams to the aeration tank is the most important experimental feature of this study. Development of the system has been described in detail (20). It is based on measurement of the components of the ideal gas equation for the inlet and exit air streams. The equation is

$$n = \frac{PV}{RT} \qquad 24$$

where terms and units are

n = pound moles of gas

P = absolute pressure of gas, pounds per square foot

V = gas volume, cubic feet

R = gas constant, 1545 pound-feet per pound-mole ^{o}Rankine

T = absolute gas temperature, ^{o}Rankine

The moles of any particular component of the gas are given by the total moles times the volumetric percentage of that component.

The quantities, P, V and T of equation 24 for both inlet and exit air can be measured by common methods with necessary precaution to achieve high accuracy, particularly in measurement of gas volume. In this study, measurements were made by the following methods:

1. Atmospheric pressure was measured by a mercury barometer. This pressure was added to gage pressures

of inlet and exit air to obtain absolute pressure.

2. Inlet air gage pressure was measured by a mercury manometer connected to the high pressure tap of the inlet air orifice.
3. Exit air gage pressure was measured by a water manometer on the precision wet test meter.
4. All temperatures were measured by mercury thermometers inserted in the air stream.
5. Inlet air volume was measured by orifices inserted between flanges in a run of 1/2" OD brass pipe. The orifice flanges were preceded by 30 inches of straight pipe and followed by 12-1/2 inches of straight pipe. Pressure taps were 1/4" diameter located 1" upstream and downstream from the orifice. Orifice discharge was computed on the basis of ASME test code formulae (2) as applied by the author (20). Orifices were calibrated against the precision wet test meter used for exit air volume measurement. The calibration was plotted in the form of SCFM/TFxPF vs $\sqrt{h}$ where

SCFM = air flow, cubic feet per minute at 60°F and 1 atmosphere (14.697) psia,

TF = temperature factor, $\sqrt{\frac{520}{^{\circ}R,\ \text{observed}}}$

PF = Pressure factor, $\sqrt{\text{abs. static pressure, psia}}$

h = orifice differential pressure, inches water.

This plot gives a straight line over the range of $\sqrt{h}$ for which the basic orifice equation is valid. Orifice diameters of 0.101", 0.14" and 0.17" were used for different flow ranges to limit $\sqrt{h}$ to that portion of the calibration plot giving a straight line relation. Figure 6 is the calibration plot for the 0.14" orifice. Orifice differential was measured by water draft tube inclined 1 vertical to 2 horizontal to give magnification of differential. This was compared with a Rouse alcohol manometer built at the Iowa Institute of Hydraulic Research and reading to 0.001" of alcohol. Maximum difference between the Rouse manometer and the orifice draft tube was 0.04" of water at 2" total differential. This is a difference of only 0.6 percent in the value of $\sqrt{h}$. Draft tube differentials were used without correction.

6. Exit air volume was measured by the wet test meter which was used as the primary standard for orifice calibration. Air flow rate was determined by timing the passage of 1 cubic foot or four revolutions of the meter sweep hand.

Measurement of air stream oxygen content involved the special application of a paramagnetic oxygen analyzer (20). The analyzer was calibrated against saturated inlet air at the upper set point and against tank gas containing 10 per cent oxygen in nitrogen at the lower set point. Since the analyzer operates on saturated air, the inlet air stream was saturated and assigned an oxygen value of 20.95 per cent by volume which

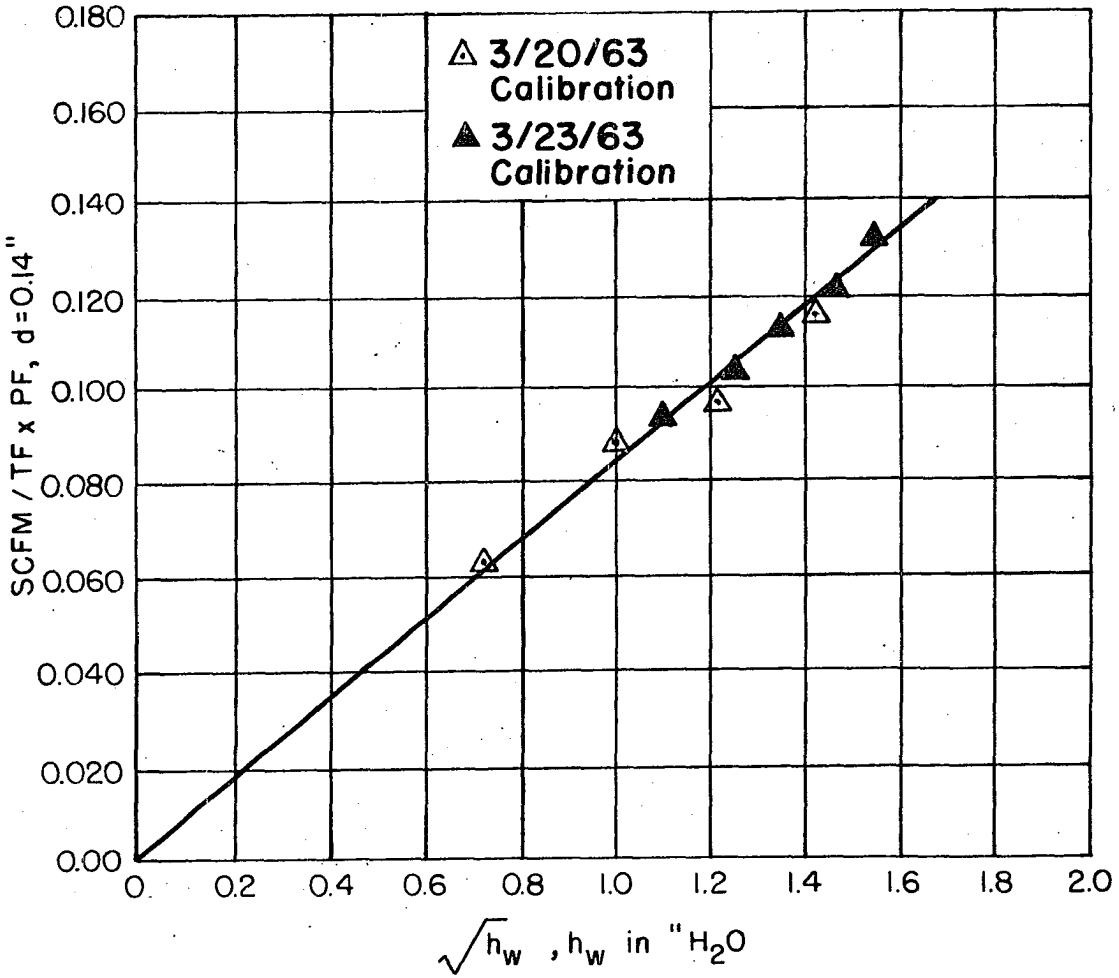


Figure 6. Calibration of 0.14" orifice

is the normal value for dry air. Thus, dry air values were indicated by the analyzer and exit air, which was saturated after passage through the saturator and the aeration tank, was corrected to dry volume by subtracting the saturated vapor pressure from the total pressure for the observed temperature in accordance with Dalton's law of partial pressures. Observed volumetric oxygen percentages applied to the pound moles of dry exit air gave the pound moles of oxygen in the exit air. The pound moles of oxygen times 32, the molecular weight of O_2, gave pounds of oxygen in the exit air stream per unit of time.

Inlet air pound-moles per unit time times 20.95 per cent gave the inlet oxygen pound-moles per unit time. Oxygen pound-moles per unit time times 32 gave pounds of oxygen per unit time in the inlet air stream.

The difference between inlet and outlet oxygen per unit time gave the rate of oxygen uptake in the aeration tank.

Figure 7 shows the portable plexiglas collection bell in place over the aeration tank. Figure 8 shows the analyzer panel. The recorder of the analyzer provides a convenient means of showing when air stream oxygen stability has been reached after a change of pilot plant operating conditions.

A typical calculation for an air run is shown on the following pages.

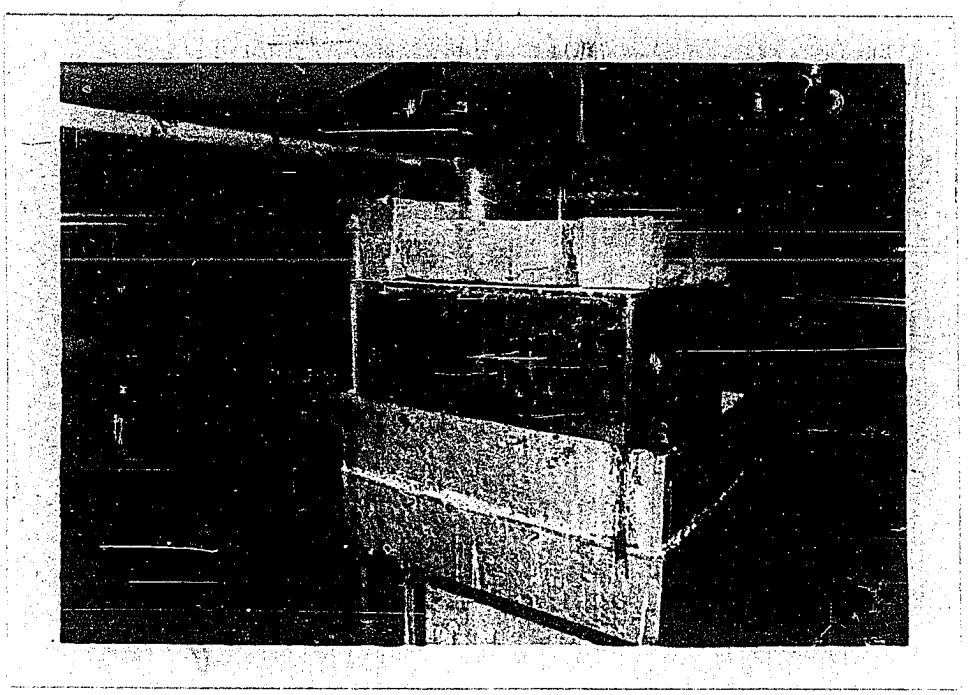


Figure 7. Exit air collection bell

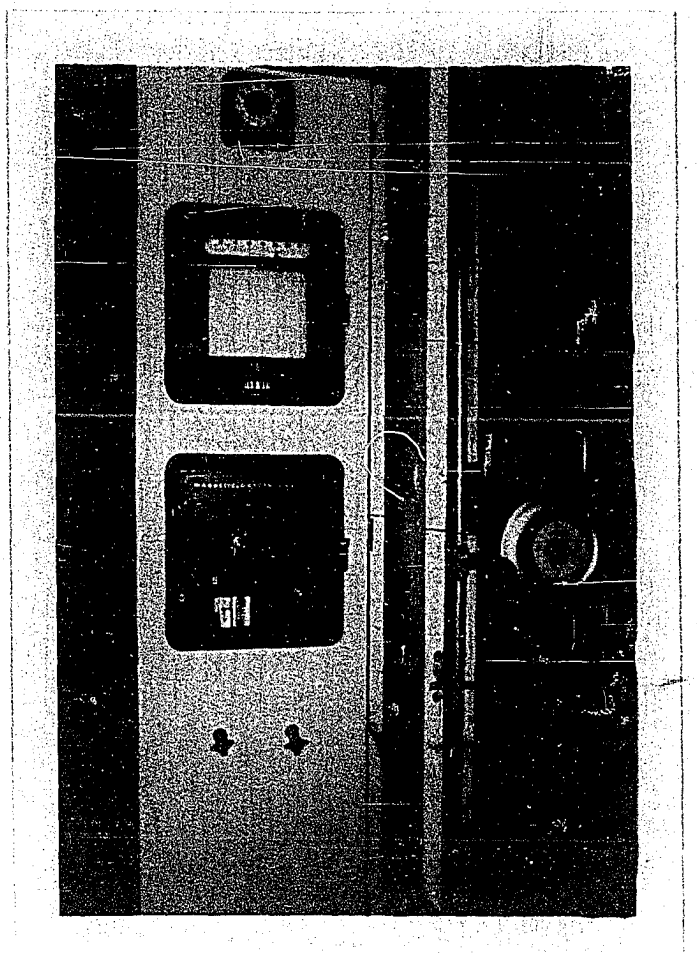


Figure 8. Oxygen analyzer panel and exit air meter

Typical calculation for oxygen uptake from air stream:

Run number	Utilometer 12
Date	April 10, 1963

Inlet air

Orifice diameter	0.14 inch
Differential head	1.125 inches
Square root of head	1.060
Barometric pressure	746.3 mm Hg 14.41 p.s.i.g.
Orifice static pressure	9.0 inches Hg 4.41 p.s.i.g.
Absolute pressure	18.82 p.s.i.a.
Temperature	82° F 542° Rankine
Temperature factor = $\sqrt{\frac{520}{°R}}$, TF	0.980
Pressure factor = $\sqrt{\text{p.s.i.a.}}$, PF	4.34
Scfm/TF x PF, from orifice calibration, $\sqrt{h}$ = 1.060,	0.0892
Standard cubic feet/min., 60°F, 14.7 psia	0.379 scfm
Moles dry air/hour = scfm x 0.1553	= 0.0589 moles/hr
Pounds oxygen/hour = moles air/hour x 0.7095 x 32	= 0.395 lb/hr

Exit air

Metered volume of air	1 cubic foot
Time	2.68 minutes
Metered rate of air, V	21.4 cubic ft/hr
Meter pressure	1.75 inches water 0.063 p.s.i.

