PERFORMANCE AND BULKING CONTROL OF INTERMITTENTLY AERATED, CONTINUOUS FLOW ACTIVATED SLUDGE SYSTEMS WITH SELECTOR

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PERFORMANCE AND BULKING CONTROL OF INTERMITTENTLY AERATED, CONTINUOUS INFLOW ACTIVATED SLUDGE SYSTEMS WITH SELECTOR

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Dissertation under the direction of Professor W. Wesley Eckenfelder

Experimental data from continuous flow activated sludge reactors with selectors, fed with a diversified, highly biodegradable, soluble feed were correlated using a simple, semi-empirical formula. Rate of the substrate removal in the selector was demonstrated to be a function of the organic loading in the initial mixing chamber. Results from both standard batch tests, and fed-batch reactor tests independently provided a satisfactory estimation of the correlation formula constants. The formula allows prediction of the substrate concentration in the selector and was used for the optimization of the selectors design for bulking control.

The formula for the reaction in the contactor provided basis for formulation of an equation for "biosorption concentration". Analysis of the equation demonstrated that an optimum sludge recycle rate exists for a set of system parameters. A nomogram allows determination of the optimum recycle rate for a given value of a single system constant. The system constant incorporates three parameters: influent concentration, return sludge concentration, and contactor nominal retention time and one experimental constant: maximum reaction rate.

The optimum recycle rate is always less than 100 percent and approaches this value as the system constant decreases. The maximum

"biosorption concentration" achievable in any completely mixed selector-reactor system is one fourth the initial substrate concentration. The substrate concentration in the contactor at the optimum sludge recycle rate is equal to one-half of the influent concentration.

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Advisor

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CHAPTER I

INTRODUCTION

The most common problem associated with activated sludge separation is the development of filamentous microorganisms which protrude from the floc and interfere with settling by bridging or creating a diffuse floc structure. This condition, commonly referred to as bulking, manifests itself in deterioration of the sludge zone settling velocity. Therefore, a secondary clarifier may not provide enough surface area to adequately separate sludge from the effluent. This results in the presence of an excessive concentration of suspended solids in the effluent, a condition which in itself may violate a discharge permit limit of the treatment facility. If bulking is not controlled, the system may lose an increasing portion of its total biomass inventory, resulting in a decrease in treatment efficiency. In extreme cases complete culture washout may occur with total system failure (no treatment).

In recent years, a great amount of research has been directed toward identifying the causes of excessive filamentous growth. It has been observed that bulking commonly occurs in completely mixed activated sludge systems operating at a low food to microorganism (F/M) ratio (Tomlinson and Chambers, 1979; Strom and Jenkins, 1984). There is also evidence that the presence of a spatial or time-varying

gradient of substrate concentration in the system tends to produce sludge with a good, stable settling characteristic.

An example of time varying substrate concentration is a batch reactor or any of its full-scale variants, such as a sequencing batch reactor. A spatial substrate concentration is present in any sludge system other than a completely mixed, single tank system. Any reactor with some type of compartmentalization or baffle arrangement has a spatial concentration gradient. A hybrid system which encompasses both spatial and temporal substrate gradients is known as an intermittently aerated, continuous flow system.

Such a treatment system consists of a single tank (Figure 1.1) with continuous introduction of wastewater into a baffled prereact zone (PRZ). An operational cycle of three to six hours consists of aeration, sedimentation and decantation periods. This system was found to be a cost effective and reliable treatment alternative for small and medium size waste streams (Goronszy and Barnes, 1980). In general, pilot and full scale plants of this type generate stable, well settling sludge (Goronszy, 1984, Bell and Hardcastle, 1984). This is most likely associated with the discussed above development of a time- or spatial-varying substrate concentration gradient in the aeration tank. The intermittent system is unique in that it combines both these features to create a substrate gradient. During air off periods most of the incoming waste accumulates to a high concentration in the PRZ; this gradually diminishes at the onset of the aeration period. The

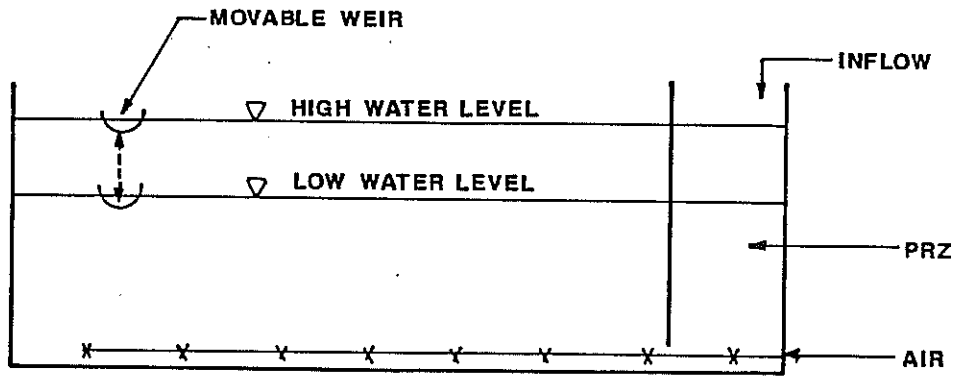


Figure 1.1. Schematics of Intermittently Aerated, Continuous Inflow Treatment System.

incoming wastes are contacted in the PRZ with the recycled sludge before entering the main tank. The substrate concentration in the PRZ depends on its size and sludge recycle ratio (controlled at a full scale plant by the size of the opening in the baffle and by the agitation intensity). There are limited data available on the design of initial mixing chambers for bulking control,and of that none was collected from intermittently aerated systems. For domestic sewage the soluble substrate concentration in the main tank is low throughout the aeration cycle due to the relatively low soluble substrate fraction and due to the substrate uptake in the PRZ. For a high strength soluble feed, the substrate concentration can vary during the cycle. In such cases cycle timing and design of PRZ might require optimization for compliance with the effluent quality requirements.

The oxygen uptake rates in intermittent systems show significant time and spatial variations (Bell and Hardcastle, 1984), making optimization of the aeration process crucial for the cost-effectiveness of the system. This is further complicated by the fact that the total oxygen requirement must be provided during a fraction of the total cycle time.

Recently, more emphasis has been placed on the control of nitrogen concentration in treated wastewater. Low loaded, intermittent systems are capable of providing full nitrification. By manipulating aerobic and anoxic sequences in the cycle, significant denitrification can be achieved as well. Full scale data demonstrate that intermittent systems can achieve 85 percent removal of total nitrogen (Goronszy and Barnes, 1982, Arora et al., 1985).

Although previous investigations indicated that the intermittently aerated, continuous inflow system was an attractive treatment alternative, especially for small to medium waste flows, almost all the available data on the system are of a general performance nature. This study was undertaken to develop detailed, kinetic information on the specific aspects of the system, such as PRZ performance, bulking control, oxygen demand profiles and kinetics of nitrification-denitrificaiton.

Initial experiments on a large scale (5,200 gal max volume) intermittently aerated pilot plant were performed in spring of 1984. The work concentrated in two areas:

- Effect of PRZ on the mixing characteristic of the reactor.
- Hydraulics of the decanting system.

A compartmentalized model of the pilot plant mixing pattern was developed which successfully correlated data from tracer tests. The PRZ was considered as a first compartment with a dead zone. The main aeration tank was conceptually divided into three compartments. Recirculation between the four compartments was then allowed in the modeling scheme. The resulting set of five differential equations was solved numerically for the test parameters.

Tests on the hydraulics of the decanting system led to the development of a relationship between decantation rate, weir loading, and maximum sludge level required to avoid sludge entrainment during the decantation.

A theoretical model for substrate removal during the aeration cycle was also developed. To arrive at the substrate profile, the model took into account continuous influent flow and changing reactor volume during the cycle.

Further expermentation on the pilot plant however, was hampered by continuous maintenance problems, unstable influent flow and varying influent composition (municipal wastewater). It also became obvious that despite the successful hydraulic modeling, the baffle with underflow creates too complex a mixing pattern to warrant development of a meaningful substrate removal model. Therefore, the subsequent work was performed with parallel units operating under controlled laboratory conditions. A synthetic, soluble feed was utilized for the study.

The laboratory set-up was modified in respect to the full-scale system by completely separating the PRZ from the main aeration tank. Also, in most cases, the PRZ was continuously aerated. In this paper, only results obtained from work performed in the laboratory are reported.

Specific objectives of the study were:

1. To define the relationship between PRZ operating parameters and PRZ performance in terms of substrate removal efficiency.
2. To optimize PRZ design for bulking control.
3. To determine the relationships between the ability of activated sludge to rapidly remove substrate from the

solution (biosorption), activated sludge system configuration, and bulking control.

4. To establish if and in what quantities activated sludge is capable of removing soluble substrate without an external electron acceptor (molecular or bound oxygen).
5. To investigate the effect of PRZ design on rates of nitrification- denitrification.
6. To determine time and spatial oxygen uptake rate profiles during cycle sequences. Separate carbonaceous and nitrogenous components of oxygen uptake rates and define the effect of PRZ design on oxygen uptake rate (OUR) profiles.

