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AN INVESTIGATION OF THE DYNAMIC RESPONSE OF THE ACTIVATED
SLUDGE PROCESS

New Jersey Institute of Technology

D.ENG.SC.

1980

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AN INVESTIGATION OF THE DYNAMIC RESPONSE
OF
THE ACTIVATED SLUDGE PROCESS

BY
TIN-BAI PAN

A DISSERTATION
PRESENTED IN PARTIAL FULFILLMENT OF
THE REQUIREMENTS FOR THE DEGREE
OF
DOCTOR OF ENGINEERING SCIENCE
AT THE
NEW JERSEY INSTITUTE OF TECHNOLOGY

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Newark, New Jersey
1980

ABSTRACT

A new mathematical model, derived from first principles of mass balance, is proposed for describing the mass transfer response of an activated sludge aeration tank in terms of a fluidized bed. The operational data to verify the proposed model were taken from an existing conventional wastewater treatment plant and a bench scale pilot plant.

The mass transfer rate, $(K_d A_s)$, was presented by an empirical dimensionless group function as follows:

$$\frac{(K_d A_s) D_p}{D_v} = f A_c \left(\frac{u D_p}{v}\right)^a \left(\left(\frac{v}{D_v}\right) (\text{MLSS})\right)^b$$

or

$$\text{Sh} = f(A_c/A_s) (\text{Re}_p)^a (\text{Sc}(\text{MLSS}))^b$$

where the definition of each parameter is given in the Glossary.

It was found that the coefficient of the dimensionless group function, f , and exponential indices of the dimensionless group function, a and b , were very approximately to unit. The activated sludge process, in terms of a fluidized bed model, can be described by this dimensionless group function. This is the first effort to apply this approach to the understanding of the activated sludge process.

The other investigation of this study was to predict the characteristics of an existing plant, based on the associated properties of a bench scale pilot plant. It was found that two distinct first order responses could be utilized to predict the operating BOD removal in the aeration tank of the activated sludge process. BOD removal in the initial portion of the aeration tank was found to be independent of the Mixed Liquor Suspended Solids concentration, but strongly dependent on the Mixed Liquor Suspended Solids concentration in the later stages.

APPROVAL OF DISSERTATION
AN INVESTIGATION OF THE DYNAMIC RESPONSE
OF
THE ACTIVATED SLUDGE PROCESS
BY
TIN-BAI PAN
FOR
DEPARTMENT OF CIVIL AND ENVIRONMENTAL ENGINEERING
NEW JERSEY INSTITUTE OF TECHNOLOGY
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GLOSSARY

a	Empirical exponential index, dimensionless.
A_c	Area of the aeration tank normal to flow, cm^2 .
A_s	Surface area of microbial mass, cm^2 .
b'	Empirical exponential index, dimensionless.
B	Kinetic constant, dimensionless.
BOD	Biochemical oxygen demand, mg/L.
BOD_5	Biochemical oxygen demand at 5 days, mg/L.
BOD_L	Ultimated biochemical oxygen demand, mg/L.
C	Concentration (BOD_5) in the reactor, mg/L.
C_o	Initial concentration (BOD_5), mg/L.
C_s	Equilibrium surface concentration (BOD_5), mg/L.
C_t	Concentration (BOD_5) at time, t, mg/L.
dC	Change of concentration, mg/L.
dt	Change of time, hr.
D_p	Diameter of partical, cm.
D_v	Molecular diffusivities of substrate, cm^2/sec .
E	Concentration of enzyme, mg/L.
E^o	Initial concentration of enzyme, mg/L.
EI	Enzyme-inhibition complex, mg/L.
ES	Enzyme-substrate complex, mg/L.
f	Empirical coefficient, dimensionless.
I	Substrate inhibition, mg/L.
K	Specific growth rate, hr^{-1} .
K'	Reaction rate constant, hr^{-1} .

K_0	Maximum specific growth rate, hr^{-1} .
K_{+1}	Rate constant for reaction in formation of the complex, hr^{-1} .
K_{+2}	Rate constant for reaction in formation of the new substrate, hr^{-1} .
K_{-1}	Rate constant for reaction in decomposit of the substrate, hr^{-1} .
K_{-2}	Rate constant for reaction in decomposit of the complex, hr^{-1} .
K_c	Contois constant, dimensionless.
K_d	Mass transfer coefficient, cm/hr .
K'_d	Microorganism decay coefficient, hr^{-1} .
$K_d A_s$	Mass transfer rate, L/hr .
K_i	Inhibition constant, mg/L .
K_m	Substrate saturation constant, mg/L .
K_p	Kinetic constant, mg/L .
K_s	Moser constant, mg/L .
K_t	Teisser constant, mg/L .
L	Length, cm .
$MLSS$	Mixed liquor suspended solids, dimensionless.
$MLSS_0$	Initial mixed liquor suspended solids, dimensionless.
$MLSS_{av}$	Weighted average mixed liquor suspended solids, dimensionless.
$MLSS_t$	Mixed liquor suspended solids at time, t , dimensionless.

P	Substrate product, mg/L.
Q	Volumetric flow rate, L.
r	Coefficient of correlation, dimensionless.
Re_p	Partical Reynolds number, uD_p/ν , dimensionless.
S	Concentration of soluble substrate in effluent, mg/L.
S^0	Concentration of initial substrate, mg/L.
S_0	Concentration of soluble substrate in influent, mg/L.
S_c	Schmidt number, ν/D_v , dimensionless.
Sh	Sherwood number, $K_d D_p/D_v$, dimensionless.
t	Time, sec.
T	Transfer time constant, hr.
T_1	First portion transfer time constant, hr.
T_2	Second portion transfer time constant, hr.
u	Superficial fluid velocity, cm/sec.
V	Volume of reactor, L.
V_f	Volume of fluid, L.
V_m	Maximum enzyme reaction rate, mg/L.
V_p	Maximum enzyme decomposition reaction rate, mg/L.
x	Distance of reactor, cm.
X	Concentration of microorganisms, mg/L.
Y	Growth yield coefficient, dimensionless.
θ	Hydraulic residence time, day.
θ_c	Mean cell residence time, day^{-1} .

λ_m	Moser kinetic constant, mg/L.
ν	Kinematic viscosity of fluid, cm^2/sec .
σ	Standard deviation of BOD_5 , mg/L.

CHAPTER I. INTRODUCTION

The purpose of this study is to investigate reaction kinetics of the activated sludge process and explore the view that the activated sludge process may be represented as a fluidized bed.

1.1 The Activated Sludge Process

The activated sludge process is most versatile and is widely used in treatment of waste water. It can produce an effluent with any desired organic concentration from very high to very low (1)*. Recent developments in the field of industrial waste treatment have stimulated considerable interest in this process and have added much to its development (2 to 7).

This process was developed in England in 1914 by Arden and Lockett (8) and was so named because it involved the production of an activated mass of microorganisms capable of aerobically stabilizing waste (9).

The basic phenomena of activated sludge have been summarized by Buswell and Long (10): "Activated sludge flocs are composed of a synthetic gelatinous matrix, similar to that of nostoc or merismopedia, in which filamentous and unicellular bacteria are imbedded and

* () indicates the references cited, see References.

on which various protozoa and some metazoa crawl and feed. The purification is accomplished by digestion and assimilation by organisms of organic matter in the sewage and its resynthesis into the living material of the flocs. This process changes organic matter from colloidal and dissolved states of dispersion to a state in which it will settle out."

Thus, this treatment by biological processes can be simply defined as a system in which flocculated biological growths are continuously circulated and contacted with organic waste in the presence of oxygen. The oxygen is usually supplied from air bubbles injected into the sludge-liquid mass. The process involves an aeration step followed by a solid- liquid separation step from which the separated sludge is recycled back for admixture with the waste. The aeration step may be considered in three functional phases (11, 12):

1. A rapid adsorption of waste substrate by the active sludge.
2. Progressive oxidation and synthesis of the adsorbed organics and organic concurrently removed from solution.
3. Further aeration results in oxidation and dispersion of the sludge particles.

The activated sludge must be kept in suspension during the period of contact with the waste substrate. It is different from other biological processes, such as the trickling filter, only in that no medium is used to support the active sludge. Simply speaking, the trickling filter, only in that no medium is used to support the active sludge. Simply speaking, the trickling filter is an adsorption in a fixed bed which involves passing the fluid phase through a stationary bed of the solid adsorbent, but activated sludge is adsorbed on which waste substrate is transferred from a fluid to the surface of the activated sludge upon contact with the latter. Adsorption is a surface phenomenon, and good adsorbents must have both a high surface - to - volume ratio and an "active" or an "activated" surface, i.e., a surface relatively free of adsorbed materials. A fluidized bed has more contact reaction between the solute material and adsorbent than a fixed bed (13, 14).

Figure 1 shows the schematic diagram of a conventional activated sludge process. In essence, the activated sludge process is carried out in an aeration tank in which the waste stream is intimately mixed with a voraciously feeding microbial mass. The process involves a primary settling tank, an aeration tank and a secondary settling tank. Gross particles are removed in

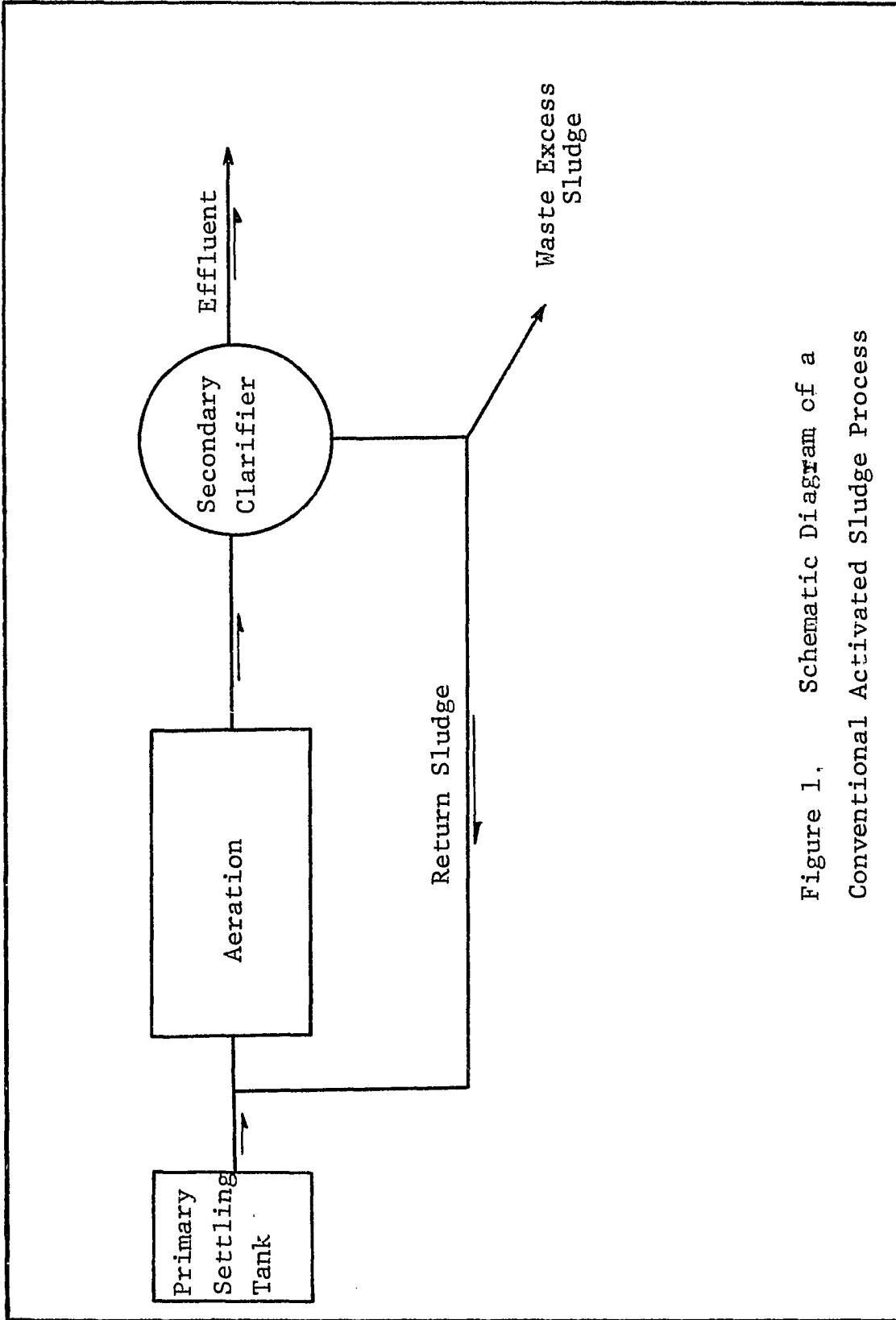


Figure 1. Schematic Diagram of a Conventional Activated Sludge Process

the primary tank. Aeration and agitation are along the aeration tank. Since much of the pollutant material has been transferred to the microbial mass in the aeration tank, it is necessary to remove the microbial floc. This removal of organisms is accomplished in the secondary tank. The basic steps involved in the activated sludge process are as follows:

1. Mixing the return activated sludge with the raw sewage. The returned activated sludge is thoroughly mixed with waste substrate. This is usually accomplished by adding the returned sludge to the settled primary sewage at the inlet end of the aeration tank. Agitation provides rapid and adequate mixing. In some cases small mixing chambers, with agitation, are provided, but it is not common practice.

2. Aeration of mixed liquid. Aeration is in the aeration tank and accomplishes two objectives: (a) keeping the sludge in suspension by agitation; and (b) supplying the required oxygen for biological oxidation. Air is generally added by one of the methods known as "diffused air" (or "pressure aeration") and "mechanical aeration." In the mechanical aeration system air is entrained from the atmosphere by the type of surface aerators or is introduced in the tank bottom by the type of turbine aerators. In the diffused air system, air

under low pressure is supplied by blowers and forced through porous materials which break up the air into fine bubbles. Air requirements are governed by the BOD loading, the quality of the activated sludge, the solids concentration, and the desired efficiency in BOD removal. The basic air requirement is that there must be sufficient air added to the waste water to maintain in it at least two ppm of dissolved oxygen under all conditions of loading in all parts of the aeration tanks, except immediately beyond the inlets.

3. Separation of activated sludge from the mixed liquor. Before the sewage treated in the aeration tank can be disposed of by discharging into receiving water, the activated sludge must be removed. This is done in secondary or final settling tanks.

The cycle of sludge removal from the final settling tank is much more important than with primary tanks. Some sludge is being removed continuously to be used as returned sludge in the aeration tanks. The excess sludge must be removed before it loses its activity because of the death of the aerobic organism resulting from lack of oxygen at the bottom of the tank. It is possible, where facilities are available, to reactivate returned sludge in separate reaeration tanks before adding to the sewage. However, it is better to retain the activity of the

sludge by prompt withdrawal from the tank.

4. Return of the proper amount of activated sludge for mixture with the sewage. The amount of the desired sludge returned to the aeration tank must be sufficient to produce the desired purification in the available aeration time and yet low enough to give economical air utilization. Because of variations in character and concentration of the sewage and the type of plant, the returned sludge must be determined for each plant by trial operation.

5. Treatment and disposal of excess activated sludge. The most common method is to pump the excess sludge to the influent end of the primary sedimentation tank where it is settled with the solids in the raw sewage. The activated sludge settles readily, and because of the large flocculant character of the sludge particles, it tends to remove some of the non-settleable solids in the sewage, thus reducing the organic and solids load on the aeration tank.

A number of variations in carrying out the above steps have been developed to meet different conditions. This has resulted in using the term "conventional activated sludge" for the original activated sludge process. A schematic diagram of the basic process is shown as

Figure 1. The wastes enter the aeration tank after being mixed with returned sludge. Diffused aeration along one side of the tank produces aeration and mixing as the wastes flow along the tank. The microorganisms aerobically stabilize the organic matter in the aeration tank and flow into the secondary settling tank. Sedimentation allows the activated sludge to flocculate and settle out, producing a clear effluent of low organic content. A portion of waste sludge is returned to the aeration tank as seed, with the excess sludge being wasted to the digester either directly or through the primary tank.

Many modifications of this basic process have involved over the years, primarily due to operating experience and to meet special operational needs. Among the various modified activated sludge processes are:

1. Step Aeration Activated Sludge Process
2. Contact Stabilization Activated Sludge Process
3. Complete Mix Activated Sludge Process
4. Modified Aeration Activated Sludge Process
5. Two-stage Activated Sludge Process
6. Pure Oxygen Activated Sludge Process
7. High Rate Aeration Activated Sludge Process
8. Kraus Activated Sludge Process
9. Extended Aeration Activated Sludge Process

In this paper, experimental data are taken from a conventional activated sludge treatment plant and a contact stabilization activated sludge laboratory system. The conventional activated sludge process has been discussed before. The contact stabilization process is shown as Figure 2. This process utilized both the rapid adsorption of soluble and particulate organic matter by the activated sludge, and the later slow oxidation of this organic material by the biomass. In the variation of this process shown in Figure 2A, the waste water is held for approximately one hour in the adsorption step, (i.e. contact tank). The sludge is then separated from the treated waste water and held for an additional several hours in a separate "Stabilization" tank for the oxidation step. By concentrating the sludge before oxidation, total aeration tank volumes are reduced by approximately 50 percent over a conventional process. Total air requirements are approximately the same as in the conventional activated sludge process and are usually equally divided between the contact and stabilization tanks. In the investigations reported here, the process of Figure 2B was utilized. This process is similar to that of the conventional activated sludge process.

1.2 The Fluidized Bed Process

Fluidized beds are an important and widely used means

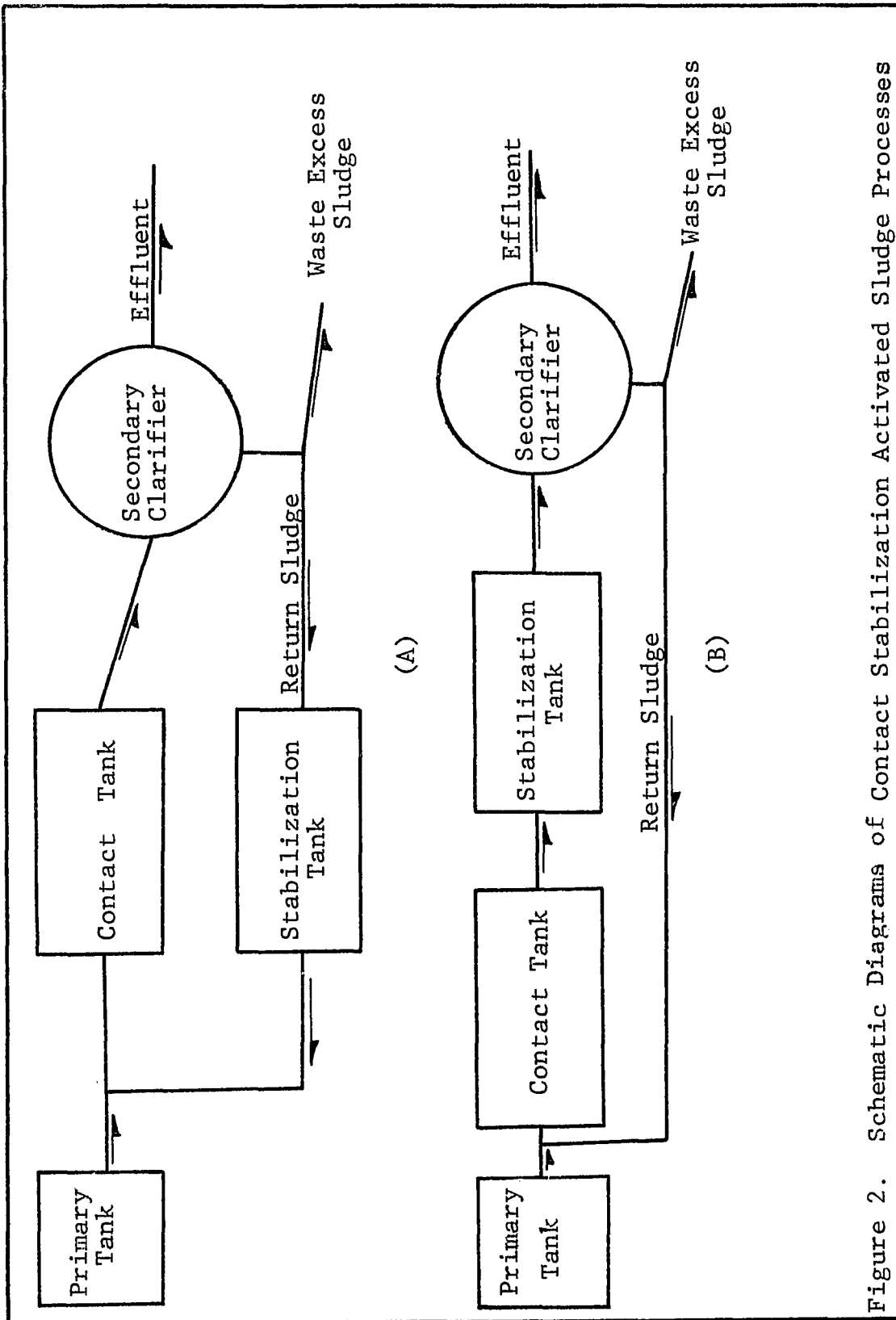


Figure 2. Schematic Diagrams of Contact Stabilization Activated Sludge Processes

for both heat and mass transfer in industrial processes. At the beginning of the twentieth century in water treatment facilities, the rapid sand filter method of water purification came into use. Backwashing of the filters, in which water was run into the filters from below, was a process in which bed particles were suspended in the moving fluid, and, by agitation of particles against each other, cleaning of absorbed floc material from bed particles and flushing away of waste material were accomplished. Backwashing of sand filters places them in a fluidized bed model.

Fluidized beds on a commercially significant, but not large, scale were first used in the mid-1920's (14). It was shortly before World War II, however, that employment of large scale fluidized beds appeared on the industrial scene. The Houdry Process, in operation since 1937, was utilized for production of high octane aviation gasoline from light oils and kerosene. The search for an improvement to this process gave rise to an extension known as Thermoform Catalytic Cracking Process (TCC). In later modifications of the process catalyst pellets were moved from one section of the reactor to another by gas lift. At the same time workers at Esso Research and Engineering, in co-operation with investigators at Massachusetts Institute of Technology, found that a

completely pneumatic system consisting of fluidized beds and air transport lines could operate suitably. The process became known as the Fluid Catalytic Cracking Process (FCC).

Although published references to the phenomenon that is known as fluidization go back as far as 1878 (15), the FCC process can be considered as the beginning of large scale applications of fluidized bed technology.

The great amount of technology available from existing fluidization practice, if available for application to wastewater treatment, could have a significant effect on approaches to wastewater (such as dimensionless group function in this study).

A fluidized bed can be described as a mass of particles held in suspension by moving stream of fluid so that no particle physically supports any other bed particle. The following are important in understanding fluidized bed phenomena (13, 14).

1. At minimum fluidization, the pressure drop through the bed equals the bed weight divided by the column cross-section, and the individual particles are lifted off each other.
2. In liquid-solid systems an increase in flow

rate above minimum fluidization usually results in a smooth, progressive expansion of the bed. Gross flow instabilities are damped and remain small, and large scale bubbling or heterogeneity is not observed under normal conditions.

3. Gas-solid systems generally behave in quite a different manner with an increase in flow rate beyond minimum fluidization. Large instabilities, with bubbling and channeling of gas, are often observed. At higher flow rates, agitation become more violent and the movement of solids becomes more vigorous.

4. Both gas and liquid fluidized beds are considered to be dense-phase fluidized beds as long as there is a fairly clearly defined upper limit or surface to the bed. However, at a sufficiently high flow rate the terminal velocity of solids is exceeded, the upper surface of the bed disappears, entrainment becomes appreciable, and solids are carried out of the bed with the fluid stream. In this state, there is a disperse, or dilute-phase fluidized bed, with pneumatic transport of solids.

Many studies have been reported in the literature concerning transfer in fluidized beds, and reporting empirical correlations involving dimensionless group

(16, 17). Models derived from first principles of mass balance have been presented for prediction of the mass transfer response of fluidized beds (18).

In the present study, the activated sludge process was regarded as a fluidized bed, since the particles in the activated sludge process are likewise suspended by the fluid. It can be assumed that no particle physically supports any other bed particle. This is the basic phenomenon of a fluidized bed. Although the fluid medium actually consists of two phases (water and air) in the activated sludge process, single continuous fluid phase in the development of a modified mathematical model will be treated. The detail of the proposed model and the mass transfer coefficient for the activated sludge process will be described in Chapter II.

1.3 Biological Reaction Kinetics

Research on the activated sludge process has been, for the most part, concentrated on microbiological aspects. During recent years, substantial research efforts have been directed toward a more complete understanding of factors affecting the activated sludge process and other biological wastewater treatment process.

Significant progress has been made in formulation, design, and operational procedures on a fundamental or

rational basis (19, 20, 21) rather than through empirical attempts. The result of these early research efforts was to indicate the importance of the ratio between mixed liquor suspended solids (MLSS) and biochemical oxygen demand at 5 days (BOD_5) (22).

Since publication of the previous design criteria, however, additional insight has been developed to explain the fundamentals of the process. Some contributions were influential in establishing microorganism specific growth rate as a process parameter (23, 24, 25). Mathematically, microorganisms specific growth rate is related to sludge age or mean cell residence time (θ_c), a term that is used with increasing frequency in the waste treatment literature (26, 27). A related factor is the hydraulic residence time, θ (28).

The mean cell residence time (θ_c) is determined by calculating the total mass of microorganism in the aeration tank and dividing them by the rate at which microorganism are intentionally and unintentionally wasted from the process. The quantity wasted includes those microorganism wastes purposely, as well as those lost in the effluent. When selecting a mean cell residence time, consideration must be given to both soluble

and particulate carbonaceous matters in the effluent.

In addition to the preceding definition of θ_c , Lawrence and McCarty (29) have developed a relationship that can be used to calculate aeration basin mixed liquor suspended solids concentration:

$$X = \frac{Y(S_o - S)}{1 + K_d \theta_c} \left(\frac{\theta_c}{\theta} \right) \quad (1)$$

in which Y = the growth-yield coefficient (mass of microorganisms/mass of substrate utilized); S_o = the concentration of soluble substrate in influent; S = the concentration of soluble substrate in effluent; K_d = the microorganism-decay coefficient and θ = the hydraulic residence time. This formula can be employed for establishing a design value for the aeration basin MLSS (X). Note that X varies inversely with values assumed or selected for the hydraulic residence time. Equation (1) applies only for the case of a completely mixed process operating at steady state conditions. For other systems such as plug flow, the design equation must be modified accordingly. However, the mathematical derivation involved for a plug hydraulic regime is rather difficult.

The principal reasons that invalidate the use of Equation (1) can be attributed to: 1. filamentous sludge; 2. oxygen transfer limitations in the aeration basin; and 3. presence of toxic substance that inhibit microbial growth. Therefore, in

the development of Equation (1), it has been assumed that microorganisms will grow adequately. They must be allowed to remain in the system long enough to reproduce. This period depends on their growth rate, which is related directly to the rate at which they metabolize or utilize the waste. This requires proper control of the biological environment. The environmental conditions can be controlled by pH regulation, temperature regulation (this is not practically possible in a large operating plant), nutrient or trace element addition, oxygen addition or exclusion, proper mixing, and the exclusion of toxic materials.

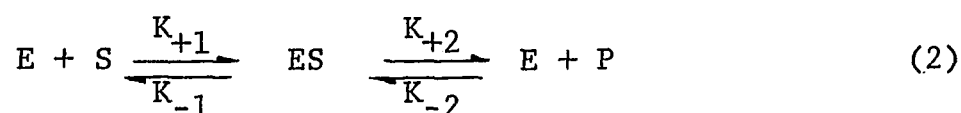
Hydraulic residence time (θ) selection has generally been on the basis of desired degree of treatment and characteristics and strength of wastewater. Although the foregoing considerations indicate that a broad range of hydraulic residence times can be employed while meeting the mean cell residence time design criterion, certain constraints must, on occasion, be imposed. Factors such as: 1. Process resistance to the imposition of shock loads; 2. Kinetics of degradation of degradation of large polymeric molecules; 3. Maximum aeration capacity per unit volume; 4. Maximum agitation intensity per unit volume; 5. Process stability; and 6. Concentration of mobile ciliates must be considered when establishing constraints for hydraulic residence time. The first five must be considered when establishing a lower level constraint on the hydraulic residence time, while the last must be considered when establishing an upper limit on the hydraulic residence time.

There is thus a widespread belief that hydraulic detention time is a factor in process operation and design. For the most part, this belief has been developed on an empirical or intuitive basis rather than on rational considerations.

A brief fundamental explanation of the activated sludge process has been presented. However, kinetic expressions will now be developed.

Any kinetic expression for the biological reaction rate must be based upon a number of simplifying assumptions. Many kinetic expressions have been formulated to fit enzyme and pure culture reactions. A few of the more common of these kinetic models will be presented for use in modeling waste treatment systems. The most widely used model is Michaelis-Menton model (30). It will be briefly described as follows:

Michaelis and Menten have formulated a very simplified model on the assumption that the substrate, S, is reversibly combined with an enzyme E, to form a complex, ES, which reversibly dissociates to form the product P, and regenerated enzyme. This model is shown as



where k_{+1} to k_{-2} are the rate constant for reaction in formation of, or decomposition of the complex.

For a batch reaction, a material balance on the substrate and complex becomes

$$\frac{dS}{dt} = \text{Rate of complexing between substrate and enzyme} = -K_{+1} S \cdot E + K_{-1} ES \quad (3)$$

$$\frac{dES}{dt} = K_{+1} S \cdot E - (K_{-1} + K_{+2}) ES + K_{-2} E \cdot P \quad (4)$$

where the four state variables, S, E, ES, and P are concentrations of their respective species. Since there are four state variables, two additional equations are needed to determine the system; these are obtained from the conservation equations on the initial enzyme and initial substrate:

$$E^0 = E + ES \quad (5)$$

$$S^0 = S + P + ES \quad (6)$$

In order to simplify this system of nonlinear equations, the reaction is assumed to proceed under equilibrium concentration of the complex. Setting Equation (4) to zero, we have

$$ES = \left(\frac{K_{+1} S + K_{-2} P}{K_{-1} + K_{+2}} \right) E \quad (7)$$

under most conditions, $S^0 \gg ES$: thus, simplifying Equation (6) to $S^0 = S + P$. Substituting Equation (6) and Equation (7) into Equation (3) yields (Appendix I)

$$-\frac{dS}{dt} = \frac{((V_m/K_m) + (V_p/K_p)) S - (V_p/K_p) S^0}{1 + ((1/K_m) - (1/K_p)) S + (1/K_p) S^0} \quad (8)$$

where

$$V_m = K_{+2} E^0$$

$$V_p = K_{-1} E^0$$

$$K_m = (K_{-1} + K_{+2})/K_{+1}$$

$$K_p = (K_{-1} + K_{+2})/K_{-2}$$

For the particular case where $K_{-2} = 0$, Equation (8) has the simplified form:

$$\frac{-dS}{dt} = \frac{dP}{dt} = \frac{V_m S}{K_m + S} \quad (9)$$

Equation (9) is called the Michaelis-Menten kinetics and the K_m is called the Michaelis-Menten or substrate saturation constant. It is important to realize that the term V_m is not a constant but depends upon the initial concentration of enzyme in the system.

Equations similar to (9) are found in the sanitary engineering literature for the rate of biological transformation of waste in the water to biological solids. Since the source of enzymes in bacteriological reactions is the bacteria themselves, V_m is replaced by

$$V_m = K_o X \quad (10)$$

where

K_o = Maximum specific growth rate.
constant, (time⁻¹)

X = concentration of microorganisms
(bacteria), mass/volume

Since S^0 and ES may be assumed to be constants, Equation (6) can be expressed as: $-dS/dt = dP/dt$. If the product (P) is assumed to be new biomass, then P is equal to X and Equation (9) becomes

$$\frac{dX}{dt} = \frac{K_o X S}{K_m + S} \quad (11)$$

Equation (11) is commonly called the Monod kinetic expression for biological synthesis. The Monod kinetics may not hold during unsteady-state conditions since the intermediate enzyme concentrations may not reach constant levels, making the steady-state assumptions of Equation (7) invalid.

The constants in Equation (11) may be given physical interpretations based on the Michaelis-Menten model. When $S \gg K_m$, the reaction kinetics are first-order in biomass and independent of substrate concentration. Physically this is understandable since the surface of bacteria is completely saturated with substrate and all internal enzymes are in the complexed state. For this condition, the rate of biosynthesis is a maximum. When $S = K_m$, the reaction rate is half of its maximum rate. When $S \ll K_m$, the reaction rate becomes first-order in substrate and biomass concentrations, or alternatively, second-order overall. Physically, we can arrive at the same result for low values of S, since very little of the substrate is available to complex with the enzyme. When S is increased, the amount of complex is immediately increased as seen in Equation (7). Since

$K_{-2} = 0$ for Michaelis-Menten kinetics, the rate of biomass production is directly proportional to the complex concentration, and hence to the substrate concentration.

Although Equation (9) indicates a one-to-one relationship between the rate of substrate consumption and product formation, this correspondence does not have to hold in practice where several parallel pathways are available for the consumption of substrate. Since a large fraction of substrate later appears as product biological cells, it is customary to relate the rate of substrate consumption to all production by

$$\frac{dX}{dt} = -Y \frac{dS}{dt} \quad (12)$$

where

$$Y = \frac{\text{units of organism formed}}{\text{units of substrate utilized}}$$

Although, the growth yield Y is usually considered as constant during modeling, it can vary over a wide range even under the most closely controlled conditions.

Before the growth yield or the constants can be used, the units for substrate and biomass concentration must be specified. In any case, it should be made clear what system of units is used in measuring S and X .

When activated sludge reactors are operated in the conventional regime, a kinetic term is included in the biomass growth rate expression to account for the decrease of biomass through

cell death or endogeneous respiration. This ratio is commonly taken as first-order in biomass concentration:

$$\left(\frac{dX}{dt}\right)_{\text{endogeneous}} = -K_d'X \quad (13)$$

combining Equation (11), (12), and (13), the net rate of biomass production becomes

$$\begin{aligned} \frac{dX}{dt} &= \text{Rate of biomass reaction} \\ &= \frac{K_o X S}{K_m + S} - K_d'X \end{aligned} \quad (14)$$

where K_d' is the endogeneous rate constant.

The rate of substrate reaction similarly becomes

$$\begin{aligned} \text{Rate of substrate reaction} &= \frac{dS}{dt} = \\ &= \frac{-K_o X S}{Y (K_m + S)} \end{aligned} \quad (15)$$

Dividing both sides of Equation 14 by X gives

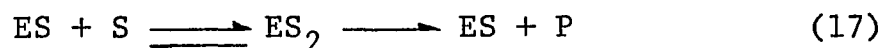
$$K = \frac{dX/dt}{X} = \frac{K_o S}{K_m + S} - K_d' \quad (16)$$

where K is called the specific growth rate and is equal to $1/\theta_c$ in the activated sludge process.