Typical air calculation (continued)

Barometric pressure	14.41 p.s.i.
Absolute pressure, saturated	14.473 p.s.i.a.
Temperature, T	70 ^{o}F 530 ^{o}R
Saturated vapor pressure 0.70^{o}F	0.360 p.s.i.
Absolute pressure, dry, P	14.113 p.s.i.a. 2038 p.s.f.a.
Moles dry air/hour, $\frac{PV}{RT} = \frac{2038 \times 21.4}{1545 \times 530}$	0.0532 mole/hr.
Oxygen content, % by volume	19.90 %
Pounds oxygen/hour = moles air/hr x.1990 x 32 =	0.339 lb/hr

Balance

Oxygen in inlet air	0.395 lb./hr
Oxygen in exit air	0.339 lb./hr
Oxygen uptake rate	0.056 lb./hr

F. Oxygen Balance

The only source of oxygen in the aeration tank, assuming no dissolved oxygen in the feed and return sludge, is from the air stream. This oxygen must satisfy the following uses:

1. Biological oxidation requirements
2. Increase of dissolved oxygen of the feed and return streams to the concentration maintained in the aeration tank.

3. Nitrification. This is a biological use but is considered separately because of its variable nature.

The objective of this work is to relate the biological oxidation requirement to the quantity and/or rate of oxygen removed from the air stream. Each component may be expressed as a total quantity over a given time or as a rate, i.e., quantity per unit time.

Relative magnitudes of the rates of oxygen use by the various components during this study were as follows:

1. The rate of supply in the air stream varied from 0.005 to 0.056 pound of oxygen per hour. All but 4 of 43 runs tabulated in the subsequent summary of operating results lie between 0.013 and 0.040 pound per hour.
2. The maximum rate of dissolved oxygen gain occurred when a total feed and return flow of 635 gallons per day gained 5.1 mg/l of dissolved oxygen. This was a rate of about 0.0001 pound of oxygen per hour or less than 0.5 per cent of the observed supply rate of 0.026 pound per hour.
3. The maximum nitrification observed was approximately 24 mg/l NO_2 -N and 1 mg/l NO_3 -N in a feed flow of 150 gallons per day. This has an oxygen equivalent of approximately 87 mg/l or 0.0005 pound of oxygen per hour or about 2 per cent of the observed supply rate of 0.023 pound per hour.

The above levels of oxygen use for dissolved oxygen gain and nitrification are maximum observed values. Since these maximum values have only minor effect on the total oxygen balance, the effect of dissolved oxygen gain and nitrification has been neglected and the rate of oxygen removal from the air has been accepted as the rate of biological oxygen uptake in this activated sludge system.

G. Waste

Originally, a synthetic waste of controllable and reproducible quality having suspended solids characteristics comparable to ordinary domestic sewage was to be used in the pilot plant. On the basis of work at the U. S. Public Health Service Robt. A. Taft Sanitary Engineering Center[1], an extended effort was made to use a waste consisting of 1314 grams of dry dog food, homogenized and dispersed in 517 gallons of tap water with 100 grams of $(NH_4)_2\ SO_4$ added to supply nitrogen in readily available form. This waste had a 5-day BOD of about 150 mg/l. However, most of the BOD was present as readily settleable suspended material which also adhered to the holding tank walls. This caused large variation in suspended solids and in the BOD of the feed during a given batch period. Moreover, the activated sludge gradually took on a gelatinous character which became exceedingly difficult to dewater.

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[1]Ettinger, M.B., Cincinnati, Ohio. Private communication.

The feed recipe was then changed to 657 grams of dry dog food and 100 grams of $(NH_4)_2\ SO_4$ in 517 gallons of settled city sewage. It was considered necessary to fortify the city sewage because of its variability due to surface and ground water dilution. This proved to be a satisfactory feed from the standpoint of reasonably uniform strength but the gelatinous nature of the sludge persisted.

Accordingly, dog food was abandoned and dry milk solids were used to fortify the city sewage. A recipe of 275 grams of dry milk solids and 100 grams of $(NH_4)_2\ SO_4$ in 517 gallons of settled sewage proved satisfactory. The 5-day BOD of this waste ranged from 200 to 284 mg/l, the waste was very treatable and the sludge characteristics were good.

BOD rate constants in the classic BOD equation were determined for feed and final effluent. The equation is

$$L_t = L\ (1 - 10^{-kt}) \qquad 25$$

where

L = total or ultimate BOD

L_t = BOD exerted in t days

k = BOD rate constant, $days^{-1}$.

The milk fortified feed was found to have a rate constant of about 0.19/day. The constant for the plant effluent was about 0.055/day. These constants give a ratio of total BOD to 5-day BOD of 1.15 for feed and 2.2 for final effluent. The 0.19/

day constant for the feed indicates a waste well suited to treatment in a short aeration period. Its rate of BOD exertion is about twice the usual value of 0.1/day for normal domestic sewage. The lower constant for the final effluent is to be expected since the nutrients are the less readily oxidizable materials remaining after the quickly oxidizable materials have been consumed in the treatment system.

H. Plant Operation and Observations

As previously noted, the pilot plant has been in operation since May 1960 except for relatively short shutdown periods when no studies were in progress. The air stream oxygen measuring system was used for another study concerning oxygen transfer characteristics before the atart of the study reported herein. This considerable background in operating peculiarities and characteristics of the plant and instrumentation was most helpful.

Initial operation for these studies was commenced with a dog food and water waste on November 5, 1962. This operation continued until January 1, 1963 when the feed was changed to city sewage fortified with dog food. Fortification of the feed with skimmed milk solids in place of dog food started on February 5, 1963 and continued for the duration of the study. Operation was continuous for the entire period from November 5,

1962, until all observations were completed about May 1, 1963.

The following routine determinations were made:

1. Mixed liquor and final effluent dissolved oxygen concentrations were determined at least once each day.
2. Mixed liquor total suspended and volatile solids concentration and 30 minute settleability for sludge volume index were determined daily.
3. Return sludge total and volatile suspended solids concentrations were determined periodically. Determinations were made daily when time permitted and were usually made whenever oxygen uptake from the air stream was measured.
4. Feed and effluent 5-day BOD were determined, in general, each time an oxygen uptake measurement was made.
5. Feed flow rate was measured daily by observing the drop in level in the holding tank. Flow rate was always measured whenever an oxygen uptake measurement was made by timing the filling of a 2100 ml jar.
6. Return sludge flow rate was usually determined daily and always whenever an oxygen uptake measurement was made. Rate was measured by timing the filling of a 2100 ml jar.
7. A number of total and volatile suspended solids determinations on feed and final effluent were made throughout the study.

In addition to these routine determinations, the following special determinations were made.

1. A series of chemical oxygen demand (COD) tests to show the relation between the COD and BOD of feed and final effluent, to show COD balance around the final settling tank and to show the COD/volatile solids relation in the mixed liquor and return sludge.
2. Ammonia, nitrite and nitrate nitrogen during a period of low loading when floating sludge typically associated with denitrification was encountered.

During the course of the study over 70 observations for oxygen uptake from the air were made. These were supplemented with respiration rate observations by means of galvanic cells in so-called "utilometers" in the later stages of the study. This work is discussed later in more detail.

A summary of routine observations made in connection with 43 oxygen uptake measurements for 3 series of runs is shown in Tables 1a, 1b and 1c. Uptake measurements not tabulated were made at earlier times during the same feed batch period or were incomplete insofar as supporting data were concerned. Feed composition for the three series was:

Series 1, runs 7 through 33: 657 grams dry dog food and 100 grams of $(NH_4)_2\ SO_4$ in 517 gallons of settled city sewage.

Series 1, runs 34 through 36: settled city sewage.

Series 2, all runs: 275 grams of dried milk solids and 100 grams of $(NH_4)\ SO_4$ in 517 gallons of settled city sewage.

Utilometer series, runs 3 through 7: same as series 2.

Utilometer series, runs 8 through 14: 400 grams of dried milk solids and 100 grams $(NH_4)_2\ SO_4$ in 517 gallons of settled city sewage.

All determinations were made in accordance with standard procedures of the American Public Health Association (1) where such standards were applicable. All final effluent BOD determinations were made on pasteurized samples using dilution water seeded with 10 ml/l of settled feed. Pasteurization was used to destroy nitrifying bacteria and avoid the effects of nitrification in the BOD determination.