CHAPTER II

LITERATURE REVIEW

Effect of Reactor Configuration on Bulking Control

Bulking has been recognized as a problem in activated sludge processes since its inception (Scott, 1928 - after Farquhar and Boyle, 1971). Pipes (1967) gave the first analysis of this condition and advanced possible causes and cures. More recent research has led to the identification of two primary causes of filamentous bulking (Eikelboom, 1977; Strom and Jenkins, 1984):

1. low dissolved oxygen (D.O.) concentration,
2. low organic loading (F/M).

Some specific, readily biodegradable industrial wastes as well as nutrient deficient and septic wastes have also been implicated (Jenkins, 1984). Farquhar and Boyle (1971) described several types of filamentous bacteria and proposed the first key for their identification based on stains. Eikelboom and Buijsen (1981) characterized 20 of the most common filament types and proposed a different procedure for their identification. Strom and Jenkins (1984) modified the identification procedure and listed the most common filament types associated with specific causes of bulking.

In low D.O. bulking it was postulated that oxygen concentration in the center of the floc is limited and filaments protruding from the floc have more favorable conditions for growth (Sezgin et al, 1978). The remedy in this case is to increase the aeration capacity.

The growth of filamentous organisms at low F/M has been shown to be suppressed by a low degree of longitudinal mixing (plug flow). Similar effects were achieved in systems equipped with an initial compartment for mixing of return sludge and influent (PRZ). Chudoba and coworkers (Chudoba et al., 1973a and 1973b) demonstrated in a series of papers that reactors simulating plug flow produced sludge with a lower sludge volume index (SVI) than completely mixed systems. They introduced the concept of the selector, an initial compartment for mixing of influent with return sludge, which promotes growth of floc-forming, nonfilamentous bacteria. Their theory explains the selection of bacterial species based on difference in growth constants in the Monod formula (discussed in more detail in the following section):

$$\mu = \mu_m S/(K_s + S) \qquad (2.1)$$

where:

μ = specific growth rate, mg/l-hr

μ_m = maximum growth rate, mg/l-hr

K_s = saturation constant, mg/l

S = limiting substrate concentration, mg/l

According to the theory, filamentous organisms are characterized by lower K_s and μ_m than floc formers, which would favor the former at the lower substrate concentrations (Figure 2.1). The theory qualitatively explains several reported cases (as discussed later) of filament suppression by creation of zones of high substrate concentration.

The methods of attaining such zones vary from baffle arrangements, which simulate plug flow, to installation of one or more chambers for precontacting return sludge and influent. A batch mode of operation (intermittent feed), which introduces time-varying concentration is theoretically equivalent to the plug flow.

Rensink (1974) reported that poorly settling sludge from an oxidation ditch treating dairy wastes was improved by intermittent feeding. His experiments with synthetic feed demonstrated that a reactor's resistance to bulking (time to the onset of bulking) increases in this order: completely mixed, plug flow, batch. In each reactor configuration the lower the overall organic load, the longer the reactor resisted bulking. In low loaded batch systems, bulking was prevented within the time frame of his study (20 to 24 days), which was less than 3 mean cell residence times. In subsequent experiments with synthetic dairy and settled domestic sewage he demonstrated that the occurrence of bulking can be prevented by either a long, rectangular or a completely mixed, round selector. Only the rectangular chamber, however, was capable of curing bulking sludge with both types of feed (Rensink, 1982).

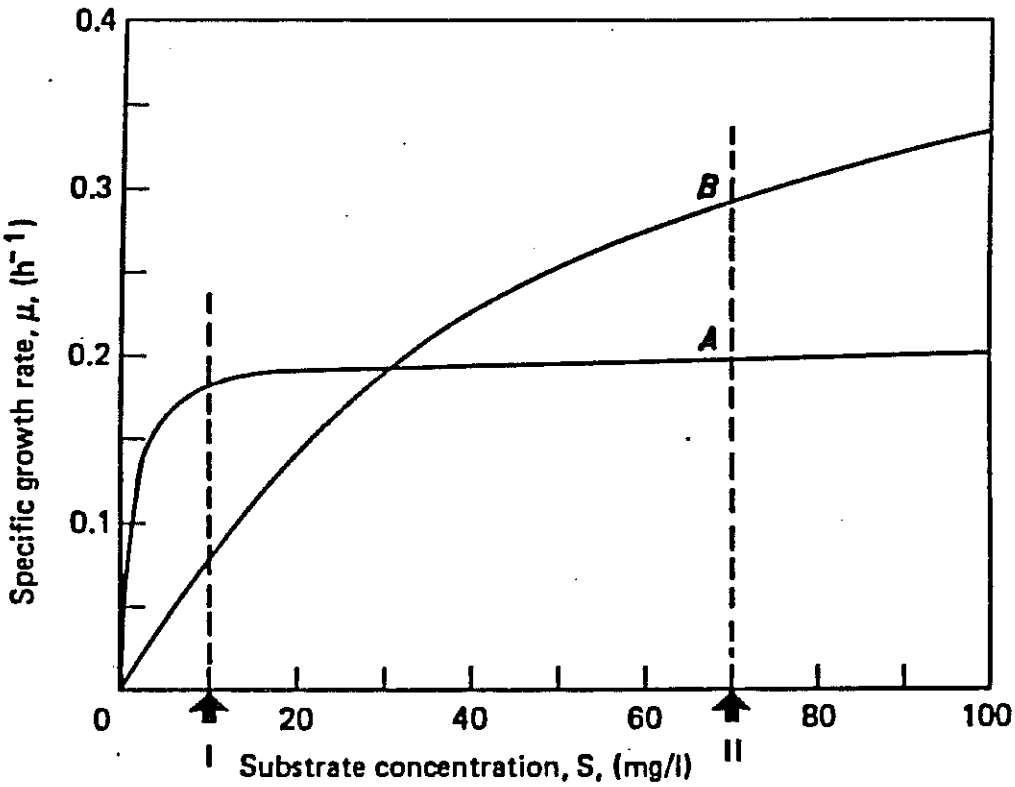


Figure 2.1. Graphical Presentation of the Principle of Selection of Microorganisms in Mixed Cultures. A-Filamentous Bacteria, B-Floc Formers (After Chudoba et al., 1973b).

In a series of papers a group of researchers (Houtmeyers et al., 1980; Verachtert et al., 1980) demonstrated that intermittently fed reactors generally produced sludges with superior settling characteristics as compared to continuously fed units. With glucose as the main substrate, continuous reactors readily developed filamentous bacteria, while settling properties of intermittently fed sludges remained good till the end of the study (20 days). Similar observations were made of systems fed with (1) nutrient broth, (2) acetate, and (3) mixture of glucose and nutrient broth. When starch was used both systems bulked, but the intermittent reactor resisted bulking for 18 days as compared to 4 days for the continuous unit. When fed casein neither system bulked during the study period (16 days). In conclusion, the authors suggested that the existence of a short exogenous phase and a long endogenous phase in the intermittent (or plug flow) mode is responsible for the "selection" of the different microorganisms. When readily available substrates are present at high concentrations the species with the higher substrate uptake rates are selected (in agreement with Chudoba's selector theory). These microorganisms can rely on accumulated material during an endogenous, starvation period.

Borgatti (1982) in his study on sequencing batch reactors fed with synthetic feed (soybean-casein digest medium with dextrose) noted that a nonaerated quiescent fill period, followed by an aeration period, suppressed filamentous growth, while an aerated-mixed fill period favored the growth of filamentous organisms. Silverstein (1982)

in an accompanying study on molasses-based feed confirmed Borgatti's findings. It should be noted that the quiescent fill period more closely simulates a theoretical batch reactor than the aerated-mix-fill sequence.

Lee et al. (1982) in their study on municipal sewage supplemented with raw sludge found that the ability of a reactor to prevent and cure bulking depended on the fraction of total reactor volume occupied by the initial compartment and, simultaneously on the compartment's soluble substrate concentration. The COD concentration gradient in the main tank resulting from its compartmentalization was not significant. The authors concluded that the initial compartment conditions - soluble COD concentration, fraction of total volume or compartment's organic loading (gCODremoved/gMLVSS-day) are critical for bulking control.

Eikelboom (1982) observed that the substrate removal process which occurs during the initial mixing stage is decisive in the selection of the microbial population. The removal process, usually termed biosorption, is a combination of biological and physico-chemical processes. These include:

1. mechanical entrapment of influent suspended solids in the open structure of the sludge floc,
2. biochemical coagulation of the colloidal fraction by exocellular bioploymers with subsequent entrapment and,
3. soluble substrate uptake with or without intermediate storage or accumulation.

To describe the conditions in the initial mixing chamber Eikelboom (1982) introduced the concept of floc loading equal to the mass flow of the available substrate per mass flow of return MLVSS (mg COD/g VSS). He concluded that biosorption efficiency of 50 to 70 percent in about 10 minutes is required in order to prevent excessive growth of filaments. The floc loading required to accomplish this goal was found to be a function of sludge characteristics rather than the type of wastewater. Best results were achieved with floc loadings between 50 to 150 mg COD/g MLSS.

Jenkins (1984) reported that, after three mean cell residence times, sludge from a reactor with a PRZ developed an ability for a higher initial uptake of soluble COD in comparison with that from a completely mixed reactor. The completely mixed reactor bulked, while the reactor with the selector did not. Borgatti (1982) in a work mentioned earlier on fill and draw reactors also found that fill conditions which discouraged filamentous bacteria also enhanced the microbial storage potential. Silverstone (1982) also found an analogous relationship.