There are other kinetic models developed since many biological reactions are susceptible to inhibitors which can change the activity within the cell. An inhibitor can take many different forms and can react with the cell in a variety of ways. Substrates or nutrients when present in the neighborhood of the cell can inhibit certain pathways and accelerate

others. Very often the permeability of the cell wall is altered, or the activity of the enzyme is changed through complex formation with the inhibitor.

One possible mechanism for substrate inhibition is the reduction of the enzyme by complexing with the excess substrate:



If equilibrium concentrations of the complexes are assumed, the rate of product formation becomes

$$\frac{dP}{dt} = \frac{V_m S (1+BS/K_i)}{S+K + S^2/K_i} \quad (18)$$

where B , K_i , and K_m are kinetic constants.

When a competitor for the enzyme is introduced, the mechanism becomes,



where the complex, EI , is biologically inactive. Again assuming equilibrium concentration of complexes, the rate of product formation becomes

$$\frac{dP}{dt} = \frac{V_m S}{K_m + S + IK_m/K_i} \quad (20)$$

The type of competitive inhibition given by Equation (19) is typically found in reactions where the product reacts with the enzyme.

In the case of noncompetitive inhibition, the presence of inhibitors on the enzyme does not prevent that complex from reacting with the substrate. A typical rate equation for noncompetitive inhibition is

$$\frac{dP}{dt} = \frac{V_m}{(1+K_m/S)(1+1/K_i)} \quad (21)$$

Whenever inhibition is present, the rate of product formation is decreased due to the lower concentration of the enzyme-substrate complex. Mathematically, this lower rate is caused by the addition of more terms to the denominator of the rate expression.

Other proposed rate equations are often variations on the Monod theme. Some examples include forms where dispersed bacteria and flocculated bacteria have their own Monod constants, and prey-predator relationships where there is a food chain with Monod kinetics. Some of the common kinetic expressions neglecting endogenous respiration are given in Table 1. Many inhibition models have been tried but generally the accuracy of the data is not sufficient to choose between them, and so the common forms of Table 1 are used. These kinetic expressions should be applied with caution under unsteady-state conditions. Since they are usually derived assuming steady-state concentrations of metabolic intermediates, it is not surprising that they fail to handle cases where flow rate and inlet substrate concentrations change with time. Additional structure can be put into the rate expressions, but the effort and accuracy necessary to evaluate

TABLE 1

Common Biological Kinetic Expressions for Cell Synthesis
(Neglecting Endogenous Respiration)

Form	Name
$K = \frac{K_o}{1 + K_m/S}$	Monod
$K = K_o (1 - e^{-S/K_t})$	Teissier
$K = \frac{K_o}{1 + (K_c X/S)}$	Contois
$K = \frac{K_o}{1 + (K_s S^{-\lambda_m})}$	Moser
$K = \frac{K_o}{1 + (K_m/S) + (S/K_i)}$	Haldane (Substrate Inhibition)
$K = \frac{K_o}{1 + (K_m/S) + IK_m/K_i S}$	Competitive Inhibition
$K = \frac{K_o}{(1 + (K_m/S))(1 + (I/K_i))}$	Noncompetitive Inhibition

the extra constants is debatable, and the Monod expression in Equation (16) still remains as the central theme in biological kinetics.

Recalling Equation (16) and applying it to the activated sludge process, it must be emphasized that S is the substrate concentration (soluble) in the reactor and is assumed the same as the effluent substrate which is usually measured by BOD_5 . This is the same as saying that BOD removal does not occur in the final clarifier (i.e., during liquid-solids separation).

It is appropriate at this time to note that all the foregoing developments were based on the following premises:

1. Substrate is soluble; i.e., only soluble BOD participates the reaction. It is assumed that settleable BOD is totally removed from the primary clarifier and only soluble BOD enters the aeration tank.

2. Those soluble substrates (food or BOD) are only removed in the reactor by the enzymatic reaction of microorganisms. This leads to another assumption that the total mass of microbial cells includes only those present in the reactor. None of the above assumptions, strictly speaking, hold in real life. As a matter of fact, there is nothing like "steady-state" or "constant" in nature, and the precise formulation of natural phenomena is extremely difficult.

1.4 A New Model Set-Up

The previous discussion has concentrated on the microbiological aspects of the activated sludge process. However, the study of bacterial growth kinetics in waste water is an inexact science (31). The common biological kinetic expressions for cell synthesis were given in Table 1. Most mathematical models of the process assume the cell yield to be constant, even though it varies considerably (32). All of the mathematical models are complicated and unsuited to use in existing plants. They must be modified by fitting the expressions to experimental data from biological reactors. For example, the transient behavior of continuous flow cultures, including activated sludge units, is not correctly described by simple growth rate equations of the Monod type (33, 34). The Monod model is also often in error for rapidly changing processes. The contact stabilization treatment process has been developed based on the ability of biomass to store substrate for later metabolism, but there have been few attempts to incorporate this phenomena into kinetic models (35). Finally, the interpretation of the BOD test, which has widely been used in kinetic studies, has been called into question by discovery that oxygen uptake by bacterial predators may be significant (36).

In this investigation, a new model is examined in which it is suggested that the activated sludge process may be regarded as a fluidized bed. A fluidized bed can be described

as a bed in which no particle physically supports any other bed particle.

Mathematical models derived from first principles are proposed. The mathematical models, which have been verified, have been proposed for describing the mass transfer response of the activated sludge process as a fluidized bed. Parameters used are flowrate, surface area, particle size, and ratio of bed volume to reactor volume (a function of mixed liquor suspended solids). The operational data were taken from an existing conventional activated sludge plant and a laboratory unit. Equipment and procedures will be discussed in the chapter on experimental work. Both the existing plant and the laboratory unit gave results which identified the concentration profile of BOD taken along the flow axis of aeration tanks. Evaluation of bed parameters of the proposed model was carried out using the measured values of BOD_5 through the tank.

It is indicated that the activated sludge process can be represented as a fluidized bed. Verified models which describe fluidized bed mass transfer phenomena appear to apply to activated sludge mass transfer processes. This is the first effort to apply this approach to understanding of the activated sludge process. A new method for process design may be provided.

CHAPTER II. THEORY

This chapter presents the development of a new mathematical model by which the activated sludge process may be treated as a fluidized bed.

A fluidized bed may be simply described as one in which no particle physically supports any other particle. Backwashing of a rapid sand filter is an example of the fluidized bed.

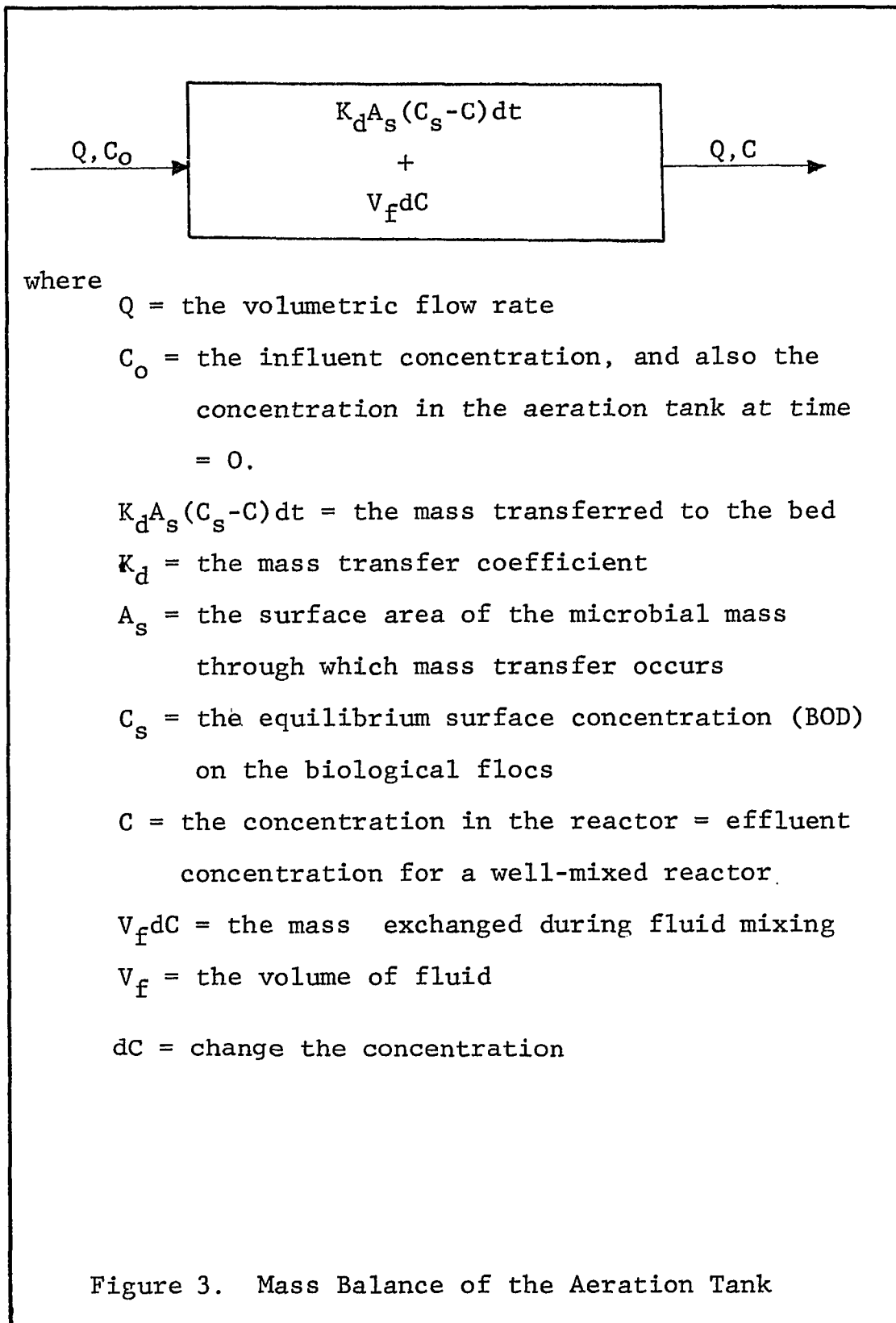
The idea of regarding the activated sludge process as a fluidized bed is based on the following phenomena.

1. The microorganisms are the solid particles in the fluidized bed.

2. The wastewater flow and the air flow represent the fluid flow in the fluidized bed. Although water and air are two phases, they will be treated as a single continuous fluid phase in the following development.

3. The particles in the activated sludge process are suspended. It is assumed that no particle physically supports any other bed particle. This is the basic phenomenon of a fluidized bed.

A mathematical model was developed based on this analogy, beginning with the schematic of an aeration tank shown in Figure 3.



A mass balance is made around the biological reactor, which includes the mass entering and leaving the fluidized bed, the mass given up to the bed solids (microorganisms), and the mass exchanged during mixing of fluid in the fluidized bed. The mass balance for the bed system is made with the following assumptions.

1. Complete back mixing.
2. The equilibrium surface concentration (BOD) on the biological flocs (C_s) and the concentration in the influent (C_o) are constant.
3. The concentration in the influent is much smaller than the equilibrium surface concentration on the biological flocs.

Referring to Figure 3, a differential mass balance around the aeration tank may be expressed as:

$$Q(C_o - C)dt = K_d A_s (C_s - C)dt + V_f dC \quad (22)$$

where $Q(C_o - C)dt$ is the total mass change, $K_d A_s (C_s - C)dt$ is the mass transferred to the bed, and $V_f dC$ is the mass exchanged during fluid mixing.

From equation (22),

$$(Q C_o - Q C - K_d A_s C_s + K_d A_s C)dt = V_f dC \quad (23)$$

$$\frac{dC}{Q C_o - K_d A_s C_s - C (Q - K_d A_s)} = \frac{dt}{V_f} \quad (24)$$

Since $K_d A_s$ is assumed constant, $dV_b/dt = V_b/t$. Multiplying the right side of Equation (28) by V/V , one obtains:

$$K_d A_s = (V_b/V) (V/t) \quad (29)$$

where V = the total volume of the aeration tank = $V_b + V_f$. Since V_b/V can be represented as MLSS in the aeration tank, and V/t is equal to Q , then, the transfer rate can be written as:

$$K_d A_s = (\text{MLSS}) Q \quad (30)$$

It is a well-known fact that the MLSS maintained is ordinarily in the range of 0.1% to 0.3% (1000 mg/l to 3000 mg/l in the aeration tank) (10). Thus, from Equation (30) it follows that:

$$K_d A_s \ll Q \quad (31)$$

Assumption 3 is equivalent to saying that

$$C_s \gg C_o \quad (32)$$

Combining Equations (27), (31) and (32), one obtains:

$$C_t = C_o - K_d A_s C_s (1 - \exp(-t/T))/Q \quad (33)$$

As t goes to infinity, C_t approaches zero. Therefore:

$$C_t = 0 = C_o - (K_d A_s C_s / Q)$$

or

$$K_d A_s C_s / Q = C_o \quad (34)$$

Based on Equation (34), Equation (33) is now written as:

The initial and boundary conditions for Equation (24) are $C=C_0$ at $t=0$, and $C=C_t$ at $t=t$. Therefore, Equation (24) becomes

$$\int_{C_0}^{C_t} \frac{dc}{Q C_0 - K_d A_s C_s - C (Q - K_d A_s)} = \int_0^t \frac{dt}{V_f} \quad (25)$$

With time, the microbial mass, and therefore A_s , will increase. However, in the development of this mathematical model it has been assumed that the group $K_d A_s$ represents a constant mass transfer rate.

With this assumption, Equation (25) may be integrated to give:

$$\ln \left[\frac{Q C_0 - K_d A_s C_s - C_t (Q - K_d A_s)}{Q C_0 - K_d A_s C_s - C_0 Q + K_d A_s C_0} \right] = \frac{-t}{T} \quad (26)$$

where T (the "transfer time") = $V_f / (Q - K_d A_s)$.

Simplifying Equation (26), the coefficient of waste substrate concentration (BOD) at any time can be obtained as follows:

$$C_t = \frac{Q C_0 - K_d A_s C_s + K_d A_s (C_s - C_0) \exp(-t/T)}{Q - K_d A_s} \quad (27)$$

Note that $K_d A_s$ is the transfer rate in the activated sludge process. This value can be equated to the rate of biological growth, or the rate of increase of suspended solids. If we define V_b as the volume of the bed (or the suspended solids), then:

$$K_d A_s = dV_b/dt \quad (28)$$

$$C_t = C_o \exp(-t/T) \quad (35)$$

Equation (35) is a simple exponential decay. A simple exponential decay equation results from assuming either a Completely Mixed Batch Reactor (CMB) or a Plug Flow (PF) Reactor (37).

For plug flow reactor, constant volume reaction, with the first order irreversible kinetics, the dynamic behavior is described as:

$$\frac{\partial C}{\partial t} = -u \frac{\partial C}{\partial x} + K' C \quad (36)$$

where

u = superficial fluid velocity

K' = reaction rate constant

x = distance of the reactor

The steady-state equation is obtained by setting the left-hand side of Equation (36) to zero,

$$0 = -u \frac{dC}{dx} + K' C \quad (37)$$

For homogeneous reaction concentration, C , the variables in Equation (37) can be separated, and the resulting expression integrated over the length, L , of the PF Reactor at which data will be taken.

$$\int_{x=0}^{x=L} \frac{dx}{u} = \int_{C_o}^{C_t} \frac{dC}{K' C}$$

or

$$L/u = (LA_c/uA_c) = V_f/Q = t = \int_{C_o}^{C_t} \frac{dC}{K'C} \quad (38)$$

where

L = length from aerator inlet at which data was taken

u = superficial fluid velocity

A_c = area of aeration tank normal to flow

V_f = volume of fluid

Q = flow rate

t = time

C_t = concentration (BOD₅) at time, t

C_o = initial concentration

dC = change of concentration

K' = reaction rate constant

C = concentration in the aerator

Equation (38) can be integrated to give:

$$K't = \ln (C_t/C_o)$$

or

$$C_t = C_o \exp K't \quad (39)$$

Although Equation (35) was developed by assuming a well mixed reactor, the PF Reactor reduces to the same form of a simple decay exponential equation when K' = -1/T in this particular case.

Equation (39) is a simple exponential decay. It will apply in the activated sludge process using BOD_5 as a parameter. The mass transfer rate ($K_d A_s$) can be represented by a dimensionless equation applicable to a fluidized bed:

$$\frac{K_d D_p}{D_v} = f \frac{A_c}{A_s} \left(\frac{u D_p}{\nu}\right)^a \left(\frac{\nu}{D_v} (MLSS)\right)^b \quad (40)$$

where

D_p = diameter of a particle

D_v = molecular diffusivities of substrate

ν = kinematic viscosity of fluid

u = superficial fluid velocity

A_c = area of the aeration tank normal to flow

$u A_c$ = flow rate (Q)

$\frac{K_d D_p}{D_v}$ = Sherwood number (Sh)

$\frac{u D_p}{\nu}$ = particle Reynolds number (Re_p)

$\frac{\nu}{D_v}$ = Schmidt number (Sc)

f, a, b = empirical constants

Equation (40) will be discussed in Chapter IV. The relationships of the dimensionless numbers expressed in Equation (40) will be obtained from the data results. Furthermore, the empirical constants (f, a, b) will be determined from the dimensionless group.

CHAPTER III. EXPERIMENTAL MATERIALS AND PROCEDURES

The basic Equation presented, $C_t = C_o - (K_d A_s C_s / Q) (1 - \exp(-t/T))$ degenerates to a simple exponential decay as Equation (35) or (39), using BOD_5 as the parameter. The data collected from the experimental investigation were used to verify the mathematical model, i.e., Equation (35). The Molitor Water Pollution Control Facility (Chatham, New Jersey), served as the source of the basic data reported in this study. Additional data came from a laboratory pilot plant. The description of the physical layout of these two facilities is given in the following paragraphs .

3.1 Molitor Water Pollution Control Facility (38)

The treatment plant site includes approximately 17 acres in Chatham and Florham Park. The completed facility which now includes primary, secondary and tertiary treatment removes approximately 95 percent of the organic impurities in the waste prior to discharge in the Passaic River. Figure 4 is a schematic diagram of the Molitor Plant. It consists of nine basic units.

A. Inlet Facilities (Comminutor, Bar Racks, Parshall Flume and Grit Removal Chamber and Mechanism).

Raw sewage enters the plant through a pipe 33 inches in diameter. Large cuttable solids are then ground by

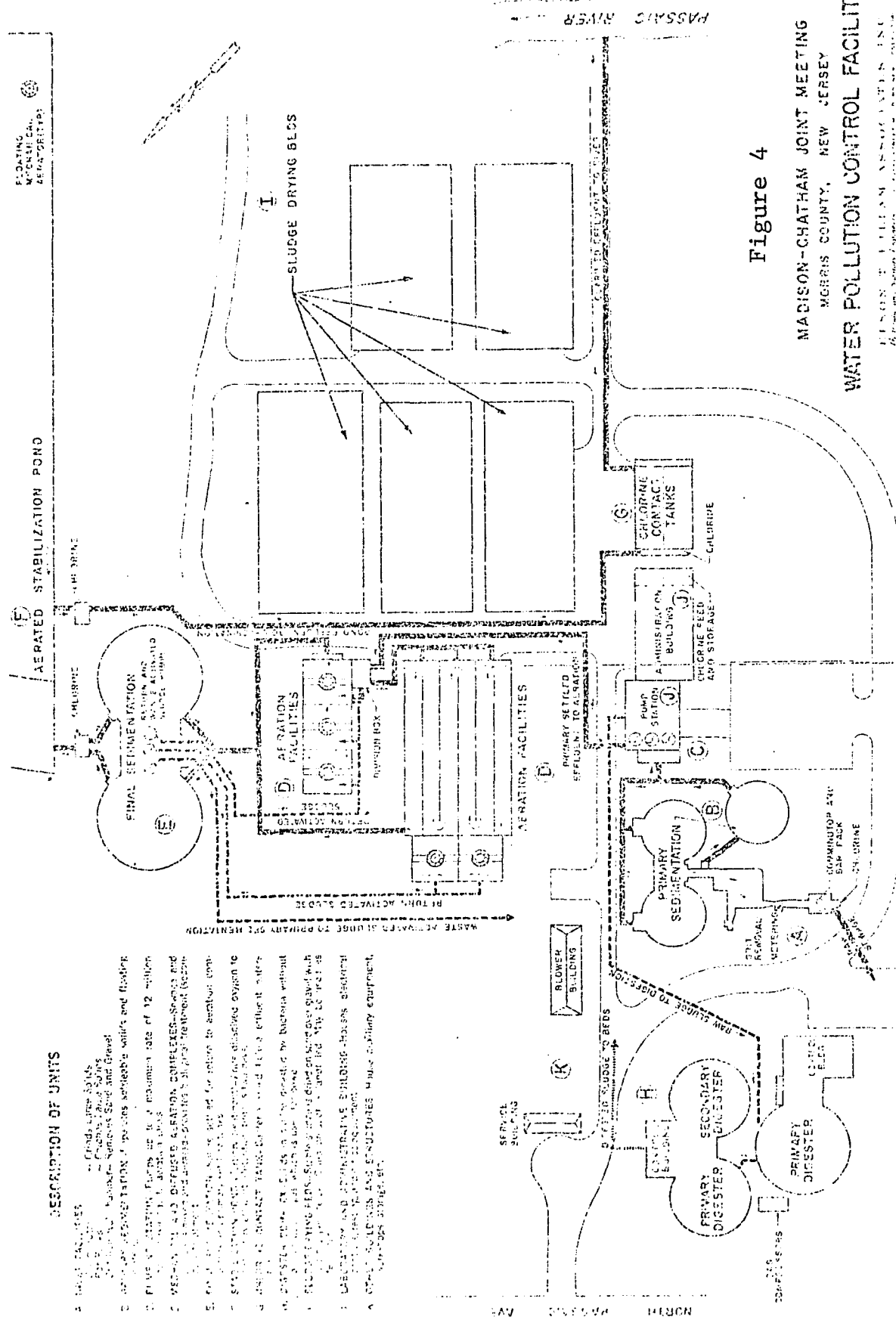


Figure 4

MADISON-CHATHAM JOINT MEETING
 MORRIS COUNTY, NEW JERSEY
 WATER POLLUTION CONTROL FACILITIES

DESIGNED BY: [unreadable]
 DRAWN BY: [unreadable]

the comminutor so that they will pass through pumps, pipes and equipment with ease. As an alternative to this mechanical operation, a bar rack is cleaned manually with a rake. The Parshall Flume measures the instantaneous rate of flow and the total flow entering the plant. The average amount of sewage entering the plant is 2.5 to 3.0 million gallons each day (MGD). The instantaneous rate where sewage enters the plant varies above and below the average, depending on the time of day.

The grit removal chamber is a tank which provides for a slower velocity of flow so that heavy solids such as stones and sand known grit will settle to the bottom. A mechanism collects the grit and removes it up a ramp to a collection hopper. The flow is then divided into three parts and flows to the primary settling tanks.

B. Primary Settling Tanks

The sewage is introduced into the center of each of the primary settling tanks. From there it flows radially to the perimeter at a low velocity so the settleable solids will drop to the bottom. The skimmer rotating at the surface of the tank removes floating materials. The sewage on the tank with a high concentration of solids (sludge) is pumped to the primary digestion tanks. The sewage with much of the settleable solids and biochemical oxygen demand removed flows over the weir plate on the

edge of the tank and then to the wet well of the new pumping station.

C. Pumping Station

The primary settled sewage is lifted normally by two pumps, sometimes working separately and sometimes together. The third is designed as a "stand-by" unit. By varying the speed of the pumps, they pump at the same rate at which flow is entering the wet well which is a chamber under the transformer pad next to the building.

The pump room in the basement also contains smaller pumps and motors for auxiliary service - sludge pumps and seal water pumps. The motor room contains the control mechanisms for all of the equipment. This includes electrical distribution (motor control center) pump speed controls and indicator and metering panel.

D. Mechanical and Diffused Aeration Complexes

The primary settled sewage is pumped to a distribution box where it is divided for flow to the two aeration complexes. Diffused aeration takes place in a series of deep, narrow tanks where primary settled sewage and sludge returned from the final settling tanks are intimately mixed. A high dissolved molecular oxygen concentration is maintained at all times. An environment is thus established in which the colloidal and truly dissolved

pollutant material can be utilized by the organisms of the activated sludge floc for food.

In the sewage, floc or groups of bacteria formed by collisions and biological actions binds them together. The floc becomes large enough so that it settles readily.

The mechanical aeration tanks provide the same features as the diffused system - mixing and providing air - by the rotation of large blades suspended on the bridges over the square tanks filled with the combined wastes.

E. Final Settling Tanks

The flow from both complexes is piped to the final settling tanks which functions similar to the primary tanks.

Between the two tanks is a pump station where three sludge pumps and their controls are located. Enough collected sludge is returned to the aeration tanks to provide the optimum amount of suspended solids in the mixed wastes. This sludge is known as return activated sludge. The remaining sludge is pumped, or wasted, to the primary settling tanks for clarification and removal to the digesters. This is known as waste activated sludge.

F. Stabilization Pond

The aerated stabilization pond provides additional treatment for the effluent from the final settling tanks (second effluent) and forms a third stage or tertiary treatment. The pond is three acres on the surface and ten feet deep and will hold the design flow (4.0 MGD) for two days. The eight million gallons of secondary treated sewage is further aerated during the detention period by four, floating, mechanical aerators. This process moves fully, cleanses the waste and provides that the natural processes of stabilization and aeration take place in the plant and not in the river.

G. Chlorine Contact Tank

Liquid leaving the stabilization pond flows to the chlorine contact tank. In this tank, chlorine in liquid form is added to the clear water from the stabilization pond. The chlorine which is put into "contact" with the treated sewage kills many of the remaining bacteria before entering the river through a large pipe known as the outfall sewer. The BOD and suspended solids in the effluent will be between 10 to 20 parts per million (ppm).

H. Digester Complex

The large digestion tanks or digesters receive the sludge from the primary settling tanks. The covered tanks exclude air (oxygen) and provide the proper environment

for the natural anaerobic bacteria which live there. These bacteria produce methane, which is held under the movable block dome of the secondary digester.

The auxiliary equipment housed in buildings attached to the digesters includes heating facilities to keep the digester contents at or near body temperature, and engines which use the methane gas to power blowers and provide the air needed for the diffused aeration tanks.

I. Sludge Drying Beds

When the solids have been completely stabilized or digested, they are transferred to the sludge beds constructed of layers of graded sand and gravel for drying. The resulting cake of dried solids is removed to the sludge storage area (next to the stabilization pond).

3.2 A Laboratory Activated Sludge Process

The Molitor Water Pollution Control Facility was the most conveniently located site for field studies. However, it was necessary to construct a laboratory scale pilot plant in which variables could be more closely controlled (39). An effort was made to closely reproduce operational characteristics of the Molitor Plant in the laboratory.

The primary waste was held for the adsorption step at the beginning in the diffused aeration tank of the

Molitor Plant and then held for the oxidation step in the same tank. In the laboratory pilot plant, these two steps were divided into two tanks due to the size of one tank was too small to hold the entire step in the aeration of mixed liquid. However, the flow line was the same as the Molitor Plant. The wastes entered the aeration tank after being mixed with returned sludge. Diffused aeration along the tanks produced aeration and mixing as the wastes flowed along the tanks. The microorganisms aerobically stabilized the organic matter in the aeration tank and flowed into the secondary clarifier. Sedimentation allowed the activated sludge to flocculate and settle out. A portion of waste sludge was returned to the aeration tanks as seed with the excess sludge and the clear effluent being wasted. In addition to the flow line, the activated sludge culture was obtained from the Molitor Plant in order to produce the same characteristics of microorganisms. Figure 5 is a schematic diagram of the pilot plant.

There are two phases (water and air) involved in the pilot plant. The air flow causes the vertical direction mixing and then the contents of wastewater are uniform in the radial direction. However, the air flow is small to mix with the concentration gradients in the longitudinal direction. Therefore, they will be treated as a

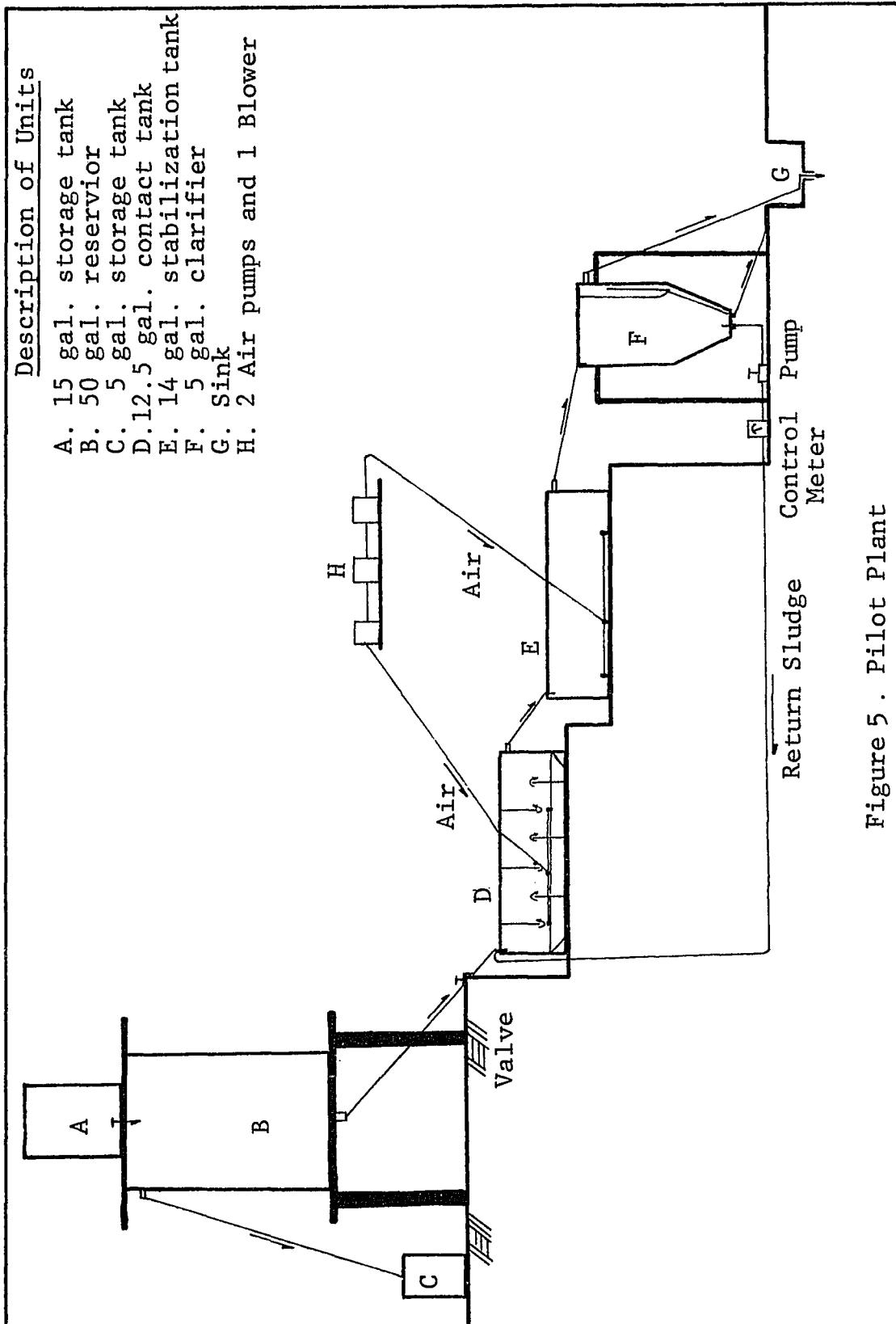


Figure 5 . Pilot Plant

single continuous fluid phase in the aeration tanks. Based on the Reynolds numbers of the pilot plant (1×10^{-4} to 5×10^{-4}), the flow was in the laminar flow region.

The components of the pilot plant were included as follows:

A. Synthetic Waste

The composition of the sucrose limiting substrate (synthetic waste) utilized in the study is given in Table 2. The sucrose concentration was 900 mg/gal, which supplied 180 mg/L of BOD_5 . This value is usually found in wastewater treatment plants. The other constituents are needed for microbial growth.

B. Seed Culture

A microorganism seed culture was obtained from the Molitor Plant.

C. Equipment

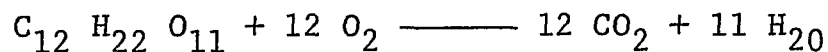
A schematic diagram of this equipment set-up is shown in Figure 5. Detailed descriptions of reservoir, aeration tanks, and settling clarifier are given below:

TABLE 2

Composition of Synthetic Waste

	<u>Substrate Constituents</u>	<u>Concentration mg/gal</u>
A	Sucrose C ₁₂ H ₂₂ O ₁₁	900
B	Ammonium Sulfate (N H ₄) ₂ SO ₄	180
C	Potassium phosphate Monobasic K ₂ H ₂ PO ₄	50
D	Magnesium Sulfate Mg SO ₄	150
E	Sodium Carbonate Na ₂ CO ₃	180
F	Sodium Nitrate Na ₂ NO ₂	54

The calculated DO demand (BOD_L, where L represents the ultimate demand) is 267 mg/L. The calculation is shown as follows:



where molecular weight of C₁₂H₂₂O₁₁ and O₂ are 342 and 32, respectively. Therefore,

$$O_2 = \frac{384}{342} \times 900 \text{ mg/gal} \times \frac{\text{gal}}{3.785 \text{ L}} = 267 \text{ mg/L}$$

The measured BOD₅ was approximately 180 mg/L. Thus, the conversion factor of BOD₅/BOD_L is 67 percent.

1. Reservoir

The synthetic waste was fed by gravity from a 15-gallon tank into a 50-gallon reservoir. The overflow from this storage reservoir fell to a 5-gallon receiving tank on the floor.

2. Aeration Tanks

Two aeration tanks were used in the experimental study. One was the contact tank and the other was the stabilization tank. Both of them were square base tanks with dimensions of 30 inch x 12 inch x 9 inch. The contact tank contained 12.5 gallons wastewater while the stabilization tank contained 14 gallons of wastewater. The contact tank was constructed of 7 pieces of plexiglass baffle, and the walls of the contact tank were inclined at 45° to direct settling sludge to the center bottom of the tank. Supplying air to the microbial mass serves to supply oxygen and to suspend the solids.

Two silent giant aquarium air pumps and one blower supplied the required air for sustaining aerobic condition in the tanks. Air diffusers on the bottom resuspended the sludge creating a dual circulatory motion in the vertical plane. Effluent from the contact tank to the stabilization tank, and then to the settling clarifier, was gravity drained in 1/4 inch Tygon tubing.