Table 1a. Pilot plant operating results, series 1

Run no.	Date	Flow rate, gal/day Feed	Return	Total	Aeration Time, hrs	5-day BOD, mg/l Feed	Effluent
7	1-1-63	198	190	388	1.57	188	15
8B	1-2-63	167	170	337	1.81	188	15
11	1-2-63	145	140	285	2.14	-	-
13	1-3-63	135	140	275	2.22	-	-
15	1-4-63	160	210	370	1.65	-	-
18	1-6-63	135	170	305	2.00	157	10
19	1-7-63	150	165	315	1.93	157	10
22	1-8-63	162	163	325	1.88	245	22
25	1-10-63	165	115	280	2.17	-	-
28	1-17-63	100	115	215	2.84	153	21
29	1-18-63	112	115	227	2.69	180	7

Table 1a. (Continued)

Run no.	Date	Flow rate, gal/day Feed	Return	Total	Aeration Times, hrs	5-day BOD, mg/l Feed	Effluent
30	1-19-63	97	100	197	3.10	137	20
31	1-20-63	183	40	223	2.74	189	16
32	1-21-63	140	132	272	2.25	160	4
33	1-22-63	167	170	337	1.81	180	10
34	2-1-63	149	290	439	1.39	127	14
35	2-2-63	150	145	295	2.07	94	4
36	2-3-63	153	167	320	1.91	90	5

Run no.	Suspended solids, mg/l: Mixed liquor Total	Mixed liquor Volatile	Return sludge Total	Return sludge Volatile	Sludge volume index	Air flow rate: Total, scfm	scfm/gal total flow
7	3154	2544	5116	4095	79	0.2200	0.8
8B	3233	2608	6276	5053	183	0.1754	0.7
11	3050	2522	5694	4620	180	0.2165	1.1
13	2948	2456	5068	4154	175	0.2025	1.1
15	3229	2658	-	-	175	0.1628	0.6
18	3193	2627	4601	3818	213	0.1680	0.8
19	2958	2470	4921	4097	243	0.1925	0.9
22	2815	2290	4720	3956	239	0.144	0.6
25	2619	2182	5795	4789	343	0.243	1.6
28	2782	2259	2007	1602	204	0.268	1.8
29	2257	2030	-	-	117	0.245	1.5
30	1797	1404	6380	5025	95	0.181	1.3
31	2535	1989	20560	17644	182	0.140	0.9
32	2087	1671	3830	3078	350	0.162	0.9
33	1279	1040	2339	1874	563	0.174	0.7
34	2095	1685	2583	2060	449	0.267	0.9
35	1933	1478	2681	2051	439	0.280	1.4
36	1689	1250	1960	1456	355	0.143	0.6

Run no.	Dissolved oxygen, mg/l: Mix.liq.	Effluent	Oxygen uptake lbs/hr	Run no.	Dissolved oxygen, mg/l: Mix.liq.	Effluent	Oxygen uptake lbs/hr
7	0.8	0	0.0165	28	3.6	0	0.0440
8B	2.9	0	0.0155	29	2.6	0	0.0330
11	3.5	1.4	0.0185	30	2.1	1.1	0.0130
13	5.3	2.4	0.0155	31	0.8	0	0.0185
15	2.6	0.9	0.0212	32	1.9	0	0.0165
18	2.0	0.6	0.0185	33	2.0	0	0.0135
19	2.5	0.1	0.0325	34	3.2	0.9	0.0210
22	0.6	0	0.0190	35	3.9	2.3	0.0230
25	3.0	0	0.0255	36	2.0	1.0	0.0160

Table 1b. Pilot plant operating results, series 2

Run no.	Date	Flow rate, gal/day Feed	Return	Total	Aeration Time, hrs	5-day BOD, mg/l Feed	Effluent
37	2-5-63	153	147	300	2.04	228	12
38	2-6-63	147	138	285	2.15	158	8
39	2-7-63	142	142	284	2.15	83	10
40	2-8-63	160	140	300	2.03	243	4
41	2-9-63	152	130	282	2.17	203	6
42	2-12-63	183	192	375	1.63	230	37
43	2-14-63	183	183	366	1.67	209	7
44	2-14-63	184	186	370	1.65	209	7
45	2-15-63	170	182	352	1.73	217	5
46	2-16-63	176	180	356	1.71	262	5
47	2-19-63	160	185	345	1.77	222	4
48	2-20-63	140	187	327	1.87	270	4
49	2-24-63	180	160	340	1.80	-	-

Run no.	Suspended solids, mg/l: Mixed liquor Total	Mixed liquor Volatile	Return sludge Total	Return sludge Volatile	Sludge volume index	Air flow rate: Total, scfm	scfm/gal total flow
37	1785	1302	3145	2330	214	0.146	0.7
38	2492	1913	3738	2797	345	0.261	1.3
39	2945	2189	5144	3830	300	0.256	1.3
40	3235	2420	6143	4632	294	0.252	1.2
41	3470	2643	6780	5177	280	0.318	1.6
42	2014	1575	3791	3008	280	0.175	0.7
43	2165	1698	3792	3005	440	0.221	0.9
44	2090	1679	3792	3005	453	0.257	1.0
45	2212	1789	3620	2958	445	0.222	0.9
46	2384	1883	4114	3309	415	0.227	0.9
47	2766	2020	2951	2131	336	0.365	1.5
48	2424	1608	3820	2650	389	0.361	1.6
49	5197	3873	13280	10900	184	0.583	2.5

Run no.	Dissolved oxygen, mg/l: Mix.liq.	Effluent	Oxygen uptake lbs/hr	Run no.	Dissolved oxygen, mg/l: Mix.liq.	Effluent	Oxygen uptake lbs/hr
37	0.3	0	0.010	44	1.8	0	0.027
38	1.7	0	0.025	45	0.6	0	0.025
39	1.1	0	0.024	46	0.2	0	0.024
40	0.5	0	0.022	47	3.9	0.9	0.031
41	0.6	0	0.025	48	4.5	1.3	0.040
42	Trace	0	0.022	49	2.8	0.2	0.036
43	1.25	0	0.025				

Table 1c. Pilot plant operating results, utilometer series

Run no.	Date	Flow rate, gal/day Feed	Return	Total	Aeration Time, hrs	5-day BOD, mg/l Feed	Effluent
3	3-23-63	251	336	587	1.04	210	10
4	3-28-63	377	660	1037	0.59	284	13
5	3-29-63	287	348	635	0.96	210	5
6	3-30-63	295	325	620	0.98	229	2
7	3-31-63	287	317	604	1.01	229	2
8	4-1-63	273	277	550	1.11	220	12
9	4-2-63	305	325	630	0.97	233	9
10	4-6-63	287	297	584	1.04	283	21
11	4-8-63	363	372	735	0.83	200	9
12	4-10-63	363	355	718	0.85	203	11
13	4-13-63	374	195	569	1.07	229	10
14	4-16-63	187	140	327	1.86	-	-

Run no.	Suspended solids, mg/l: Mixed liquor Total	Mixed liquor Volatile	Return sludge Total	Return sludge Volatile	Sludge volume index	Air flow rate: Total, scfm	scfm/gal total flow
3	3209	2047	-	-	-	0.342	0.8
4	3882	2850	5547	4091	243	0.408	0.6
5	3145	2259	4770	3414	289	0.494	1.1
6	2250	1703	-	-	418	0.441	2.2
7	2038	1449	-	-	459	0.415	1.0
8	2204	1590	3748	2786	441	0.423	1.1
9	2195	1757	3361	2519	438	0.433	1.0
10	4059	2796	7330	5050	81	0.443	1.1
11	4061	2836	8092	5929	150	0.433	0.8
12	4219	3087	7811	5764	218	0.379	0.8
13	4306	3087	11085	7417	227	0.396	1.0
14	4216	3095	8379	6134	232	0.315	1.4

Run no.	Dissolved oxygen, mg/l: Mix. liq.	Effluent	Oxygen uptake lbs/hr	Run no.	Dissolved oxygen, mg/l: Mix. liq.	Effluent	Oxygen uptake lbs/hr
3	3.2	0	0.026	9	4.4	0	0.020
4	0.6	0	0.005	10	1.0	0	0.028
5	5.1	0.7	0.026	11	3.9	0.1	0.023
6	5.0	Trace	0.008	12	2.4	0	0.056
7	4.4	0	0.020	13	3.2	0	0.034
8	5.0	0	0.016	14	1.2	0	0.031

V. FINDINGS AND SUBSEQUENT WORK

A. Reaction Time

McKinney (14) considers reaction time, or effective time, in a system with sludge return to be the total displacement time of the aeration and final settling tank based only on the rate of feed. All McKinney's equations and his demonstrated application of them use this definition of the quantity, t. It is important that the value of t be consistent with the actual time during which the reactions are completed in order that the total change of any characteristic can be converted to a rate of change.

If t is a function of total aeration tank and final sedimentation tank volume, then biological oxidation and/or metabolism must proceed in the final settling tank as well as in the aeration tank. If this is the case, there should be a significant difference in such factors as COD or BOD through the final settling tank because the volume (and displacement period) of the final settling tank is 2.35 times that of the aeration tank. Furthermore, it would be reasonable to expect that the BOD and COD of the final effluent would be less than that of the mixed liquor filtrate due to continued metabolism in the final tank.

Two observations were made to determine the nature of BOD or COD change through the final settling tank.

1. COD determinations were made on mixed liquor, final effluent and return sludge so that a COD balance around the final tank could be made. COD was used in preference to BOD because of the high dilution factors and consequent inherent error in mixed liquor and return sludge BOD determination.
2. COD and/or BOD of several mixed liquor filtrate and effluent samples were determined and compared.

Tables 2 and 3 show COD values and the COD balance around the final settling tank. These data show a disappearance of less than 10 per cent of the mixed liquor COD at lower volatile solids levels and about 20 per cent disappearance at higher solids levels in the final settling tank. This apparent work done in the final settling tank is certainly not sufficient to conclude that this tank volume should be included in the calculation of effective reaction time, especially when, in this case, the resulting time would be 335 per cent of that for the aeration tank alone.

Table 4 shows COD or BOD values for simultaneously collected samples of final effluent and mixed liquor filtrate. Since the COD and BOD of the mixed liquor filtrate is equal to or less than that of the plant effluent it follows that no removal of organic material from the liquid fraction of the flow occurs in the final settling tank.

Table 2. Flow and COD for final tank COD balance

Run no.	Flow, gal/day			COD					
	Feed	Return	Total	Mixed liquor		Return sludge		Effluent	
				mg/l	lbs/day	mg/l	lbs/day	mg/l	lbs/day
43	183	183	366	2210	6.74	4300	6.56	108	0.17
45	170	182	352	2230	6.55	3900	5.92	117	0.17
46	176	180	356	2540	7.53	4550	6.83	74	0.11
48	140	187	327	2280	6.22	3590	5.60	87	0.10
51	173	175	348	5850	16.9	9340	13.60	127	0.18
52	160	175	335	5910	16.5	8880	12.82	102	0.14

Table 3. Final tank COD balance and COD/volatile solids ratio

Run no.	COD balance, lbs/day					Vol. solids, mg/l		COD/vol. solids	
	In(+)	Out(-)			Diff.	Mix.liq.	Return	Mix.liq.	Return
	Mix.liq.	Return	Effluent	Total	(+ or -)				
43	6.74	6.56	0.17	6.73	+0.01	1698	3003	1.30	1.43
45	6.55	5.92	0.17	6.09	+0.46	1789	2958	1.25	1.32
46	7.53	6.83	0.11	6.94	+0.59	1883	3309	1.35	1.38
48	6.22	5.60	0.10	5.70	+0.52	1608	2650	1.42	1.36
51	16.90	13.60	0.18	13.78	+3.12	4595	7574	1.27	1.23
52	16.50	12.82	0.14	12.96	+3.54	5054	7707	1.17	1.14

Table 4. Comparison of mixed liquor filtrate and plant effluent COD and BOD

Date	BOD, mg/l		COD, mg/l	
	Mixed liquor filtrate	Effluent, unfiltered	Mixed liquor filtrate	Effluent, unfiltered
2-28-63	3	4	106	102
3-26-63			83	106
3-27-63			75	92

In view of these data it must be concluded that the major mechanism in the final settling tank is solids separation -- not further metabolic action -- and that biological oxidation reaction rates should be calculated on the volume of the aeration tank only, since the biological reactions are essentially completed within that tank. Furthermore, residence and reaction time in the aeration tank should be a function of total feed and return flow, not feed alone. These conclusions were used in all subsequent work.

B. Oxygen Utilization-BOD Correlation

Oxygen requirements are frequently expressed as the quantity of oxygen required per unit of BOD removed. This is inferred from McKinney's simplified analysis of oxygen requirement in which he concludes that oxidation of one pound of 5 day BOD requires a constant quantity of 1.25 pounds of oxygen.

Equation 21 expresses the same concept,

$$\frac{dO}{dt} = a' \frac{dL}{dt} + bS \quad . \qquad 21$$

Division of equation 21 by S, volatile suspended solids in the aeration tank, gives

$$\frac{\frac{dO}{dt}}{S} = a' \frac{\frac{dL}{dt}}{S} + b \quad . \qquad 21a$$

If this is a valid relation, a rectangular plot of oxygen use rate per weight of solids under aeration vs. 5 day BOD removal rate per weight of solids under aeration should give a straight line. The slope of this plot should give a', the energy conversion fraction, and the intercept at

$$\frac{\frac{dL}{dt}}{S} = 0 \quad \text{should give b.}$$

Values of a' and b should, presumably, be constant for any given waste with an acclimated sludge.