Kinetics of Substrate Removal

Single Substrate Removal

Steady State: Single substrate removal in homogenous as well as in multispecies populations, has been exhaustively researched and quantified by numerous researchers. The mechanistic basis for reaction

rate expression originated in the Michaelis-Menten equation describing rate of enzyme catalyzed biological reaction. The reaction involving a single substrate was proposed to proceed in two steps:

1. reversible combination of enzyme (E) and substrate (S), and
2. irreversible decomposition of the substrate-enzyme complex (ES) into enzyme and product (P).

The reaction proceeds in the following way:

$$E + S \underset{k_2}{\overset{k_1}{\rightleftharpoons}} ES \xrightarrow{k_3} E + P$$

where k_n indicates reaction rate constant.

It can be shown that resulting expression for reaction rate (R_r) has the form

$$R_r = R_m S/(K_m+S) \qquad (2.2)$$

where:

R_r = substrate removal rate

R_m = maximum reaction rate (when all enzyme is in the form of ES)

K_m = $(k_2 + k_3)/k_1$ = concentration of S at which reaction rate is equal to one-half of R_m.

Based on experimental data, Monod (1949) developed an equation describing the specific growth rate of a bacterial culture in its exponential and declining growth rate phases:

$$(dX/dt)/X = \mu_i = \mu_{mi}S/(K_{si}+S) \qquad (2.3)$$

where:

X = biomass concentration

t = time

μ_i = specific growth rate of bacterium i

μ_{mi} = maximum growth rate or bacterium i

S = growth limiting nutrient concentration

K_{si} = half saturation constant for bacterium i

In Equation 2.3 the constant K_{si} represents the limiting substrate concentration at which the growth rate is equal to one-half of the maximum reaction rate, μ_{mi}. If one disregards the microbial endogenous decay, the growth rate of the bacteria is proportional to the substrate removal rate. Consequently, the terms: "growth rate" and "substrate removal rate" are interchangeable with a utilization of a theoretical yield coefficient Y.

$$dX/dt = Y(dS/dt) \qquad (2.4)$$

The most frequently used, and also the simplest, modification of the Monod equation is a set of two equations describing removal (growth) kinetics in two extreme cases: when the substrate concentration is small or large compared to the half saturation constant.

These equations are simplifications of the Equation 2.3 and have this form:

$$\mu_i = \mu_{mi}S/K_{si} \quad \text{when } S << K_{si} \qquad (2.5)$$

$$\mu_i = \mu_{mi} \qquad \text{when } S >> K_{si} \qquad (2.6)$$

Another linear modification of this type is referred to as Blackman kinetics and more adequately reflects Monod kinetics in the transient region where substrate concentration is comparable to the half saturation constant:

$$\mu_i = \mu_{mi}/2K_{si} \qquad \text{when } S \leq 2\ K_{si} \qquad (2.7)$$

$$\mu_i = \mu_{mi} \qquad \text{when } S > 2\ K_{si} \qquad (2.8)$$

These equations model the growth (removal) rate of a biological culture under steady state conditions.

In practical applications, however, a steady state with respect to the substrate concentration is seldom achieved. First of all, influent concentration is time-varying for most real-life systems. In a controlled laboratory settings, utilizing synthetic feed, this variation can be eliminated. However, a true steady state condition exists only in a completely mixed (CSTR) system without sludge recycle (chemostat), since even a relatively small sludge retention time in a clarifier unavoidably introduces substrate gradients into the system. Anything other than a completely mixed system, such as plug flow, baffled systems, or CSTR in series renders the steady state assumption inapplicable, at least at the microbial level. Therefore, it is apparent that for almost any practical activated sludge system elements of a dynamic microbial response should be considered.

Transient Phenomena: Two types of dynamic growth response were described in the literature: growth rate hysteresis (GRH) and available reaction potential (ARP). GRH describes the tendency of a

microbial population growth rate to lag behind the growth rate predicted from a steady-state equation upon shift-up of substrate concentration. A similar phenomenon occurs during a decrease of the substrate concentration. The GRH was first predicted theoretically by Perret (1950) and verified experimentally for both pure and mixed cultures by Storer and Guady (1969). The ARP concept was introduced by McLellan and Bush (1969) and expresses the ability of the microbial population to rapidly increase its growth rate upon an increase in the substrate concentration. The ARP generally increases with the decreasing steady-state growth rate (increasing sludge age), indicating that the smaller portion of the available enzymatic pool is actively involved in metabolic activity at lower steady-state growth rates.

It should be noted that GRH and ARP type responses are not mutually exclusive, and both phenomena were shown to occur concurrently. Actually, the response predicted from a steady-state assumption and GRH can be considered as two extreme cases while introduction of ARP in combination with GRH can account for any intermediate response, as illustrated in Figure 2.2. GRH and APR type responses were incorporated into a dynamic response model by Powell (1967 and 1969) and were found to accurately describe transient response of a multispecies natural population to shift-ups (Chi and Howell, 1976; Chase, 1977). To date, most of the research done in support of these concepts has utilized a single-substrate feed, usually glucose.

Maloe and Kjeldgaard (1966) reviewed the role of RNA levels in cells in transient response, and subsequent work (Koch and Dieppe,

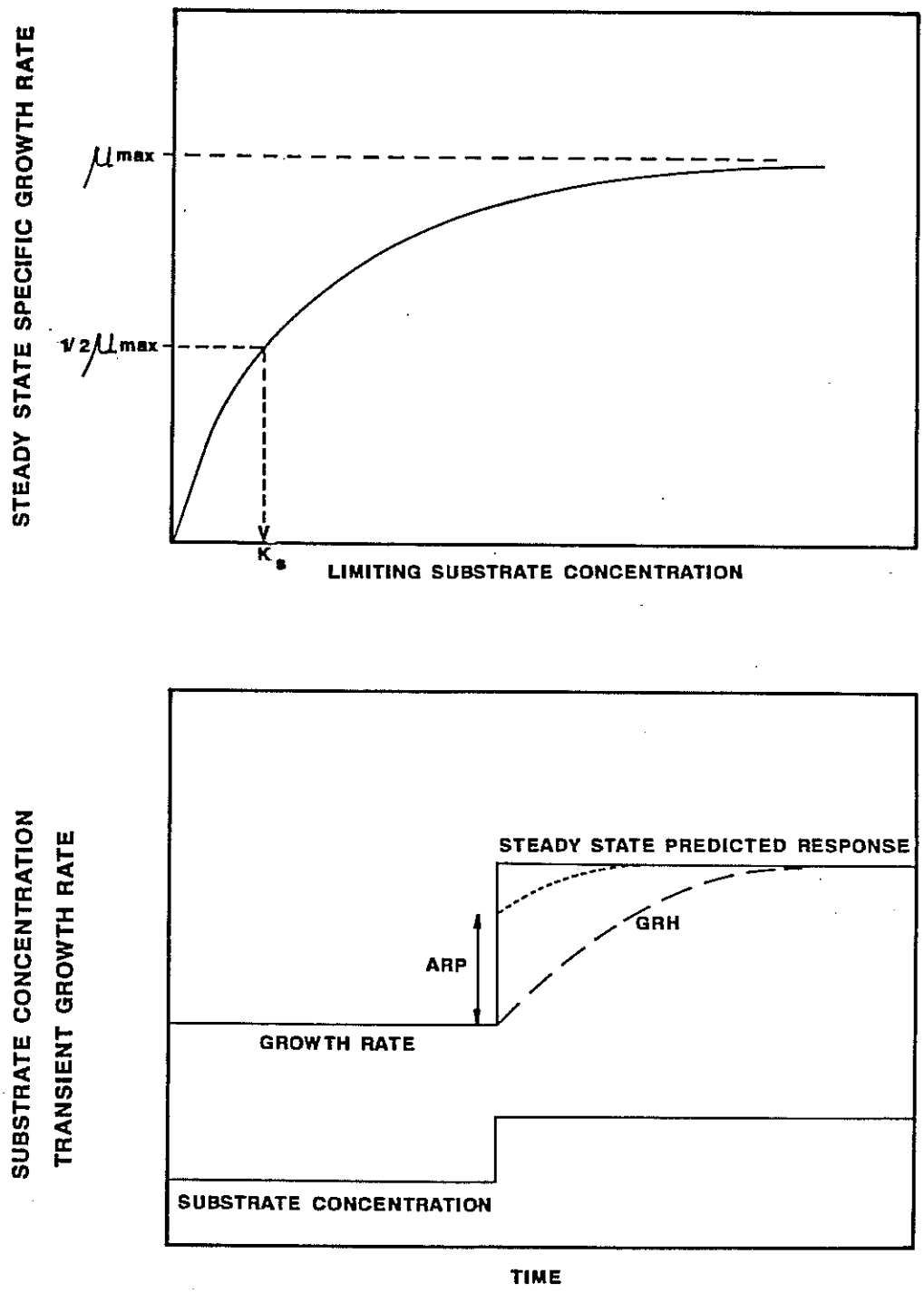


Figure 2.2. Illustration of Steady-State, GRH and ARP Type Responses to an Increase in the Limiting Substrate Concentration.

1971) demonstrated that a population developed at low steady state growth rates possesses extra RNA which is available during a transient response. Daigger and Grady (1982) concluded that both RNA and other intracellular components such as enzymes and metabolic intermediates are responsible for a culture's transient response.