3. Settling Clarifier

A radial up-flow clarifier constructed of plexiglass was used for gravity concentration of biological solids. The clarifier was a 5 liter settling cone, with a rubber stopper at the lower end. The rubber stopper was fitted with two 1/4 inch Tygon tubes. One tube was for returning settled sludge to the aeration tank and the other tube for draining clarified effluent from the liquid surface. A pump was included in the return sludge line, along with a control flow meter.

3.3 Operation of the Activated Sludge Process

Because of the large number of factors involved, the best operating procedures for each plant must be determined by experience. With this qualification, the following criteria will generally apply:

1. There must be sufficient aeration to maintain a dissolved oxygen content of at least 2 mg/L at all times throughout the aeration tank.
2. Dissolved oxygen should be present at all times in the treated sewage in the final settling tank. This can be expected when the continuous flow from the stabilization tank is provided and the air supply is continuous.
3. Activated sludge must be returned continuously from the clarifier to the aeration tank.

4. The optimum rate of effluent return will vary with each installation. For the bench scale reactor discussed here, the rates of recirculation were 50 percent and 100 percent of the influent volumetric flow.

5. The mixed liquor suspended solids is controlled between 1000 to 3000 ppm in the aeration tanks. The lower or the higher value can not perform the adequate transfer.

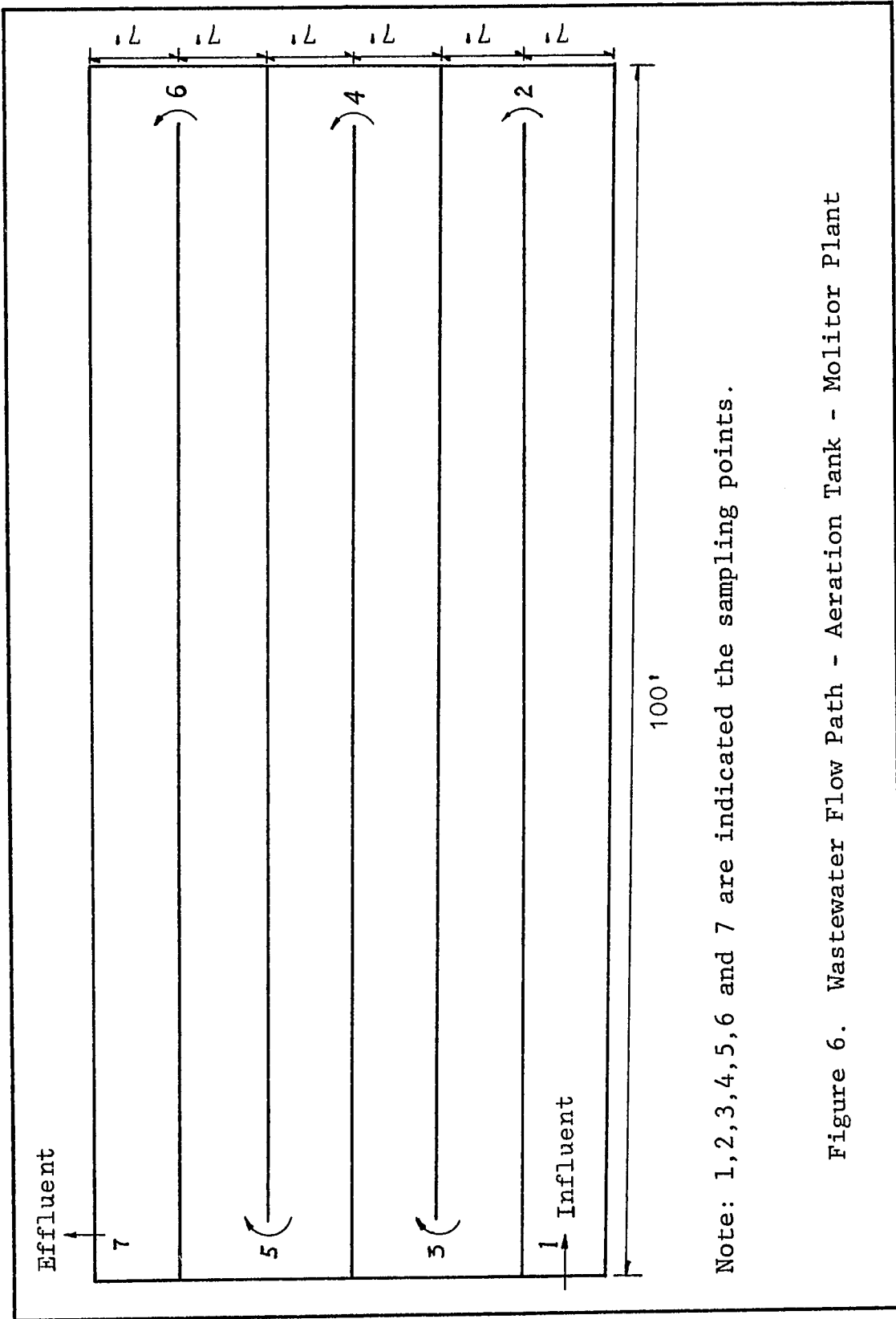
6. The suspended solids content in the aeration tanks may be controlled by the amount of sludge returned. All sludge in excess of that needed in the aeration tanks must be removed from the system. Excess sludge should be removed in small amounts continuously, or at frequent intervals, rather than in large amounts at any one time. Sludge held too long in the clarifier will become septic, lose its activity and deplete the necessary dissolved oxygen content in the tank.

7. Septic conditions in the storage reservoir will adversely affect the functioning of the activated sludge process. Pre-chlorination or pre-aeration was used to avoid septic conditions in the wastewater entering the aeration tank.

3.4 Experimental Procedure

A. Molitor Plant

Basic data taken from the Molitor Plant gave guidance as to the direction in which to proceed in the pilot plant investigations. The data collected were BOD and MLSS. They were the only two significant parameters in the activated sludge process, since the BOD indicated the amount of the waste substrate and the MLSS represented the micro-organism concentration. These parameters were used to evaluate the proposed model. The BOD and MLSS samples were collected at the Molitor Plant seven fixed points along the center line of the aeration tank. The path followed by the wastewater flow in the tank is illustrated in Figure 6. Distances along the aeration tank were converted to flow times of wastewater. Samples were collected, in every case, late in the morning. The plant influent flow rate was closely constant from about 9:30 AM to Noon on the days when samples were collected. This was determined by perusal of the chart records of the Parshall Flume measuring plant flow. In addition, a check on flow rate was made by measuring the head on the wide influent weir feeding the aeration tank. For the periods in which samples were collected it was found that the detention time was in all cases very closely seven hours. Temperature of the wastewaters in the period in



Note: 1,2,3,4,5,6 and 7 are indicated the sampling points.

Figure 6. Wastewater Flow Path - Aeration Tank - Molitor Plant

question ranged from 19.5°C to 21.5°C. BOD and MLSS determinations were made according to the 14th Edition of "Standard Methods for the Examination of Water and Wastewater" (40). Determinations of BOD values were run in triplicate at each sampling point at the Molitor Plant. Triplicate sample determinations gave good control over analytical results. BOD values mutually consistent were averaged and this average value reported and used in data treatment. Values were reported to the nearest whole number. Statistical treatment of the BOD data gave standard deviations in the range of 2 mg/L. Data which showed BOD differences from the middle value greater than 5 mg/L were not used. It is felt that this is a reasonable approach. MLSS concentration is of prime importance in operation of a treatment plant and the Molitor Plant has experienced considerable difficulty in process operation when the MLSS values were greater than 2100 mg/L or less than 1000 mg/L.

B. Laboratory Pilot Plant

After the laboratory equipment was set up, all tanks were filled with tap water in order to determine the integrity of the system. At the beginning of operation, it was necessary to ensure that all transfer tubes were free of solids plugging. The detention time was based on the flow rate. The flow rate was supplied as 24 L/hr

for most of the experimental investigations in the pilot plant. They were executed the 4.2 hours total detention time for the aeration tanks (contact and stabilization). The long detention time would be disadvantageous for the economical point. However, it was found that a detention time shorter than 1.5 hours in the aeration tanks was inadequate for the activated sludge process. The D.O. concentration was maintained at a minimum value of 2 mg/L in all parts of the aeration tanks.

Fresh seed material, collected from the Molitor Plant, was added to the aeration tanks of the laboratory plant. The reservoir was supplied with tap water from the Newark municipal supply. The synthetic substrate (see Table 2) was mixed in the reservoir. Any residual chlorine had already been removed. The flow rate was carefully controlled at a constant value.

During each run, samples from several points along the center line of the aeration tank (contact tank) were collected for analysis. Parameters of prime interest were BOD and mixed liquor suspended solids, since the BOD indicated the amount of the waste substrate and the MLSS represented the microorganism concentration. These parameters were used to evaluate the proposed model. BOD and MLSS determinations were made according to "Standard Methods for the Examination of Water and

Wastewater." Filtered and unfiltered BOD₅ samples gave essentially the same results.

CHAPTER IV. RESULTS AND DISCUSSIONS

4.1 Molitor Plant Data

The Molitor Water Pollution Control Facility served as the source of the basic data used to check the mathematical development in Chapter II (Equation 39). The data collected were BOD_5 and mixed liquor suspended solids. Points 1 to 7 are shown in Figure 6. The data appear as straight lines on semi-logarithmic plots of BOD_5 vs. time. This means that a simple exponential decay as described in Equation (35) or (39) can be represented on each plot.

The plant data are presented in Table 3. The first column of the table gives sampling date. The second column indicates the points along the length of the aeration tank where samples were taken. The distance between each point is 100 ft. The flow rate was observed to be rather steady from 9:30 AM to Noon on the days when samples were collected. There were seven fixed points and six time intervals represented. The time interval was defined by the detention time multiplied by the distance from the sample point to the influent and divided by the total length of the tank. Each time interval is 70 minutes. The third column shows the concentration C_t , represented as BOD_5 at time, t . The fourth column presents the mixed liquor suspended solids concentration, and fifth column

TABLE 3. Results of BOD₅ and Mixed Liquor Suspended Solids for the Molitor Plant

Date	Sampling Point	BOD ₅ at Time t (C _t), mg/L	MLSS _t mg/L	$\frac{MLSS_t}{MLSS_o}$
5/11/79	1	96	1287	1
	2	54	1127	0.88
	3	45	1080	0.84
	4	39	949	0.74
	5	25	929	0.72
	6	21	841	0.65
	7	15	836	0.65
5/25/79	1	150	6234	1
	2	92	5792	0.93
	3	73	3270	0.52
	4	60	3450	0.55
	5	57	2132	0.34
	6	33	2490	0.40
	7	26	1244	0.20
6/3/79	1	44	1323	1
	2	24	1181	0.89
	3	18	1080	0.82
	4	16	949	0.72
	5	14	928	0.70
	6	11	834	0.63
	7	10	716	0.54
6/4/79	1	75	1369	1
	2	46	1233	0.90
	3	34	1048	0.77
	4	26	1020	0.75
	5	20	962	0.70
	6	19	926	0.68
	7	16	862	0.63

Note:

t = time

MLSS = mixed liquor suspended solids

o = initial time

TABLE 3. Results of BOD₅ and Mixed Liquor Suspended Solids for the Molitor Plant (continued)

Date	Sampling Point	BOD ₅ at Time t (C _t), mg/L	MLSS _t mg/L	$\frac{\text{MLSS}_t}{\text{MLSS}_0}$
6/14/79	1	38	2382	1
	2	25	2339	0.98
	3	22	1616	0.68
	4	17	1553	0.65
	5	14	1169	0.49
	6	13	1145	0.48
	7	9	949	0.40
6/28/79	1	75	3383	1
	2	61	2889	0.85
	3	49	2756	0.81
	4	43	2189	0.65
	5	38	1674	0.49
	6	34	1504	0.44
	7	32	1381	0.41
7/5/79	1	73	5751	1
	2	50	4524	0.79
	3	43	3891	0.68
	4	37	3364	0.58
	5	32	3002	0.52
	6	26	2645	0.46
	7	24	2171	0.38

gives the ratio of mixed liquor suspended solids at time, t , to the initial mixed liquor suspended solid.

4.2 Laboratory Pilot Plant Data

Data were collected on the total flow rate, BOD_5 and MLSS. The detention time and the related time at each point were measured.

1. Flow Rate

The total flow rate was the combined flow from the reservoir and the returned sludge flow from the clarifier. They were measured by determining how long it took to fill up a 1 L beaker, and were checked by two gilmont flow meters. This information is shown as Table 4.

2. BOD_5 and MLSS at Each Point

The samples were collected from influent, effluent and 5 other points along the aeration contact tank (see Figure 5). These samples were analyzed for BOD_5 and MLSS according to the "Standard Methods for the Examination of Water and Wastewater."

3. Detention Time of the Aeration Tanks

The physical characteristics of the pilot plant were determined (see p. 42). The total flow rate of each sampling data was given in Table 4. The detention time can be determined by the volume of the aeration tanks divided by the total flow rate. The detention time of

TABLE 4. Composition of the Combined Flow in
the Pilot Plant

Date	Total Flow Rate L/hr	Composition
12/17/79	24	(1) + (2)
1/6/80	24	(1) + (2)
1/9/80	24	(1) + (2)
1/10/80	24	(1) + (2)
1/11/80	24	(1) + (2)
1/17/80	24	(1) + (2)
1/18/80	24	(1) + (2)
1/28/80	48	(3) + (4)
1/29/80	30	(5) + (6)
1/30/80	24	(7) + (2)
1/31/80	18	(1) + (8)
2/1/80	18	(1) + (8)

Note:

- (1) = 12 L/hr of the synthetic wastewater in which the concentrations are those given in Table 2
- (2) = 12 L/hr of the returned sludge flow
- (3) = 24 L/hr of the synthetic wastewater in which the concentrations are those given in Table 2
- (4) = 24 L/hr of the returned sludge flow
- (5) = 15 L/hr of the synthetic wastewater in which the concentrations are those given in Table 2
- (6) = 15 L/hr of the returned sludge flow
- (7) = 12 L/hr of the synthetic wastewater in which the concentrations are those given in Table 2
- (8) = 6 L/hr of the returned sludge flow

the contact tank was 118 minutes at a total flow rate of 24 L/hr, 59 minutes at 48 L/hr, 79.5 minutes at 30 L/hr, and 157 minutes at 18 L/hr.

4. Related Time at Each Point

The flow velocity in the aeration tank was assumed constant. The tank is a rectangular tank. Therefore, the related time of sampling transfer at each pint is equal to the detention time times the distance from the sample point to the influent divided by the total length of the tank.

The results are presented in Table 5. The first column of the table shows sampling date. The second column gives the points along the length of the contact tank where samples were taken. The third column indicates the related time of sampling transfer at each point. The fourth column gives the concentration C_t , represented as BOD_5 at time t . The fifth column gives the mixed liquor suspended solids concentration, and the sixth column presents the ratio of mixed liquor suspended solids at time, t , to the initial mixed liquor suspended solids - $(MLSS_t/MLSS_o)$ vs. time (t) .

TABLE 5. Results of BOD₅ and Mixed Liquor Suspended Solids for the Pilot Plant

Date	Sampling Point	Time (t), min	BOD ₅ at Time t (C _t), mg/L	MLSS _t mg/L	$\frac{MLSS_t}{MLSS_o}$
1/28/80	1	0	201	1410	1
	2	12.5	179	1306	0.93
	3	21	175	1419	1
	4	29.5	150	1257	0.89
	5	38	150	1369	0.97
	6	46.5	139	1199	0.85
	7	59	120	1109	0.79
1/29/80	1	0	110	1293	1
	2	20	41	1201	0.93
	3	33.6	37	1113	0.86
	4	47.2	33	1031	0.80
	5	60.8	30	978	0.76
	6	74.4	28	906	0.70
	7	94.4	24	843	0.65
12/17/79	1	0	85	1038	1
	2	25	36	951	0.92
	3	42	31	846	0.82
	4	59	27	801	0.77
	5	76	24	732	0.71
	6	93	21	693	0.68
	7	118	19	624	0.60
1/6/80	1	0	94	1575	1
	2	25	41	1425	0.90
	3	42	35	1338	0.85
	4	59	31	1301	0.83
	5	76	26	1262	0.80
	6	93	23	1217	0.77
	7	118	19	1104	0.70

Note:

t = time

MLSS = mixed liquor suspended solids

o = initial time

TABLE 5. Results of BOD₅ and Mixed Liquor Suspended Solids for the Pilot Plant (continued)

Date	Sampling Point	Time (t), min	BOD ₅ at Time t (C _t), mg/L	MLSS _t mg/L	$\frac{MLSS_t}{MLSS_0}$
1/9/80	1	0	94	1827	1
	2	25	39	1699	0.93
	3	42	33	1534	0.84
	4	59	28	1506	0.82
	5	76	21	1381	0.76
	6	93	20	1327	0.73
	7	118	15	1289	0.71
1/10/80	1	0	89	1437	1
	2	25	38	1352	0.94
	3	42	31	1257	0.87
	4	59	28	1183	0.82
	5	76	26	1104	0.77
	6	93	20	1059	0.74
	7	118	16	986	0.69
1/11/80	1	0	92	1385	1
	2	25	40	1296	0.94
	3	42	36	1207	0.87
	4	59	30	1131	0.82
	5	76	27	1078	0.78
	6	93	22	1004	0.72
	7	118	20	932	0.67
1/17/80	1	0	93	1348	1
	2	25	39	1228	0.91
	3	42	34	1179	0.87
	4	59	30	1102	0.82
	5	76	26	1025	0.76
	6	93	23	978	0.73
	7	118	19	915	0.68
1/18/80	1	0	104	2001	1
	2	25	57	1763	0.88
	3	42	42	1609	0.80
	4	59	35	1521	0.76
	5	76	30	1408	0.70
	6	93	26	1317	0.66
	7	118	20	1187	0.59

TABLE 5. Results of BOD₅ and Mixed Liquor Suspended Solids for the Pilot Plant (continued)

Date	Sampling Point	Time (t), min	BOD ₅ at Time t (C _t), mg/L	MLSS _t mg/L	$\frac{\text{MLSS}_t}{\text{MLSS}_0}$
1/30/80	1	0	191	1979	1
	2	25	65	1789	0.90
	3	42	50	1697	0.86
	4	59	41	1593	0.80
	5	76	35	1510	0.76
	6	93	28	1403	0.71
	7	118	24	1343	0.68
1/31/80	1	0	92	1637	1
	2	33	39	1528	0.93
	3	56	33	1471	0.90
	4	79	28	1403	0.86
	5	101	24	1356	0.83
	6	124	20	1311	0.80
	7	157	17	1259	0.77
2/1/80	1	0	124	1210	1
	2	33	40	1076	0.89
	3	56	31	1029	0.85
	4	79	27	968	0.80
	5	101	24	920	0.76
	6	124	21	835	0.69
	7	157	18	738	0.61

4.3 Comparison Results of the Molitor Plant and the Pilot Plant

Figures 7 to 13 are plots of BOD_5 concentration vs. time, and Figures 14 to 20 are plots of mixed liquor suspended solids at time, t , to the initial mixed liquor suspended solids vs. time for the Molitor Plant. Figures 21 to 32 are plots of BOD_5 concentration vs. time, and Figures 33 to 44 are mixed liquor suspended solids at time, t , to the initial mixed liquor suspended solids vs. time for the pilot plant. Each straight line on the semi-logarithmic plots of BOD_5 vs. time indicate a simple exponential decay equation as Equation (35) or (39). The initial portion of $\log BOD_5$ vs. time on each figure does not follow this general agreement. This deviation may be due to the Biosorption process occurring at the beginning of the aeration tank (41, 42). The Biosorption process causes the rapid reduction of the waste substrate concentration which is measured by BOD_5 in this study. Except the initial point of $\log BOD_5$ vs. time, the other points are determined by the Method of Least Squares to find out the slope and intercept of Equation (35). The intercept is the back calculated initial BOD_5 concentration. The slope divided by 2.303 is the slope of $\log BOD_5$ vs. time. The intercept and slope of the individual data are written down on each Figure (see Figure 7 to 13 and 21 to 32).

The relationship of the ratio of the mixed liquor suspended solids to the initial mixed liquor suspended solids vs. time appears to be linear in a semi-logarithmic scale. Thus, the mixed liquor suspended solids at any time can be obtained once the initial mixed liquor suspended solids is given. Since the $\log (MLSS_t/MLSS_o)$ vs. linear time declines in a linear manner, the weighted average value of the mixed liquor suspended solids at any time t can be easily determined. This value for each experimental run is calculated by the summation of mixed liquor suspended solids at time t multiplied by the time interval and divided by the detention time. Therefore, a value of $K_d A_s$, which is a function of mixed liquor suspended solids, can be determined for each experimental run. The straight line plots of $\log (MLSS_t/MLSS_o)$ vs. linear time in which the Biosorption phenomenon is evident are displayed in Figures 14, 15, 16, 17, 20, 33, 34, 35, 39, 42 and 43. Plots in which the Biosorption phenomenon are not significant are presented in Figures 18, 19, 36, 37, 38, 40 and 41. The semi-logarithmic plots of $MLSS_t/MLSS_o$ vs. time agree well with the calculated straight lines.

On each figure the calculated correlation coefficient is reported. The correlation coefficient, r , is a measure of the success with which the calculated line of best fit represents the individual data points. The absolute values

of r for the pilot plant are greater than those derived for the Molitor Plant. The environmental conditions in the pilot plant were carefully controlled. These parameters were pH, temperature, dissolved oxygen (D.O.) and the Food/Organisms (F/M) ratio. Analysis of significant parameters in the mass transfer expressions is more properly carried out with data resulting from the pilot plant operation.

4.4 Significant Parameters in the Mass Transfer Equations

The purpose of this section is to indicate the significant parameters of Equation (22), $Q(C_o - C)dt = K_d A_s (C_s - C)dt + V_f dc$. Integration of Equation (22) gives Equation (35), $C_t = C_o \exp - t/T$. The data results using BOD_5 as the major parameter gave an exponential decay of the type of Equation (35). It was found in every case that $\ln BOD_5 (C_t)$ vs. linear time, t , appeared as a straight line plot. The slope and intercept of the straight line were $-1/T$ and $\ln C_o$, respectively. T is the transfer time constant and C_o is the initial concentration. There are three terms when the transfer rate $K_d A_s$ is treated as a single term. The others are C_s , the equilibrium surface concentration and V_f , the fluid volume. Theoretically, $K_d A_s$ can be obtained from Equation (30). However, in reality the MLSS and, therefore, $K_d A_s$ change along the aeration tank. The average value of MLSS is simply

determined because the MLSS declines in an exponential manner along the tank. The second unknown V_f can be determined by $V_f = T (Q - K_d A_s)$. $T = -1/\text{slope}$ in each plot of $\ln C_t$ vs. t . Note that T is equal to $t/(\ln C_t - \ln C_0)$ from Equation (35). From Equation (34), one can obtain C_s as follows:

$$C_s = C_0 Q / K_d A_s$$

Table 6 presents the values of all of the parameters of Equation (22). The first column gives the sampling date. The second column presents the calculated initial concentration, C_0 , and the third column gives the time constant T . The fourth column gives the average value of mixed liquor suspended solids, $MLSS_{av}$. The fifth column gives the transfer rate in the activated sludge process, $K_d A_s$. The sixth column gives the volume of fluid, V_f . The seventh column gives the concentration on the biological flocs, C_s . The eighth column gives the flow rate, Q . The ninth column gives the calculated concentration of BOD_5 at the detention time of the aeration tank, $C_t (C_t = C_0 \exp (-t/T))$.

The results presented in Table 6 indicate a general agreement between the experimental and calculated concentrations of BOD_5 in the entire process, except for the initial portion. This deviation is due to the

TABLE 6. Values of the Significant Parameters for the Pilot Plant

1	2	3	4	5	6	7	8	9
Date	C_o mg/L	T hr	MLSS _{av} mg/L	$K_d^{A_s}$ L/hr	V_f L	C_s mg/L	Q L/hr	C_t mg/L
12/17/79	41.6	2.38	756	0.0181	57	55026	24	18.2
1/6/80	50.0	2.00	1226	0.0294	48	40793	24	18.7
1/9/80	50.3	1.61	1403	0.0337	39	35822	24	14.8
1/10/80	47.5	1.85	1115	0.0268	44	42601	24	16.4
1/11/80	48.7	2.11	1069	0.0257	51	45557	24	19.2
1/17/80	47.2	2.16	1034	0.0248	52	45648	24	19.0
1/18/80	69.2	1.56	1435	0.0344	37	48223	24	19.6
1/28/80	202.2	1.95	1210	0.0581	94	167107	48	122.0
1/29/80	46.8	2.35	980	0.0294	71	47755	30	24.0
1/30/80	80.1	1.55	1504	0.0361	37	53278	24	22.5
1/31/80	48.2	2.44	1340	0.0241	44	35970	18	16.5
2/1/80	45.8	2.69	912	0.0164	48	50219	18	17.3

Note: C_o = back calculated initial
BOD₅ concentration

MLSS_{av} = weighted average value of the mixed liquor suspended solids

$K_d^{A_s}$ = transfer rate

C_s = equilibrium surface BOD₅ concentration on the biological flocs

Q = flow rate

T = transfer time constant

V_f = volume of the fluid

C_t = BOD₅ concentration at the
detention time

Biosorption process occurring at the beginning of the aeration tank (41, 42). The Biosorption process causes the rapid reduction of the waste substrate concentration, which is measured by BOD_5 in this study.

Comparing values of $K_d A_s$ and Q in Table 6, it is clear that the assumption that $K_d A_s \ll Q$ is valid. By comparing values of C_s and C_o , assumption 3 (p. 32) used to derive the equations also appears to be valid (namely, $C_o \ll C_s$).

4.5 Dimensionless Group

In this section Equation (40) will be discussed. The dimensionless numbers in Equation (40) are the Sherwood number (Sh), the particle Reynolds number (Re_p) and the Schmidt number (Sc). The relationships of the dimensionless numbers using the laboratory results can be obtained from the following:

1. The normal surface area of the aeration tank in the laboratory unit (A_c) = 566 cm².

2. Assume the shape of organism is spherical with taking an average value for the diameter (44):

$$D_p = 2 \times 10^{-4} \text{ cm}$$

3. The kinematic viscosity of wastewater (ν) used in this study is approximately equal to that of potable water, i.e., $\nu = 10^{-2} \text{ cm}^2/\text{sec}$.

4. The velocity of wastewater (u) in the aeration tank is equal to the flow rate (Q) divided by normal area of the tank (A_c). They are as follows:

Q , L/hr	A_c , cm^2	u , cm/sec
18	566	0.0088
24	566	0.0118
30	566	0.0147
48	566	0.0236

5. The molecular diffusivity of synthetic waste (D_v) is a function of: temperature, square root of molecular weight of solvent (wastewater), viscosity, and (molar volume)^{-0.6} (45). The molecular diffusivity of waste (sucrose is the major substrate) can be assumed as: (see reference 41 in Table 23-1 and reference 47, 48).

$$D_v = 5 \times 10^{-6} \text{ cm}^2/\text{sec}$$

6. When a constant flow rate (Q) is chosen by 24 L/hr, the Reynolds number;

$$Re_p = u D_p / \nu = 2.36 \times 10^{-4}$$

7. The value of $K_d A_s$ can be determined from Equation (30). At $Q = 24$ L/hr, the relationship between $Sh A_s$ and Sc (MLSS) at constant Reynolds number ($Re_p = 2.36 \times 10^{-4}$) is listed as following table.

Date	MLSS _{av} , L/L	$K_d A_s$, L/hr	$Sh A_s$, cm ²	Sc (MLSS _{av})
12/17/79	7.56×10^{-4}	0.0181	0.201	1.512
1/6/80	1.23×10^{-3}	0.0294	0.332	2.448
1/9/80	1.40×10^{-3}	0.0337	0.373	2.808
1/10/80	1.16×10^{-3}	0.0268	0.297	2.232
1/11/80	1.07×10^{-3}	0.0257	0.284	2.136
1/17/80	1.03×10^{-3}	0.0248	0.275	2.070
1/18/80	1.44×10^{-3}	0.0344	0.382	2.868
1/30/80	1.50×10^{-3}	0.0361	0.400	3.006

Equation (40) can be expressed as

$$\ln (Sh A_s) = \ln (A_c f (Re_p)^a) + b \ln (Sc (MLSS)) \quad (41)$$

From Equation (41), one can determine the type equation ($Y = b X + g$) by using the "least square" method. The slope and intercept are:

$$b \text{ (slope)} = 1.004$$

$$g \text{ (intercept)} = \ln (A_c f (Re_p)^a) = -2.019$$

8. With a constant MLSS = 10^{-3} L/L and variable flow rate (Q). The table of Q, $K_d A_s$, Sc(MLSS), Re_p , and $Sh A_s$ is given as follows:

Q, L/hr	MLSS, L/L	$K_d A_s$, L/hr	Sc(MLSS)	Re_p	$Sh A_s$, cm^2
18	10^{-3}	1.8×10^{-2}	2.00	0.000177	0.200
24	10^{-3}	2.4×10^{-2}	2.00	0.000236	0.267
30	10^{-3}	3.0×10^{-2}	2.00	0.000295	0.334
48	10^{-3}	4.8×10^{-2}	2.00	0.000472	0.533

Equation (40) can be expressed as

$$\ln(Sh A_s) = \ln(A_c f (Sc MLSS)^b) + a \ln(Re_p) \quad (42)$$

From Equation (42), one can determine the type equation ($Y = aX' + j$) by using the "least square" method. The slope and the intercept are:

$$a(\text{slope}) = 0.999$$

$$j \text{ (intercept)} = \ln (A_c f (Sc MLSS)^b) = 7.024$$

$$f = e^{7.024} / (566 \times 2.00)^{1.004} = 0.98$$

From the above empirical constants (f, a, b), Equation (40) can be written as:

$$Sh = 0.98 (A_c / A_s) (Re_p)^{0.999} (Sc MLSS)^{1.004} \quad (43)$$

Equation (43) is the empirical dimensionless group function for the activated sludge process in terms of the fluidized

bed model. The relationship of Equation (43) can be further verified by the theoretical dimensionless group function as follows:

Substituting $Q = A_c u$ into Equation (30), yields

$$K_d = (A_c/A_s) u \text{ (MLSS)} \quad (44)$$

Multiplying both sides of Equation (44) by D_p and $1/D_v$, with v/v of right side, one obtains:

$$K_d D_p/D_v = (A_c/A_s) (u D_p/v) (v/D_v) \text{ (MLSS)} \quad (45)$$

Recalling the definitions of the Sherwood number, the particle Reynolds number, and the Schmidt number, Equation (45) can now be expressed as:

$$\text{Sh} = (A_c/A_s) \text{Re}_p \text{Sc} \text{ (MLSS)} \quad (46)$$

It can be seen that Equations (46) and (43) are identical in terms of dimensional analysis. Thus, Equation (40) is capable of correlating the laboratory data.

4.6. Relationships of Parameters

The adequate transfer conditions in the bed system depend on the transfer time and the mixed liquor suspended solids. The former relates the activity of bacteria digesting ability to the substrate of the wastewater in the aeration tank. The latter relates the amount of area available to the substrate of the wastewater for transfer. If the transfer time is too

short, it would not have sufficient time for bacteria growth. As a result, the consumption of substrate by the bacteria could not achieve the stable condition as predicted by the principle of mass transfer. For example, the flow rate at 48 L/hr in the laboratory study was found to be too short for the stable bacteria growth in the aeration tank (see Figures 28 and 40). In addition, the MLSS should be controlled between 1000 to 3000 ppm in the aeration tank (9). The lower or the higher values do not permit adequate transfer (see Figures 8 and 15). Dynamic bacterial populations could not adequately utilize the waste substrate. The exact value should be based on bed conditions.

The relationships of the parameters of interest were shown in Figures 45 to 48. The following explanation is presented.

The transfer rate of waste substrate ($K_d A_s$) vs. the volume of fluid in the aeration tank (V_f) is shown in Figure 45. When $Q = 24$ L/hr, the type equation of the transfer rate can be found as:

$$K_d A_s = 0.0632 - 0.000785 V_f$$

The volume of fluid in the aeration tank (V_f) vs. the mixed liquor suspended solids (MLSS) is shown in Figure 46. When Q is less than 24 L/hr, the type equation of the volume of fluid can be found as:

$$V_f = 86.87 \exp (-\text{MLSS}/1808)$$

The equilibrium surface concentration on the biological flocs (C_s) vs. the mixed liquor suspended solids (MLSS) is shown in Figure 47. When the initial concentration (C_o) is 50 mg/L and below, the type equation of the equilibrium surface concentration can be found as:

$$C_s = 94.171 \exp (-MLSS/1435)$$

The transfer time (T) vs. the mixed liquor suspended solids (MLSS) is shown in Figure 48. The type equation of the transfer time can be found as:

$$T = 4.185 \exp (-MLSS/1603)$$

4.7 Reproducing Response of the Molitor Plant

A study to predict the characteristics of the existing plant based on the associated properties of the bench scale pilot plant is given in this section.

The calculated average transfer time (V/Q) of the bench scale pilot plant was found to be 2.1 hours. The corresponding MLSS from Figure (48) is 1100 mg/L. The detention time of the Molitor Plant is designed for 7 hours. The volume of the aeration tank is 7' x 9' x 600' (37800 ft³). The flow rate is 3.5 MGD. For the first portion of $\ln C_t$ vs. t curve (Figure 7), the measured time constant is equal to $T_1 = 80/.693 = 115 \text{ min} = 1.93 \text{ hours}$. The calculation of $T_1 = V_f/Q = 37800 \text{ ft}^3/3.5 \text{ MGD} = 1.94 \text{ hours}$. The calculated time

constant T_1 is very closely equal to the measured time constant. The expression $T_1 = V_f / (Q - K_d A_s) \cong V_f / Q$ is essentially a description of reactor size and flow rate. Since $K_d A_s \ll Q$, $V_f / Q = T_1$ is independent of the MLSS. For the second portion of the $\ln C_t$ vs. t curve, the following ratios apply:

$$\frac{T_2 \text{ (Molitor)}}{T_1 \text{ (Molitor Calculated)}} = \frac{T_2 \text{ (Pilot)}}{T_1 \text{ (Pilot Calculated)}}$$

$$\frac{T_2 \text{ (Pilot)}}{T_2 \text{ (Molitor Measured)}} = \text{MLSS}_{\text{(Molitor)}} / \text{MLSS}_{\text{(Pilot)}}$$

Therefore, $T_2 \text{ (Molitor)} = 1.94 \times 7/2.1 \times \text{MLSS}_{\text{(Molitor)}} / 1100$

The time constant, T_2 , is proportional to the MLSS. From the above discussion, it appears that it is possible to scale up from pilot plant data to the full sized plant.

In the following tables are listed the calculated C_t and the measured C_t for each date.

Table 7. The Calculated BOD₅ and the Measured BOD₅ for Each Date.