Figure 9 is a plot of $\frac{dO}{dt}/S$ vs $\frac{dL}{dt}/S$ for all runs using milk fortified waste for which complete BOD and solids data were taken. This includes 23 runs over a period of about 6 weeks. It is obvious that correlation is of very low order.

Examination of the data expressing the BOD removed per unit of oxygen uptake without regard to volatile solids content of the aeration tank shows equally uncorrelated data. These values are shown in Table 5.

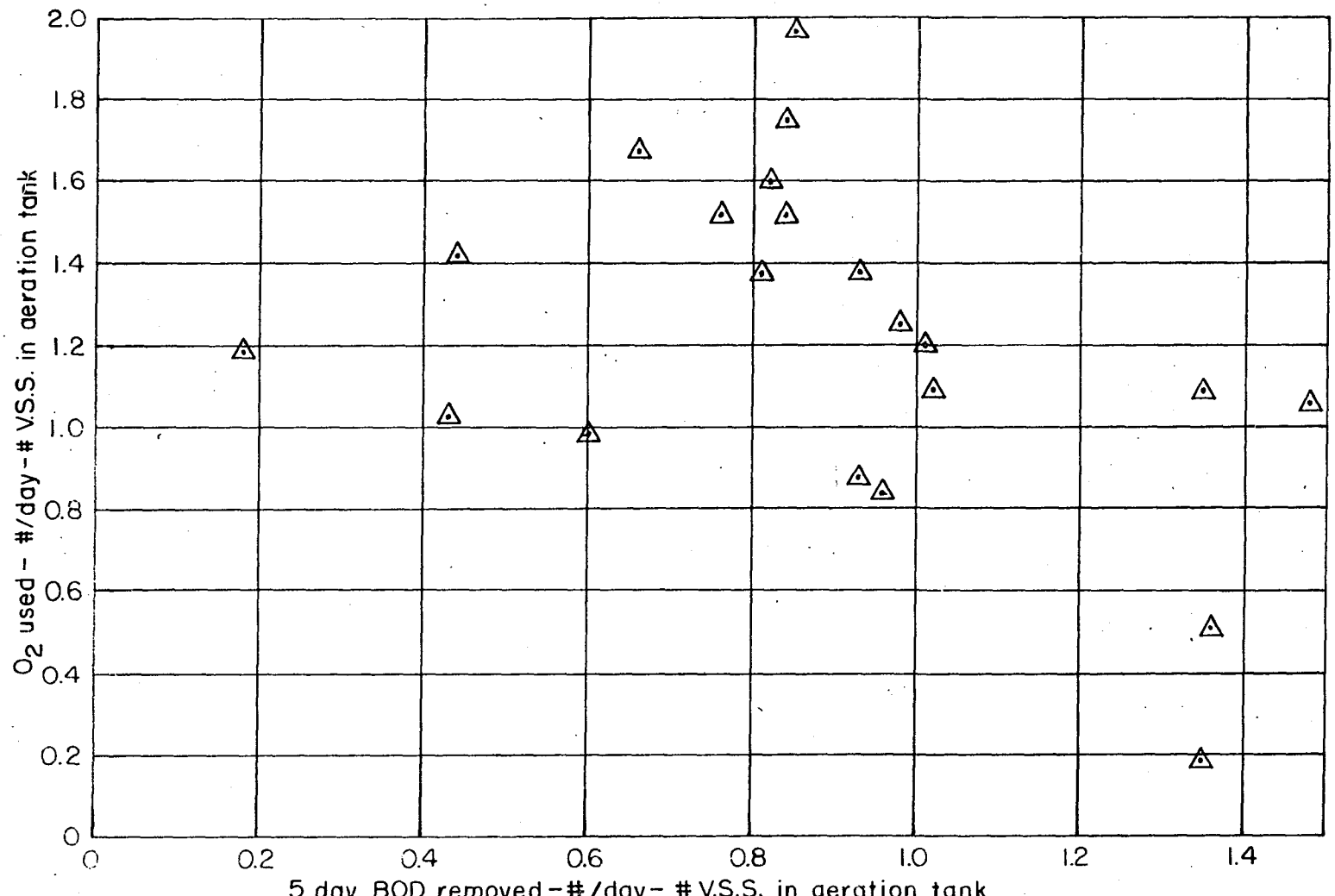

Figure 9. 5 day BOD removed v.s. oxygen uptake

Table 5. BOD removed-oxygen uptake

Run no.	BOD, lb/day Applied	BOD, lb/day Removed	% BOD removed	lbs BOD/ 1000 cu.ft. aeration tank	O_2 uptake, lb/day	lb O_2/ lb BOD rem.
			Series 2			
37	.290	.275	94.8	87	.240	0.87
38	.192	.184	95.7	58	.600	3.26
39	.098	.086	96.6	29	.576	6.72
40	.325	.320	98.5	97	.528	1.65
41	.258	.250	96.8	77	.600	2.40
42	.350	.294	83.8	105	.528	1.80
43	.318	.308	96.8	95	.600	1.95
44	.321	.310	96.6	96	.648	2.09
45	.308	.300	97.2	92	.600	2.00
46	.384	.377	98.3	115	.576	1.53
47	.296	.291	98.3	89	.745	2.56
48	.315	.310	98.3	94	.960	3.10
			Series Utilometer			
3	.439	.418	95.3	132	.624	1.49
4	.893	.851	95.3	268	.120	0.14
5	.502	.490	97.6	151	.624	1.27
6	.513	.508	98.8	154	.192	3.78
7	.538	.533	98.8	161	.480	0.90
8	.501	.473	94.4	150	.384	0.81
9	.593	.570	96.2	178	.408	0.72
10	.676	.626	92.6	203	.672	1.07
11	.605	.578	95.6	181	.552	0.95
12	.615	.581	94.5	184	1.345	2.31
13	.715	.683	95.7	214	.815	1.19

These data indicate the existence of factors other than BOD removal or volatile solids in the aeration tank which affect the rate of oxygen uptake and utilization by the system.

C. Oxygen Utilization-BOD and Active Mass Correlation

Having failed to establish correlation between oxygen uptake and BOD or between BOD and volatile mixed liquor solids, attention was directed toward equation 19

$$dO/dt = k_5 L_r + k_e M_a \qquad 19$$

for the mixed liquor with t defined as aeration displacement time for combined feed and return flows. This is a differential equation identical in form to the equation encountered in nuclear work for expressing the total activity resulting from two independent activities where the independent activities are decreasing as the result of first order decay. The constants, k_5 and k_e, are analogous to the decay constants of the two constituents and L_r and M_a are analagous to the amounts of the respective constituents present. The rate of oxygen use, dO/dt, is analogous to the total radioactivity (disintegration per unit time) of the two components. The total activity, dO/dt, is the sum of the individual activities, k_5L_r and k_eM_a. L_r and M_a are expressed as their respective oxygen equivalents and are dimensionally consistent with 0. Equation 19 may be written in the following form:

$$\frac{dO}{dt} = (k_5L_r)_o\, e^{-k_5t} + (k_eM_a)_o e^{-k_et} \qquad 26$$

in which $(k_5L_r)_o$ and $(k_eM_a)_o$ denote the activity or rate of oxygen uptake of each member at t = 0. Equation 26 gives the total oxygen uptake rate of a sample at any time t after the

sample is removed from the stream.

If the log of activity, or log dO/dt, is plotted against time and if the rates of oxygenation or values of k_5 and k_e, are different, a curve that is concave upward is produced. The curvature is the result of the decreasing significance of the more rapidly decaying component with the passage of time. After sufficient time, the slow decaying component will predominate and a straight line plot will result. If this straight line is extrapolated back to t = 0 and subtracted from the total curve, the residual will be a second straight line which represents activity versus time of the fast decaying component. The values of the oxygenation constants, k_5 and k_e, may then be determined from the time required for the activity of the respective components to reduce to a given fraction of original value. Usually the time for activity to reduce to 1/2 original value is used. This is the half life of radioactivity or of any exponential decay process. The amount of a component present at any time, t, is obtained by dividing the activity of that component at the given time by the oxygenation constant of the component.

If more than two components are present, subtraction of the extrapolated straight line of the slowest decaying component produces a second curve. This curve may be treated the same as the original curve and the process repeated until all components are identified by their straight line semi-log plots.

In practice, it is difficult to resolve a system with more than three components and even systems with only two components are not well defined unless the decay rates differ by a factor of two or more.

Complete description of the process of resolving multiple component exponential decay systems is given by Friedlander and Kennedy (9) and Overman and Clark (18).

If dO/dt of the aeration tank mixed liquor can be measured, equation 19 and the above described method of resolution provide a method of evaluating the oxygenation constant k_5 and k_e and the concentration of nutrient L_r and active mass M_a, in terms of oxygen equivalents. The galvanic cell oxygen analyzer developed at the University of North Carolina, School of Public Health, Department of Sanitary Engineering and described by Nancy and Westgarth (17) provides a means of measuring the rate of oxygen utilization. Two such cells made at the University of North Carolina together with reaction vessels were available and were used for this measurement. The equipment consists of two 2100 ml glass reaction vessels located in a water bath for temperature control with a galvanic cell installed in each vessel. Current output was indicated by a microammeter for each cell. The complete assembly is called an oxygen utilometer. Figure 10 shows the setup.

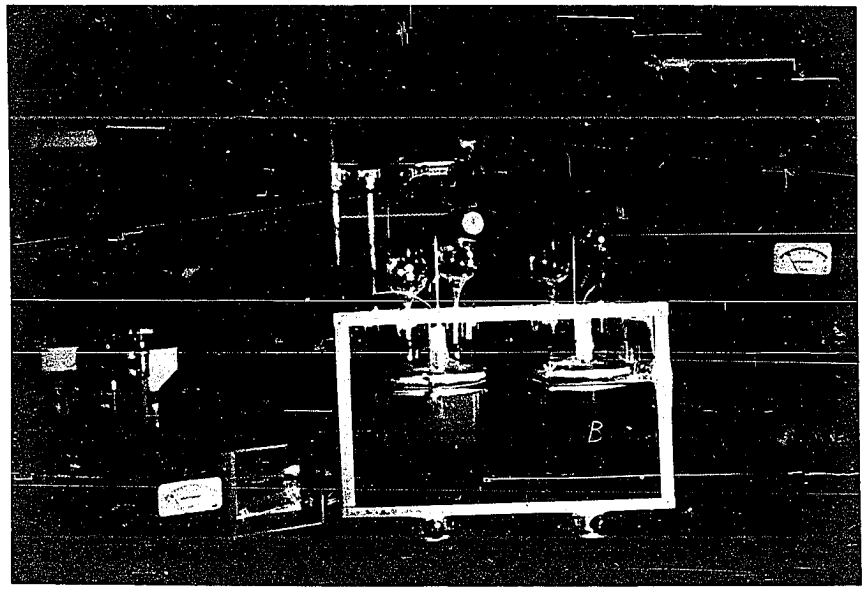

Figure 10. Oxygen utilometer

The method of use was as follows:

1. Residual current output of the cells was reduced to a minimum value by immersion in a sodium sulfite solution (zero dissolved oxygen).
2. The vessels were filled with tap water, the water temperature was adjusted to 20°C and the water was aerated with the cells in the vessels until the cell current output showed a steady value.
3. 750 ml of water was siphoned from the jar and 750 ml of mixed liquor added through a large funnel temporarily inserted in one of the thistle tube openings so the sample could be added without delay and readings started. Zero time was taken as the sample was withdrawn from the aeration tank and initial cell current readings were usually made in 30 to 45 seconds.
4. Cell current output and time were observed. Frequency of observation varied from one each 30 seconds during the first few minutes after the sample was added when the oxygen utilization rate changed rapidly to one each 5 to 10 minutes in the later part of the run when the oxygen utilization rate was low and changed only slightly. Dissolved oxygen of the mixture was calculated from the net cell current output (observed less residual reading) and the temperature vs $\frac{\text{microampere}}{\text{mg/l}}$ calibration for the cell as described below.