Storage Phenomena: So far in this paper the transient phenomena were discussed in terms of dynamic growth response. It is now apparent that another type of transient response, storage, plays an important role in cultures exposed to a time or space-varying substrate gradient. Three different types of a soluble substrate storage have been identified.

Extracellular accumulation on cells' membrane has been demonstrated to account for some of the immediate removal of dichlorophenol (Schneider, 1987), and ethoxylated nonylphenols (Patoczka, 1987) from the water phase. This phenomenon requires affinity of the substrate to lipids, the main structural components of the cell membrane. The extracellular accumulation was modeled as a physical adsorption (Freundlich isotherm) in both of the above quoted works. The adsorption process is independent from the actual utilization of the substrate by the cell and was shown to be reversible (Schneider, 1987). The rapid physical adsorption of nonylphenol ethoxylates was followed by a slow biodegradation of the adsorbed species (Patoczka, 1987).

Some microorganisms were shown to possess the capacity for an intracellular accumulation of the original substrate in an unchanged

form (Rickenberg et al., 1956; Kepes and Cohen, 1962). This type of response is possible only for simple types of low molecular weight substrates, such as glucose, which can be transported through the membrane in an unchanged form. The accumulation capacity was shown to be a Monod type function of the extracellular substrate concentration (Rickenberg et al., 1956) and to proceed as a first order reaction with the rate constant independent of the external substrate concentration (Kepes and Cohen, 1962).

Intracellular storage of a partially modified substrate is the third kind of the storage-type phenomena (Porges, et al., 1955). Most of this intracellular storage phenomenon is attributed to an accumulation of polysaccharide storage polymers (glycogen) and lipids (poly-B-hydroxybutyrate-PHB). These intermediates can subsequently enter metabolic pathways. Accumulation capacity was shown to be dependent on substrate nature and microbial species present. The higher the proportion of carbohydrates in the substrate, the higher was the storage rate and capacity (Takii, 1970, 1977a and b). When proteins or complex synthetic organic compounds are present storage is not likely to occur since the rate limiting step in assimilation is the initial substrate break-down. Under this condition, build-up of the storage polymer precursors would not occur.

In recent years a lot of research effort has concentrated on the role of a system's operational mode (feeding pattern) in microbial adaptation and species selection, based on the storage capacity concept.

In particular, this research was concerned with means of prevention of activated sludge bulking.

A highly structured, mechanistic model of the dynamics of microbial growth on soluble substrate was presented by Daigger and Grady (1982). This model divides the cellular mass into five components (a synthetic component, a structural component, precursor molecules for those components, enzymes and storage products) and proposes rate expressions governing concentration of these constituents. The model incorporates all processes important in governing microbial growth. The resulting model complexity precludes application in its entirety to any practical problem. However, either growth type or storage type transient response can be described conceptually through a proper selection of the rate constants i.e. by equating some of them to zero.

Multicomponent Substrate Removal

Transient Phenomena and Storage Phenomena described in the Single Substrate Removal Section also apply, at least qualitatively, to multicomponent substrate situations. Steady-state kinetics, however, require a different approach. A direct extension of the Monod equation (Formula 2.3) is difficult to justify, due to the complex interactions (competitive, inhibitory) between individual substrates, particularly in a heterogenous population.

Despite this, the mechanistic basis for the initial attempts at modeling substrate removal in a CSTR activated sludge system orginate

from the Monod equation (Formula 2.3). This formula was first applied for the design of the CSTR activated sludge system by Lawrence and McCarty (1970) and is referred to in the literature as Lawrence-McCarty model:

$$R_r = YR_mS/(K_s + S) \tag{2.9}$$

where:

R_r = rate of substrate utilization

At a low limiting substrate concentration ($S<<K_s$) Equation 2.9 is reduced to a first order relationship, while at high concentration ($S>>K_s$), the reaction proceeds at a zero order rate. Eckenfelder and Ford (1970) proposed that in most practical applications substrate concentration in a completely mixed biological system is low and first order approximation is adequate:

$$(S_o - S_e)/Xt = KS_e \tag{2.10}$$

where:

S_o = influent substrate concentration

S_e = effluent substrate concentration

t = hydraulic retention time in CSTR activated sludge system

K = reaction rate constant

As pointed out previously, there are serious doubts about applicability of Monod based equations for the modeling of reaction kinetics in a heterogenous sludge treating multicomponent wastewater. First, each substrate component can be expected to have a different rate constant for different bacterial species; and second, sequential substrate removal was demonstrated to occur, indicating that some types

of substrate (particularly simple ones like glucose, galactose, gluconic acid and manitol) exhibit a repressive effects on metabolism of co-substrates (Gaudy et al., 1963, Chian and DeWalle, 1975).

A less fundamental approach was proposed by Tebbutt and Christoulas (1975). They applied a concept of a retardation (or response) factor, initially introduced by Fair, Moore and Thomas (1941) (after Fair et al., 1968). The reaction rate is modeled as a function of the degree of the treatment accomplished, reflecting decrease of the treatability as the reaction proceeds:

$$-(dS/dt) = K(S/S_o)^n X S_o \qquad (2.11)$$

Tebbutt and Christoulas (1975) demonstrated that a satisfactory fit of batch experimental data can be obtained using that correlation. For a special case, when n=0, the rate expression reduces to a first order reaction, as proposed of Eckenfelder and Ford (1970) - Formula 2.10.

Another type of kinetic formula with a retardation factor was proposed by Grau et al., (1975):

$$-(dS/dt) = K(S/S_o)^n X \qquad (2.12)$$

They provided an intuitive explanation of the relationship arguing that it reflects "decrease of the removal rate caused by a reduced number of components and thus the decrease in total substrate concentration with time." Because the exponent in Formula 2.12 has to fit any pattern of gradual components diminutation, it is not limited

to integral values. Benefield and Randall (1977) concluded that Equation 2.12 with n=1 is the preferred, most flexible way of correlating experimental data from both plug flow and completely mixed reactor systems. A similar equation was proposed earlier by Adams, et al. (1975).

$$-(dS/dt) = KSe/SoX \qquad (2.13)$$

These rate expressions, derived from the first or zero order approximations of the Monod equation (with or without a retardation factor), are widely used for experimental data correlation and design of activated sludge plants. For a correlation of a particular data set, the equation which provides best data fit is selected. The design of the reactors, particularly CSTR is usually based on either Equation 2.10 or 2.13.

However, as was pointed out by Buch (1984), the practice of applying rate constants obtained from batch tests for design of a CSTR system frequently results in oversizing the system. If the reaction rate calculated from a batch test at substrate concentration, S, equal to the design effluent concentration, S_e, is used, then the resulting CSTR will certainly fulfill the expectations. The desired effluent concentration is usually close to complete removal, and the CSTR will assure achievement of almost complete removal, an effect called by Buch (1984), "a self-fulfilling prophecy". The CSTR has, in most cases, the capability of operating at a much higher reaction rate, if enough substrate is provided. The reaction rate is then controlled by the organic loading and not by an extension of kinetic concepts taken from chemical engineering into the biochemical process.

The concept of relating biological reaction rate in CSTR to the overall, applied organic loading was proposed by Suschka (1980). He showed that a number of data sets published in the literature can be satisfactory correlated by the following equation:

$$W = W_{max}L/(K_W+L) \qquad (2.13)$$

where:

W = specific substrate removal rate

W_{max}= specific maximum substrate removal rate

L = applied organic loading

K_W = "constant equal to the substrate removal rate when W=one-half of W_{max}" (actually, this should be "loading at which the removal rate is equal to one-half of W_{max}")

The resulting correlations were much better than those obtained by the original authors with the classical models.

Lump Parameters and Metabolic Byproducts

In the modeling of the kinetic relationships during biooxidation, reaction rates should ideally be specified as functions of the individual substrate concentrations. For multicomponent substrate systems, the complexity of quantitative determination of individual substrate concentrations for any but the simplest synthetic feed mixture is enormous. Furthermore, most individual substrates undergo multistage, partially extracellular enzymatic breakdown, exponentially increasing complexity of the analytical task. Monitoring of the

individual components and the concentrations of their intermediate breakdown products in a natural wastewater, such as municipal sewage, becomes an impossible proposition. Even if the analytical data were available, construction of a kinetic model incorporating a myriad of components and reaction rates, let alone possible inhibitory and/or synergistic effects, is without any potential practical application.

For this reason kinetic relationships in wastewater treatment are expressed in terms of lumped parameters such as BOD_5, COD or TOC. Consequently, these relationships are of a semiempirical nature, at best, and any success in presenting overall reaction rate in terms of a simple, nth-order equation can be considered as accidental.

Use of lumped parameters for monitoring reaction kinetics is also burdened with errors arising from superimposition of rising concentrations of metabolic byproducts on diminishing substrate concentrations (in batch tests). Of the three common lumped parameters, BOD_5 has the advantage of reflecting only the biodegradable part of organic matter released during the biodegradation. COD and TOC, incidently more rapidly and accurately determined than BOD_5, would indiscriminately account for the refractory organic matter as well.