Date: 5/11/79

First Portion: $T_1 = 1.94$ hours

$t = 1.2$ hours

$C_0 = 96$ mg/L

$C_t = 96 \exp(-1.2/1.94) = 51.7$ mg/L

Second Portion: MLSS = 1000 mg/L

$T_2 = 1.94 \times 7/2.1 \times 1000/1100 = 5.88$ hours

$C_0 = 51.7 / (\exp -1.2/5.88) = 63$ mg/L

t , hr	C_t (calculated) mg/L	C_t (measured) mg/L
0	63	96
1.17	52	54
2.33	42	45
3.50	35 $\sigma^* = 2.2$	39 $\sigma = 2.4$
4.67	28	25
5.83	23	23
7.00	19	15

*

σ = standard deviation of BOD₅

Date: 5/25/79

First Portion: $T_1 = 1.94$ hours

$t = 1.2$ hours

$C_o = 150$ mg/L

$C_t = 150 \exp(-1.2/1.94) = 80.8$ mg/L

Second Portion: MLSS = 3500 mg/L

$T_2 = 1.94 \times 7/2.1 \times 3500/1100 = 20.8$ hours

$C_o = 80.8 / (\exp -1.2/20.58) = 86$ mg/L

$t, \text{ hr}$	C_t (calculated) mg/L	C_t (measured) mg/L
0	86	150
1.17	81	92
2.33	77	73
3.50	73	60
	$\sigma = 2.0$	$\sigma = 2.6$
4.67	69	57
5.83	65	33
7.00	61	26

Date: 6/3/79

First Portion: $T_1 = 1.94$ hours

$t = 1.2$ hours

$C_o = 44$ mg/L

$C_t = 44 \exp(-1.2/1.94) = 23.7$ mg/L

Second Portion: MLSS = 1000 mg/L

$T_2 = 1.94 \times 7/2.1 \times 1000/1100 = 5.88$ hours

$C_o = 23.7 / (\exp -1.2/5.88) = 29$ mg/L

$t, \text{ hr}$	C_t (calculated) mg/L	C_t (measured) mg/L
0	29	44
1.17	24	24
2.33	20	18
3.5	16 $\sigma = 1.8$	16 $\sigma = 1.6$
4.67	13	14
5.83	11	11
7.00	9	10

Date: 6/4/79

First Portion: $T_1 = 1.94$ hours
 $t = 1.2$ hours
 $C_o = 75$ mg/L
 $C_t = 75 \exp(-1.2/1.94) = 40.4$ mg/L

Second Portion: MLSS = 1060 mg/L
 $T_2 = 1.94 \times 7/2.1 \times 1060/1100 = 6.23$ hours
 $C_o = 40.4 \exp(-1.2/6.23) = 49$ mg/L

t, hr	C_t (calculated) mg/L	C_t (measured) mg/L
0	49	75
1.17	41	46
2.33	34	34
3.5	28	26
	$\sigma = 2.1$	$\sigma = 2.1$
4.67	23	20
5.83	19	19
7.00	16	16

Date: 6/14/79

First Portion: $T_1 = 1.94$ hours

$t = 1.2$ hours

$C_o = 38$ mg/L

$C_t = C_o \exp(-t/T_1) = 20.47$ mg/L

Second Portion: MLSS = 1600 mg/L

$T_2 = 1.94 \times 7/2.1 \times 1600/1100 = 9.4$ hours

$C_o = 20.47/\exp(-1.2/9.4) = 23$ mg/L

t, hr	C_t (calculated) mg/L	C_t (measured) mg/L
0	23	38
1.17	21	25
2.33	18	22
3.5	16	17
	$\delta = 1.4$	$\delta = 1.7$
4.67	14	14
5.83	13	13
7.00	11	11

Date: 6/28/79

First Portion: $T_1 = 1.94$ hours
 $t = 1.2$ hours
 $C_o = 75$ mg/L
 $C_t = 75 \exp(-1.2/1.94) = 40.4$ mg/L

Second Portion: MLSS = 2250 mg/L
 $T_2 = 1.94 \times 7/2.1 \times 2250/1100 = 13.23$ hours
 $C_o = 40.4/\exp(-1.2/13.23) = 44$ mg/L

t, hr	C_t (calculated) mg/L	C_t (measured) mg/L
0	44	75
1.17	40	61
2.33	37	49
3.5	34 $\sigma = 1.8$	43 $\sigma = 2.1$
4.67	31	38
5.83	28	34
7.00	26	32

Date: 7/5/79

First Portion: $T_1 = 1,94$ hours

$t = 1,2$ hours

$C_o = 73$ mg/L

$C_t = 73 \exp(-1,2/1,94) = 39,3$ mg/L

Second Portion: MLSS = 3600 mg/L

$T_2 = 1,94 \times 7/2,1 \times 36/11 = 21,16$ hours

$C_o = 39,3 / (\exp(-1,2/21,16)) = 42$ mg/L

t, hr	C_t (calculated) mg/L	C_t (measured) mg/L
0	42	73
1,17	40	50
2,33	38	43
3,50	36 $\delta = 1,5$	37 $\delta = 2,1$
4,67	34	32
5,83	32	26
7,00	30	24

From the above data, it appears that the measured C_t values are closely reproduced by the calculated values when the Mixed Liquor Suspended Solids are less than 2000 mg/L.

It appears that the two distinct first order responses demonstrated can be utilized to predict responses of activated sludge plants before construction. The first curve is independent of the amount of MLSS and closely follows the response predicted from the theoretical values derived using the plant parameters of flow rate, aeration tank volume and influent BOD. The second portion can be represented in terms of the full size plant and data derived from the bench scale pilot plant.

CHAPTER V. CONCLUSIONS

As a result of the investigations reported here, there is a means by which mass transfer responses of an activated sludge aeration tank can be reasonably predicted in the design stage. Alternative designs can be evaluated and compared and decisions made concerning expected operating results before the plant is constructed.

Strong support has been developed for representation of the activated sludge process as a fluidized bed. Significant advances in wastewater treatment will probably be in the area of process development. New hardware development will allow some forward movement, but process modification offers much greater promise of major improvement. Much technology now exists as a result of developments in fluidization engineering and there is promise that some of this technology can be directly applied to wastewater treatment.

Equations which have been shown to apply to mass transfer phenomena in fluidized beds have been shown to apply to the systems examined in the study reported here. A descriptive model, derived from first principles, has been introduced. The basic equation presented, $C_t = C_o - \frac{K_d A_s C_s}{Q} (1 - \exp - \frac{t}{T})$ degenerates to a simple exponential decay, using BOD_5 as the parameter. The data appear as a straight line on a semilogarithmic plot of BOD_5 vs. time. The majority of the data do follow this straight line.

In the early portions of the response curves plotted as semi-logarithmic presentations, there is deviation from the linear. This portion appears to represent the well known phenomenon of Biosorption.

In the Equation $C_t = C_o - (K_d A_s C_s / Q) (1 - \exp(-t/T))$, when $C_o \cong K_d A_s C_s / Q$, $C_t = C_o - C_o + C_o \exp(-t/T) = C_o \exp(-t/T)$. This is the measured response:

$$K_d A_s C_s / Q = C_o$$

$$C_s \cong \text{constant, when MLSS} = \text{constant.}$$

If this is the case,

$$K_d A_s = C_o Q / C_s. \text{ ---- MLSS} = \text{constant}$$

$$K_d A_s \propto Q \text{ ---- MLSS} = \text{constant}$$

$$K_d A_s \propto C_o \text{ ---- MLSS} = \text{constant}$$

The larger the flow rate, the greater the mass transfer. The larger the influent concentration, the greater the mass transfer. Since MLSS = constant, the surface area will be closely constant and the mass transfer coefficient will be greater.

The mass transfer coefficient can be correlated using the same dimensionless functions as a fluidized bed. Equation (40), $K_d A_s D_p / D_v = f A_c (u D_p / \nu)^a ((\nu / D_v) (\text{MLSS}))^b$, is the empirical dimensionless group function. It was found that the empirical constants (f, a, b) were very closely equal to 1. Therefore, Equation (40) could be expressed the dimensionless group as that of the theoretical dimensionless Equation (46), $Sh = (A_c / A_s) Re_p Sc (\text{MLSS})$.

CHAPTER VI. FUTURE WORK

Strong supporting evidence has been presented for the view that the activated sludge wastewater treatment process may be represented as a fluidized bed. The results and conclusions given in the present work are based on data obtained from an operating municipal water pollution control facility and a bench scale laboratory pilot plant. The properties of the pilot plant influent were closely those of the municipal plant. Municipal wastewater is a complex mixture of biodegradable substances and a representative value of molecular diffusivity which can be applied to municipal wastewater is not now known. Molecular diffusivity must be known before the Schmidt Number, ν/D_v , can be determined. At present, it is not possible to apply the large amount of existing fluidized bed technology to the activated sludge process because of the lack of knowledge of the molecular diffusivity of wastewater.

Future work should be concerned with synthetic wastewaters in which the substrates are substances of known molecular diffusivities. Such investigation will allow determination of all of the terms in the proposed models. Determination of activated sludge process responses into previously published and well accepted curves descriptive

of fluidized bed mass transfer phenomena. The curves thus generated should be regarded as adequate demonstration of the validity of the representation of the activated sludge process as a fluidized bed.

The question of a useable Schmidt Number for municipal wastewater should be addressed. A number of operating plants should be examined in the manner set forth in this study. The resulting determinations of Sherwood Numbers, Reynolds Numbers and Mixed Liquor Suspended Solids will give Schmidt Numbers for the wastewater of the plants tested. It should be possible to thus determine a value or range of values of Schmidt Numbers of utility to the designer.

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APPENDIX I Derivation of Equation (8)

$$\begin{aligned}
-\frac{ds}{dt} &= K_{+1} S E - K_{-1} ES \\
&= \frac{K_{+1} S E (K_{+2} + K_{-1})}{K_{-1} + K_{+2}} - K_{-1} ES \\
&= \frac{K_{+1} K_{+2} S E}{K_{-1} + K_{+2}} - K_{-1} ES + \frac{K_{-1} K_{+1} S E}{K_{-1} + K_{+2}} \\
&= \frac{K_1 K_2 S E}{K_{-1} + K_{+2}} - \frac{K_{-1}}{K_{-1} + K_{+2}} \frac{(K_1 + K_{+2}) ES}{E} - K_{+1} S E \\
&= \frac{K_{-1} K_{+2} S E}{K_{-1} + K_{+2}} - \frac{K_{-1}}{K_{-1} + K_{+2}} K_{-2} P E \\
&= \frac{(E + ES) \left(\frac{K_{+1} K_{+2} S - K_1 K_{-2} P}{K_{-1} + K_{+2}} \right) E}{E + ES} \\
&= \frac{(E+ES) K_{+1} K_{+2} S - (E + ES) K_{-1} K_{-2} P}{K_{-1} + K_{+2}} \\
&= 1 + \frac{ES}{E} \\
&- \frac{K_{-1} (E+ES) (S+P) K_{-2}}{K_1 + K_{+2}} + \frac{(E+ES) (K_{+1} K_{+2} + K_{-1} K_{-2}) S}{K_{-1} + K_{+2}} \\
&= \frac{1 + \frac{K_{+1} - K_{-2}}{K_{-1} + K_{+2}} S + \frac{K_{-2} (S+P)}{K_{-1} + K_{+2}}}{1 + \left(\frac{K_{+1}}{K_{-1} + K_{+2}} - \frac{K_{-2}}{K_{-1} + K_{+2}} \right) S + \left(\frac{K_{-2}}{K_{-1} + K_{+2}} \right) S^0} \\
&- \frac{K_{-1} K_{-2} E^0 S^0}{K_{-1} + K_{+2}} + \left(\frac{K_{+1} K_{+2} E^0}{K_{-1} + K_{+2}} + \frac{K_{-1} K_{-2} E^0}{K_{-1} + K_{+2}} \right) S \\
&= \frac{1 + \left(\frac{K_{+1}}{K_{-1} + K_{+2}} - \frac{K_{-2}}{K_{-1} + K_{+2}} \right) S + \left(\frac{K_{-2}}{K_{-1} + K_{+2}} \right) S^0}{1 + \left(\frac{1}{K_m} - \frac{1}{K_p} \right) S + \frac{1}{K_p} S^0} \\
&= \frac{((V_m/K_m) + (V_p/K_p)) S - (V_p/K_p) S^0}{1 + ((1/K_m) - (1/K_p)) S + (1/K_p) S^0}
\end{aligned}$$

APPENDIX II

BOD₅ and Mixed Liquor Suspended Solids
Data for the Molitor Plant

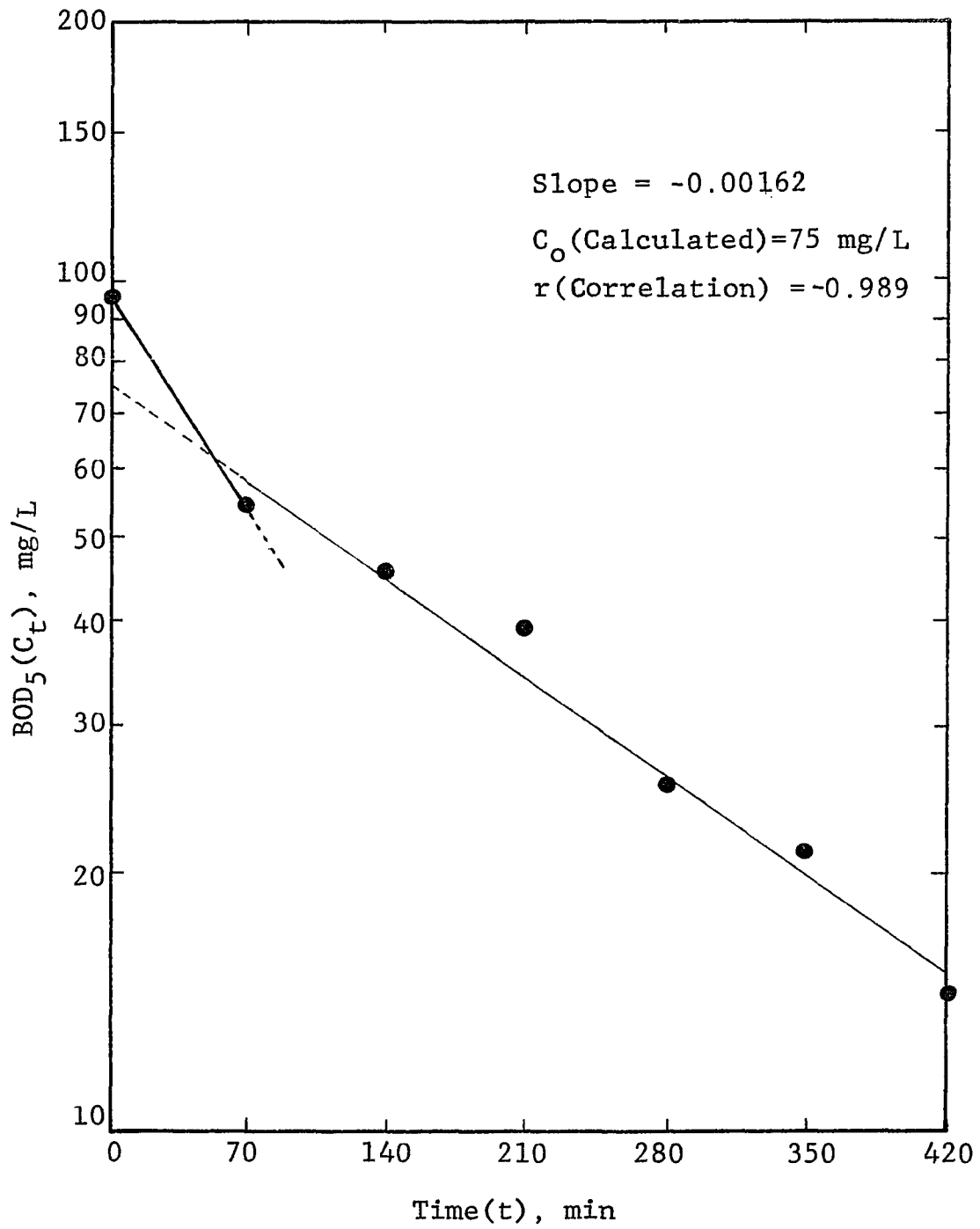


Figure 7. Molitor Plant (Data of 5/11/79)

BOD₅(C_t) vs. Time (t)

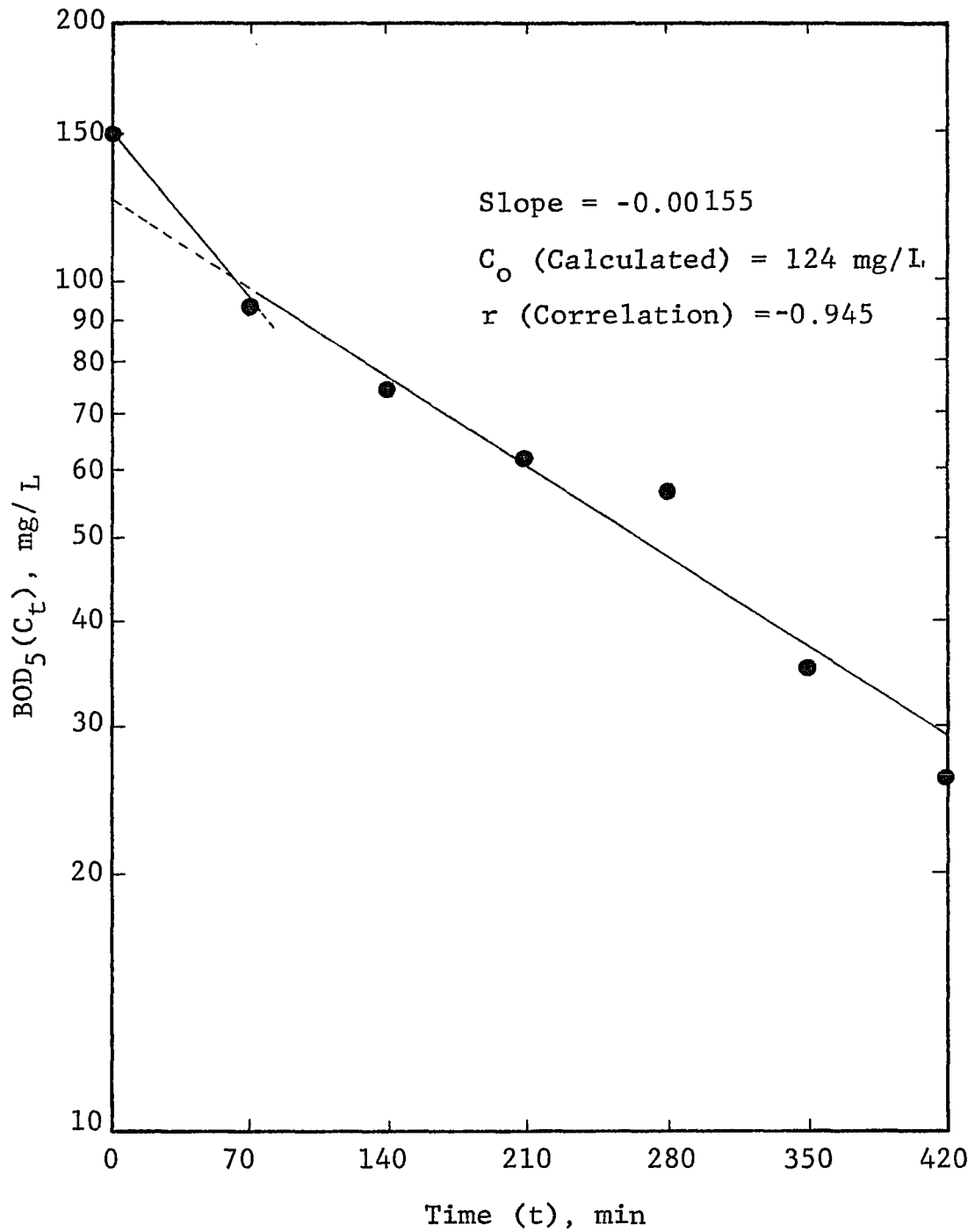


Figure 8. Molitor Plant (Data of 5/25/79)

$BOD_5(C_t)$ vs. Time (t)

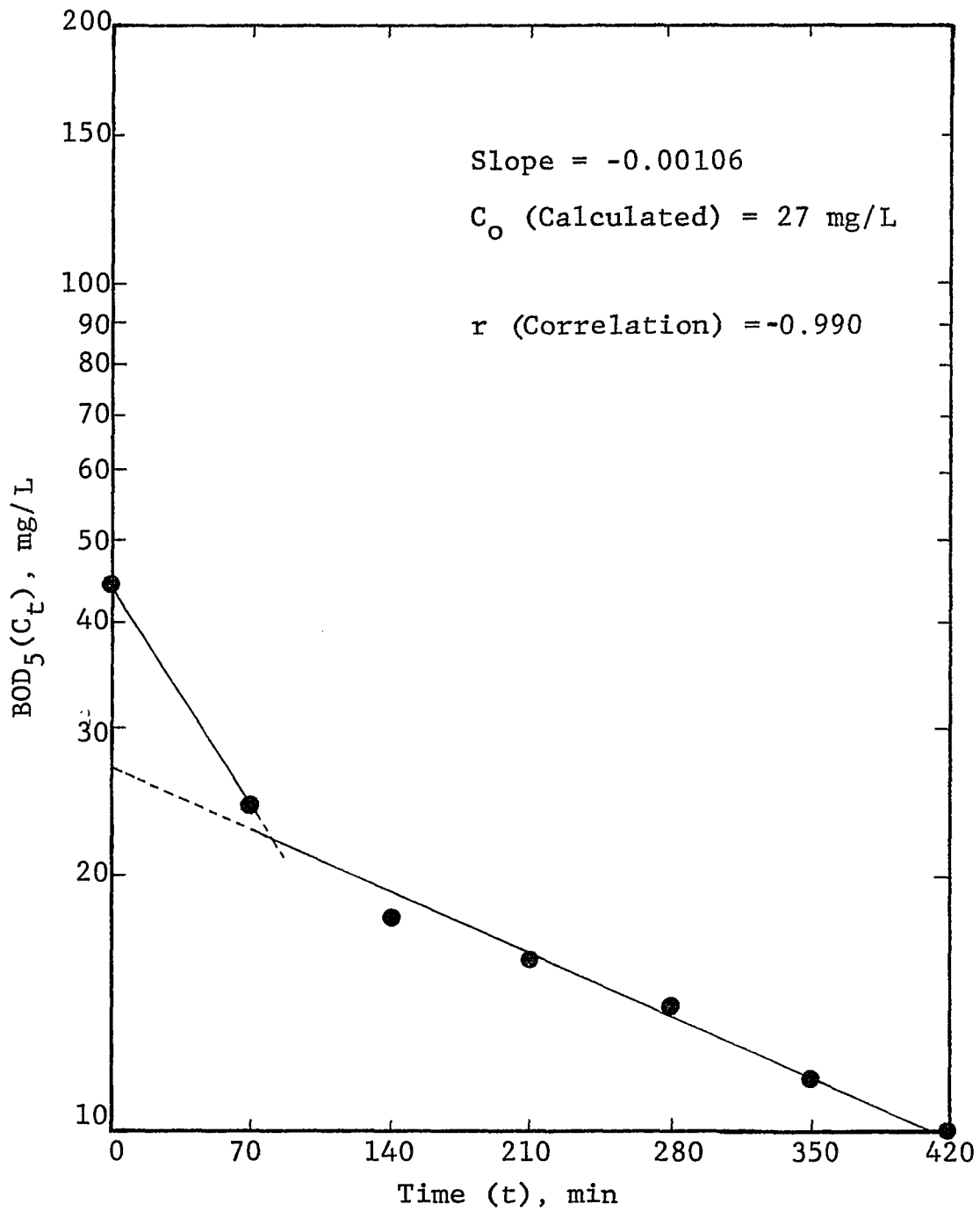


Figure 9. Molitor Plant (Data of 6/3/79)

$BOD_5(C_t)$ vs. Time (t)

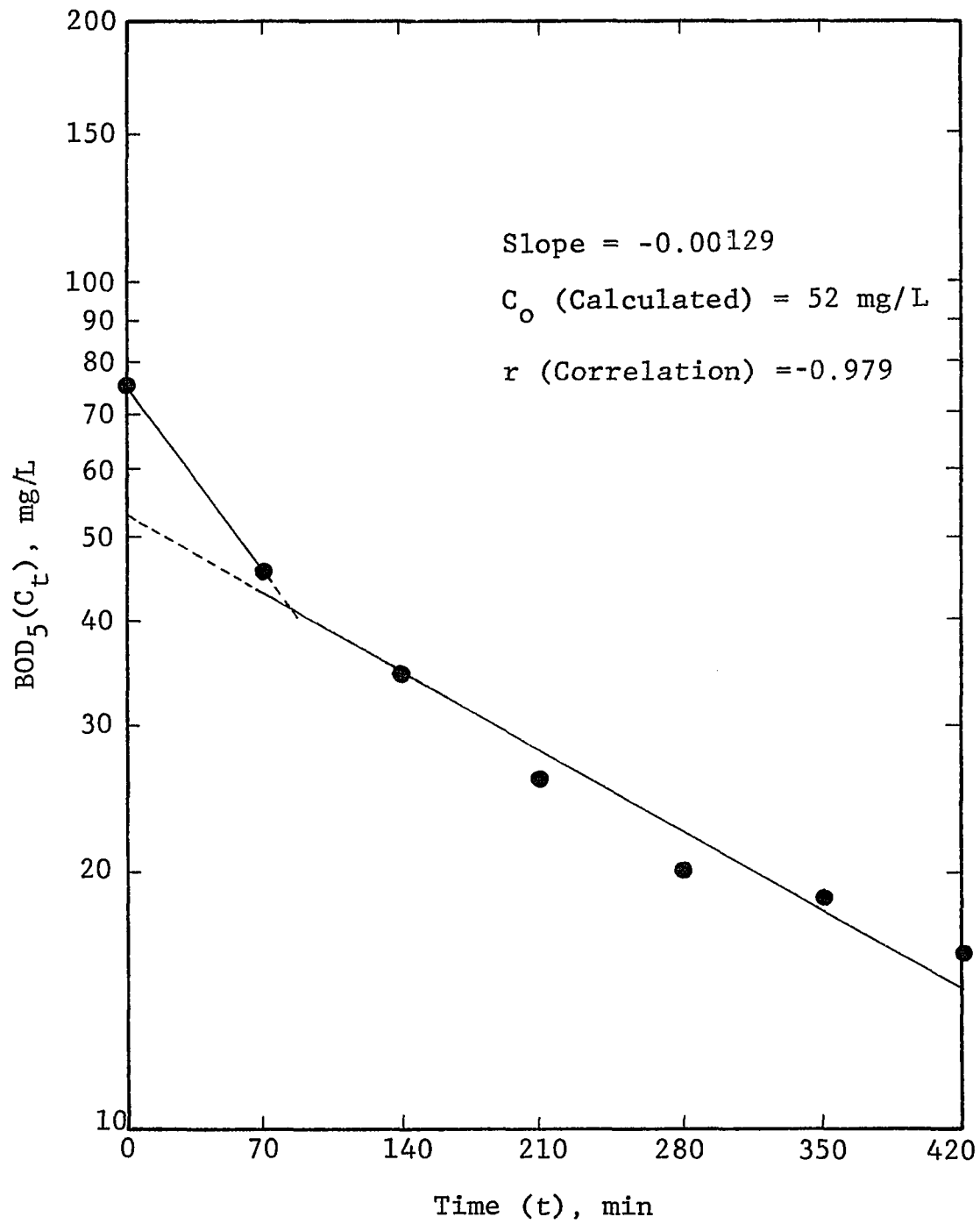


Figure 10. Molitor Plant (Data of 6/4/79)

BOD₅(C_t) vs. Time (t)

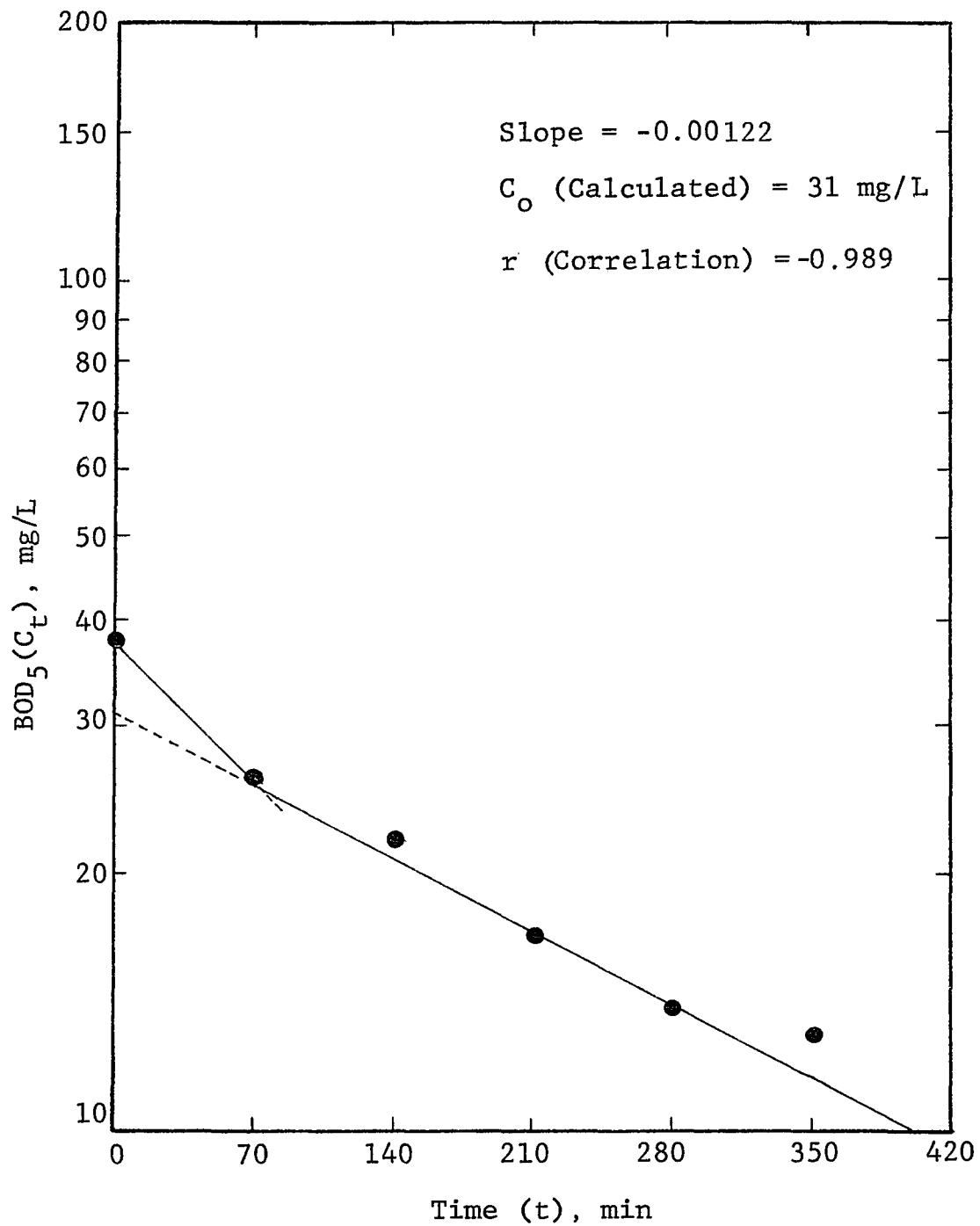


Figure 11. Molitor Plant (Data of 6/14/79)

 $BOD_5(C_t)$ vs. Time (t)

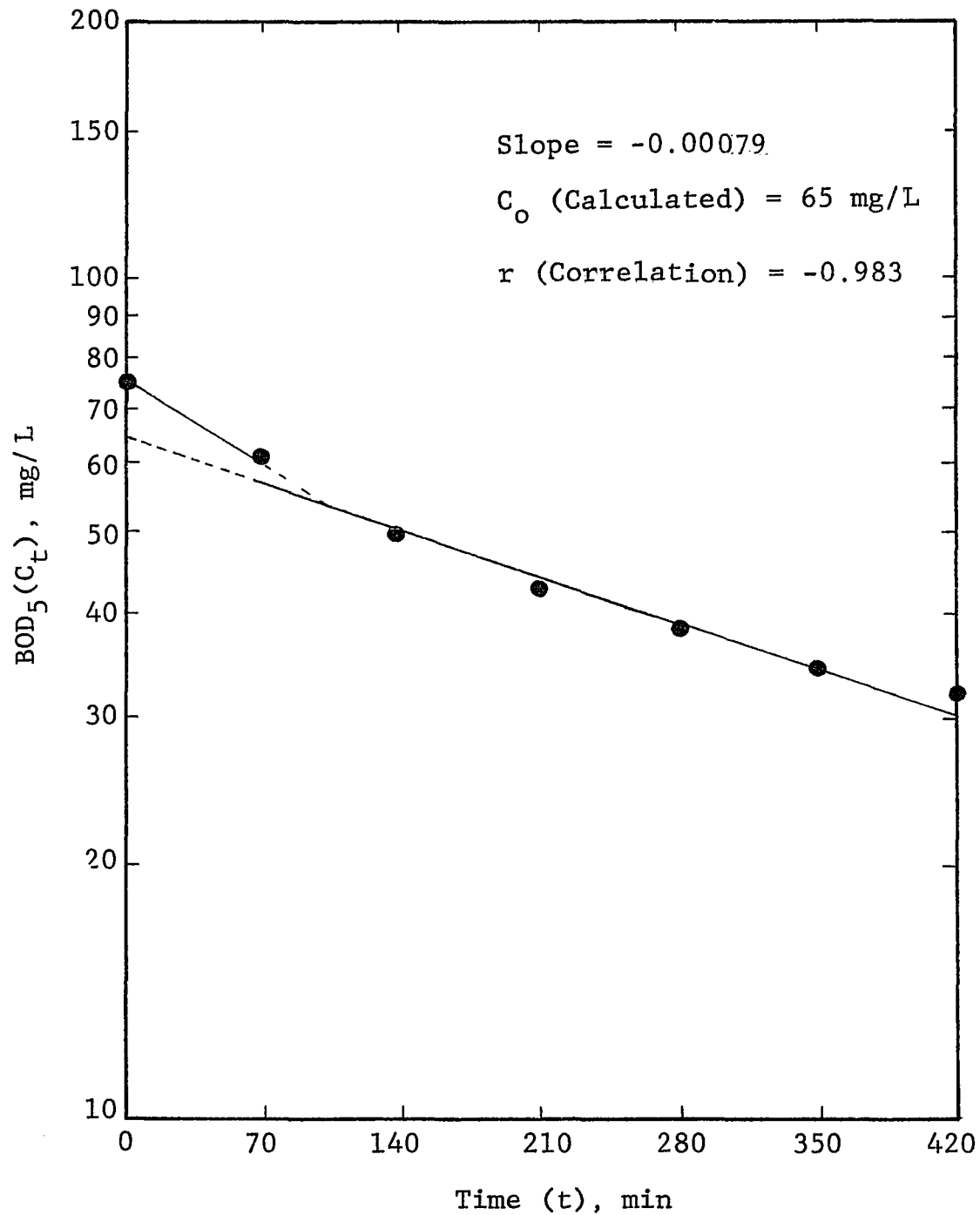


Figure 12. Molitor Plant (Data of 6/28/79)