5. When the dissolved oxygen content of the mixture fell to less than 1 mg/l the mixture was reaerated. Readings were, of course, suspended during reaeration and resumed following reaeration. After a condition of slow, steady decay had been reached, continuous rate of use observations were suspended and observations made periodically over about a 24 hour period. The mixture was aerated continuously between these observations.

Cells were calibrated after installing new membranes by comparing net current output against dissolved oxygen of tap water at various temperatures. Observed values were plotted as $\log \frac{\text{microamperes}}{\text{mg/l of D.O.}}$ v.s. temperature. Figure 11 is a typical calibration plot for cell A. Recalibration after a period of use showed the calibration plot to have the same slope but lower values of $\log \frac{\text{microamperes}}{\text{mg/l of D.O.}}$ for given temperatures. Therefore, new calibration for a given run was determined simply by observing net current output, dissolved oxygen and temperature in a tap water sample and drawing the new calibration line through this point parallel to the original calibration line.

Dissolved oxygen of the mixture as determined from cell current output was plotted against time. The total rate of oxygen use, dO/dt, in units of mg/l hr during any time interval was determined by the slope of this plot multiplied by the

ratio, $\frac{\text{vessel volume}}{\text{sample volume}}$. This plot is curved, concave upward, when the rate is decreasing and becomes a straight line when the rate is constant over a given time interval. Average slope over time intervals of 1 or 2 minutes was used in the early stages when the rate was decreasing. Longer intervals of 5 to 25 minutes were used when the rate was steady over longer intervals.

The log of the calculated rate of oxygen use, log mg/l hr, was plotted against time. Observations were continued until the slope of the slow decaying, straight line portion of this plot was established. This required from 18 to 24 hours.

The plot of log mg/l hr vs. time is the plot of the differential relationship represented by equation 19. It can be resolved into its components and the oxygenation constants k_5 and k_e, representing the oxygen used per hour per mg/l of fast and slow decaying component present respectively, can be determined from the slope and half-life of the component plots. The intercept of the component plots at $t = 0$ will be the use rate of that component at the moment the sample was withdrawn from the aeration tank, in other words, the condition within the tank. Values of use rate at $t = 0$ divided by the respective constants will give the total oxygen demand of that particular component in the aeration tank.

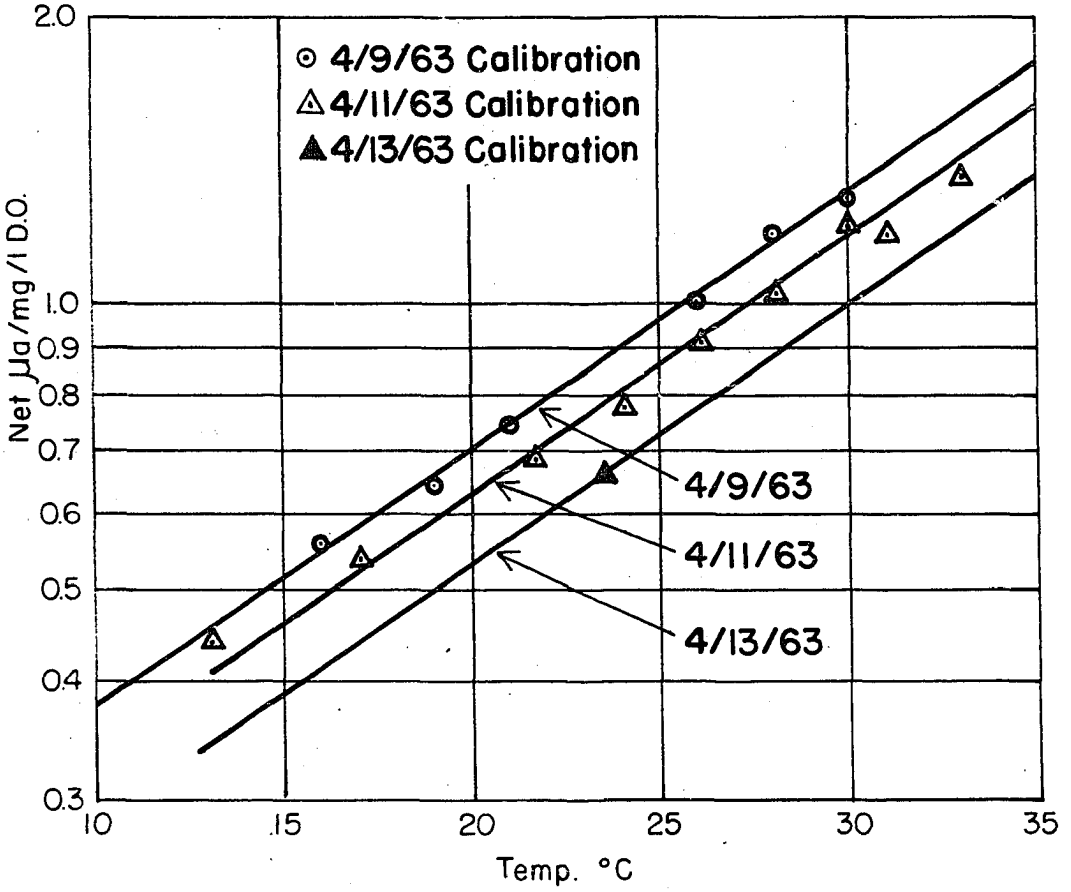

Figure 11. Calibration of galvanic cell A

Figures 12a and 12b are plots of log mg/l hr vs time and its resolution for a typical run (No. 12) on a mixed liquor sample. This shows that the mixed liquor sample consisted of two components; one with an oxygenation constant, k_5 of 9.2 mg/l hr O_2 per mg/l oxygen demand of component, and the other with an oxygenation constant, k_e, of 0.031 mg/l hr O_2 per mg/l oxygen demand of the component. The total rate of oxygen demand at t = 0 was 100 mg/l hr of which 43 mg/l hr was due to the high rate component and 57 mg/l hr to the slow rate component. These constants and rates correspond to total oxygen demand at t = 0 of 5 mg/l for the high rate component and 1840 mg/l for the low rate component.

If equation 19 is a complete and valid expression for the oxygen utilization rate in an activated sludge system, the total uptake rate given by Figures 12a and 12b should correspond with the rate of removal of oxygen from the air since this is the only source of oxygen and other uses, as previously shown, constitute only a negligible part of the total oxygen utilization. The measured rate from the air was 0.056 pound/hour or 25,400 mg/hour. The total waste flow rate (combined return sludge and feed) was 718 gallons per day or 113.2 liters per hour. Thus, the actual oxygen uptake mass from the air was 225 mg/l.