Chudoba (1985) classified refractory organic compounds produced by activated sludge into three categories:

1. Compounds excreted by microorganisms in order to establish a proper concentration gradient across the cellular membrane. Concentration of the products released without presence of substrate was found to be temperature independent with COD

in the range from one to eight mg/l depending on sludge type (Parkins and McCarty, 1981). The fact that the excreted organics concentration is independent of the sludge concentration (in range 7 to 2,390 mg/l TSS) indicates that their release is a result of osmotic equilibrium across the cell membranes.

2. Compounds produced as a result of substrate metabolism and bacterial growth (excluding intermediates, the formation rate of which is small at the relatively large sludge ages encountered in wastewater treatment, and which are biodegradable). In non-proliferating (low loaded) batch systems a linear relationship between the initial substrate concentration and refractory compounds concentration was demonstrated (Chudoba, 1985) according to the formula:

$$S_r = 0.00712S_o + 2.69$$

where:

S_r = concentration of refractory compounds, mg/l COD

S_o = initial substrate concentration, mg/l COD

In proliferating systems with an intensive microbial replication the refractory compounds are produced in much higher quantities.

3. Compounds released during the lysis and degradation of the microorganisms. The mass balance provided by authors indicated that during aerobic stabilization about 15 to

25 mg of refractory COD per gram of biomass degraded is released.

Process Kinetics in Intermittently Aerated Systems

Organic Substrate Removal

In intermittent systems treating municipal wastes the duration of the cycle is governed mostly by hydraulic considerations (Goronszy, 1979). With the overall F/M and MLSS set by the designer's preference (typically 0.05 g BOD_5/g MLSS-day, and 5,000 mg/l, respectively), vessel capacity becomes a function of mass inflow of organic substrate. Frequency of decanting (total cycle time) is then determined by the inflow rate. Decantation rate, and therefore the duration of the decant period is restricted by possible entrainment of settled solids. The duration of the settling period is dependent on the sludge settling characteristic and must be of sufficient length to allow the sludge blanket to fall below a fixed level. The third cycle period - aeration time - is therefore determined by the hydraulic constraints discussed above. In intermittent plants treating domestic sewage, soluble substrate concentration during the aeration cycle is low and does not change substantially. Therefore, effluent BOD_5 is not considered a factor in establishing duration of the aeration period.

Reasons for low and stable soluble BOD_5 concentration in the main tank through most of the aeration period are manifold. The intermittent plants operate at low organic loading (0.05 g BOD/g

MLSS-day) and the aeration tank has a large dilution capacity. Alternately, up to 70 percent of the organic matter in raw domestic sewage is in suspended and colloidal forms (Hunter and Heukelekian, 1965) which are rapidly incorporated into the floc structure. This process is probably completed during initial mixing with the recycle sludge in the PRZ (about 10 min retention time). A significant portion of the soluble substrate can be expected to be removed in the PRZ, further diminishing potential for substrate concentration variations in the main tank.

During sedimentation and decantation periods, when mixing is provided neither in the main tank nor in the PRZ, the bulk of the influent is contained in the PRZ. Irvine and Busch (1979), showed that a substantial amount of the soluble substrate can be taken up by the sludge under anoxic conditions. Utilization of bound oxygen in the form of nitrates for carbon oxidation (denitrification) can account for these observations but capability of sludges to remove soluble substrate without an external electron acceptor cannot be discounted. Separation of these two mechanisms requires additional work. By removal of substrate without molecular or bound oxygen supply we mean an accumulation and storage process without utilization of anaerobic metabolic pathways (no strict anaerobes present).

In intermittent systems working under higher organic loading than discussed, substrate build-up during air-off phase may be more significant and may necessitate longer aeration periods in order to deplete the substrate concentration to the required level. Potential

for this build up would be increased for wastes with lower TKN:COD ratio, since less bound oxygen would be available for facultative heterotrophs during the anoxic phase.

Nitrogen Removal

In the intermittent systems alternating aerobic and anoxic periods facilitate the occurrence of nitrification-denitrification. The process provides in time what a traditional single-sludge system for biological nitrogen removal achieves in space, with the mixed liquor divided between aerobic and anoxic basins.

Domestic wastewater and many industrial wastes contain nitrogen primarily as organic or ammonia nitrogen. Enzymatic breakdown of organic nitrogen with release of free ammonia (ammonification) usually progresses rapidly and is not the rate limiting step in nitrification (Argaman 1982).

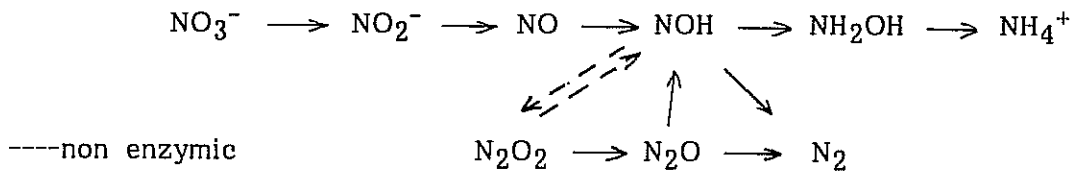
Oxidation of ammonia into nitrates is a two step process. First Nitrosomonas oxidize NH_4^+ to NO_2^-, and then Nitrobacters complete the process by oxidizing NO_2^- to NO_3^-. This first step is under most conditions rate limiting and therefore nitrite concentration in nitrifying sludge is usually minimal (USEPA, 1975). The saturation constant in the Monod type expression for the rate limiting step i.e. oxidation of ammonia to nitrite is less than 1 mg NH_4^+-N/l, and therefore the reaction proceeds as zero order to a very low substrate (NH_4^+) concentration (Hung and Hopson, 1976). Nitrification rate is influenced by a number of process conditions. Oxygen concentration below 1 mg/l decreases overall rate (Hung and Hopson, 1974).

Nitrification rate depends on pH, with optimum usually in the range eight to nine. Inhibition of nitrification occurring at low pH is significant in view of the fact that chemotrophic nitrifiers utilize inorganic carbon (alkalinity) as a carbon source.

Heterotrophic microorganisms in activated sludge grow by utilizing organic carbon sources and molecular or bound oxygen as an electron acceptor. Many species which can utilize nitrates as a source of bound oxygen contribute to the dissimilatory nitrate reduction (denitrification) which proceeds in the following stages as proposed by Christensen and Harremoes (1977).

$$NO_3^- \longrightarrow NO_2^- \longrightarrow NO \longrightarrow N_2O \longrightarrow N_2$$

Assimilatory denitrification involves reduction of nitrates to ammonia used in cell synthesis. According to Painter (1970) the most likely pathways of assimilatory and dissimilatory denitrification are as follows:



Heterotrophic denitrifying organisms can utilize a variety of organic carbon sources. In multiple sludge nitrification-denitrification systems an external carbon source (usually methanol)

was added in the anoxic stage. Single sludge systems eliminate the necessity for addition of external carbon by utilizing organics present in the wastewater or endogenous decay of the biological sludge. When the raw wastewater is used as a carbon source denitrification rates are about one third (at 20°C) of those found in methanol-based systems. When endogenous decay of the organism is used as a carbon source, denitrification rates further fall by about 50 percent and the organism's endogenous decay becomes the rate limiting step (USEPA, 1975).

In intermittent systems substrate entering the aeration tank in the last part of the aeration period and endogenous decay become carbon sources available for denitrification. During the air off phase, wastewater incoming to the PRZ provides a carbon source for denitrifying organisms contained in the PRZ. Furthermore, part of the incoming wastewater leaves the PRZ through the baffle underflow, and can diffuse into the sludge blanket in the main tank, enhancing denitrification.

Incorporation of an anoxic mixing period within the aeration sequence provides another means of facilitating denitrification of accumulated ammonia. Data from full scale intermittent plants show that the duration of the cycle periods can be optimized to achieve 85 percent total nitrogen removal (Goronszy and Barnes, 1982). Bell and Hardcastle (1984) reported that in a laboratory scale intermittent

reactor without a separate PRZ almost complete nitrification and denitrification were achieved under various cycle combinations.

Oxygen Uptake Rates

Specific oxygen uptake rates (SOUR) in the intermittent system show significant time and spatial variations (expected for a system with cyclical operation and a concentration gradient). Bell and Hardcastle (1984) found that in a reactor without a PRZ the SOUR were high at the start of the aeration and after 30 to 60 min were rapidly reduced to low levels. The authors attributed the high initial SOUR to the accumulation of the organic substrate during the anoxic period.

Available data indicate that, for readily biodegradable wastewater with a high TKN:COD ratio, significant and rapid changes in SOUR can be expected. Since economic delivery of the required oxygen is crucial for cost effectiveness of the system, ability to model the SOUR profile during a cycle becomes important. Significant factors influencing the SOUR profile are expected to be the PRZ design and TKN:COD ratio.

CHAPTER III

DEVELOPMENT OF THE EXPERIMENTAL WORK

The experimental work was carried out in three phases. During the first phase in progress from May to October 1984, four laboratory reactors were operated at average organic loading of 0.08gBOD_5/ gMLVSS-day, with a nominal feed concentration of 150 mg/l BOD_5. After initial experimentation with feed composition and operation in a continuous aeration mode, three reactors were switched to an intermittent aeration mode. The fourth reactor served as a control (continuously aerated, no PRZ). Out of three intermittently aerated reactors two were equipped with PRZs of different sizes. Cycle time was 3 hr with the reactors kept in an anoxic condition for about 50 percent of the cycle time. The reactors were operated in this mode for about three months (Phase I). During that time the effect of PRZ on the reactors' resistance to bulking was monitored. During the track studies, the oxygen uptake rates related to nitrification-denitrification in the intermittently aerated reactors were investigated.