$BOD_5(C_t)$ vs. Time (t)

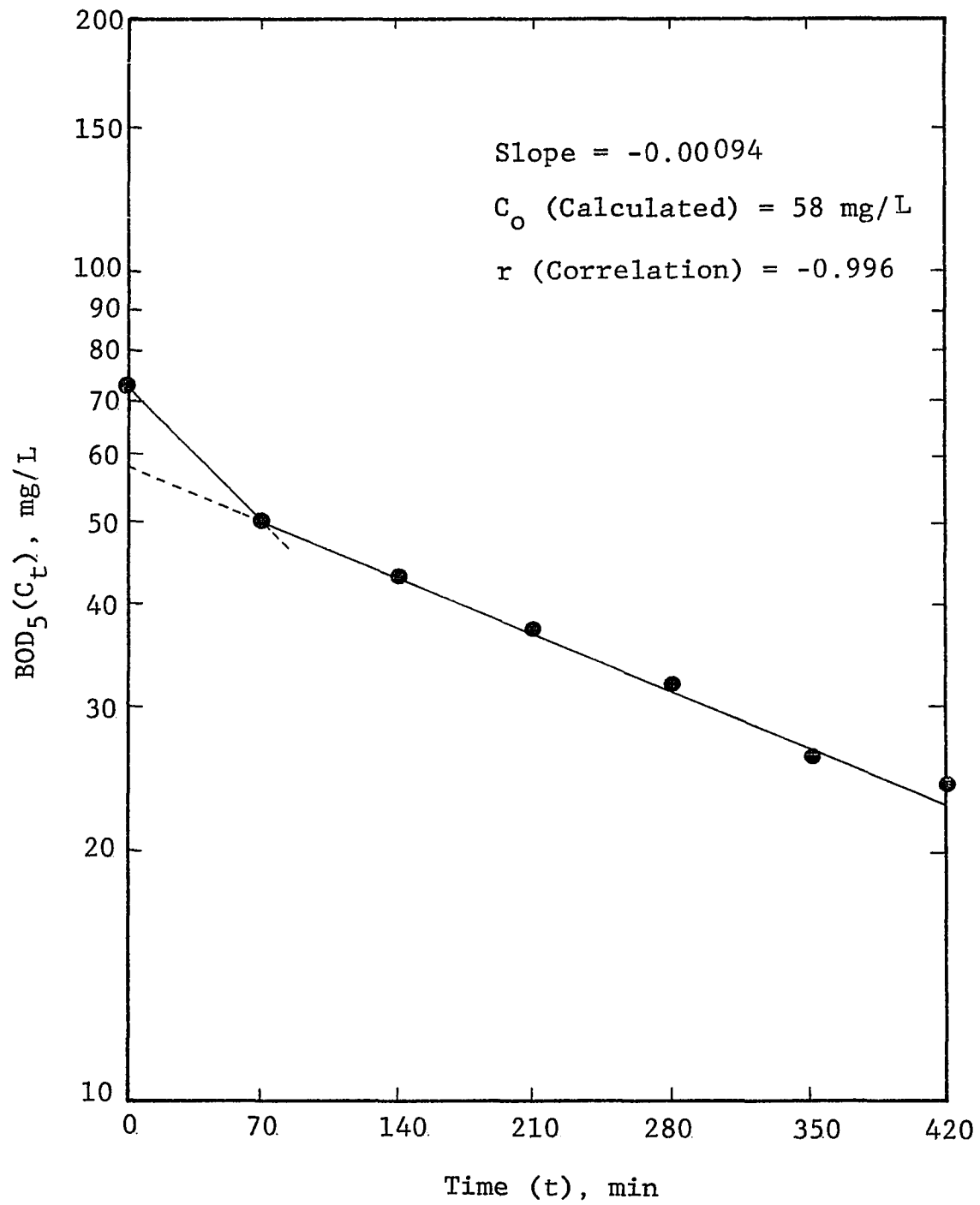


Figure 13. Molitor Plant (Data of 7/5/79)

BOD₅(C_t) vs. Time (t)

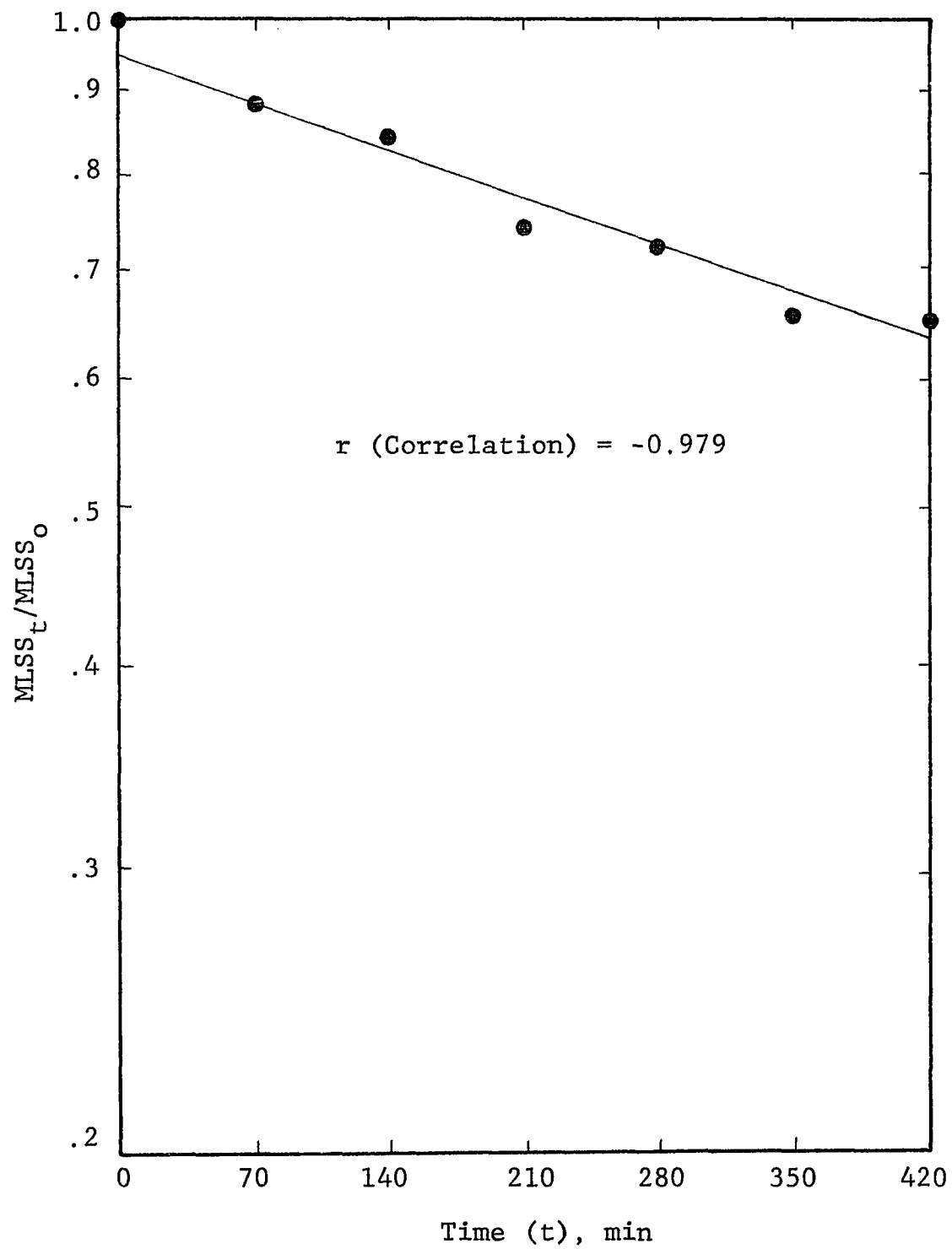


Figure 14. Molitor Plant (Data of 5/11/79)

$MLSS_t/MLSS_0$ vs. Time (t)

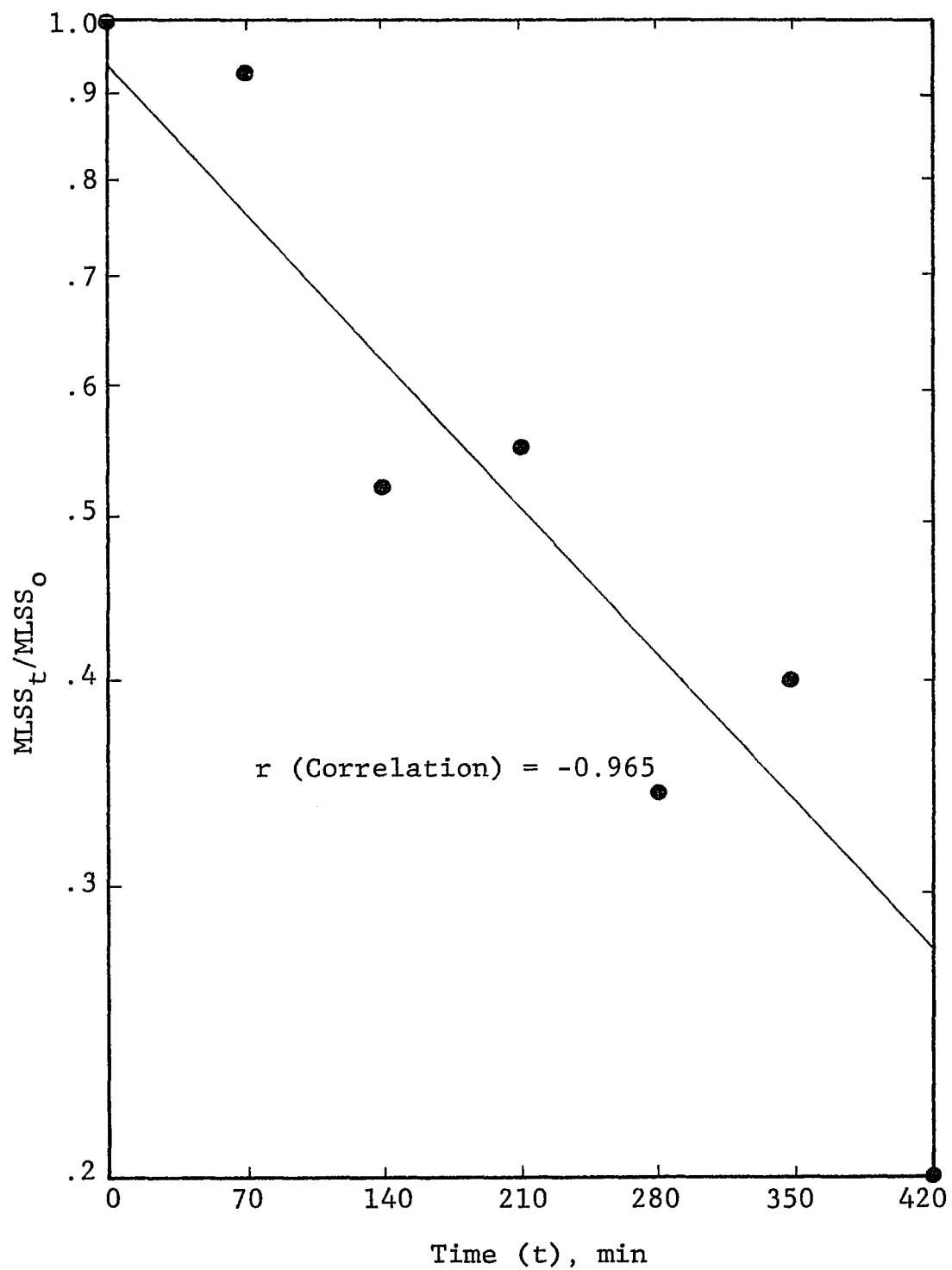


Figure 15. Molitor Plant (Data of 5/25/79)

$MLSS_t/MLSS_0$ vs. Time (t)

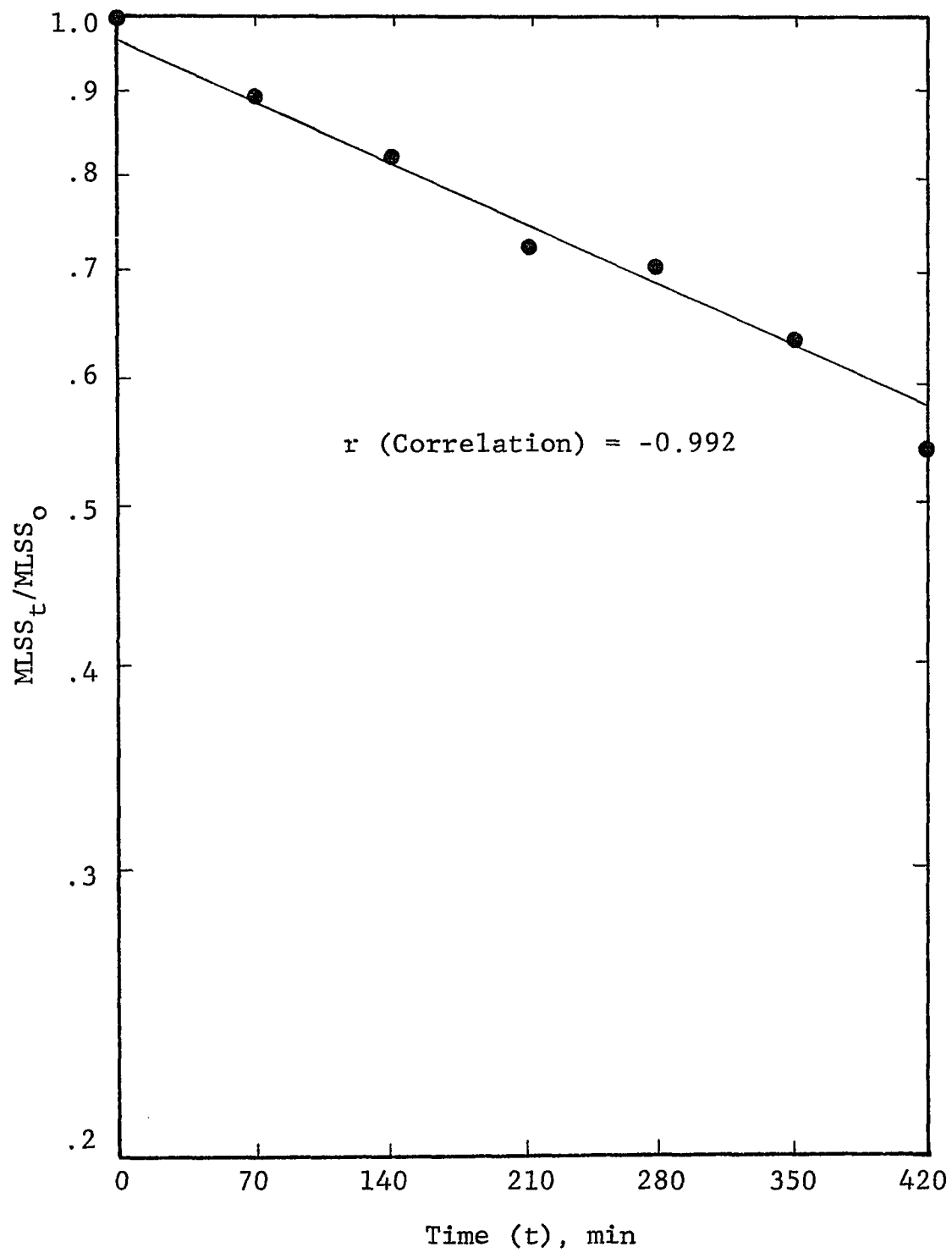


Figure 16. Molitor Plant (Data of 6/3/79)

MLSS_t/MLSS₀ vs. Time (t)

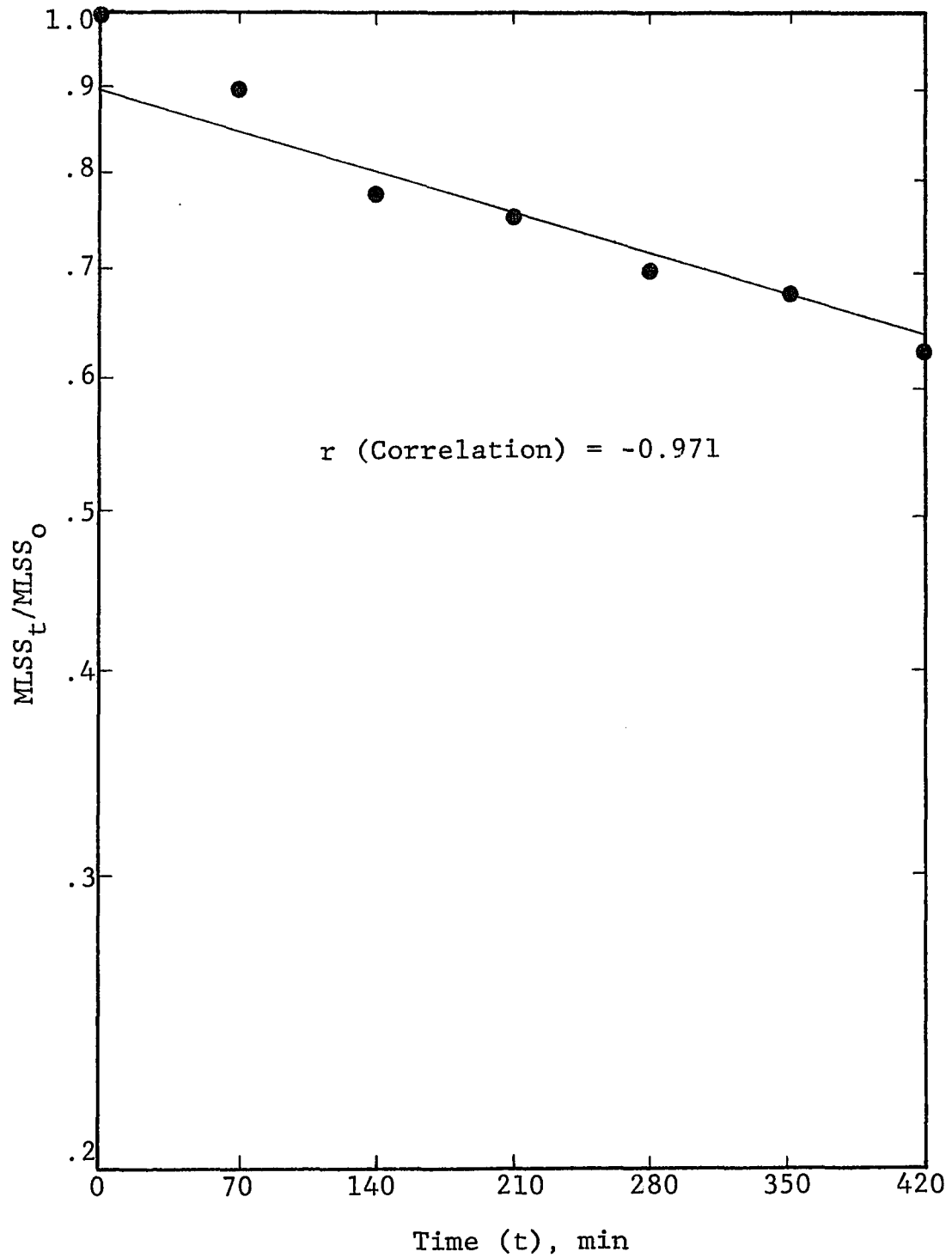


Figure 17. Molitor Plant (Data of 6/4/79)

$MLSS_t/MLSS_0$ vs. Time (t)

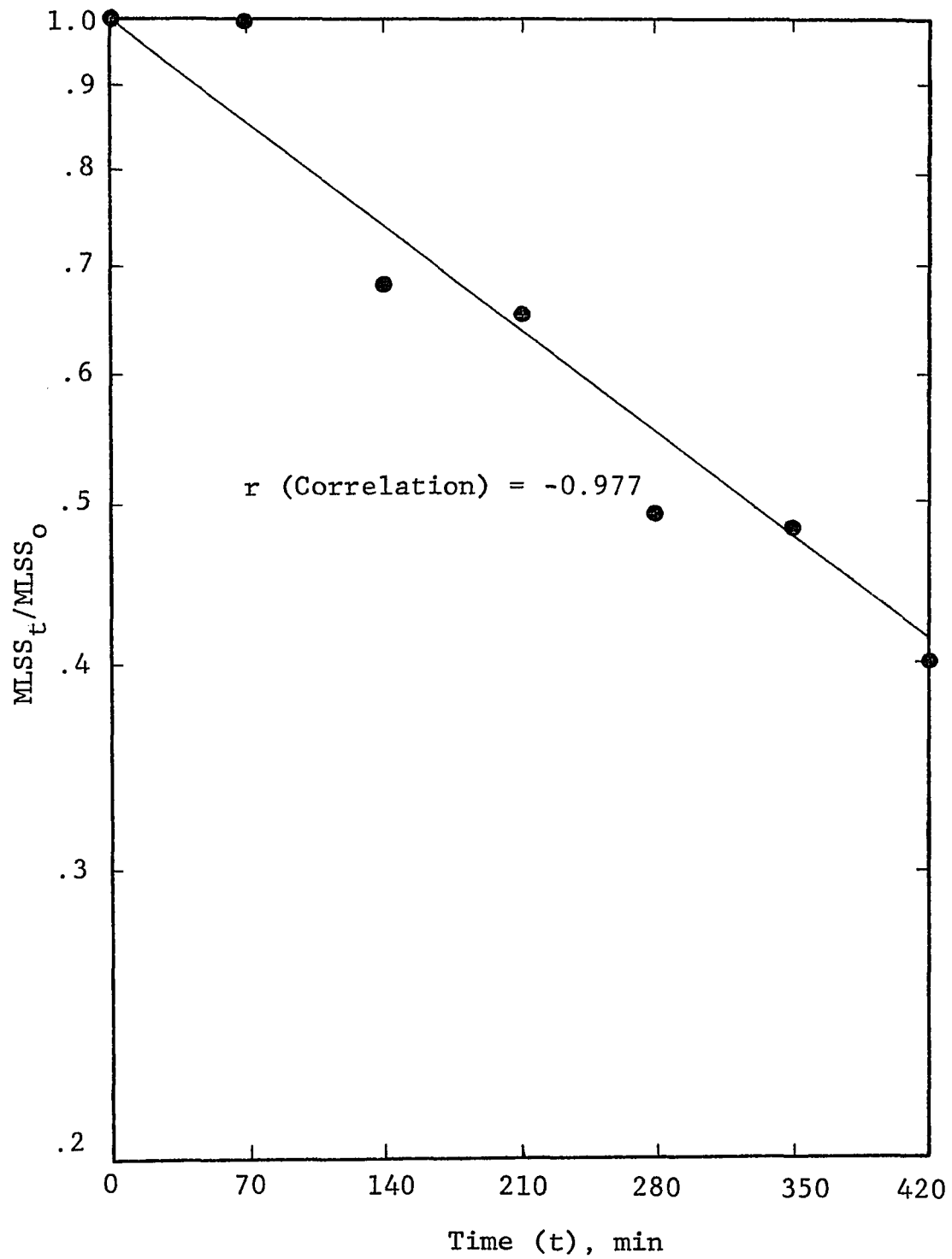


Figure 18. Molitor Plant (Data of 6/14/79)

MLSS_t/MLSS₀ vs. Time (t)

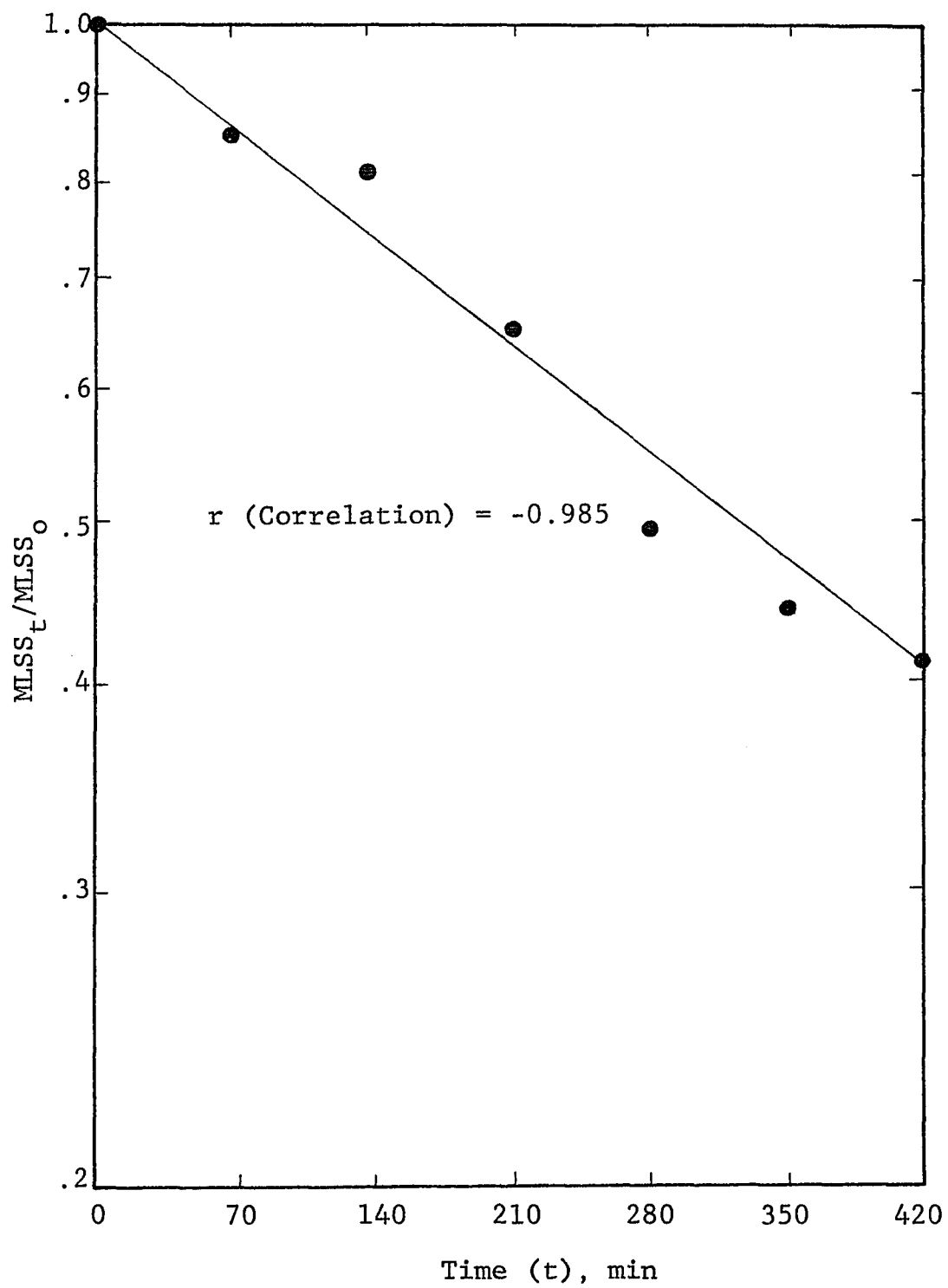


Figure 19. Molitor Plant (Data of 6/28/79)

$MLSS_t/MLSS_0$ vs. Time (t)

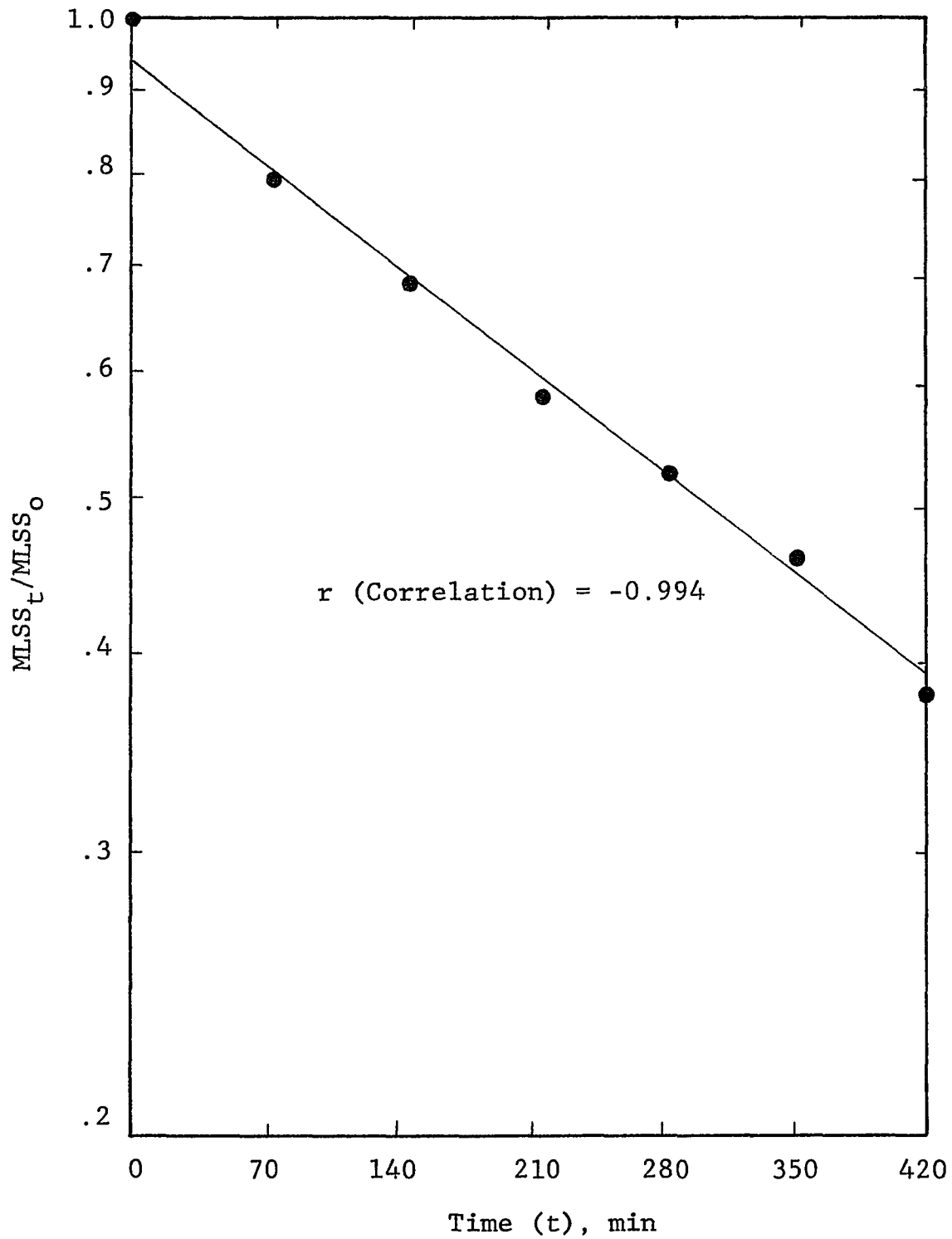


Figure 20. Molitor Plant (Data of 7/5/79)

$MLSS_t/MLSS_0$ vs. Time (t)

APPENDIX III

BOD₅ and Mixed Liquor Suspended Solids
Data for the Pilot Plant

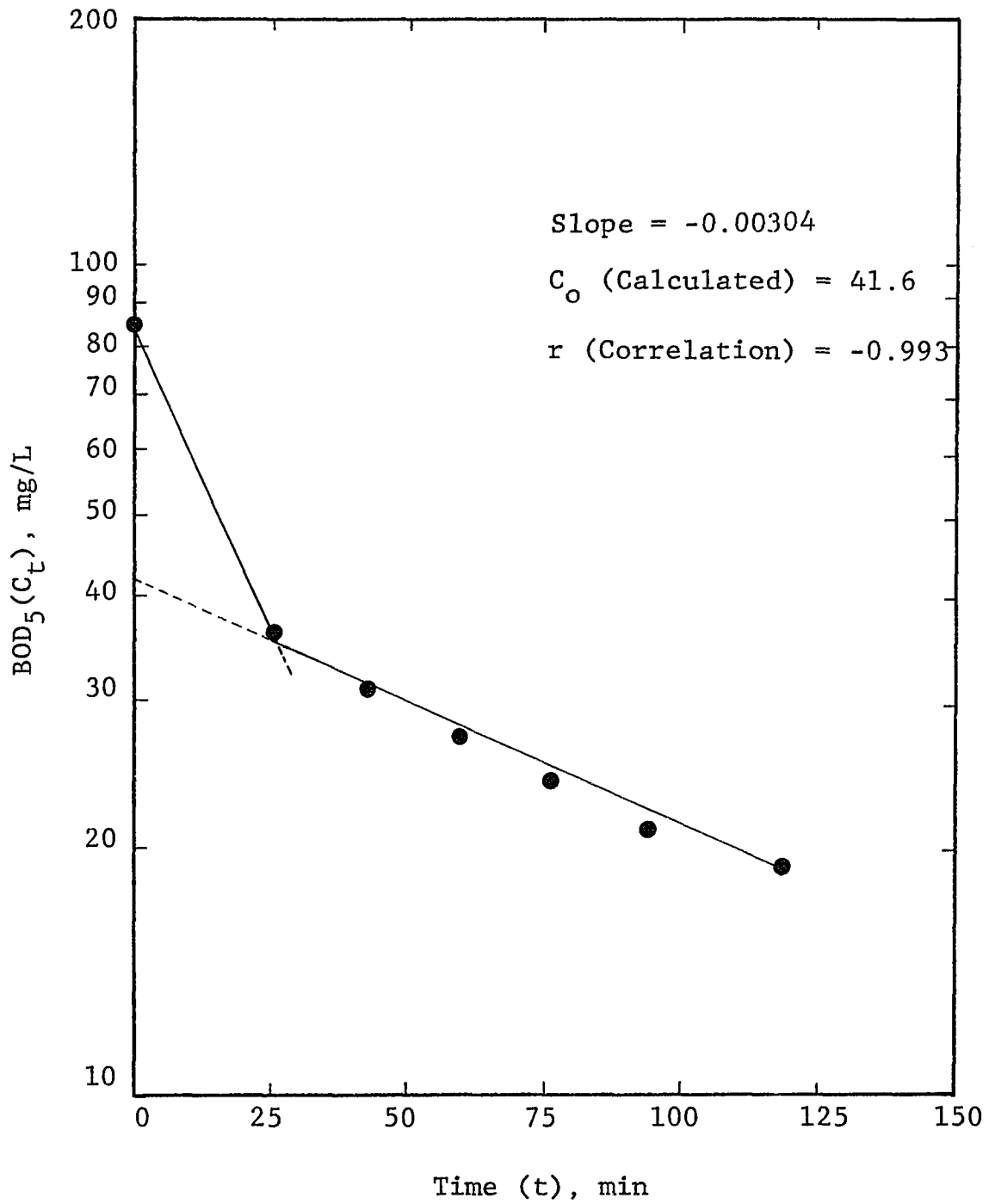


Figure 21. Pilot Plant (Data of 12/17/79)
BOD₅(C_t) vs. Time (t)