Equation 19 can be integrated to give equation 20

$$O = (k_5 L_r + k_e M_a)t \qquad 20$$

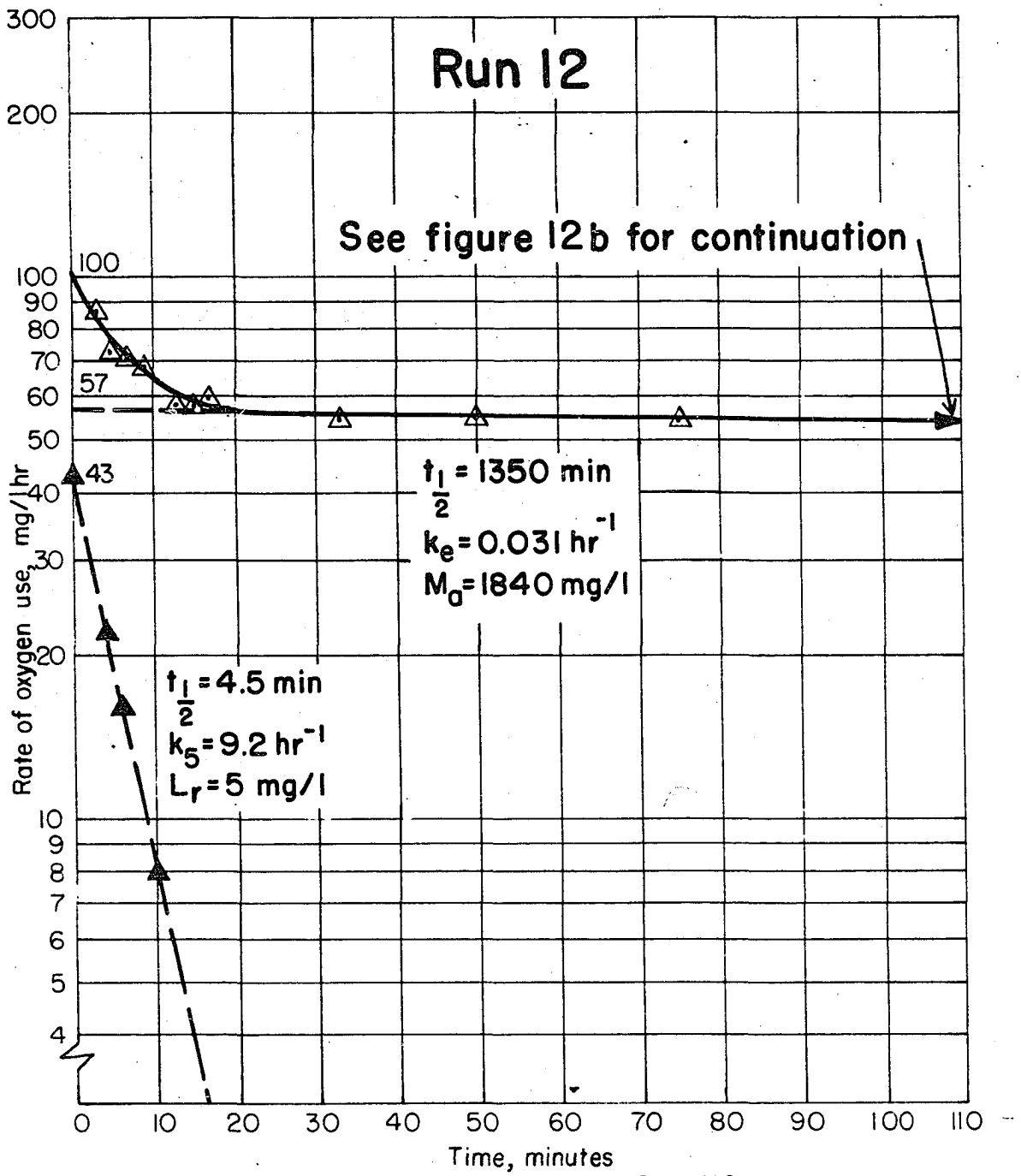

Figure 12a. Mixed liquor oxygen use rate, t=0 to 110 minutes

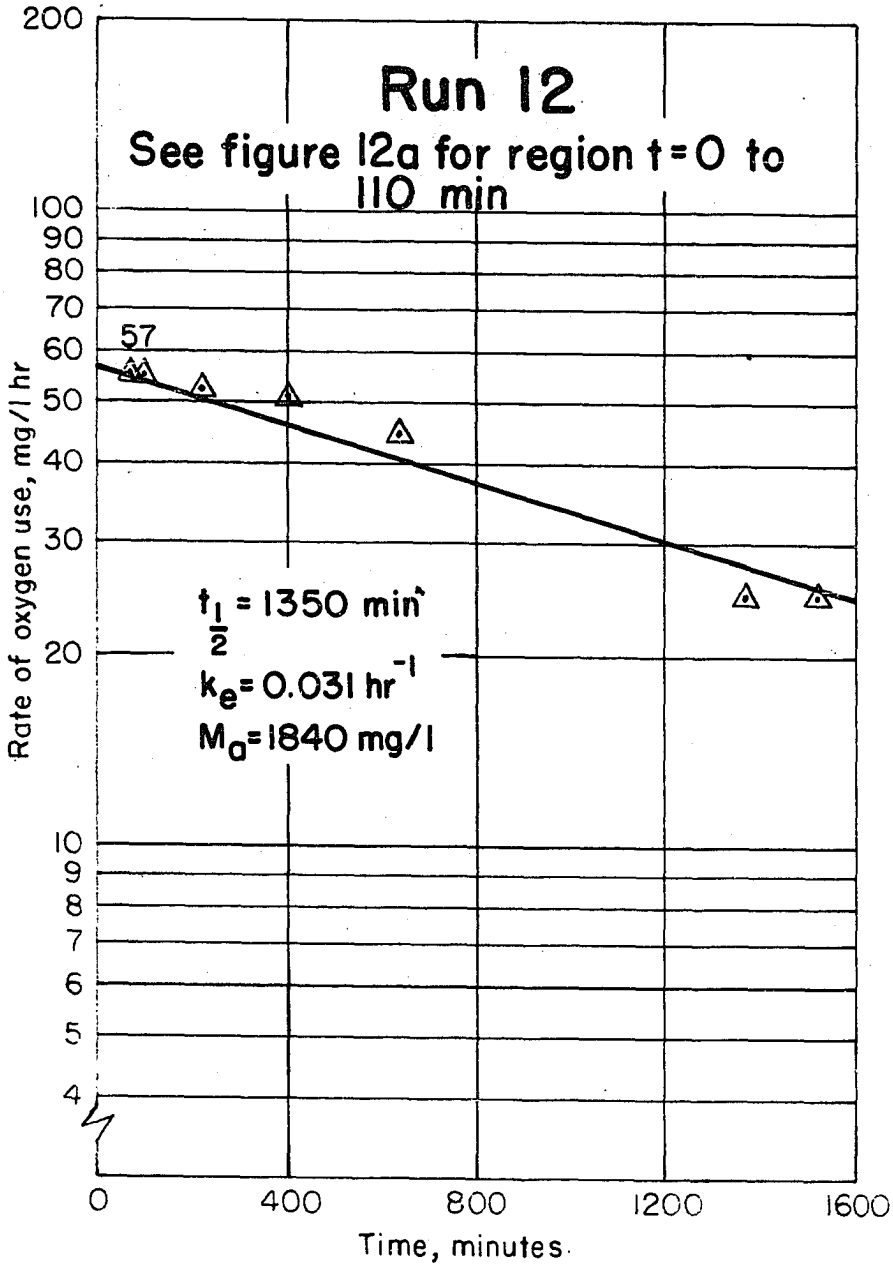

Figure 12b. Mixed liquor oxygen use rate, t=0 to 1600 minutes

which is the total oxygen, expressed as mg/l with other items in units used in the above discussion, which is required to meet the demands imposed by the two components exerted over time, t. In this case t is aeration time. For run No. 12, Figures 12a and 12b, total oxygen use would apparently be:

$k_5 L_r$ = 43 mg/l hr

$k_e M_a$ = 57

Total rate = 100 mg/l hr

Aeration time = 0.85 hour

Mass of oxygen use in mixed liquor

100 x 0.85 = 85 mg/l

The observed uptake of 225 mg/l is 265 per cent of that predicted from the properties of the mixed liquor and the aeration period.

Table 6 shows similar disagreement in similar runs between oxygen uptake predicted by equations 19 and 20 and that measured by uptake from the air.

D. Oxygen Utilization Balance Around Aeration Tank

Discrepancies between observed oxygen uptake rates or quantities and these predicted by equations 19, 20 or 21 forced reconsideration of the events affecting oxygen uptake. This was done by considering total oxygen uptake as a mass balance of oxygen demand of the various components in the influent and effluent of the aeration tank.

The aeration tank influent stream is the combined return

Table 6. Comparison of measured oxygen uptake and uptake predicted by equation 19

Run no.	Total flow liters/hr	Aera time hrs.	Measured O_2 uptake Rate lb/hr	Rate mg/hr	mg/l hr	Mass mg/l	k_5L_r, mg/l hr	Predicted O_2 uptake k_eM_a, mg/l hr	Total rate mg/l hr	Mass mg/l
9	99.4	0.97	0.020	9060	94	91	22	29	51	49
11	115.8	0.83	0.023	10400	108	90	20	43	63	52
12	113.2	0.85	0.056	25400	265	225	43	57	100	85
13	89.6	1.07	0.034	15400	161	172	50	62	112	120
14	51.5	1.86	0.031	14100	147	274	74	46	120	223

sludge and feed flow. The effluent stream is the mixed liquor leaving the tank. The oxygen demand of each component and the total demand of each stream can be determined from resolution of log mg/l hr vs time plots as described above. The decrease in total oxygen demand from influent to effluent can be interpreted as demand which is satisfied by oxygen supplied from the air stream.

Figure 13a and 13b are plots of log mg/l hr vs time from utilometer observations of a sample of feed plus return sludge from run no. 12. Return sludge and feed were combined in a 500 ml sample in proportion to their respective rates of flow, allowed to stand 1 minute after mixing to approximate the unaerated contact period in the aeration tank drop pipe, then subjected to the same oxygen use rate test in the utilometer as previously described for mixed liquor.

It will be noted that the curved part of the plot resolves into two high constant components in addition to the low constant component. More components are to be expected in view of the more complex nature of the nutrient in the feed. The oxygen demands of the various components in the influent for this run were:

$$L_r' = 73 \frac{mg}{l\ hr} \times \frac{hr}{42} = 2 \text{ mg/l}$$

$$L_r'' = 150 \frac{mg}{l\ hr} \times \frac{hr}{4.9} = 31 \text{ mg/l}$$

$$M_a = 57 \frac{mg}{l\ hr} \times \frac{hr}{0.028} = 2040 \text{ mg/l}$$

Total oxygen equivalent = 2073 mg/l .

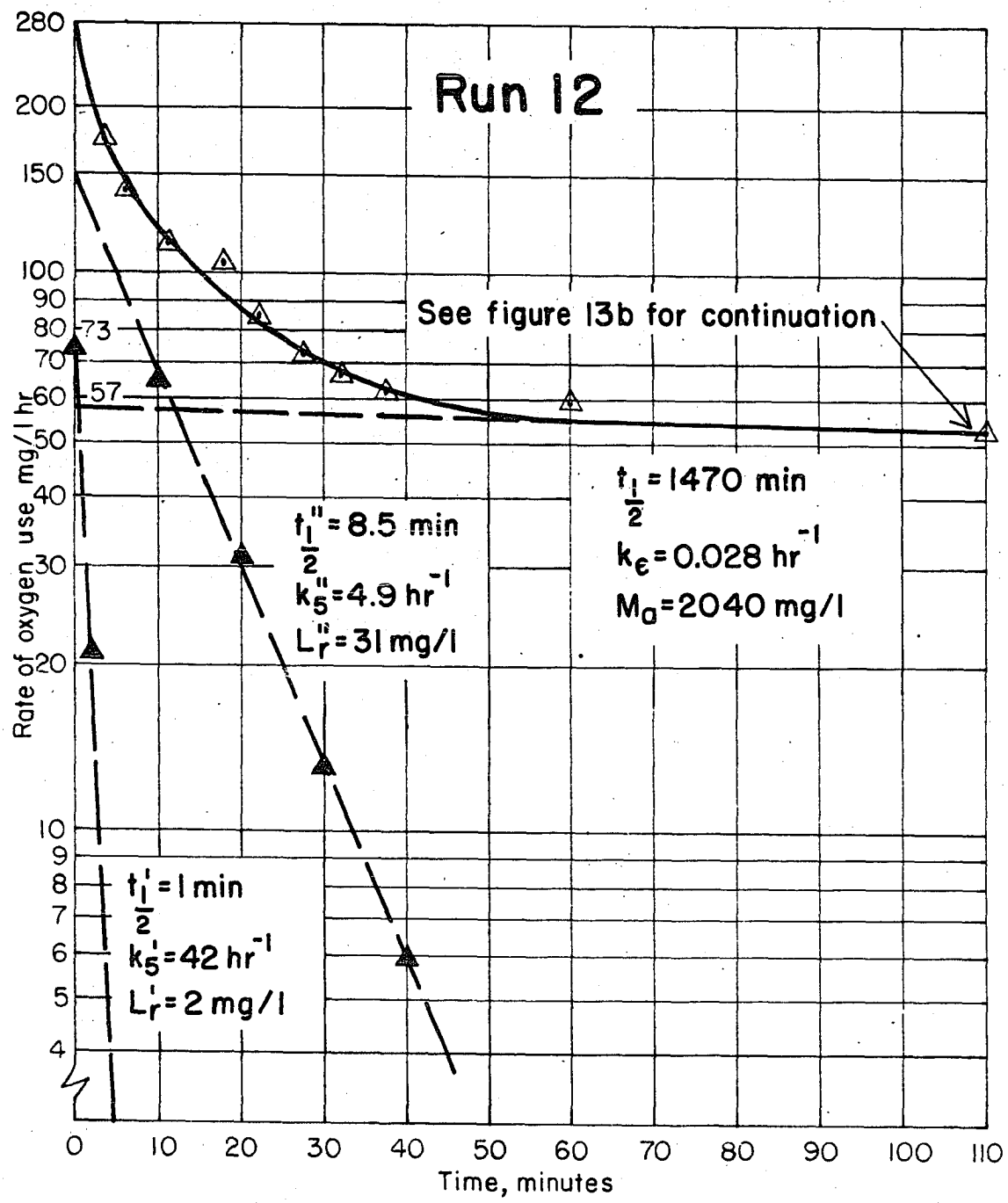

Figure 13a. Return sludge and feed oxygen use rate, t=0 to 110 minutes

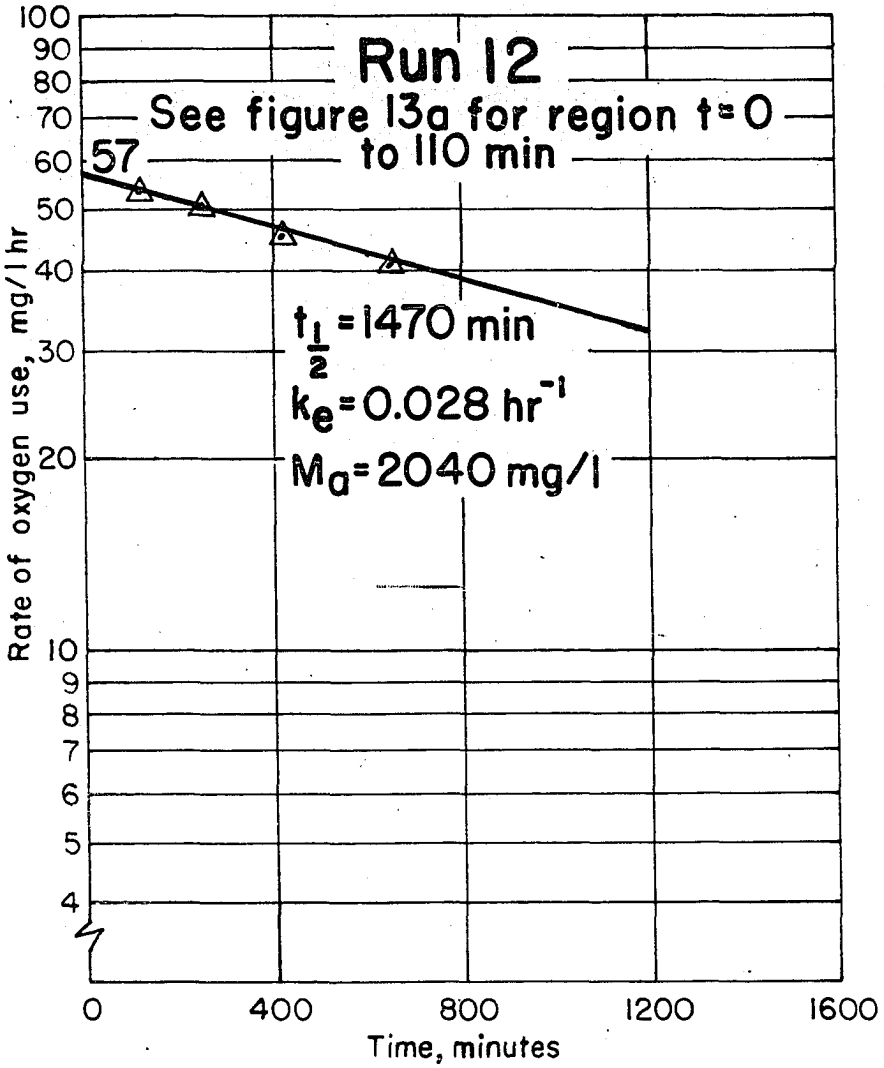

Figure 13b. Return sludge and feed oxygen use rate, t=0 to 1600 minutes

Since the aeration tank is completely mixed, mixed liquor samples from the tank are of the same composition as the effluent. As previously noted, the oxygen equivalents of the mixed liquor components for run no. 12 were:

$$L_r = 43 \frac{mg}{l\ hr} \times \frac{hr}{9.2} = 5 \text{ mg/l}$$

$$M_a = 57 \frac{mg}{l\ hr} \times \frac{hr}{.031} = 1840 \text{ mg/l}$$

Total oxygen equivalent = 1845 mg/l .

The difference between the aeration tank influent and effluent oxygen demands was 2073 - 1845 = 228 mg/l. This agrees with the observed oxygen uptake of 225 mg/l from the air.

Table 7 shows results of similar analyses for the same runs shown in Table 6. It is noted that in all cases, the mixed liquor and effluent oxygen demand was less than that of the return sludge plus feed and that this difference corresponded to the oxygen taken up from the air with a maximum discrepancy of 6 percent (runs no. 11 and 13).

This balance is on a mass basis, not a rate basis. The total oxygen used was taken up during the mean residence time in the aeration tank. Therefore, the average uptake rate was the total uptake divided by the aeration time as shown in Table 8. This also shows good agreement between oxygen uptake rates calculated from liquid stream balance and those calculated from air stream balance.

Table 7. Comparison of measured oxygen uptake and uptake predicted by oxygen demand balance around aeration tank

Run no.	Return sludge + feed (influent)			Oxygen demand, mg/l	Mixed liquor (effluent)			Uptake by balance mg/l	Uptake from air mg/l	Ratio bal./air
	L''_r	L'_r	Ma	Total	L_r	Ma	Total			
9	29	36	557	622	4	527	531	91	91	1.0
11	25	0	964	989	9	895	904	85	90	0.94
12	31	2	2040	2073	5	1840	1845	228	225	1.01
13	25	52	1655	1732	6	1565	1571	161	172	0.94
14	72	0	1240	1312	Trace	1040	1040	272	272	1.00

Table 8. Oxygen uptake rate

Run no.	Aeration time, hrs	Oxygen uptake By balance Mass, mg/l	By balance Rate mg/l hr	By air Mass, mg/l	By air Rate mg/l hr	Rate ratio balance/air
9	0.97	91	94	91	94	1.00
11	0.83	85	102	90	108	0.95
12	0.85	228	271	225	265	0.99
13	1.07	161	151	172	161	0.94
14	1.86	272	146	274	147	0.99

E. Oxygenation Constants

The previously described procedure produces measured values for the oxygenation constants k_5 and k_e. k_5 is the rate of oxidation of that part of the BOD used for energy. k_3 of equations 12, 14 and 15 is the total rate of oxidation of all BOD for both synthesis and energy. If a constant synthesis/energy ratio of 2/1 is assumed, k_3 will be three times k_5. Thus the procedure also provides a basis for estimating the value of k_3.

Table 9 shows oxygenation constant results from 11 runs. The slow rate constant, k_e, was not determined for runs prior to number 9 because observations were continued only long enough to define the early concave portion of the curve of log mg/l hr v.s. time. Resolution for these runs was done by

Table 9. Oxygenation constants, $hour^{-1}$

Run no.	Mixed liquor		Return sludge + feed		
	High rate k_5	Slow rate k_e	High rate k_5'	High rate k_5''	Slow rate k_e
3A	4.2	-	-	-	-
3B	4.4	-	-	-	-
4	1.7	-	-	-	-
5	5.1	-	-	-	-
6	4.2	-	1.9	-	-
8	5.9	-	3.6	-	-
9	5.9	0.055	8.4	1.2	0.032
11	2.3	0.048	1.2	None	0.029
12	9.2	0.031	42.0	4.9	0.028
13	8.3	0.040	10.2	2.2	0.035
14	13.9	0.044	2.1	None	0.036
Av.	5.9	0.044	9.9[a]	2.7	0.032

[a]Average without run number 12 = 4.6

horizontal extrapolation from the point of tangency which was reached in about 60 minutes. Figures 12a and 13a show that the slow rate component has very small change during this time interval. The approximation does not introduce appreciable error in the resolution for the high rate component.

Statistically, 6 of 11 of the mixed liquor k_5 values and 6 of 7 of the return sludge plus feed highest rate k_5' values lie within 90 percent confidence intervals about the average values. In the case of the k_e values, all but one of the mixed liquor values and all of the return sludge plus feed values lie within 90 percent confidence intervals.

These results show an average value of 5.9 $hour^{-1}$ for k_5 of the mixed liquor. Estimated value of k_3, therefore, is 3 times 5.9 or 17.7 $hour^{-1}$. These values are in reasonable agreement with McKinney's reported values of .5 $hour^{-1}$ for k_5 and 15 $hour^{-1}$ for k_3 (14).

F. Predicted BOD Reduction

Actual BOD of the plant effluent was compared with that predicted by equation 15a with the following conditions applied to the terms of equation 15a:

1. Ultimate BOD of feed and effluent were determined by multiplying the 5 day BOD by 1.15 and 2.2 for feed and effluent respectively. These are the factors determined from the BOD exertion curves as previously discussed.
2. Time, t, was taken as aeration tank displacement time for combined feed and return sludge streams rather than total aeration and final tank displacement volume as proposed by McKinney.

3. Value of k_3 was taken as 17.7 or 3 times average value of k_5 for the mixed liquor samples tabulated in Table 9.

Results of this comparison are shown in Table 10.

Table 10. Comparison of measured and predicted effluent BOD

Run no.	Aeration time, t, hours	Measured BOD, mg/l Feed, L		Effluent, L		Predict. effluent BOD[a]	% ultimate BOD reduction	
		5-day	Ultimate	5-day	Ultimate		Meas.	Predt.
3A	1.04	210	241	10	22	13	84	90
3B	1.04	210	241	10	22	13	84	90
4	0.59	284	326	13	29	30	91	91
5	0.96	210	241	5	11	14	95	94
6	0.98	229	263	2	4	15	98	94
8	1.11	220	253	12	26	13	90	95
9	0.97	233	268	9	20	16	92	94
11	0.83	200	230	9	20	15	91	93
12	0.85	203	223	11	24	13	89	94
13	1.07	229	263	10	22	14	92	95

[a]Predicted from $L_1 = \frac{L_o}{k_3 t+1}$

G. Effect of Return Sludge Condition

It will be noted that the slow rate component constitutes by far the greatest source of oxygen demand in each stream. This component presumably represents the endogenous

respiration of the active mass. Since the greatest share of active mass in the aeration tank comes from the return sludge, the condition of the return sludge may have an important affect on the oxygen balance around the aeration tank. Although outside the primary objective of this study, this question was considered sufficiently important to explore the effect of a relatively long period of return sludge under anaerobic conditions compared with a normal period of residence in the final settling tank.

Two samples of return sludge and feed were collected. The first sample of return sludge was combined with feed in the ratio of their flow rates (43% return sludge and 57% feed) immediately upon collection, allowed to stand 1 minute and subjected to the utilometer test for rate of oxygen uptake as previously described. The second sample of return sludge was allowed to stand for four hours before combining with the feed which was refrigerated during the standing period. This sample was also allowed to stand 1 minute after combination then subjected to the oxygen uptake rate test procedure. Approximate residence time of the sludge in the final settling tank was 1.75 hours. The return sludge samples were therefore, subjected to anaerobic periods of about 1.75 and 5.75 hours before combination with feed and oxygen uptake rate determination.

In both cases, typical exponential decay curves similar to Figures 13a and 13b were obtained. In the case of the

fresher sample, the curve resolved into 4 components; 3 components were obtained from the curve for the longer standing sample. Summary of the results is as follows:

Fresh sample:

Component No. 1, $L' = 1300 \frac{mg}{1\,hr} \times \frac{hr}{83.3} = 16$ mg/1

2, $L'' = 145 \frac{mg}{1\,hr} \times \frac{hr}{8.32} = 55$ mg/1

3, $L''' = 41 \frac{mg}{1\,hr} \times \frac{hr}{0.55} = 75$ mg/1

4, $M_a = 14 \frac{mg}{1\,hr} \times \frac{hr}{0.024} = 583$ mg/1

Total oxygen demand ---------- 729 mg/1

Sample after standing:

Component No. 1, $L^s = 150 \frac{mg}{1\,hr} \times \frac{hr}{10.4} = 14$ mg/1

2, $L'' = 43 \frac{mg}{1\,hr} \times \frac{hr}{0.95} = 45$ mg/1

3, $M_a = 18 \frac{mg}{1\,hr} \times \frac{hr}{0.075} = 240$ mg/1

Total oxygen demand ---------- 299 mg/1

This shows significant difference in the mixtures of different periods of anaerobic sludge condition. It was not possible to apply these different sludge conditions to the pilot plant but it may be deduced that the much lower active mass component of the longer standing sludge would cause change in operating characteristics and oxygen uptake rate.

H. Active Mass

The procedure for determining the oxygen demand of the various components of the influent and effluent streams provides a direct estimate of the active mass concentration.

It is assumed that the slow rate component of oxygen uptake is due to endogenous respiration of the active mass. Therefore, the demand of this component at t = 0 is the oxygen equivalent of the active mass in the given stream. The concentration of the active mass can then be estimated from the oxygen/mass ratio of 1.42.

Table 11 is a summary of these measurements and estimates for active mass in the mixed liquor for those runs in which the endogenous respiration was defined. Mixed liquor volatile suspended solids and ratio of estimated active mass to volatile suspended solids are shown for comparison.

Table 11. Estimate of active mass in mixed liquor

Run no.	Active mass, mg/l		Vol. susp. solids, mg/l	Ratios
	O_2 Equiv.	Solids		$\frac{\text{Active mass}}{\text{vss}}$
9	527	371	1757	0.21
11	895	630	2836	0.22
12	1840	1295	3087	0.42
13	1565	1100	3087	0.36
14	1040	732	3095	0.24

VI. DISCUSSION AND CONCLUSIONS

A. Reaction Time

The development of a concept and mathematical model for the kinetics of an activated sludge system depends on a definition of time during which a reaction occurs in order to express reaction rates. The studies have produced the following facts concerning relative amounts of nutrient or oxidizable material removed in the aeration and final settling tank:

1. The amount of COD removed in the final settling tank was less than 10 per cent of that in the incoming mixed liquor with mixed liquor volatile solids concentration between 1600 and 1900 mg/l, a relatively normal level. COD removal in the final tank increased to about 20 per cent when mixed liquor volatile solids were at an abnormally high level of 4500 to 5000 mg/l. The volume and displacement period of the final settling tank is 235 per cent of the aeration tank.
2. Both the COD and BOD of the liquid portion (filtrate) of the mixed liquor were found to be the same or less than the COD and BOD of the plant effluent. This demonstrated that no additional nutrient re-

moval from the flowing through stream occurs in the final settling tank.