A more comprehensive experiment based on experience gained from Phase I was designed for the subsequent phases of the study. The work now concentrated on modeling PRZ performance and bulking patterns. There was less emphasis on overall reactor performance and patterns of

OUR and nitrification-denitrification rates. In addition to operation of continuous flow and batch-fed reactors, a battery of batch-type tests were performed on sludge generated in these reactors. The reactors' feed composition remained the same as in Phase I. However, its strength was increased from 180 to 420 mg/l COD, with the overall organic loading increasing to 0.15gBOD_5/gMLVSS-day. The feed strength was doubled in order to facilitate testing of a wider range of operating regimes in the PRZ, particularly substrate concentrations.

During Phase II six intermittently aerated reactors without PRZs were maintained, with the objective of determining reproducibility of the performance of reactors run at identical operating parameters. The performance data showed little statistical difference between the reactors. Sludge settling characteristics followed the same trend in all the reactors and batch tests showed slight variations in the sludges biosorption capacity. This information lent confidence that performance data obtained in the subsequent tests would be a function of the operating regime of reactors and not a result of random variations.

In Phase III, five out of a total of six continuous flow reactors were equipped with PRZs operating at different floc loads and contact times. Monitoring of the continuously operated reactors was supplemented by several types of batch tests. Among these were tests

for kinetics of the substrate removal under aerobic and anaerobic conditions, fed batch reactor tests and tests for biosorption capacity. Special tests related to the kinetics of substrate uptake under anaerobic conditions were also conducted.

CHAPTER IV

METHODOLOGY OF THE LABORATORY WORK

Reactor Configuration and Operation - Phase I

Reactor Design and Feed Preparation

Experiments were conducted at the Environmental and Water Resources Engineering Department laboratory of Vanderbilt University. In Phase I four bench scale reactors with baffled sections for clarification of the effluent were used (Figure 4.1) with a total reactor volume of 19.5 l and an aerated volume of 14.5 l.

Initially, reactors were fed with substrate prepared from homogenized and settled canned dog food. However, after a few days of operation an accumulation of nonbiodegradable fat was observed in the reactor. This variable, and difficult to define nonbiodegradable influent fraction, which was physically trapped in the reactors, would have eventually fouled the reactors, disrupting the mass balance and obscuring the reactors' performance. Therefore, it was decided to switch the reactors to a synthetic, completely soluble feed consisting of nutrient broth (BBL Microbiology Systems), yeast extract and dextrose (D-glucose), supplemented with nutrients.

Feed concentrate was prepared as needed (approximately once per month) and stored in a refrigerator. The feed concentrate composition is shown in Table 4.1. Fresh feed solution was made up daily from the

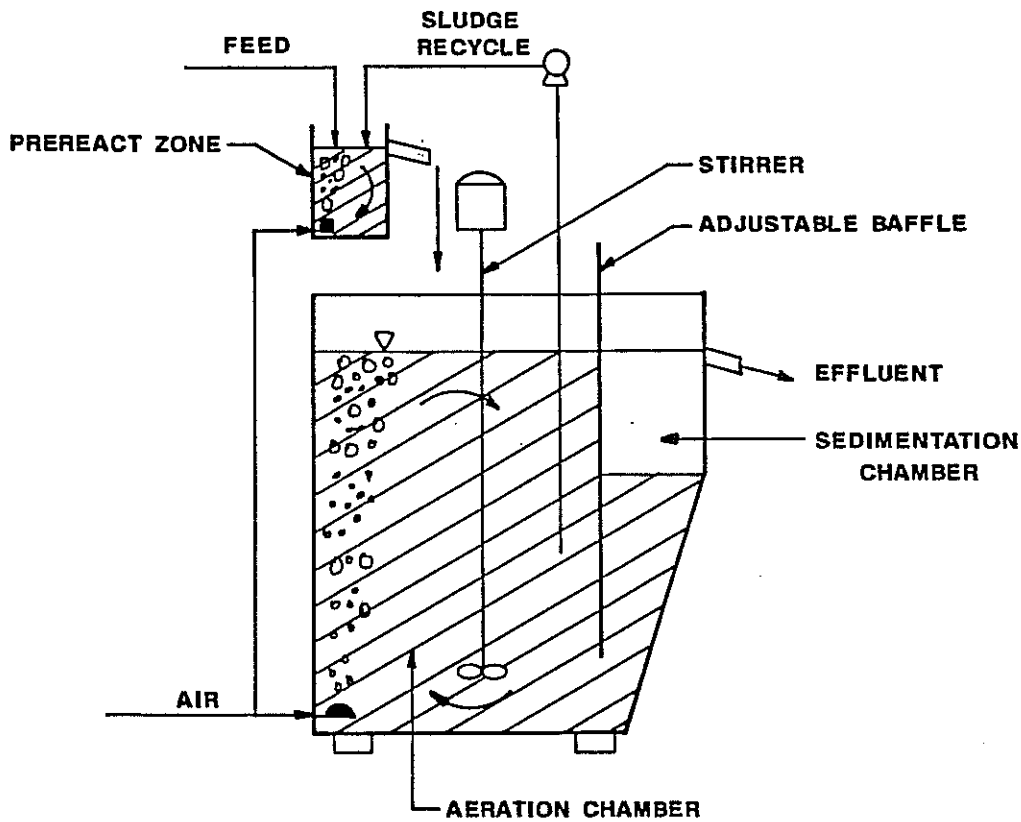


Figure 4.1. Schematic of Continuous-Flow, Completely Mixed Reactor with Prereact Zone (PRZ) and Underflow BAffle (Phase I).

TABLE 4.1

FEED CONCENTRATE COMPOSITION

Substrate	Concentration (g/l)
Nutrient Broth	125
Glucose	44.9
Yeast Extract	10.7
$(NH_4)_2 SO_4$	23.5
KH_2PO_4	4.4
KOH	1.2

concentrate with addition of Ca^{+2}, Mg^{+2}, Mn^{+2}, Fe^{+3}, and $NaHCO_3$ solutions. The raw influent final composition and analysis results are given in Table 4.2. The feed was pumped into the reactors using Masterflex peristaltic pumps at a rate of 26.6 l/day. The fresh feed BOD_5 was 150 mg/l, but due to the biodegradation in the feed tank and tubing, the average BOD_5 of the reactors influent was 118 mg/l.

MLVSS in all reactors was maintained at a nominal concentration of 2,000 mg VSS/l. It should be noted that for MLVSS measurements, the reactors' baffles were lifted and consequently the 2,000 mg/l MLVSS value is based on the total tank volume. The resulting organic loading was:

$$F/M = \frac{26.6 \text{ l/day. } 118 \text{ mg } BOD_5/l}{2000 \text{ mg VSS/l . } 19.5} = 0.08 \quad \frac{gBOD_5}{g \text{ VSS-day}}$$

Reactor Operation

After the change in feed, during the initial acclimation period, the experimental system was operated in a continuous aeration mode with completely mixed reactors preceded by prereact zones (PRZ).

The PRZs were built as small, separate compartments in which recycled sludge was contacted with the influent (Figure 4.1). The PRZs were continuously aerated to achieve complete mixing and to maintain DO concentrations above 3 mg/l. The return sludge was pumped directly from the aeration tank. This arrangement assured a stable concentration of return sludge. The PRZs' effluent was discharged through an overflow to the main aeration tank. The design floc loadings in the PRZs of the three reactors were 200, 70 and 20 mg COD/g VSS and were controlled by the return sludge flow rate. Contact time

TABLE 4.2

RAW INFLUENT COMPOSITION - PHASE I

Parameter	Concentration (mg/l)
Organic Substrates	
Nutrient Broth	155
Glucose	56.1
Yeast Extract	13.4
Nutrients	
NH_4^+ (as N)	6.2
PO_4^{+3} (as P)	1.3
Metals	
Mg^{+2}	2
Mn^{+2}	0.02
Fe^{+3}	0.05
Ca^{+2}	10
Alkalinity (as $CaCO_3$)	125
Analysis[a]	
BOD_5	150
COD	208
TKN	26
pH	7.4[b]

[a]Measured values in freshly prepared feed. The remaining concentrations were calculated by chemical addition.

[b]pH standard units.

(hydraulic retention time), including sludge recycle, in all three PRZs was set at 10 min by selecting proper PRZ volumes. The fourth reactor was operated as a control unit without a PRZ.

After four weeks of operation in continuous aeration mode, three reactors (Nos. 1, 2 and 3) were switched to an intermittent aeration mode. Cycle time was 3 hr with a 75 min aeration phase and 105 min air off phase. With this distribution of air on/off time, the reactors were kept in an aerobic condition for about 50 percent of the cycle time. Since the major objective of the research was to determine factors affecting sludge settling characteristics, the PRZ from reactor 3 was removed to provide an intermittently aerated, control unit. Operational parameters of the reactors are given in Table 4.3.

Return sludge in reactors Nos. 1 and 2 was continuously pumped directly from the aeration tanks. During the air off phase, the contents of the reactors were mechanically stirred to provide a uniform return sludge concentration.