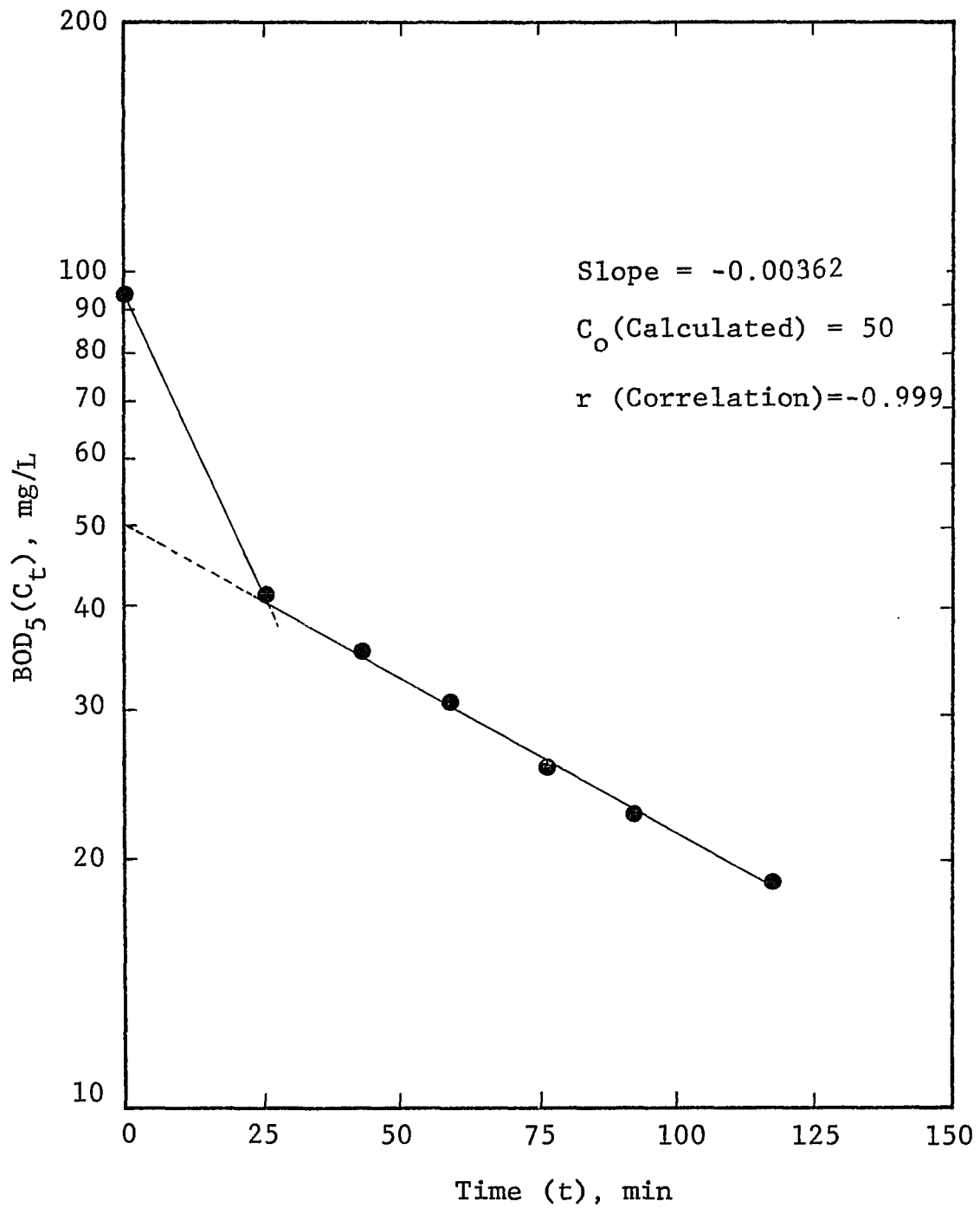


Figure 22. Pilot Plant (Data of 1/6/80)

BOD₅(C_t) vs. Time (t)

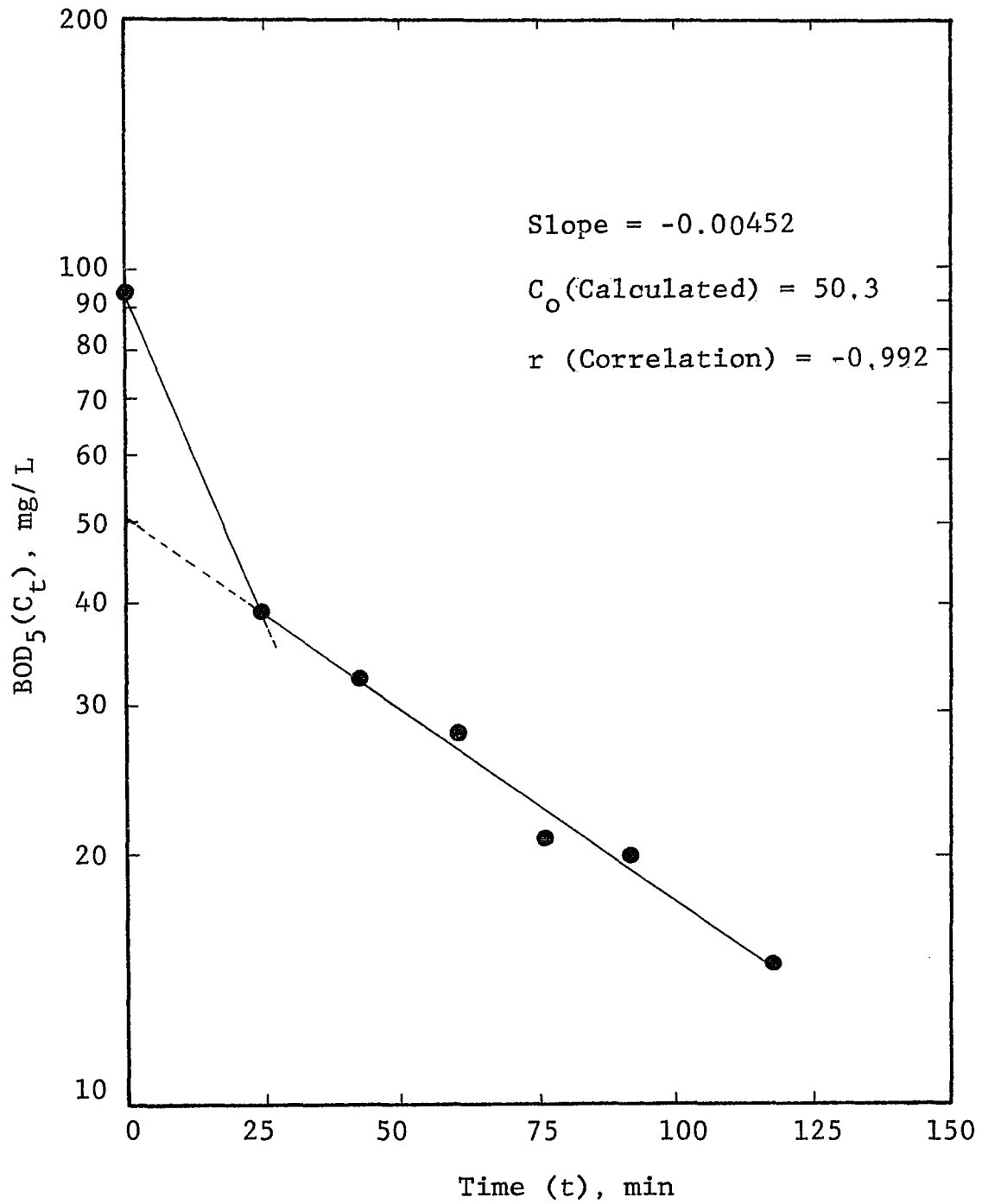


Figure 23. Pilot Plant (Data of 1/9/80)

BOD₅(C_t) vs. Time (t)

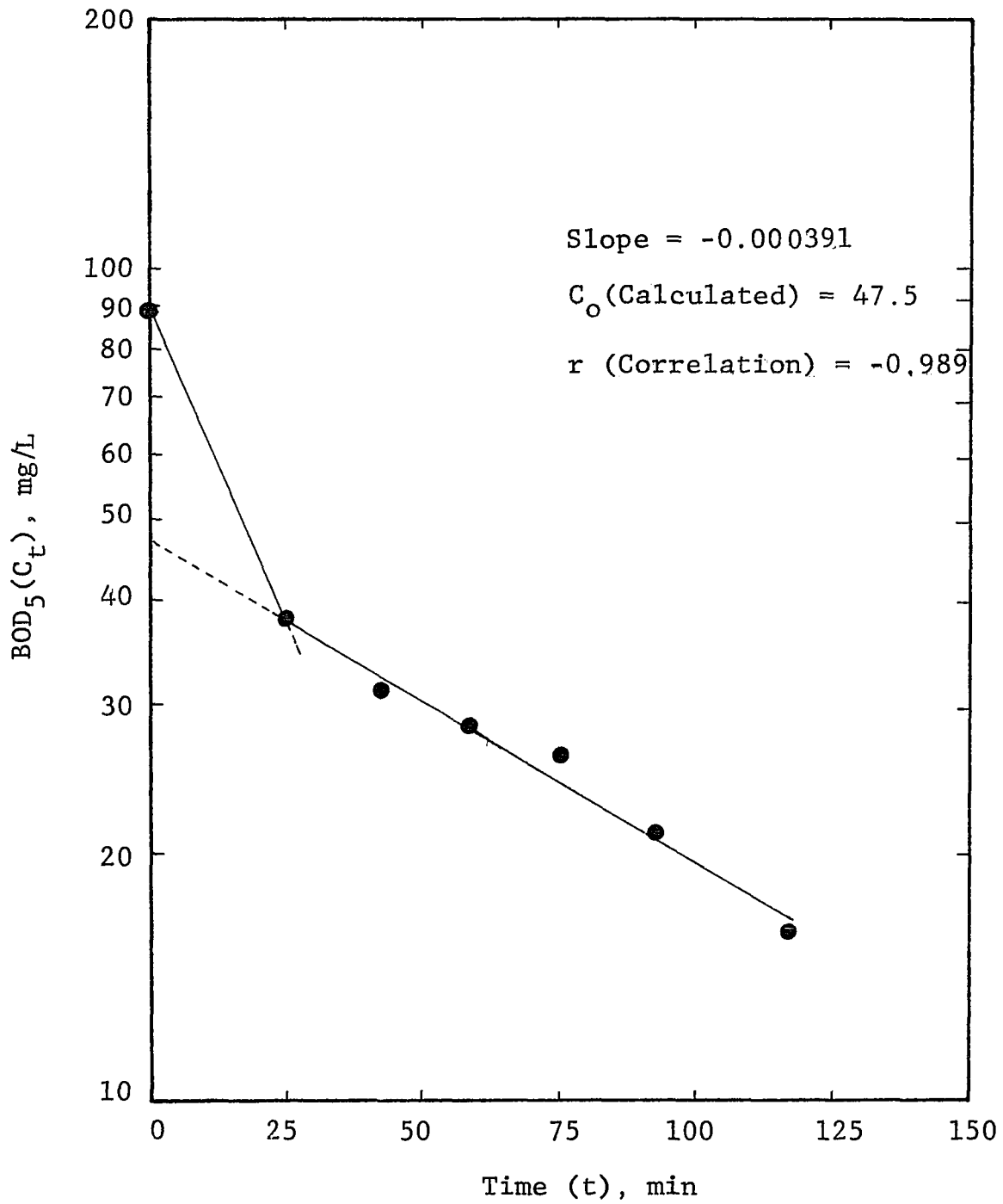


Figure 24. Pilot Plant (Data of 1/10/80)
 BOD₅(C_t) vs. Time (t)

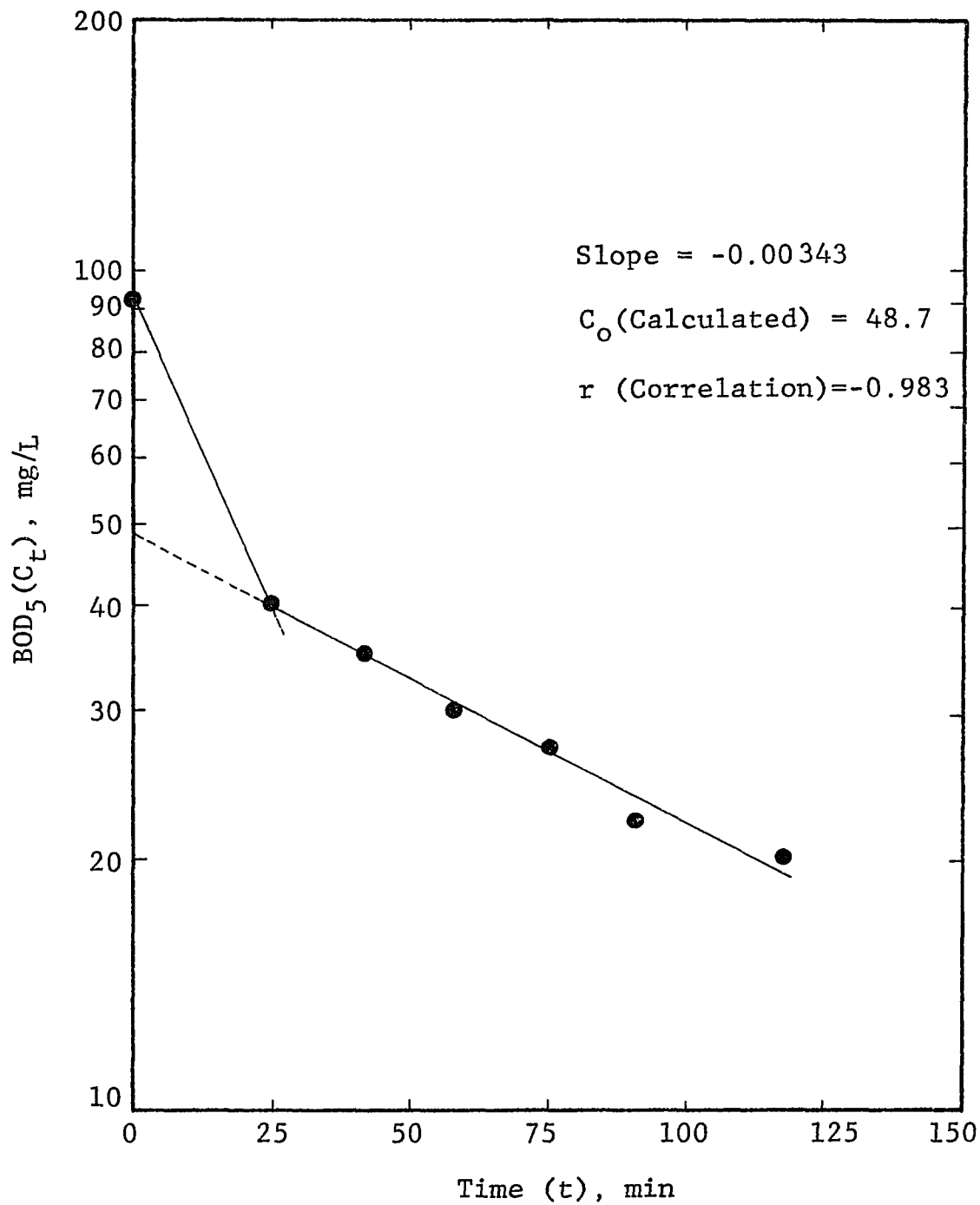


Figure 25. Pilot Plant (Data of 1/11/80)

BOD₅(C_t) vs. Time (t)

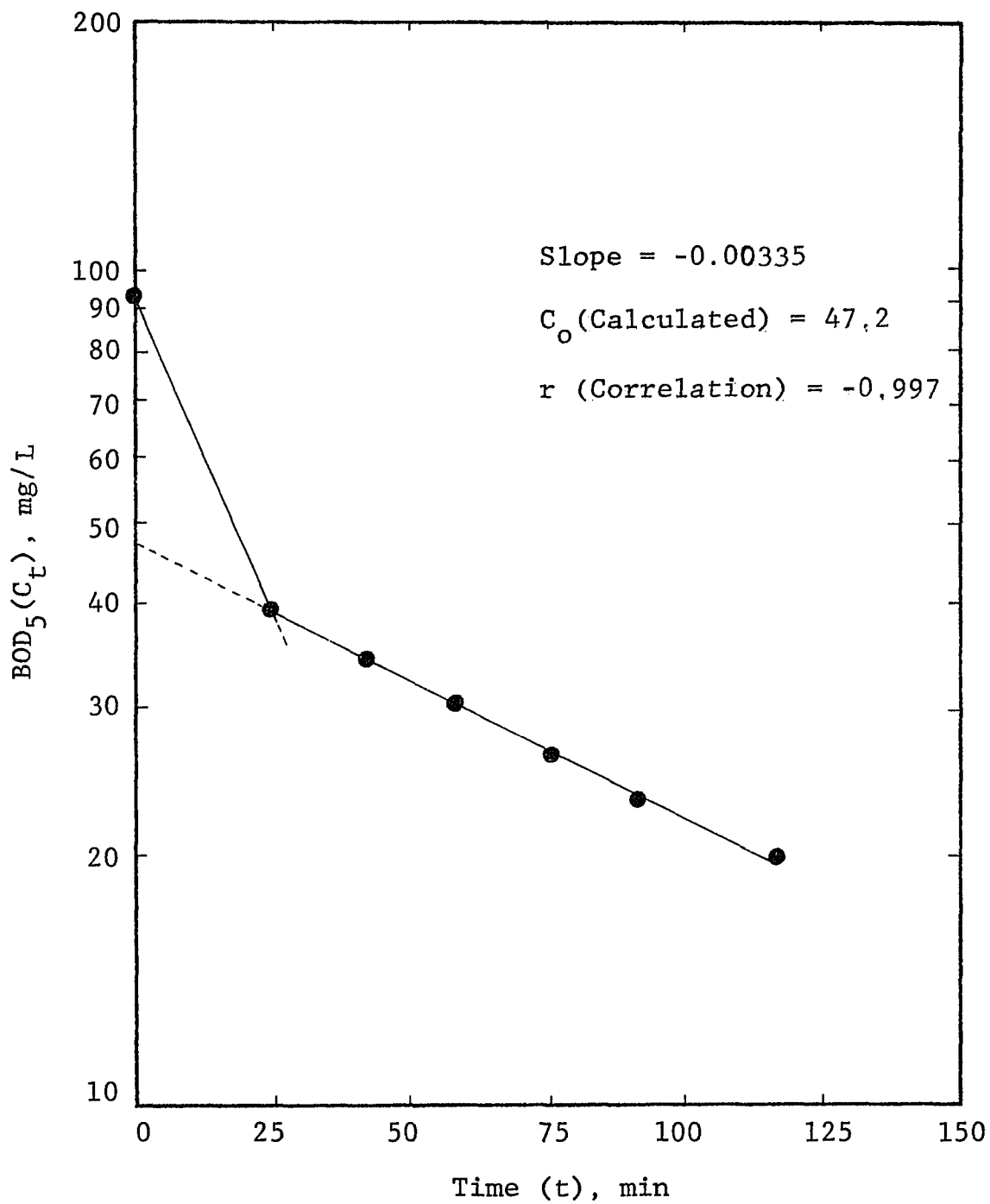


Figure 26. Pilot Plant (Data of 1/17/80)

$BOD_5(C_t)$ vs. Time (t)

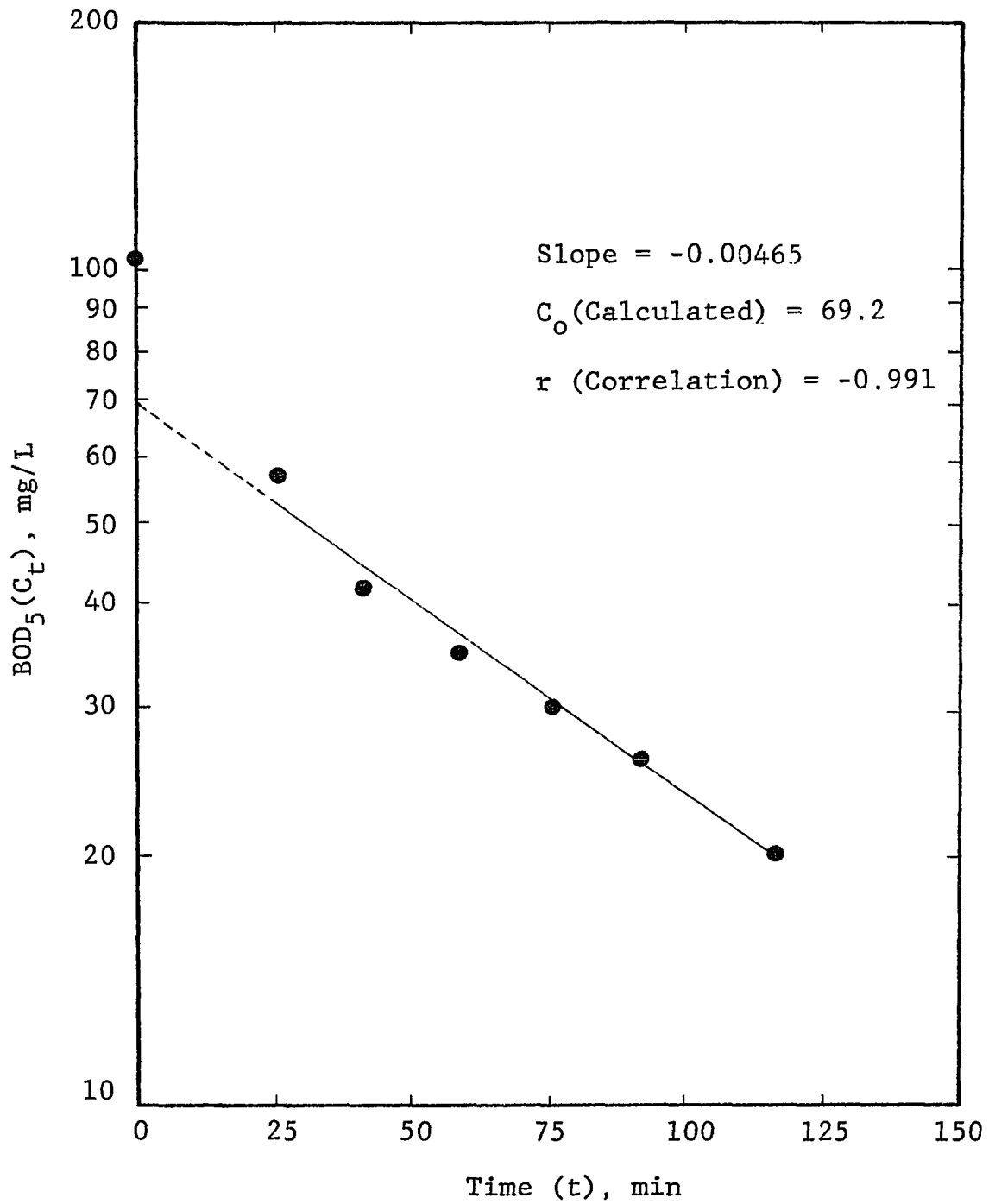


Figure 27. Pilot Plant (Data of 1/18/80)

BOD₅(C_t) vs. Time (t)

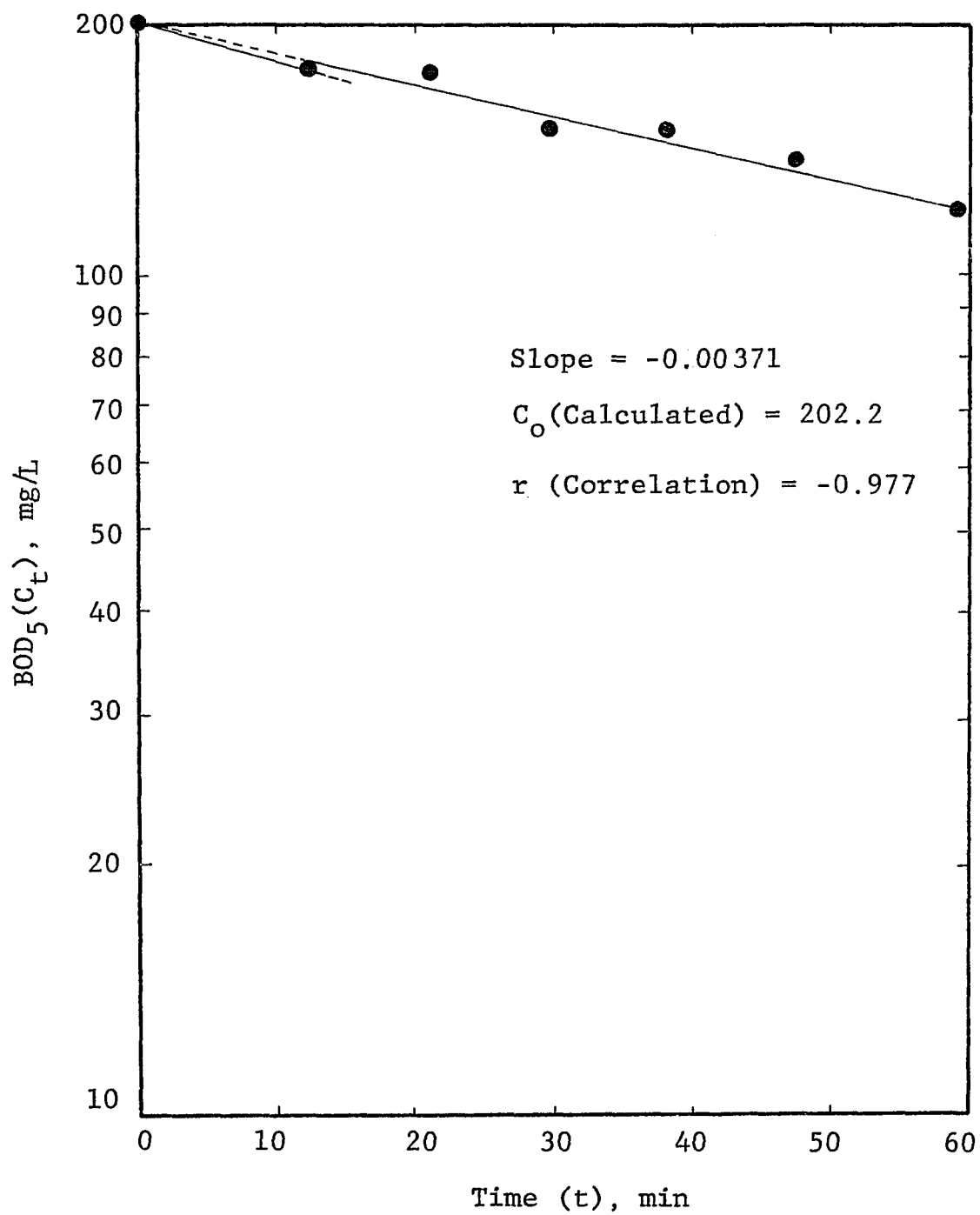


Figure 28. Pilot Plant (Data of 1/28/80)

BOD₅(C_t) vs. Time (t)

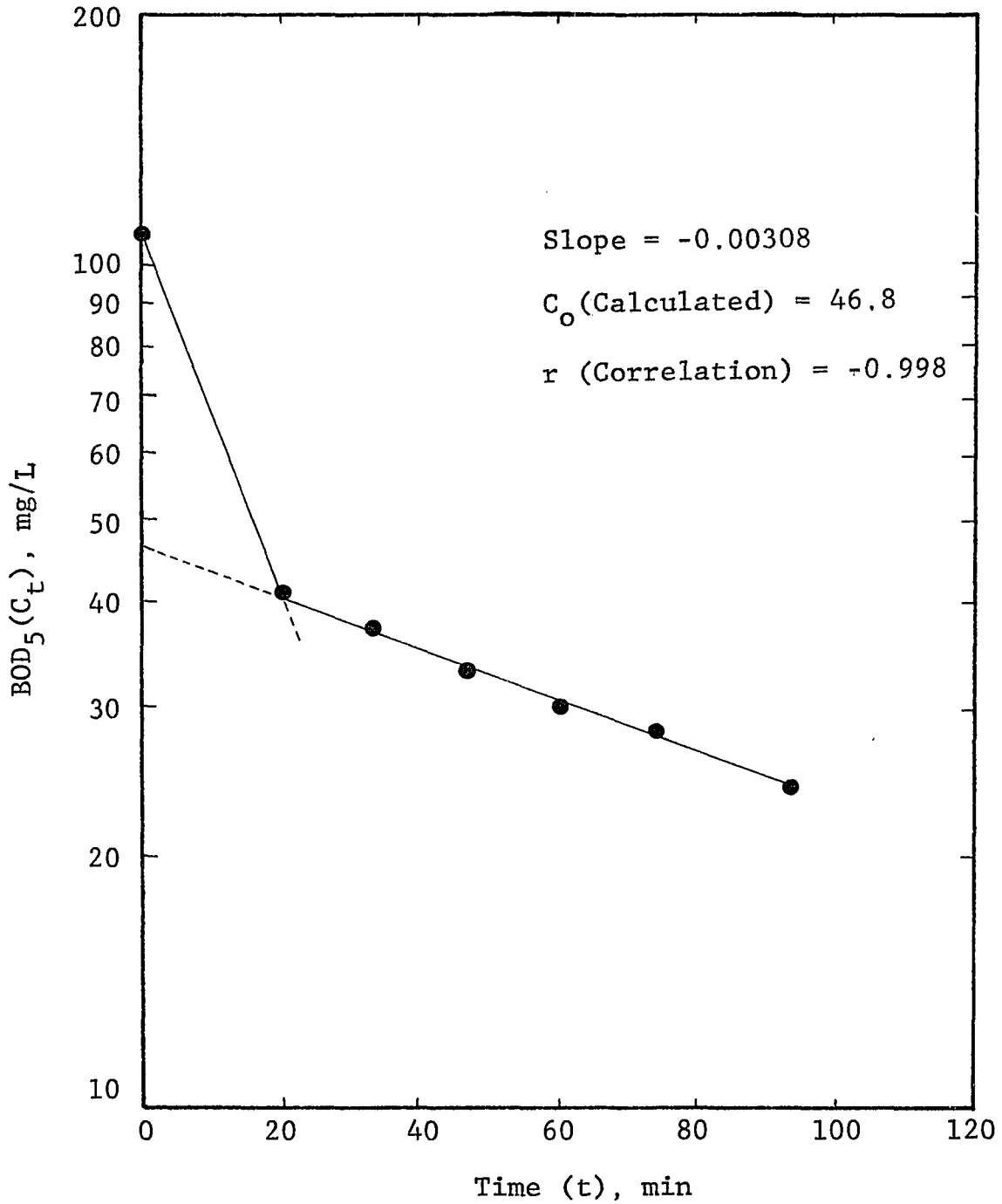


Figure 29. Pilot Plant (Data of 1/29/80)

BOD₅(C_t) vs. Time (t)

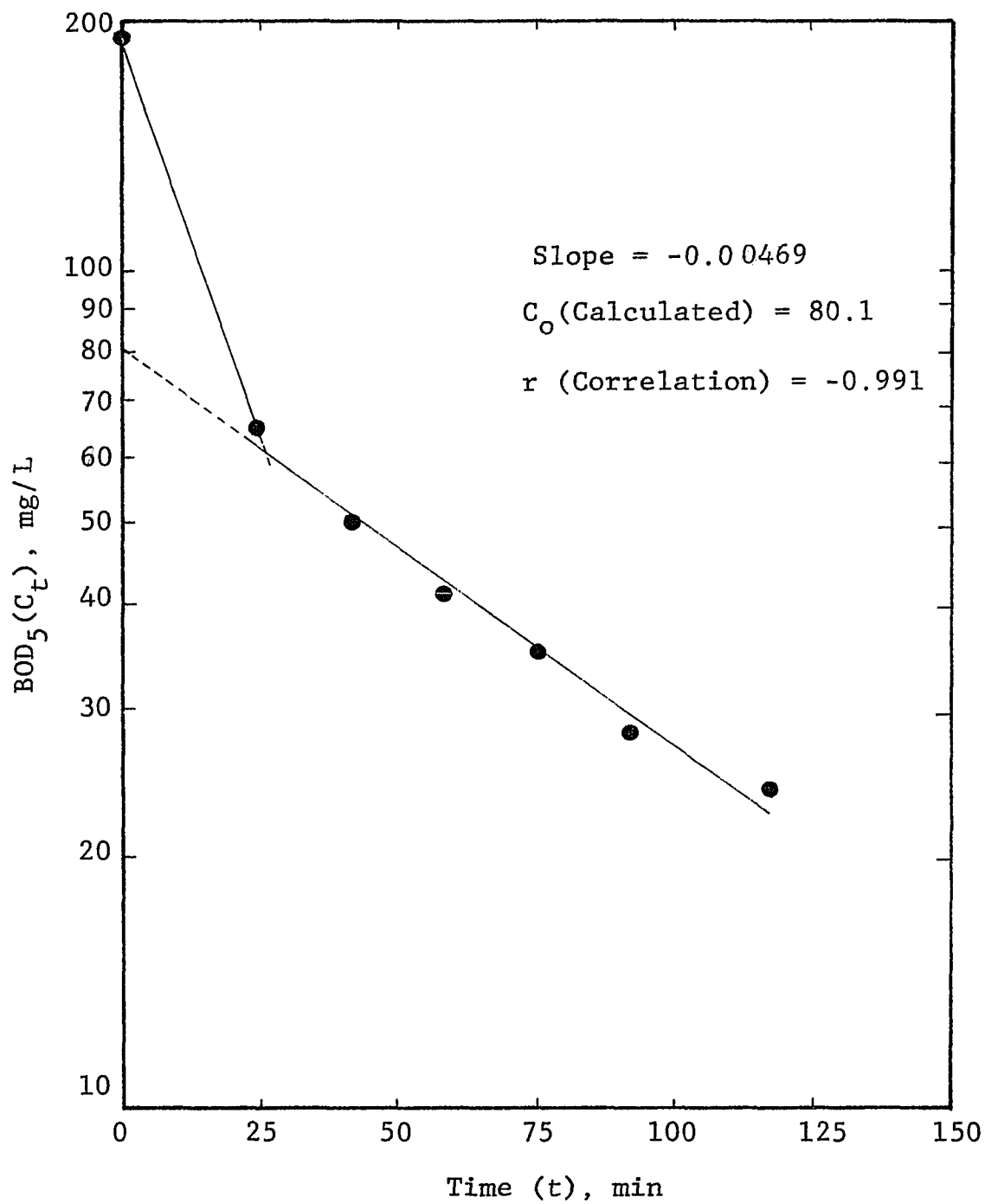


Figure 30. Pilot Plant (Data of 1/30/80)

$BOD_5(C_t)$ vs. Time (t)