These facts lead to the conclusion that metabolism is essentially completed within the aeration tank and the sole function of the final settling tank is the separation of solids from the mixed liquor and their concentration for sludge return. Thus, process reaction time should be taken as the mean residence time in the aeration tank. This time is the displacement period of the combined feed and return sludge.

Expressions for rates of reaction must, therefore, be based on the mean residence time or displacement time of the total influent stream, return sludge plus feed, in the aeration tank only.

B. Predicted vs Measured BOD Removal

Table 10 shows that there is good agreement between per cent removal of ultimate BOD predicted from equation 15a and that measured by the standard dilution and incubation procedure. This is conditioned on the following definitions of terms in equation 15a

1. L_o is the ultimate BOD of the feed, mg/l
2. t is the aeration period, hours, based on combined feed and return sludge flow
3. k_3 is the declining growth BOD removal rate constant,

hour^{-1}, taken as three times the average of measured values of k_5, the high rate oxygenation constant, of the mixed liquor.

It is important to note that the value of k_3 is estimated from the value of k_5 for mixed liquor, not the feed and return sludge stream. Table 9 shows that the oxygenation constants, k_5, for the mixed liquor and the feed-return sludge stream may be different. This difference is not only in value of the constants but also in the multi-valued character of the apparently more complex feed-return sludge stream. Therefore, in order for equation 15a to be reliable, it must be used with k_3 values for mixed liquor. This means that application of equation 15a to a given waste requires some method of measuring k_3 for that waste when it is dispersed in the mixed liquor of a completely mixed aeration tank.

The agreement between measured and predicted BOD removal is strong support for the presumed interrelation between k_3 and k_5. It also indicates the correctness of the k_5 values determined in this study. This, in turn, supports the method of oxygen demand resolution which led to the values of k_5.

The importance of correct interpretation of reaction time in applying theoretical relationships is demonstrated in equation 15a. Use of McKinney's definition of t as total aeration and final tank displacement time for feed flow only

would have given values of t 5 to 6 times larger than those on which results of Table 10 are based. This would reduce predicted effluent BOD to about 20 per cent of the tabulated values.

It may be concluded, therefore, that final effluent BOD of completely mixed systems may be predicted from equation 15a with sufficient accuracy for design and operating purposes provided the restrictions on t and k_3 noted herein are observed. This conclusion is made only for the range of reaction times covered in this study, namely, 0.6 to 1.1 hours.

C. Oxygen Balance

These studies show that expressions for rate or mass of oxygen uptake such as equations 19 and 20 are valid for a single stream. However, the studies also show that oxygen actually taken up in an operating system can only be accounted for by a balance of total oxygen demand between entering and leaving streams. This picture of rate at which oxygen must be supplied produces substantially different results than those given by equation 19, which, according to McKinney, completely defines the rate at which oxygen must be supplied. In fact, Table 6 shows that an aeration system for this pilot plant capable of supplying only requirements predicted by equation 19 would be grossly undersized. It must be concluded that equation 19 is applicable only to a batch system,

a condition rarely encountered in practice.

Eckenfelder's equation (21) has been shown to have no correlation for a continuous system. Here again, this equation defines only batch conditions and takes no account of balance between the oxygen demand of influent and effluent streams, hence it is of little practical significance for a continuous system. Similarly, McKinney's simplified picture of oxygen required to satisfy a unit of BOD has little value when demand balance in a given system is considered.

The results of these studies, therefore, force the conclusion that oxygen use in a continuous system can be completely described only on the basis of oxygen demand balance around the aeration tank, not simply on the basis of rate of demand exertion by the tank contents. It has been demonstrated that different treatment of the return sludge produces greatly different demand values for the influent stream and these different values must be reflected in a difference in oxygen use within the aeration tank to produce the same results.

It follows that a practical definition of oxygen requirement which is suitable for design is still lacking. Without such a definition the engineer is still forced to rely on empirical relations to design his aeration systems. However, the methods used in this study can be applied to operation control as noted below.

D. Active Mass

Table 11 shows the great difference between active mass and mixed liquor volatile suspended solids. Furthermore, Table 11 indicates that the ratio of active mass to volatile solids is not constant and may vary widely. Such variation is probably affected by the total solids level in the mixed liquor. The variation of active mass to volatile solids ratio is not due to differences in the ratio of volatile to total solids since the later ratio exhibited very little variation throughout these studies.

In view of the fact that a small and variable portion of mixed liquor volatile solids is active mass and since only active mass contributes to oxygen demand and use by the solids in the system, estimates of oxygen use based upon volatile solids under aeration such as equation 21 do not appear well founded.

E. Operational Control of Aeration

The method of determining total oxygen demand of the different streams by resolution of the oxygen uptake rate curves as described herein has been shown to give mass balance results that are consistent with the oxygen used from the air stream. This procedure can be a valuable operating tool since it provides a direct measure of oxygen used in an

operating system without reliance on assumed oxygen transfer factors which can vary widely with dissolved oxygen and solids concentration.

As an alternative or in conjunction with differential analysis of the influent and effluent streams, the air stream oxygen measuring system used in these studies can be readily adapted to any aeration installation in which provision of a gas-tight cover over the aeration tanks is feasible.

With techniques and instrumentation for determination of oxygen uptake available for an operating system, it becomes possible to isolate and control those factors affecting oxygen use such as return sludge rate, suspended solids concentration and sludge residence time in the final tank. The operating engineer may then select such values of these parameters as give minimum oxygen requirement for the required degree of treatment.

A partial adoption of the air oxygen instrumentation is possible in those cases where complete collection of all air coming from the aeration tank is not feasible. This might include long, conventional spiral flow tanks. In such cases, samples of exit air can be collected by gas bells at intervals along the tank. These samples can be piped to one centrally located oxygen analyzer. Air supply along the tank can then be regulated according to oxygen content of the exit air to avoid undersupply or oversupply to different regions

of the tank.

F. "Plug" Flow Systems

This development has been based on completely mixed flow characteristics in the aeration tank. However, the conclusions concerning the definition of total oxygen use on the basis of demand balance between influent and effluent streams are valid for "plug" flow as well as completely mixed flow. This is true since the definition is based on conditions external to the tank which are independent of conditions within the tank. The results are average oxygen use through the tank. In the case of a completely mixed tank, the oxygen use will be uniform throughout the tank. In the case of a "plug" flow tank, there will be substantial variation from the average rate in the different regions of the tank but the total oxygen which must be supplied to the tank must be determined by demand balance around the tank.

VII. SUMMARY AND RECOMMENDATIONS

The work of recent investigators in the field of microbiology and sanitary engineering has contributed much knowledge and insight into the mechanisms of biological waste treatment. This greater understanding has led to development of theoretical models of the activated sludge process. These models allow a rational design approach based on characteristics of the waste and proposed system. This rationalization has been necessary as the sanitary engineer has been called upon to solve an ever increasing variety of industrial and municipal waste water problems.

Some important aspects of newly developed theoretical concepts need verification in full scale operation or in continuous flow pilot plants where full scale environment can be achieved. The amount of oxygen which must be supplied to the system is a vital factor of design and operation. It is important that theoretical expressions for oxygen use be consistent with actual events within the operating system. The principal objective of these studies was to compare measured quantity and rate of oxygen uptake in a continuous flow pilot plant with the quantity and rate of oxygen uptake predicted from theoretical expressions of various investigators. It was hoped that modifications in the theory or mechanistic concepts could be suggested to account for any discrepancies which might be found.

During the course of the studies it was also possible to investigate such items as McKinney's equation for predicted effluent BOD, his definition of reaction time and the assumed relation between the oxygenation constant, k_5, and the BOD removal constant, k_3. These investigations were made by comparing measured and predicted results. Here also, modifications or redefinition of various factors necessary to correlate predictions with measurements were desired.

The pilot plant and associated equipment at the State University of Iowa P. F. Morgan Sanitary Engineering Laboratory was well suited to these studies. The unique application of oxygen measuring apparatus provided an accurate measurement of actual oxygen use by balance between the inlet and exit air streams. This measurement has not been made on prior research. The size and arrangement of the pilot plant permitted continuous flow and maintenance of biological environment that was the same as that to be expected in a full scale plant.

The most important finding of this study is that oxygen use in a continuous flow system must be defined on the basis of oxygen demand balance around the aeration tank. It cannot be completely defined by conditions within the tank. The oxygen demand of the active mass is by far the largest component of demand in the return sludge and feed influent stream and the mixed liquor effluent stream. The demand of

the active mass in the return sludge is affected by prolonged periods of anaerobic standing and this standing may, in turn, effect the oxygen demand balance around the aeration tank. This finding is applicable to both completely mixed flow and "plug" flow systems.

The oxygen demand resolution procedure described herein is a unique but reliable method of measuring the oxygen demand of the components of each stream. The procedure also yields values for the oxygenation constants and the rates of demand exertion of the various components. The information derived from the procedure should be useful for operation control as well as research. For example, the true active mass concentration must be known to determine BOD to organism loading factors. As demonstrated in this study, this loading factor may be materially greater than that based on volatile solids. The resolution procedure can be used to measure active mass.

It was found that biological reaction is essentially complete in the aeration tank. This forces definition of reaction time as the mean residence time or displacement period of the total return sludge and feed flow in the aeration tank.

Equation 15a, proposed by McKinney

$$L_r = \frac{L_o}{1 + k_3 t} \qquad 15a$$

was found to be reliable for predicting effluent BOD in a completely mixed system subject to the following conditions:

1. t is the aeration displacement period, hours, of the total return sludge plus feed flow.
2. The value of k_3 is taken as three times the value of the high rate oxygenation constant, k_5, of the mixed liquor as measured by the oxygen demand resolution procedure.

The k_5 and k_3 values for the mixed liquor and the feed plus return sludge do not appear to be equivalent. This means that equation 15a cannot be used with k_5 and k_3 values based on feed or feed plus return sludge without more knowledge of the relationship between these oxygenation constant values and those of the mixed liquor.

The instrumentation for oxygen balance between inlet and exit air streams proved to be sufficiently sensitive and accurate for this study. The components of the system are commercially available or readily fabricated. The system can be applied to full scale plants to provide valuable information for operation control.

The following recommendations are made as a result of these studies:

1. Aeration systems must provide the oxygen required to satisfy the difference in oxygen demand between the return sludge plus feed aeration tank influent

stream and the mixed liquor aeration tank effluent stream and not only to satisfy the apparent rate of oxygen uptake of the tank contents. The oxygen demand balance will be substantially affected by such variables as active mass concentration and return sludge condition. The effect of these variables cannot be determined precisely at this time and the designer must rely on prior experience to proportion the aeration system. Conservative design of the aeration system is recommended to insure that the system can handle unexpected loads or operating conditions.

2. The oxygen demand resolution procedure provides a useful research tool by which the affect of variables on oxygen use and plant performance can be studied. Specific studies recommended are:
   a. Are there practically useful relations by which the oxygen demand of aeration tank influent and effluent streams may be predicted for design and operation purposes?
   b. What is the effect of return sludge condition on oxygen demand balance around the aeration tank?
   c. How can the value of k_3 in the mixed liquor be determined for various feed and return sludge compositions so that equation 15a becomes a useful design relation?

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