Reactors Configuration and Operation - Phases II and III

Reactor Design and Feed Preparation

In Phases II and III, the reactor feed was prepared using the same feed concentrate as in Phase I (Table 4.1). The feed strength, however, was doubled with fresh feed BOD_5 increased to nearly 300 mg/l. The parameters of the raw influent used in Phases II and III are summarized in Table 4.4.

TABLE 4.3

OPERATIONAL PARAMETERS OF THE REACTORS
PHASE I

Parameter	Reactor 1	2	3	4
Mode of Aeration				
Aeration Basin	intermittent	intermittent	intermittent	continuous
PRZ	continuous	continuous	--	--
Sludge Recycle	continuous	continuous	--	--
Influent BOD_5-mg O_2/l[a]	118	118	118	118
Influent COD-mg O_2/l[b]	182	182	182	182
Reactor volume - l	19.5	19.5	19.5	19.5
PRZ volume - l	0.27	1.14	--	--
Influent flow rate - l/day	26.6	26.6	26.6	26.6
Sludge Recycle Rate - l/day	13.2	132	--	--
Design PRZ floc loading mg COD/g VSS	200	20	--	--
Temperature ^{o}C	24	24	24	24
Dissolved Oxygen, mg/l	6.5-7.5	6.5-7.5	6.5-7.5	6.5-7.5

[a] As measured in the point of inflow to the reactor. Concentration of the freshly prepared feed in the feed tank was 150 mg O_2/l.

[b] As measured in the point of inflow to the reactor. Concentration of the freshly prepared feed in the feed tank was 208 mg O_2/l.

TABLE 4.4

RAW INFLUENT COMPOSITION - PHASES II AND III

Parameter	Concentration (mg/l)
Organic Substrates	
Nutrient Broth	310
Glucose	112.2
Yeast Extract	26.8
Nutrients	
NH_4^+ (as N)	12.4
PO_4^{+3} (as P)	2.6
Metals	
Ca^{+2}	20
Mg^{+2}	4
Fe^{+3}	0.1
Mn^{+2}	0.04
Alkalinity (as $CaCO_3$)	250
Analysis[a]	
BOD_5	298
COD	420
TKN	52
pH	7.4[b]

[a]Measured values. The remaining concentrations were calculated by chemical addition.

[b]pH standard units.

In order to minimize substrate biodegradation in the feed tank and feed lines noted during the previous phases, the feed tank was refrigerated and small diameter tubes (1/16 in.) were used for the feed lines. The feed lines were cleaned or replaced about once a week.

Construction of the continuous flow reactors used during Phases II and III was modified. The traditional set-up, used in Phase I, consisted of an underflow baffle separating the aeration basin and the clarification chamber (Figure 4.1). In such a configuration, the fraction of sludge remaining in the clarification chamber varied, and was difficult to quantify. This complicated control of the return sludge concentration and, consequently, the floc load. In the new design, the underflow baffle was replaced with a solid baffle with a large window, covered with a filtering cloth (Figure 4.2). Replacement of sedimentation by filtration for the separation of the sludge and the effluent stabilized MLSS concentrations in the aeration chambers and simplified the reactor's maintenance and sampling procedures. The filtering cloth was cleaned by brushing whenever a differential head of 0.5 cm or more developed between the aeration and effluent chambers, usually once every 3 days.

MLVSS in the aeration tank was kept at 2,000 mg VSS/l. The resulting organic loading was therefore:

$$F/M = \frac{14.5\ \text{l/day} \cdot 300\ \text{mg/l}\ BOD_5}{2{,}000\ \text{mg VSS/l} \cdot 14.5\ \text{l}} = 0.15\ \frac{gBOD_5}{g\ \text{VSS-day}}$$

Aeration basins in all the reactors used in Phases II and III were aerated intermittently. The cycle time was 3 hr with a 75 min aeration

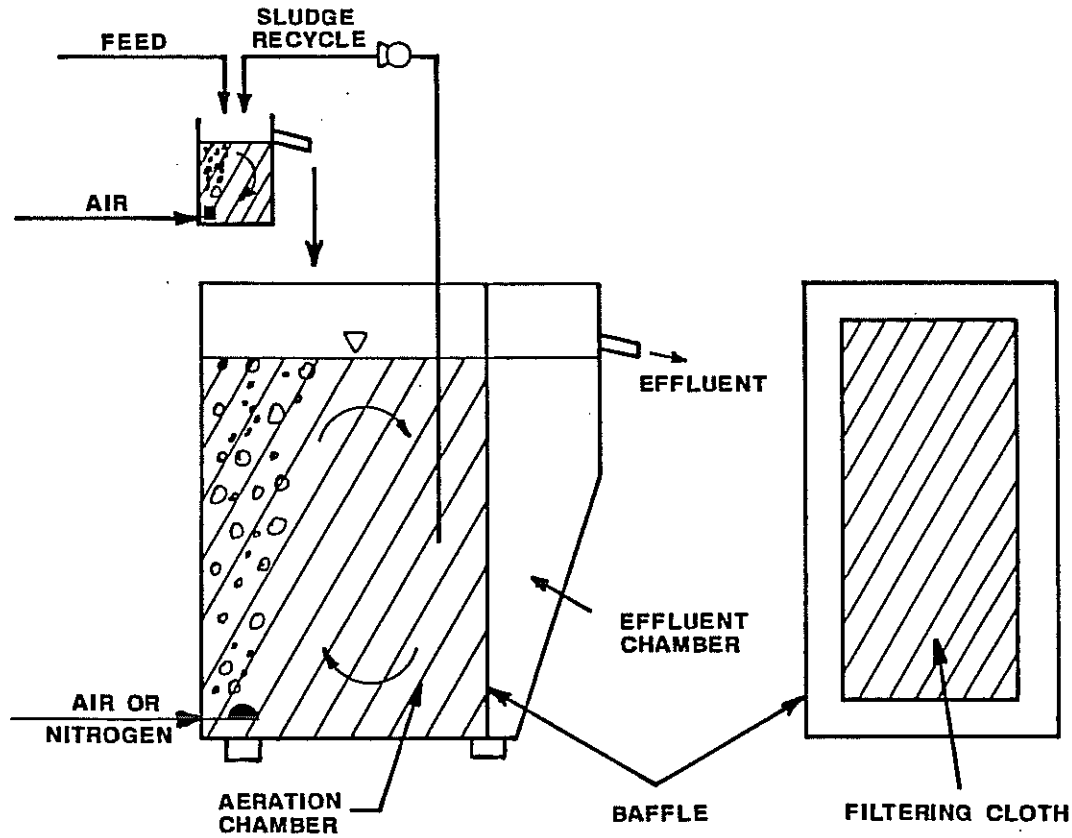


Figure 4.2. Schematic of Continuous-Flow, Completely Mixed Reactor with Prereact Zone (PRZ) and Baffle with Filtering Cloth (Phases II and III).

period and 105 min air off period which provided 50/50 distribution of aerobic and anoxic conditions. During the air off time, the aeration chamber contents were stirred with nitrogen gas from a cylinder. The nitrogen was delivered every 2 min in 15 sec pulses. This arrangement provided adequate mixing of the activated sludge, while minimizing nitrogen consumption. The flow of the gases into the aeration chambers was directed by electromagnetic valves controlled by timers.

Reactor Operation

Phase II: The objective of Phase II of the study was to evaluate statistical differences in the performance and sludge settling characteristic of the reactors attributable to stochastic, rather than systematic variations. Consequently, the reactors' configurations were identical, and operating parameters were kept as uniform as possible. Six continuous flow, intermittently aerated reactors without PRZ were operated during a four week period from May 8 to June 3, 1985. The reactors were seeded with activated sludge from the Nashville Metro treatment plant. Prior to addition to the reactors, the sludge was screened through a 20 mesh sieve.

Phase III: The main experimental effort commenced on July 25, 1985 and continued through October 2, 1985. During this period, six continuous flow reactors were operated, five of them equipped with PRZs. Four of the PRZs were continuously aerated with continuous sludge recycle, while the fifth remained quiescent during the air-off cycle. Three of the continuously aerated PRZ's operated at a hydraulic retention time (HRT) of 8 min and floc loads of 30, 60 and

150 mgCOD/gVSS. In the fourth continuously aerated PRZ, the HRT was 16 min and the floc load was 60 mgCOD/gVSS. The fifth, intermittently aerated PRZ operated at 8 min HRT and a floc load of 30 mgCOD/gVSS. During the air-off phase, the sludge recycle to the PRZ was also discontinued in this reactor. A schematic of the experimental configuration for Phase III is shown in Figure 4.3. Table 4.5 presents design operational parameters of the reactors.

In addition to the six continuous reactors, a 5 l, batch fed reactor was also maintained. The reactor was fed once a day with concentrated feed at the same overall F/M as the continuous flow reactors. The reactor was aerated intermittently with the same cycle as the remaining reactors. Twice a day, the batch reactor's content was allowed to settle and 2.5 l of supernatant was discarded and replaced with tap water. In this way the HRT in the batch fed reactor nominally corresponded to the HRT in the continuous reactors (24 hr).