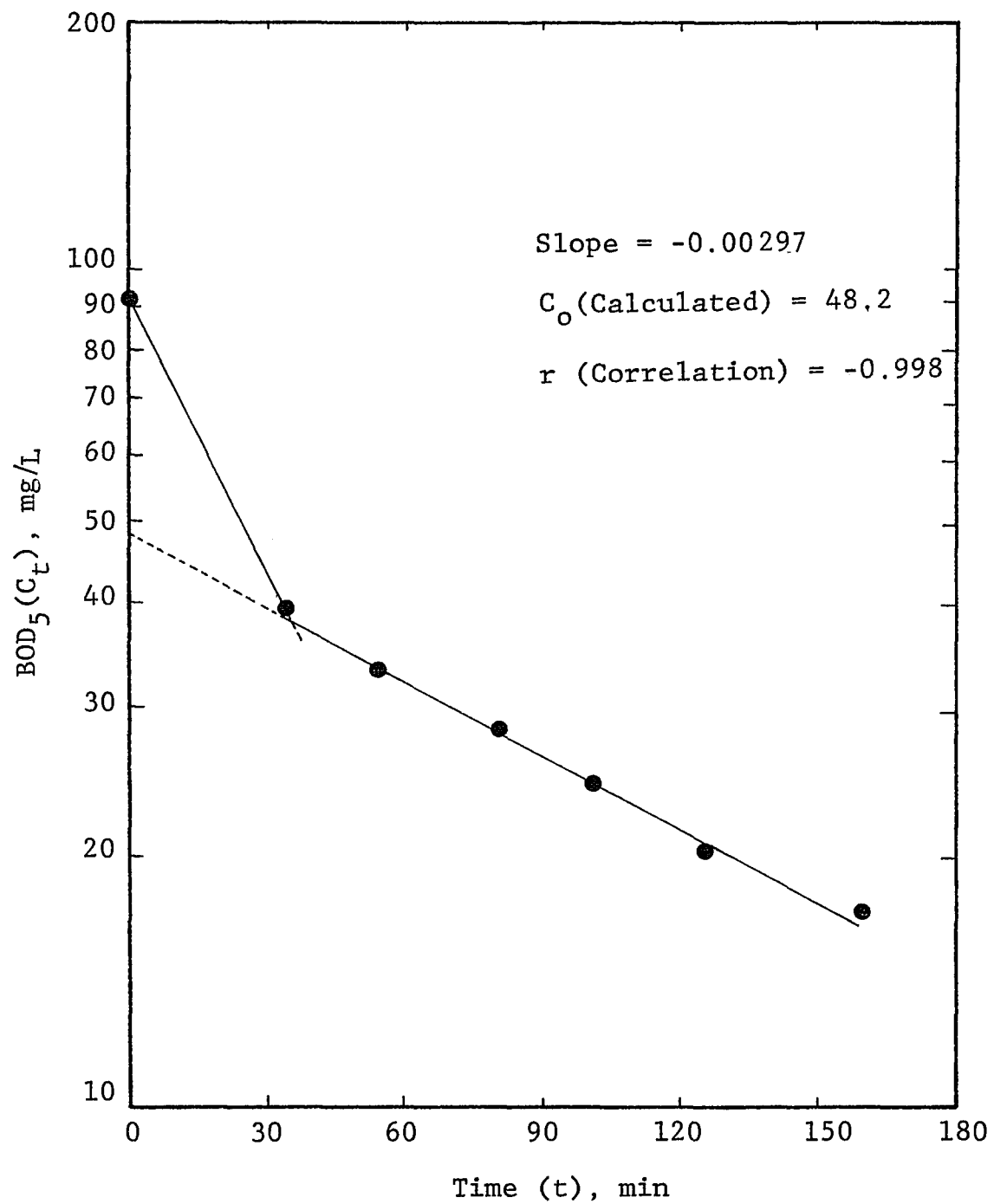


Figure 31. Pilot Plant (Data of 1/31/80)
BOD₅(C_t) vs. Time (t)

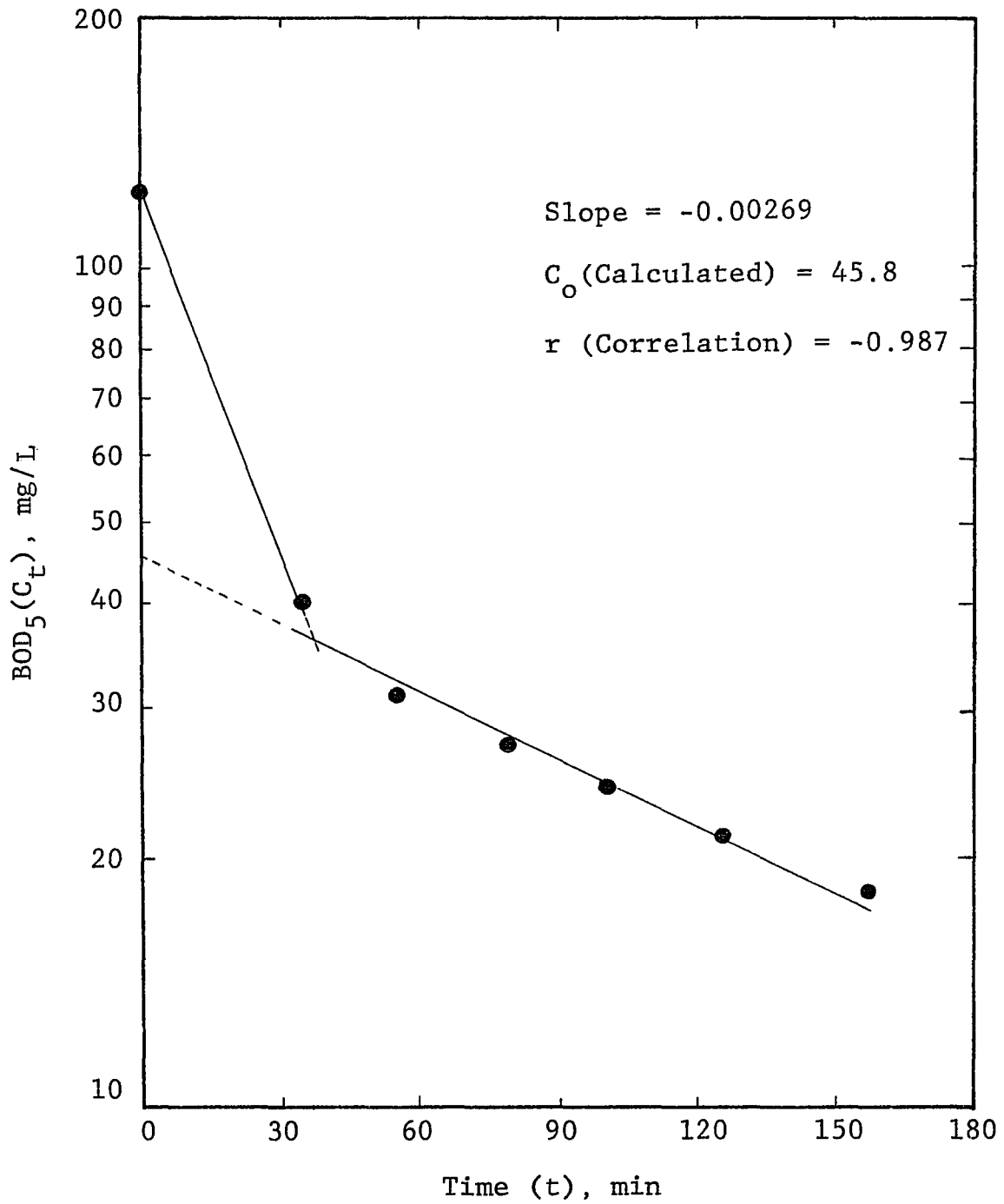


Figure 32. Pilot Plant (Data of 2/1/80)

BOD₅(C_t) vs. Time (t)

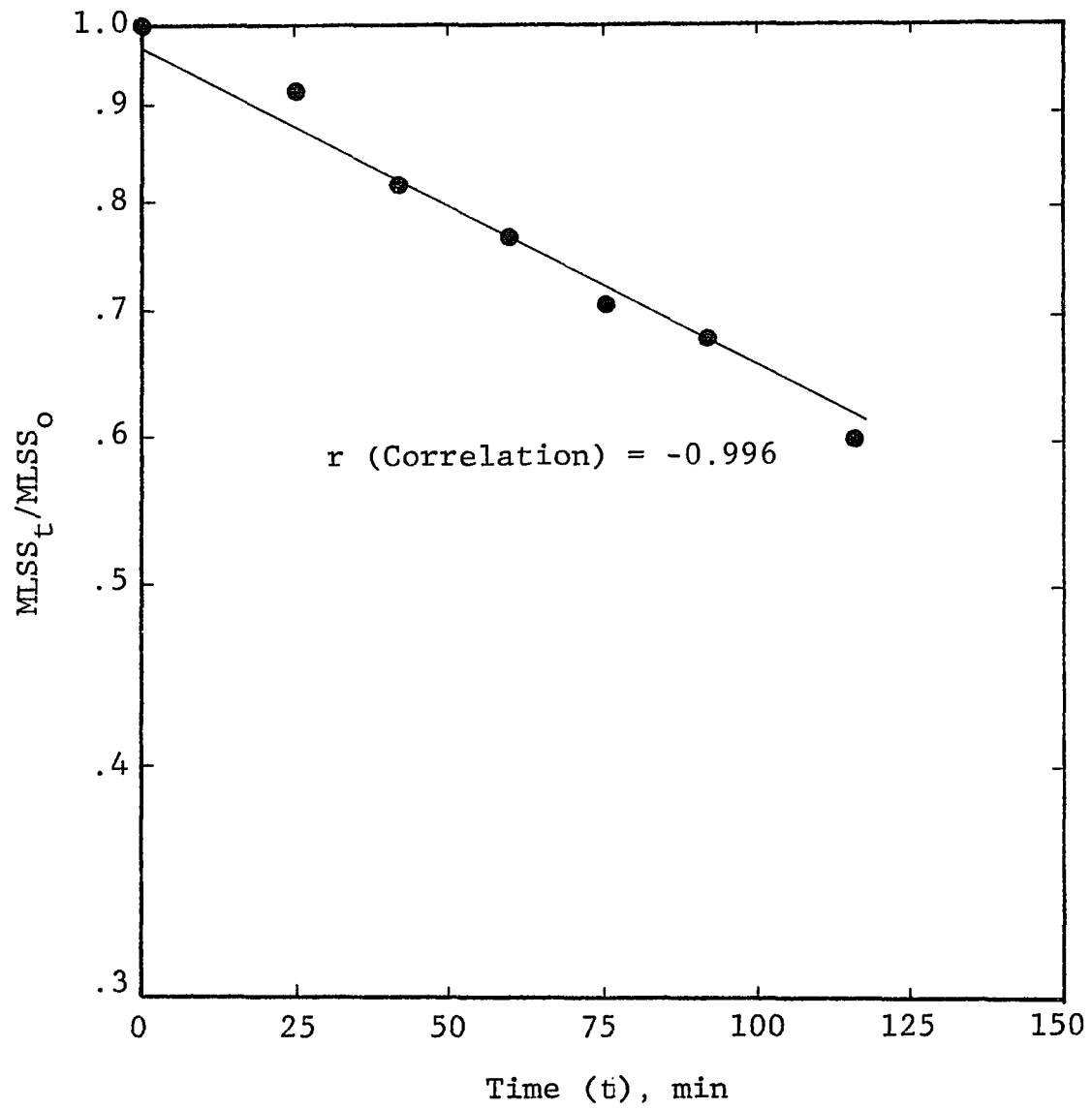


Figure 33. Pilot Plant (Data of 12/17/79)

$MLSS_t/MLSS_0$ vs. Time (t)

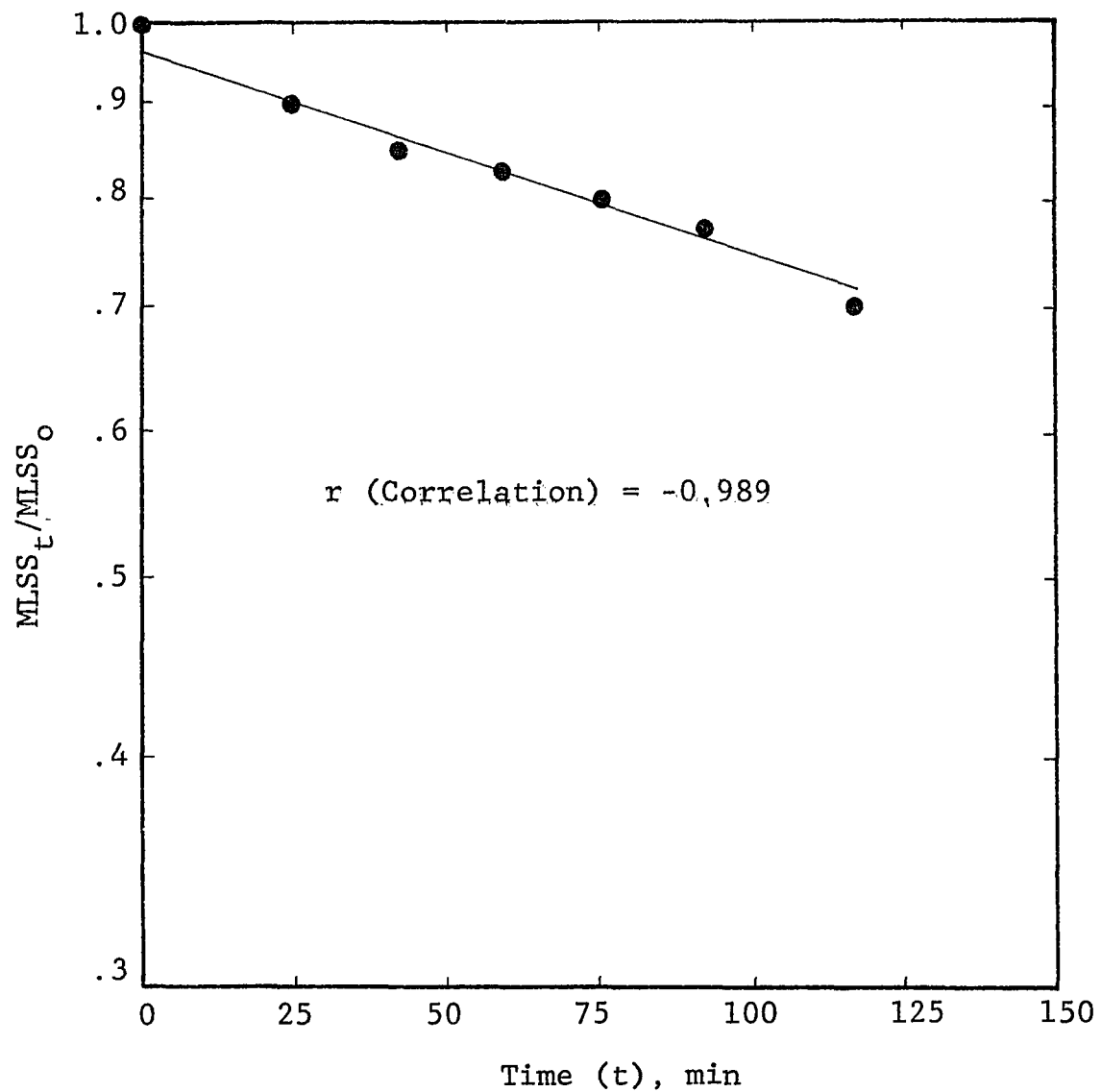


Figure 34. Pilot Plant (Data of 1/6/80)

$MLSS_t/MLSS_0$ vs. Time (t)

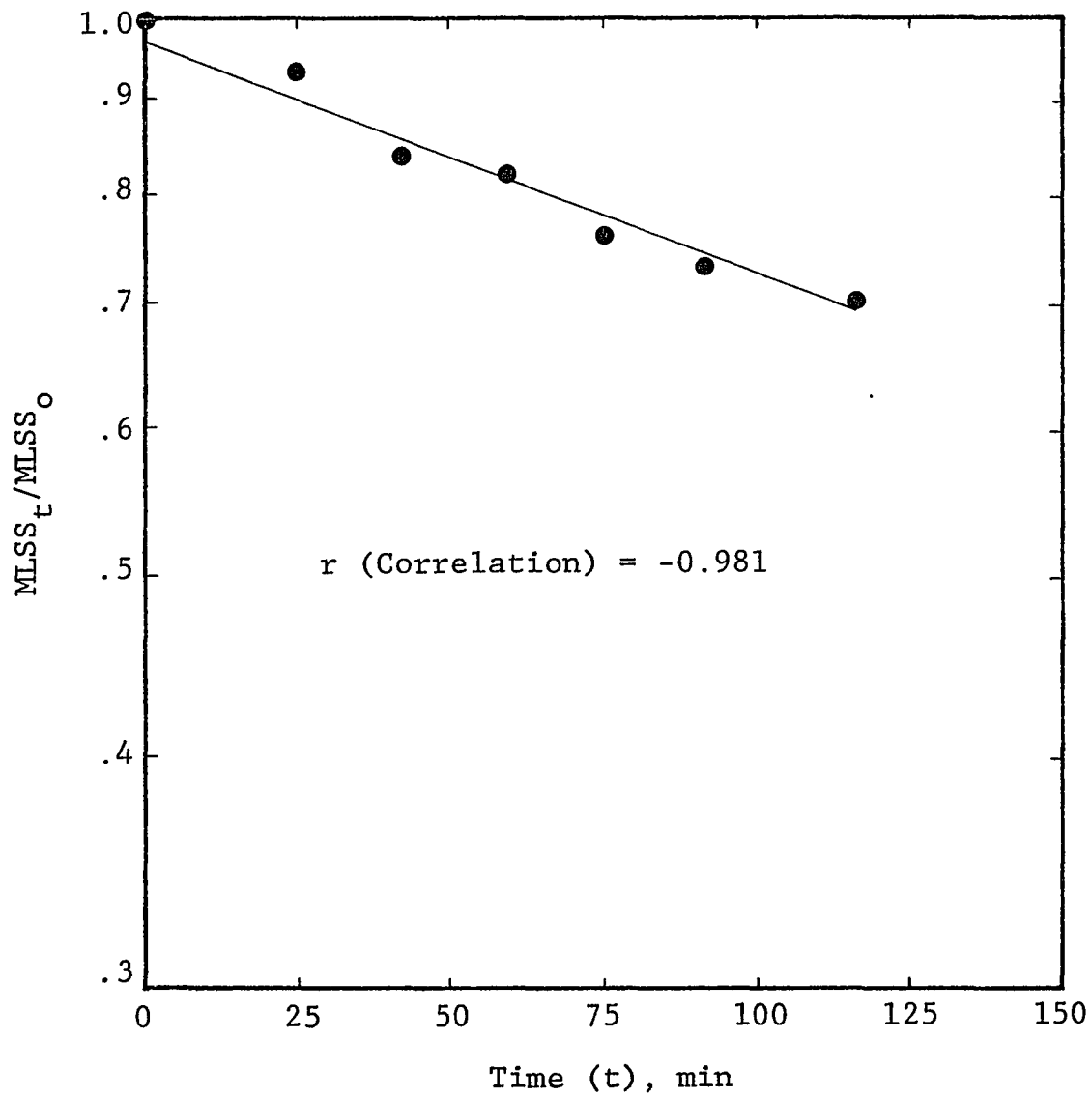


Figure 35. Pilot Plant (Data of 1/9/80)

$MLSS_t/MLSS_0$ vs. Time (t)

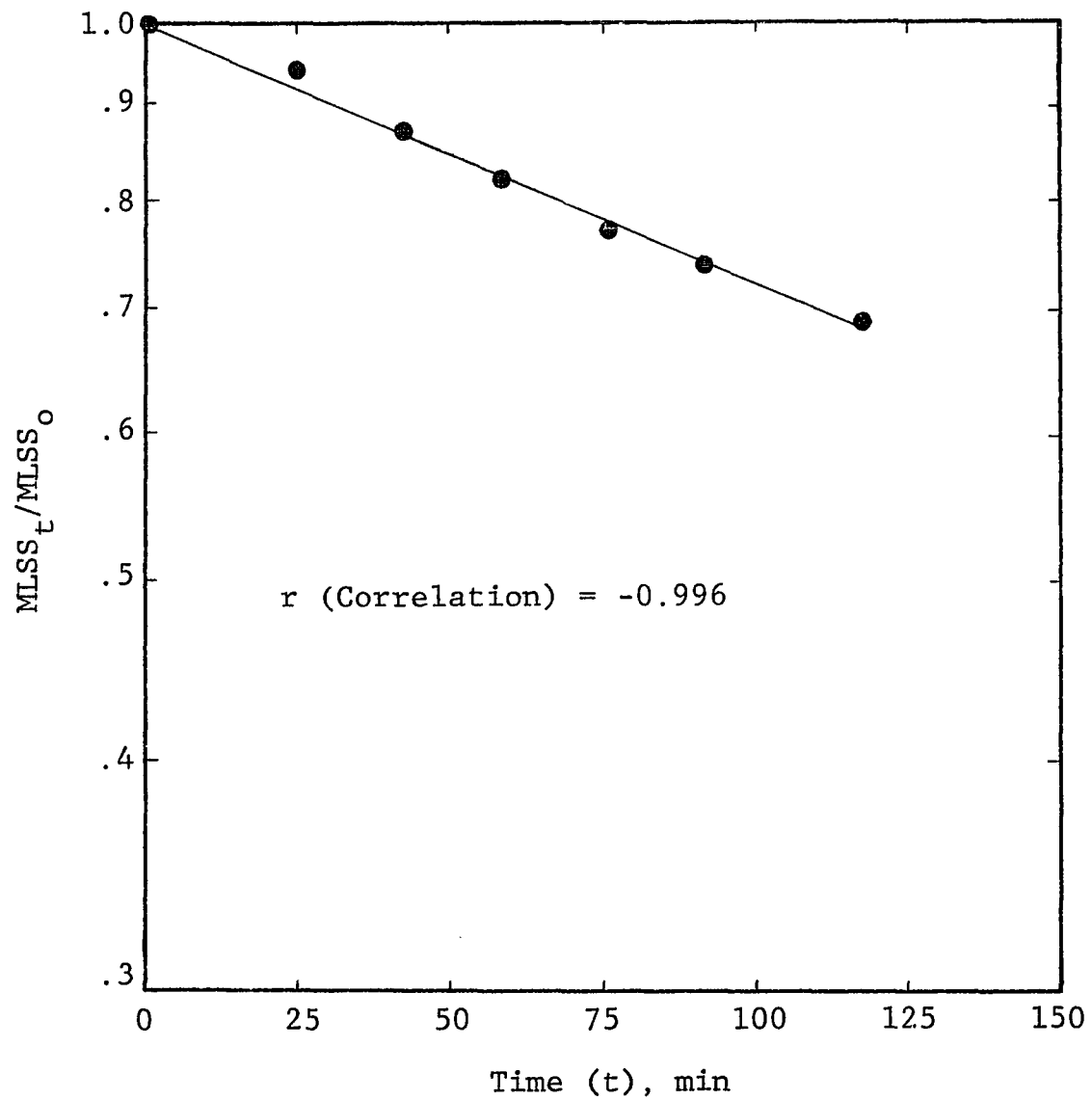


Figure 36. Pilot Plant (Data of 1/10/80)

$MLSS_t/MLSS_0$ vs. Time (t)

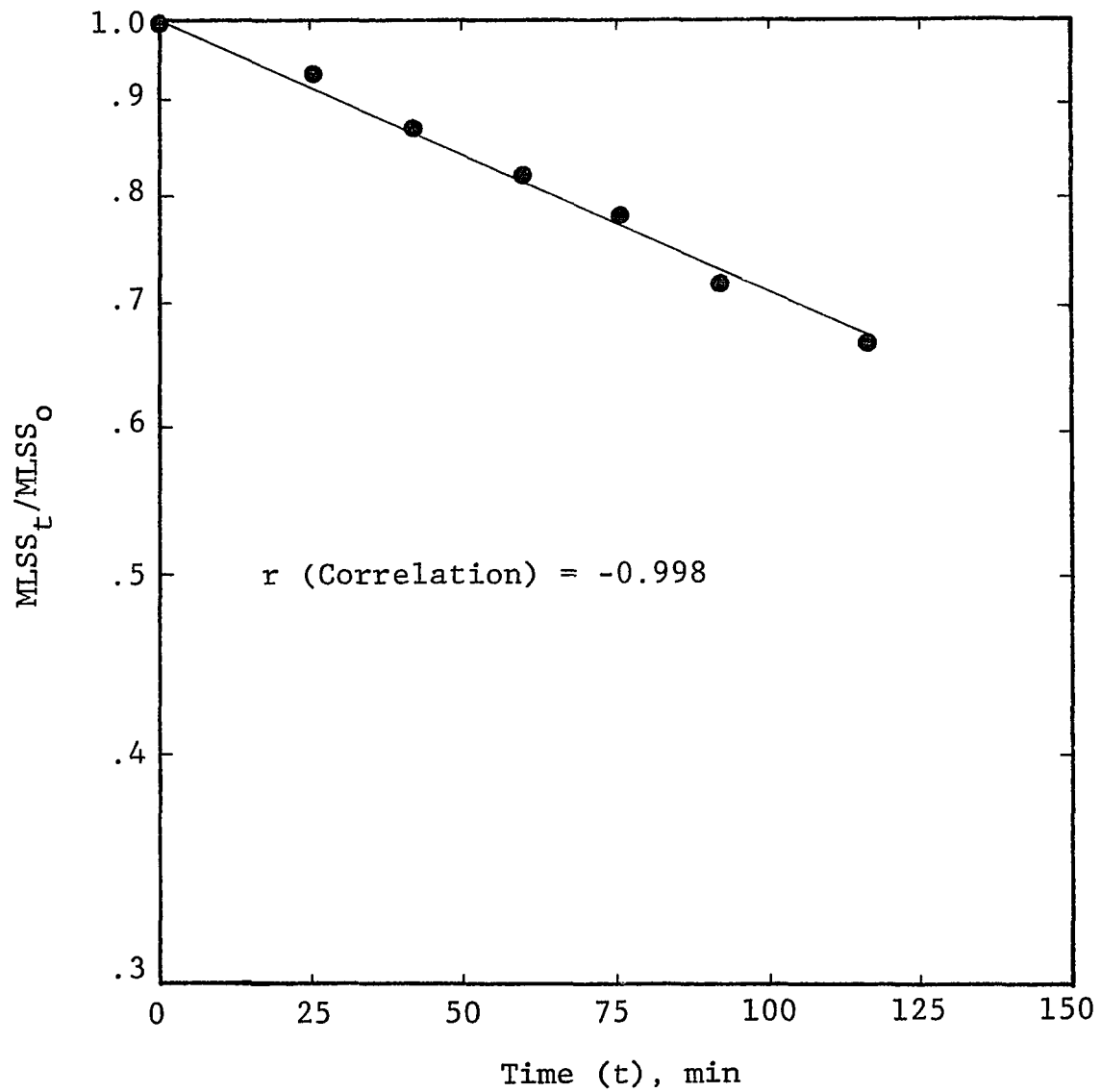


Figure 37. Pilot Plant (Data of 1/11/80)

$MLSS_t/MLSS_0$ vs. Time (t)

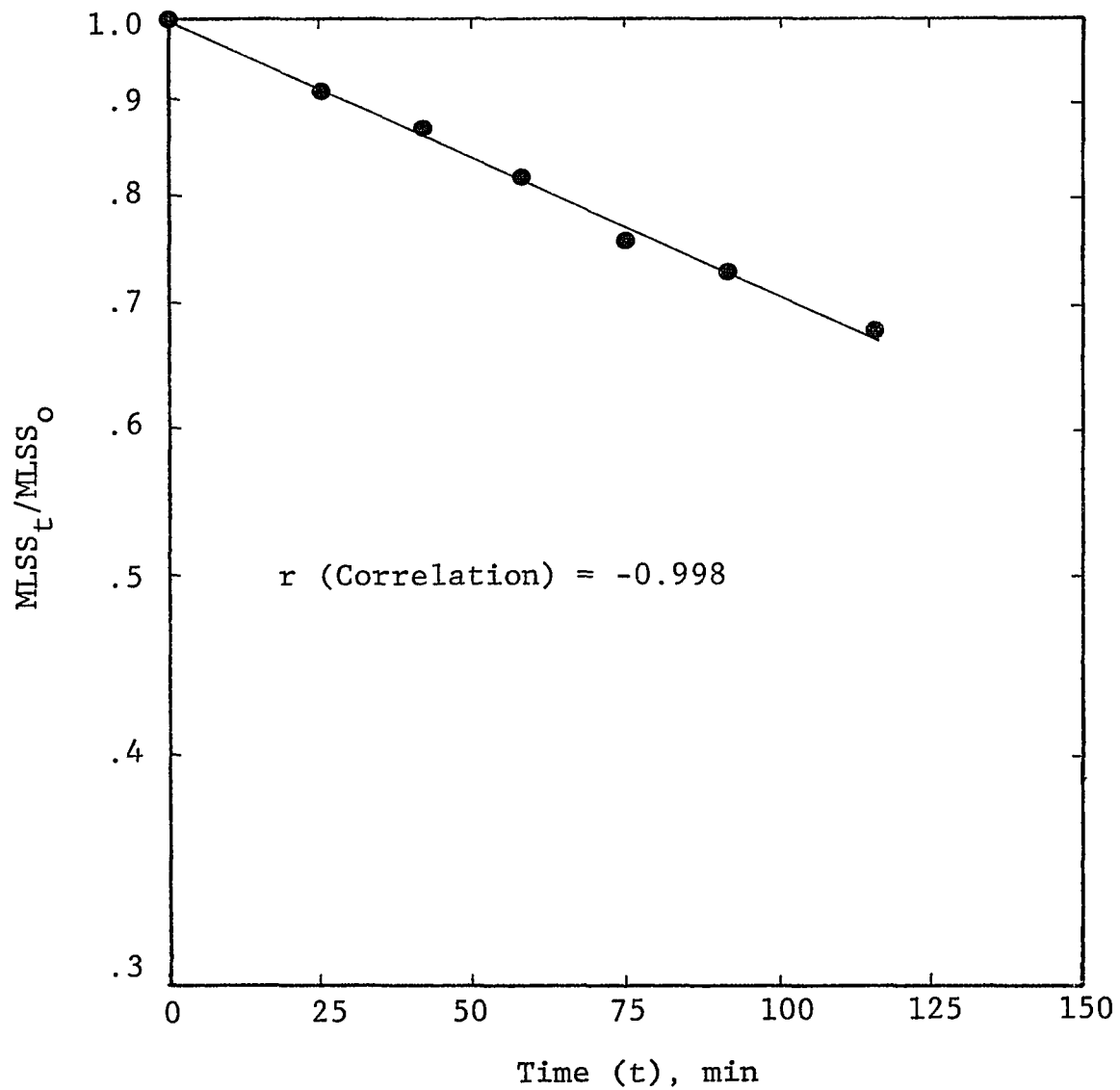


Figure 38. Pilot Plant (Data of 1/17/80)

$MLSS_t/MLSS_0$ vs. Time (t)

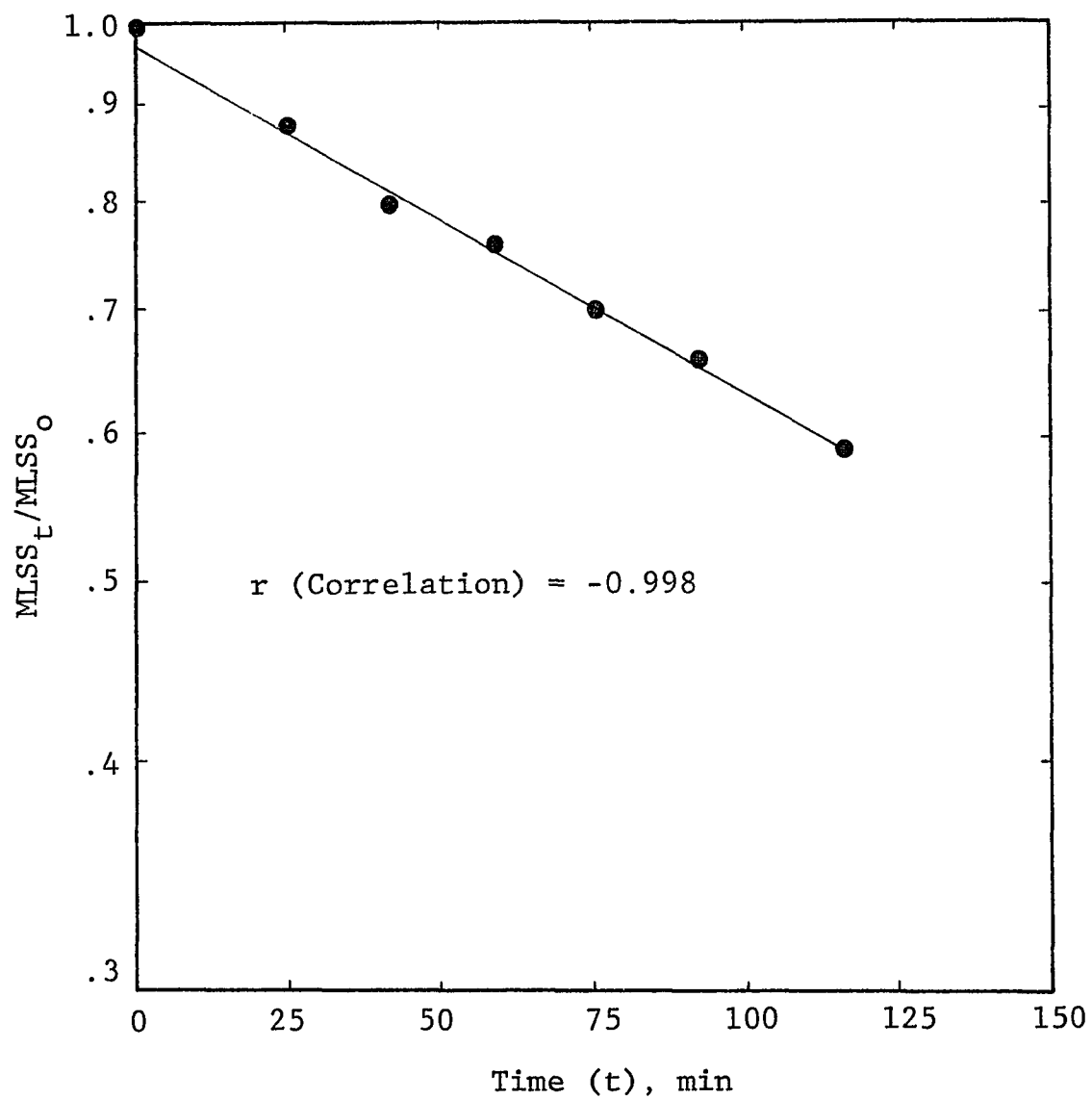


Figure 39. Pilot Plant (Data of 1/18/80)

$MLSS_t/MLSS_0$ vs. Time (t)

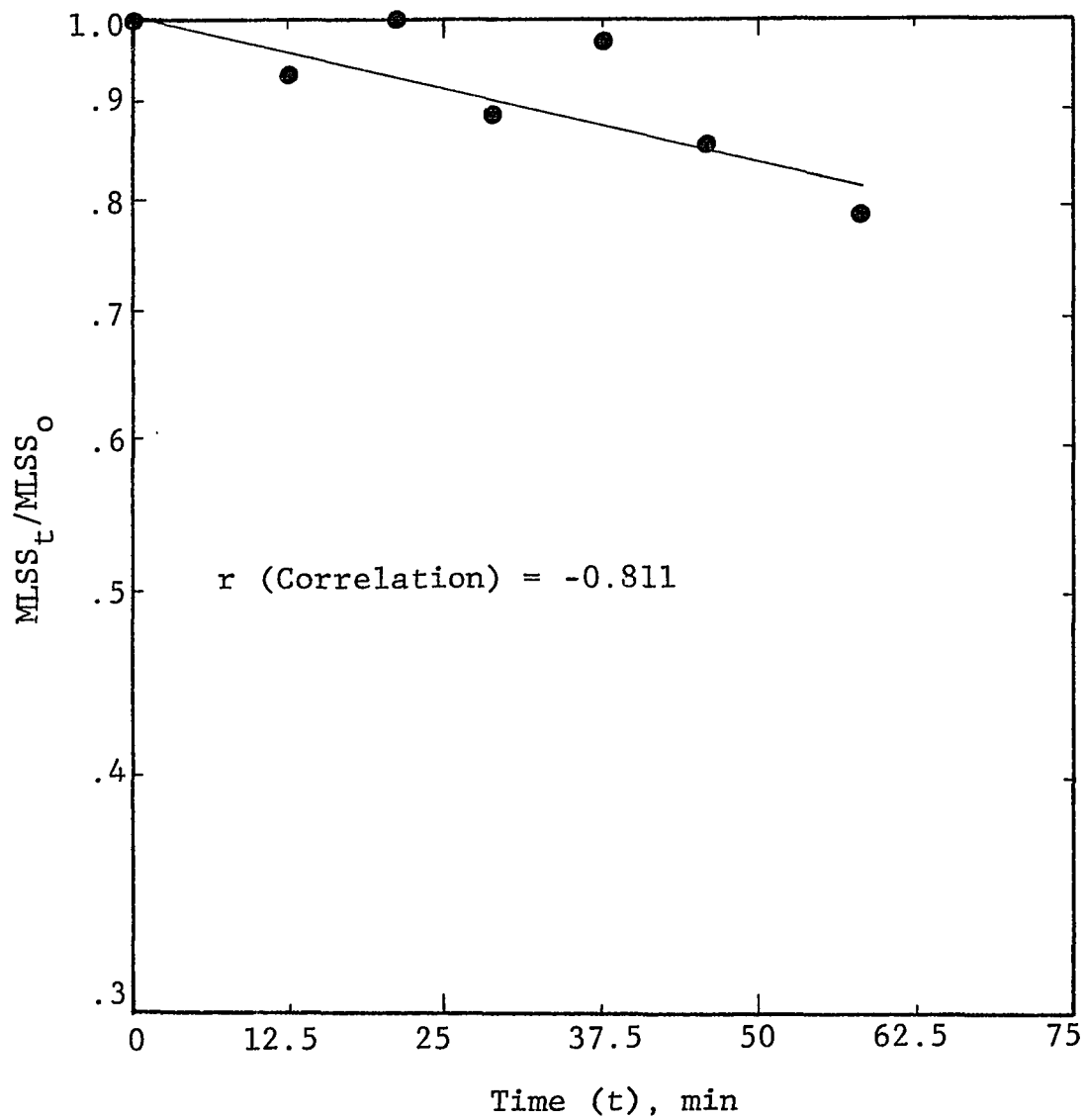


Figure 40. Pilot Plant (Data of 1/28/80)

$MLSS_t/MLSS_0$ vs. Time (t)

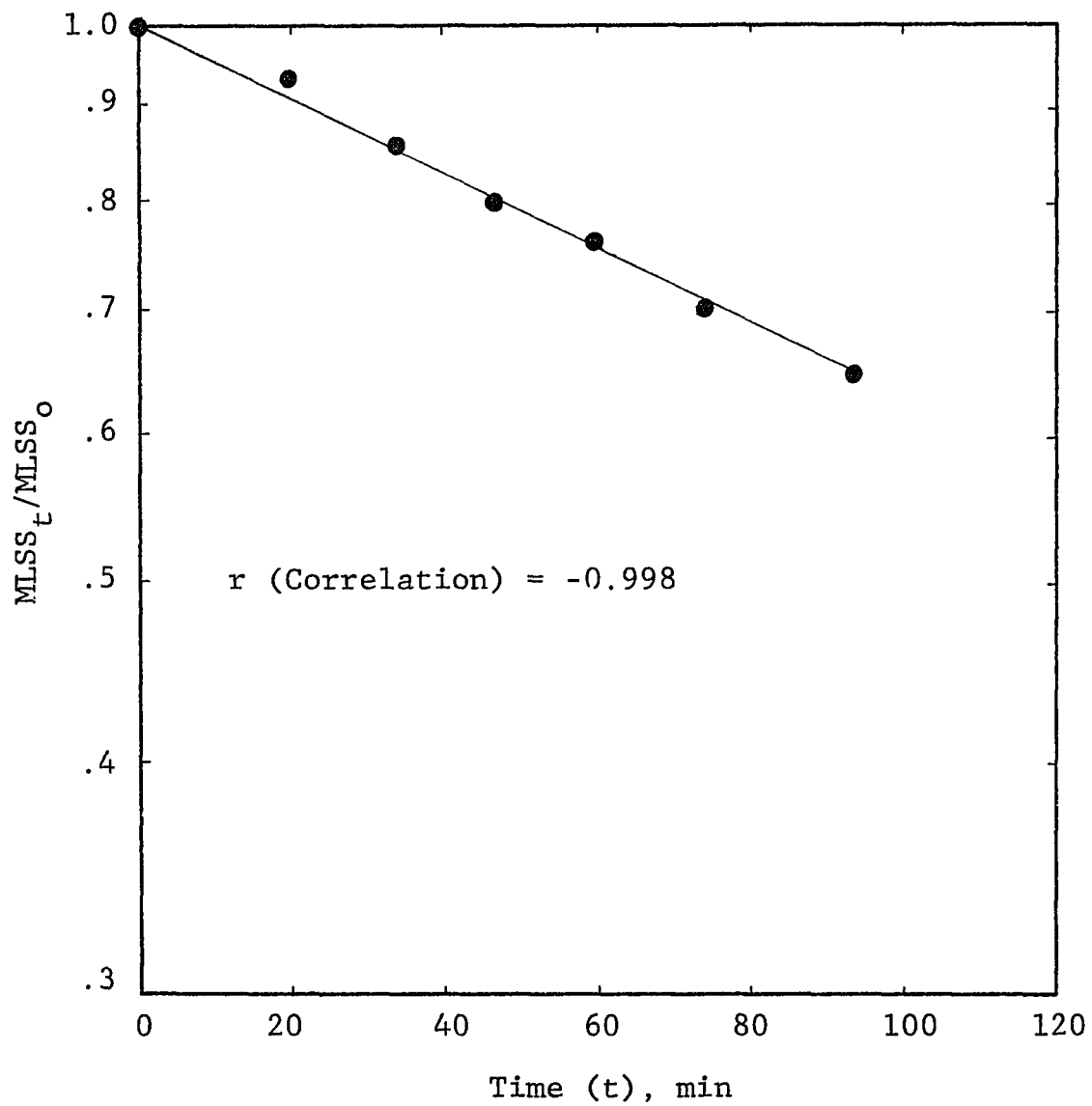


Figure 41. Pilot Plant (Data of 1/29/80)

MLSS_t/MLSS₀ vs. Time (t)

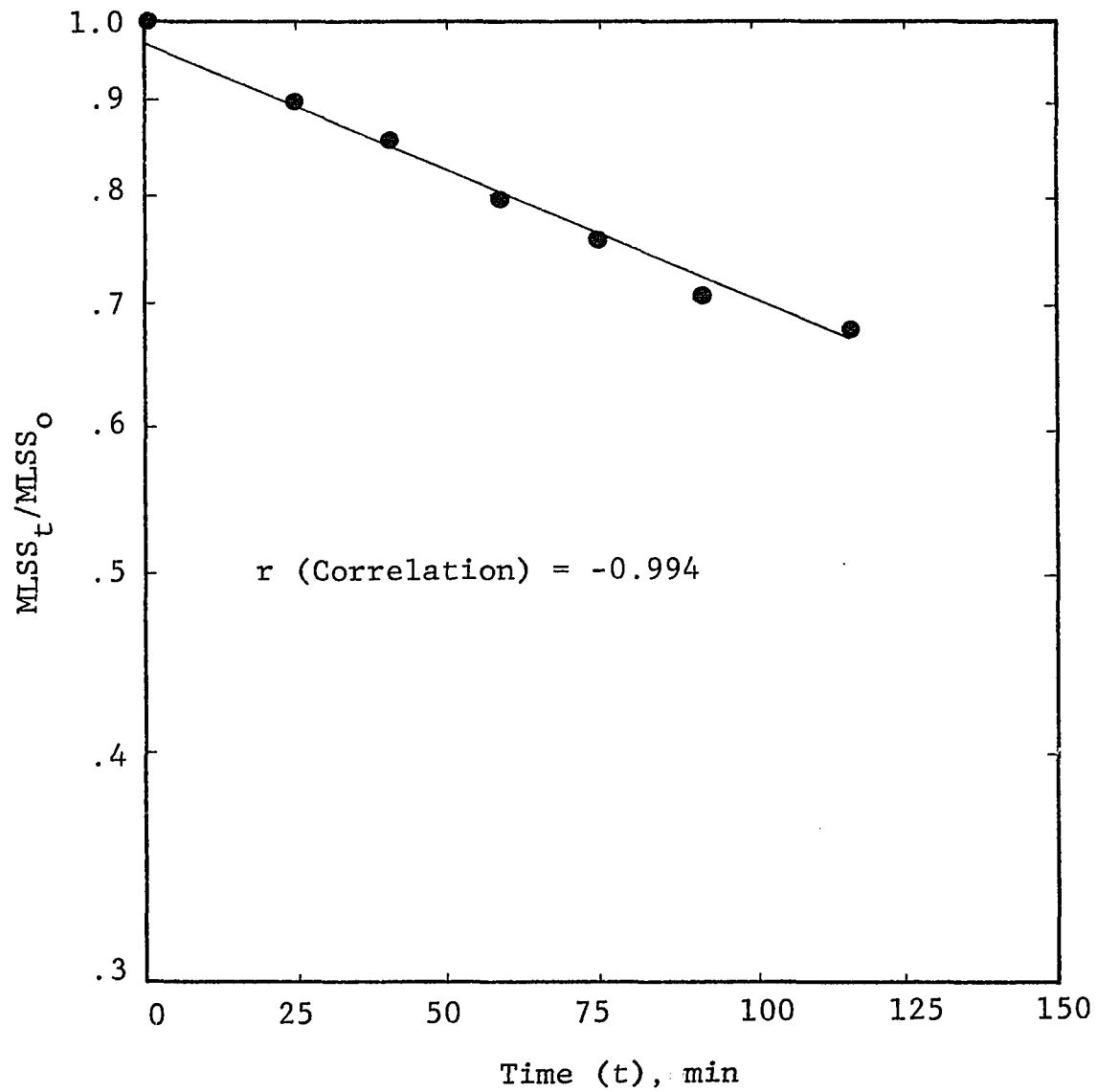


Figure 42. Pilot Plant (Data of 1/30/80)

$MLSS_t/MLSS_0$ vs. Time (t)

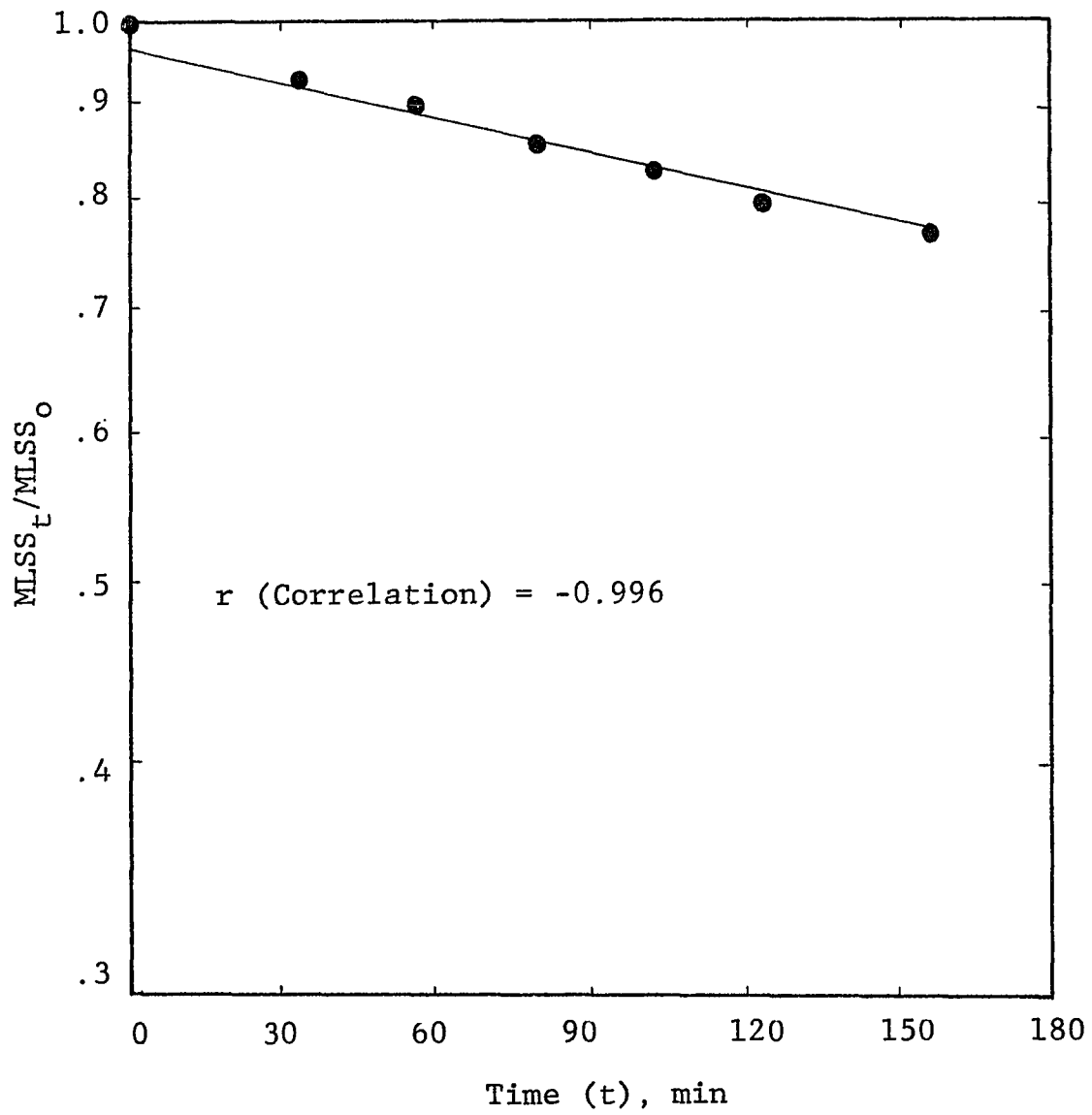


Figure 43. Pilot Plant (Data of 1/31/80)

$MLSS_t/MLSS_0$ vs. Time (t)

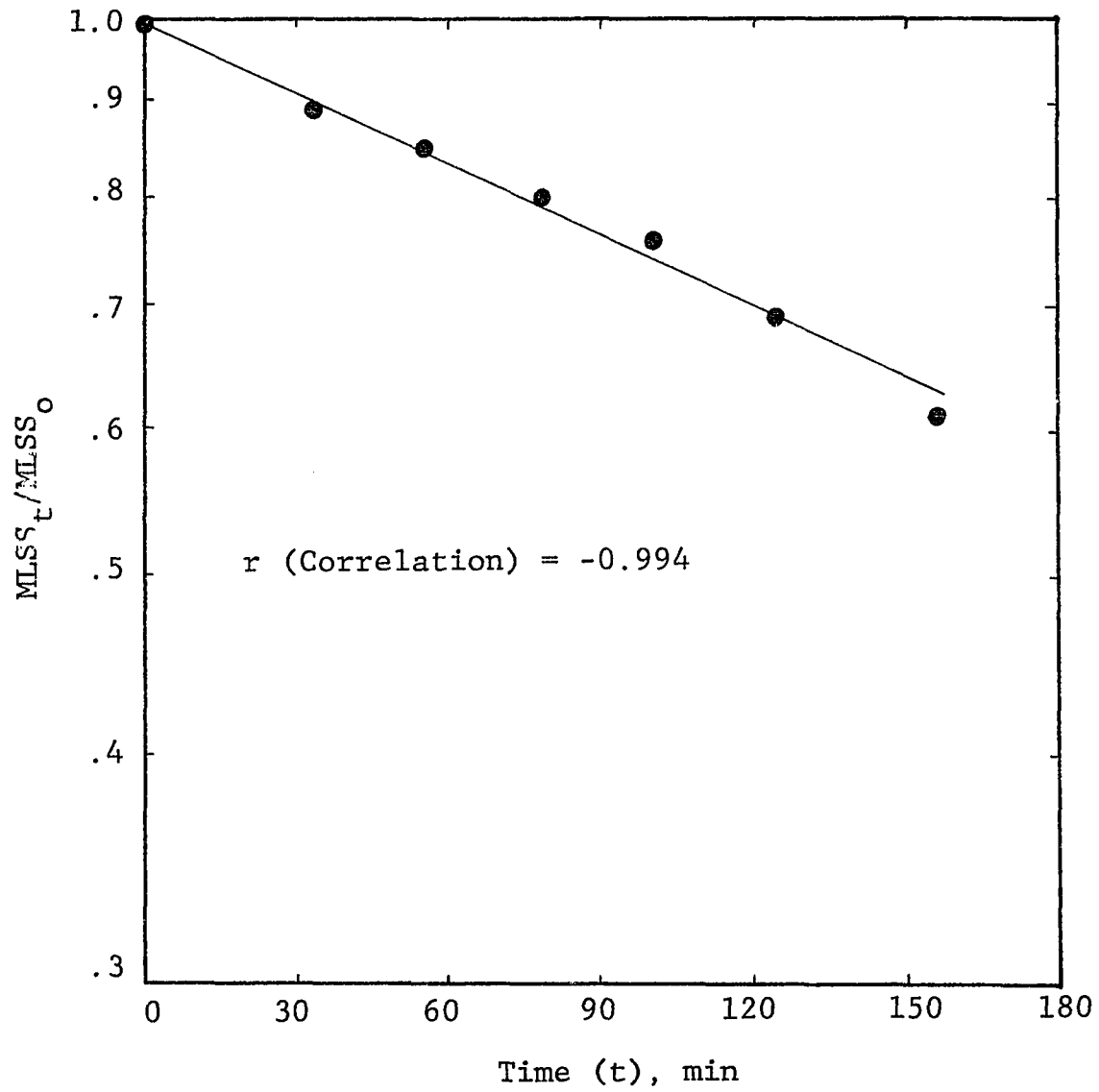
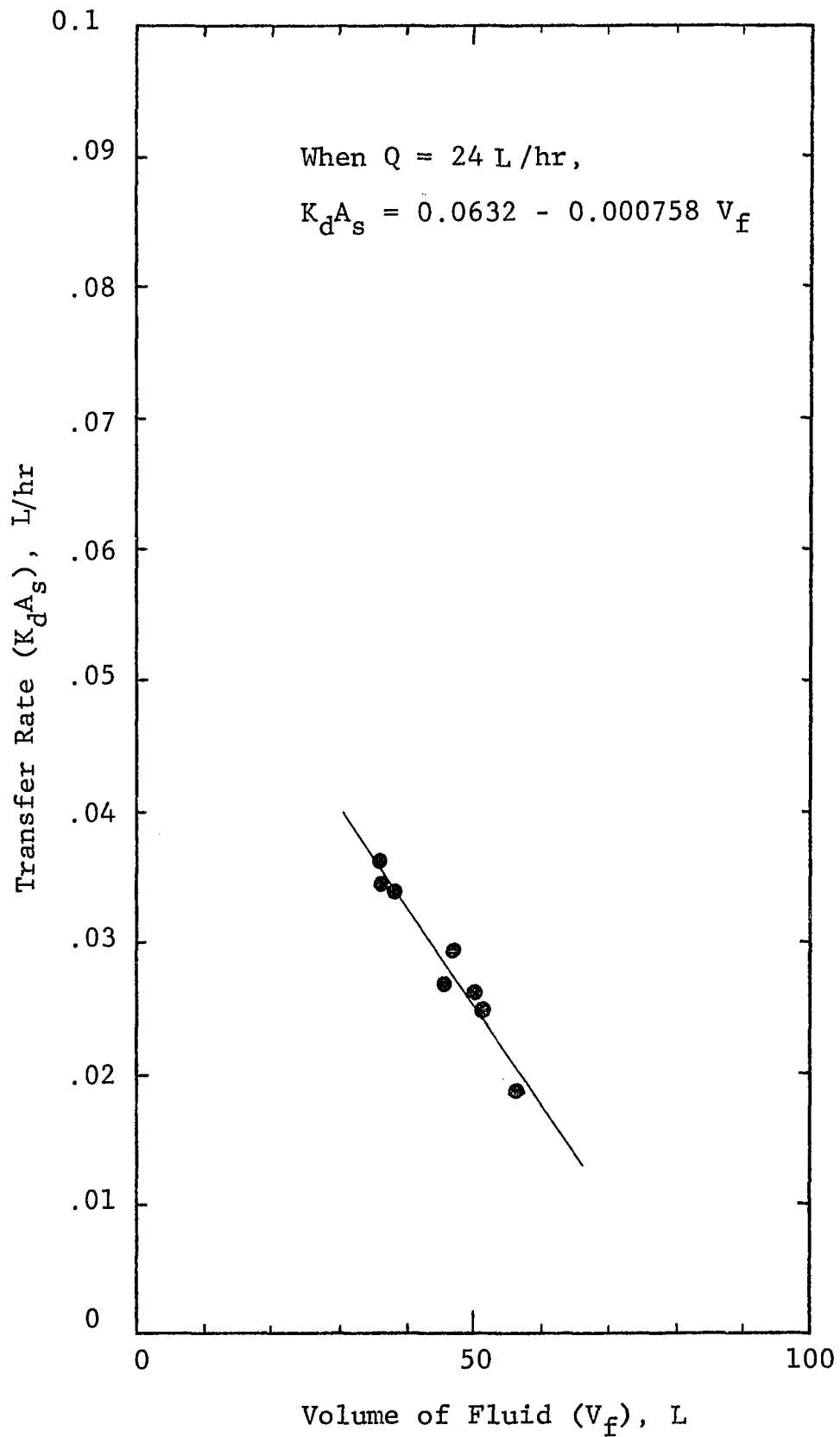


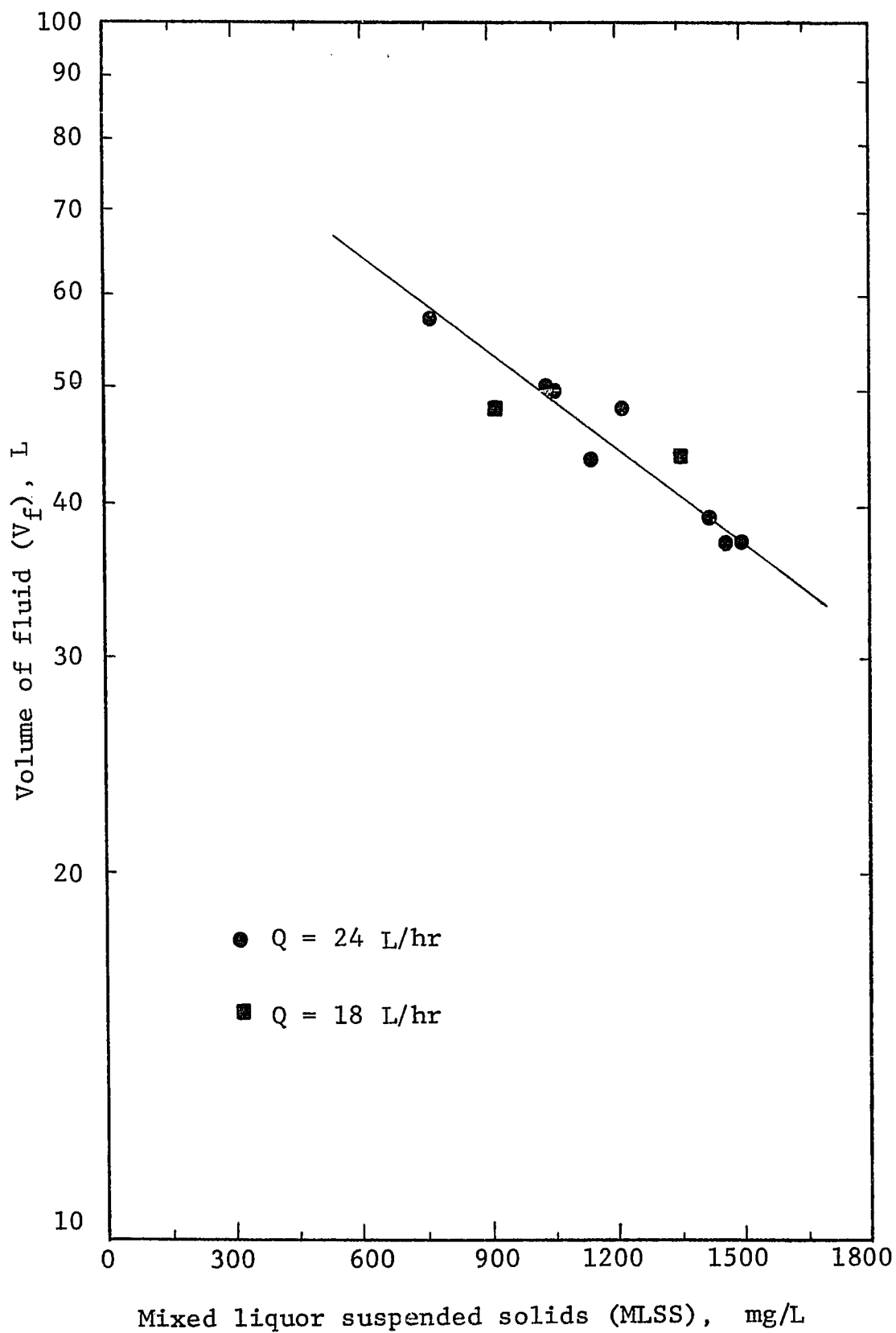
Figure 44. Pilot Plant (Data of 2/1/80)

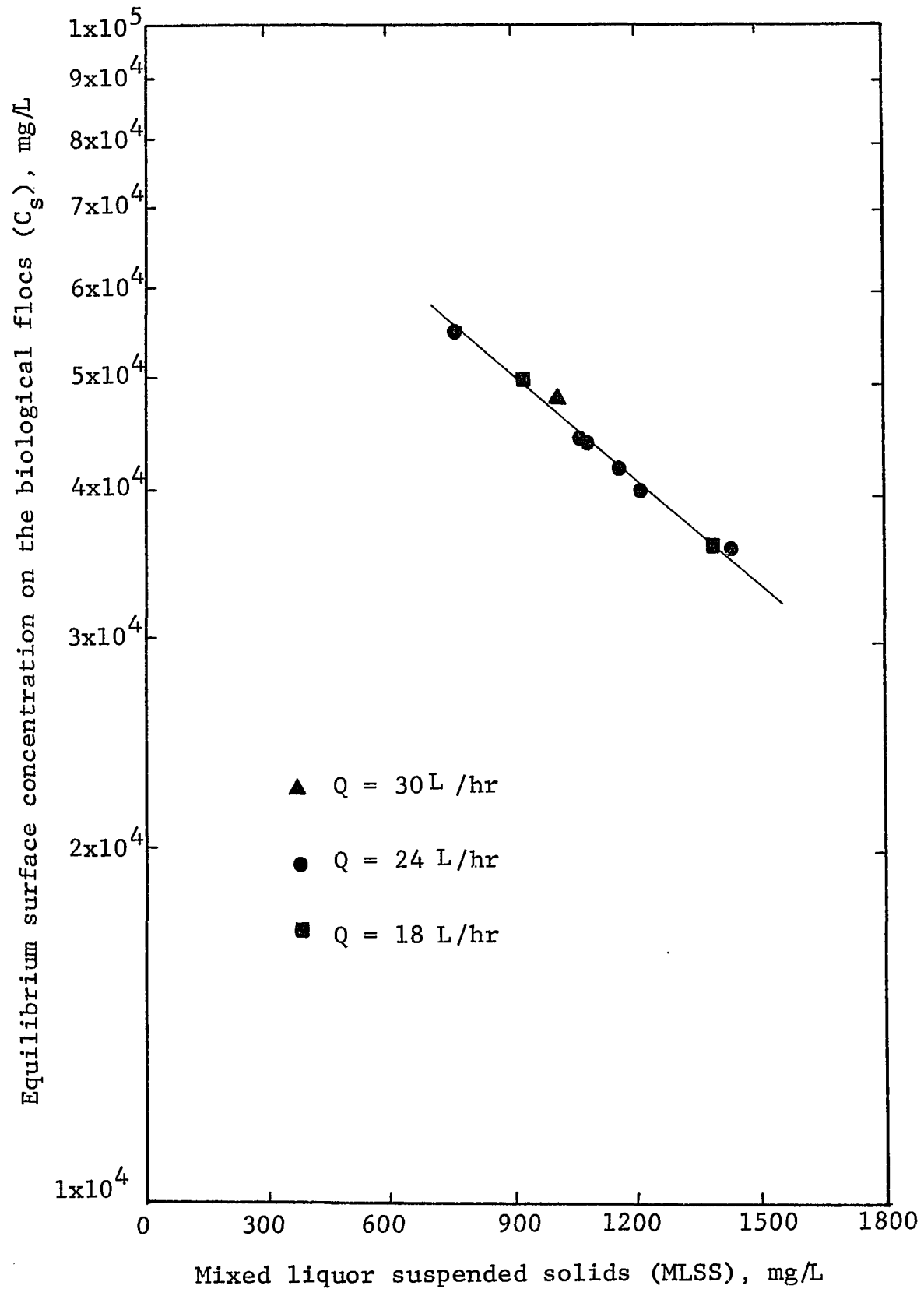
MLSS_t/MLSS₀ vs. Time (t)

APPENDIX IV

Relationships of Parameters for
the Pilot Plant

Figure 45. $K_d A_s$ vs. V_f - Pilot Plant

Figure 46. V_f vs. MLSS - Pilot Plant

Figure 47. C_s vs. MLSS - Pilot Plant

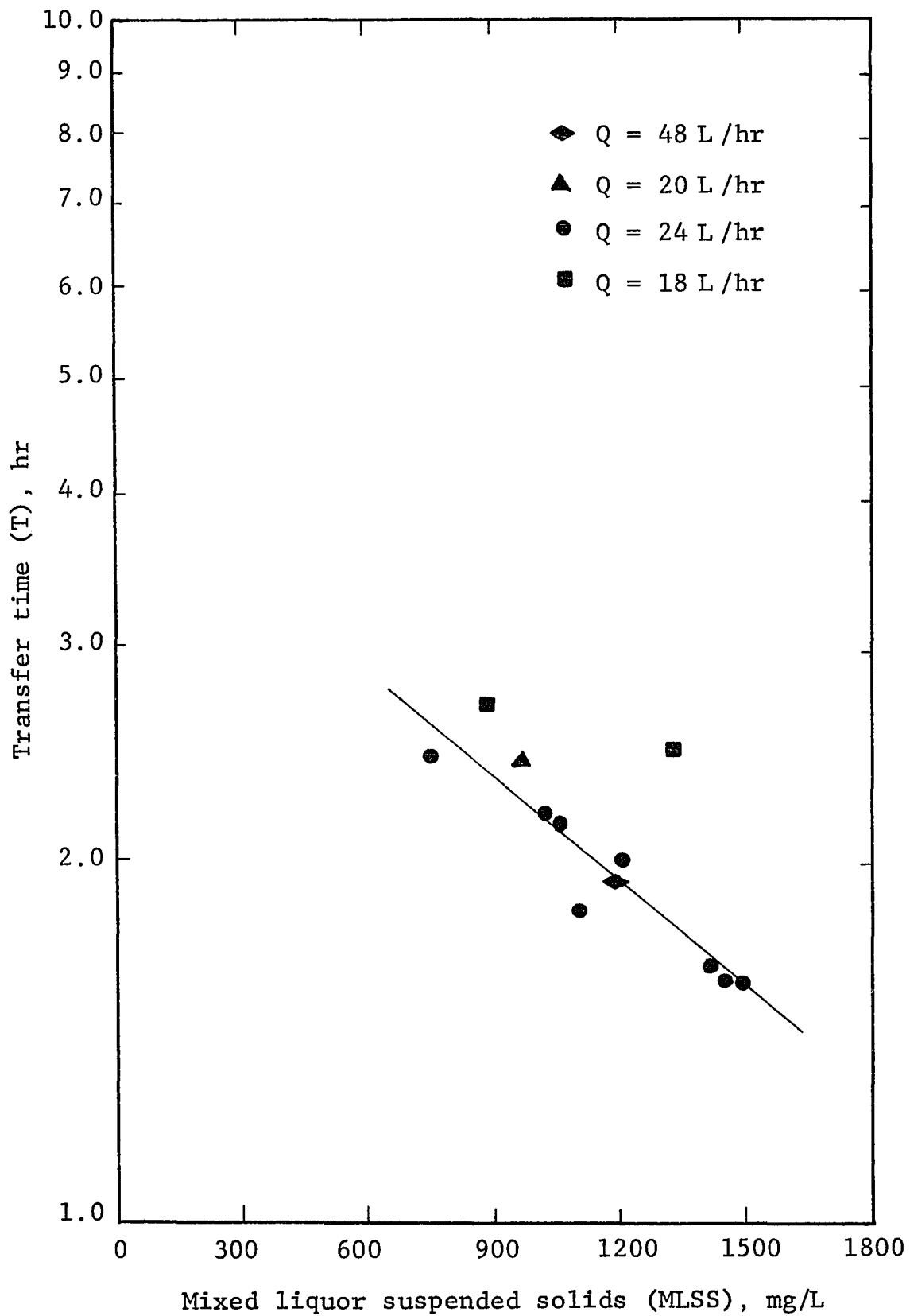


Figure 48. T vs. MLSS - Pilot Plant