Analytical Procedures

Chemical Oxygen Demand

For chemical oxygen demand (COD) measurements, the semimicro, closed reflux, titrometric method recently endorsed by Standard Methods, (1985) was adopted. The procedure requires a small sample volume (5 ml) and significantly reduces material, bench space, and labor requirements compared to the traditional open reflux method while also allowing a larger number of samples to be processed simultaneously. During the implementation of the semi-micro procedure, several problems were encountered which warrant a more detailed discussion.

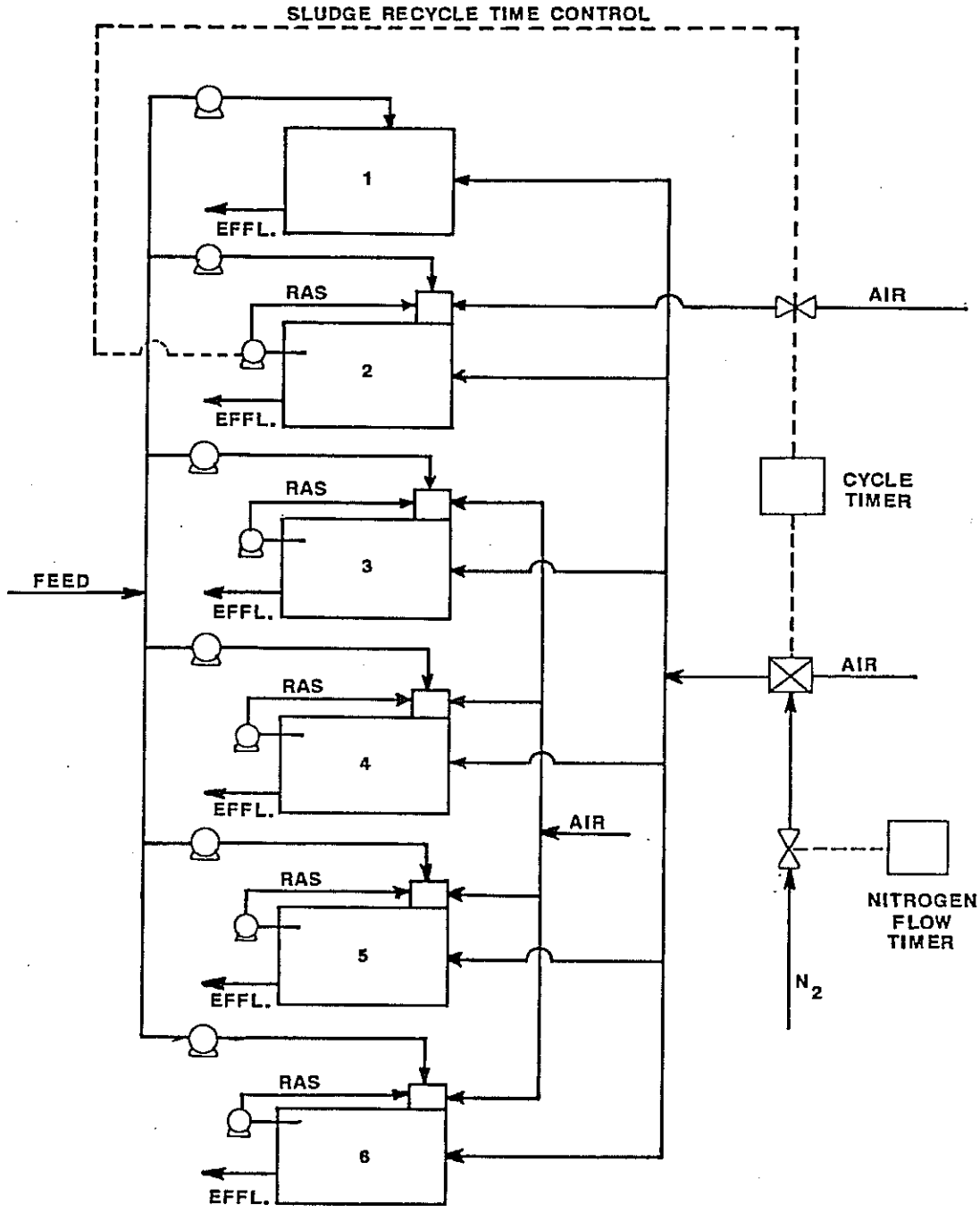


Figure 4.3. Schematics of the Experimental Set-Up (PhaseIII).

TABLE 4.5

DESIGN OPERATIONAL PARAMETERS OF THE REACTORS - PHASE III

Parameter	Reactor 1	2	3	4	5	6	7
Mode of Feed Addition	Continuous	Continuous	Continuous	Continuous	Continuous	Continuous	Batch
Mode of Aeration							
Aeration Basin	Intermittent	Intermittent	Intermittent	Intermittent	Intermittent	Intermittent	Intermittent
PRZ	N/A	Intermittent	Continuous	Continuous	Continuous	Continuous	N/A
Sludge Recycle	N/A	Intermittent	Continuous	Continuous	Continuous	Continuous	N/A
Influent BOD_5, mg/l	298	298	298	298	298	298	[a]
Influent COD, mg/l	420	420	420	420	420	420	[a]
Reactor volume, l	14.3	14.2	14.4	14.4	14.4	14.5	5
PRZ volume, ml	--	970	970	440	1,100	295	--
Influent Flow Rate, ml/min	9.9	10.5	10.6	10.3	10.6	10.2	--
Sludge recycle rate, ml/min	--	111	111	44.3	44.3	26.6	--
HRT in Reactor, hr	24	24	24	24	24	24	24
HRT in PRZ, min	--	8	8	8	20	8	--
MLVSS	2,000	2,000	2,000	2,000	2,000	2,000	2,000
Floc Load, mg COD/g VSS	--	20	20	50	50	80	--

[a]Fed with feed concentrate.

The COD samples were digested in 20 ml capped, glass vials in two heating blocks with a total number of 44 slots. The temperature was maintained throughout the digestion step at the required value of 150°$\pm$ 2°C.

Two sets of reagent solutions were used allowing measurements in two COD concentration ranges: 50 to 800 mg/l (high range) and 10 to 160 mg/l (low range). Each COD sample was analyzed in duplicate and the average value was reported. If the discrepancy between duplicates exceeded 0.2 ml of the titrant (equivalent to about 3.2 mg/l COD in the low range and 16 mg/l in the high range), the analysis was repeated. In almost all such cases, the repeated analyses indicated that the lower COD value from the questionable first set of duplicates was correct. This suggested that the spurious higher results were caused by contaminants exerting a COD demand. The small sample volume in this semi-micro procedure made the measurement highly vulnerable to even trace organic contaminants from the glassware or atmosphere. Therefore, if the spread between the duplicate values was too high and the sample was not available for a repeated analyses, the lower COD from the duplicate was reported.

During the procedure, the organic matter in the sample was oxidized by a boiling mixture of an excess amount of potassium dichromate ($K_2Cr_2O_7$) with sulfuric acid. To aid in a complete oxidation of organics a small amount of silver sulfate (Ag_2SO_4) was added as a catalyst. Interferences caused by the presence of halides were eliminated by complexing with mercuric sulfate ($HgSO_4$) which was

added to the digestion solution. After the digestion, the remaining unreduced $K_2Cr_2O_7$ was titrated with ferrous ammomium sulfate (FAS) to determine the amount of dichromate consumed. With each set of samples, a distilled water blank was carried through the digestion step in triplicate.

The Standard Methods calculation of results is based on the following formula:

$$\text{COD, } mgO_2/l = \frac{8{,}000(A - B)M}{V} \qquad (4.1)$$

A = ml FAS used for blank titration

B = ml FAS used for sample titration

M = molarity of FAS, measured by titration with standard $K_2C_2O_7$ solutions

V = sample volume

The molarity of FAS was determined by titration against a standard $K_2Cr_2O_7$ solution.

Alternatively, the COD results were calculated from titration results of the COD standard-potassium hydrogen phthalate (KHP). A stock standard solution of KHP (1,000 mgO_2/l) was prepared monthly. Two freshly made KHP dilutions were analyzed with each set of COD samples. The dilutions were placed at the low and high end of the COD range thus allowing an independent calculation of the constants B and M in Formula 4.1. Values of B and M obtained by the two methods usually

compared well. The second method, utilizing calibration with COD standards, was chosen for reporting COD results in this study.

In a series of a preliminary experiments, a number of split samples consisting of various feed dilutions were analysed using low and high COD range reagents. The sample concentrations were selected so that they were in the overlapping part of low and high COD range (50 to 160 mg/l). The results indicated that the low range method underestimated COD by about 10 percent with respect to the high range results. Consequently, serial dilutions of feed stock solution (420 mg/l O_2) were prepared and analysed in the low and high COD ranges. The resulting concentrations of the stock solution as calculated from digestion at different dillutions are presented in Figure 4.4. This figure demonstrates that samples digested at concentrations below 600 mg/l using high range COD reagents gave almost a constant COD concentration of the feed stock, while in the low range the recovery decreased with the increasing concentration of the sample over the entire range. It was therefore assumed that the measurements done in the high COD range provide a true COD of the sample. With that assumption, the results of the first and subsequent sets of serial dilution tests and other split sample tests performed throughout the investigation were expressed in terms of a correction factor (Figure 4.5). From Figure 4.5 it can be inferred that with decreasing residual concentration of the chromic acid in the digestion step, the completeness of organics oxidation also decreases. This effect is

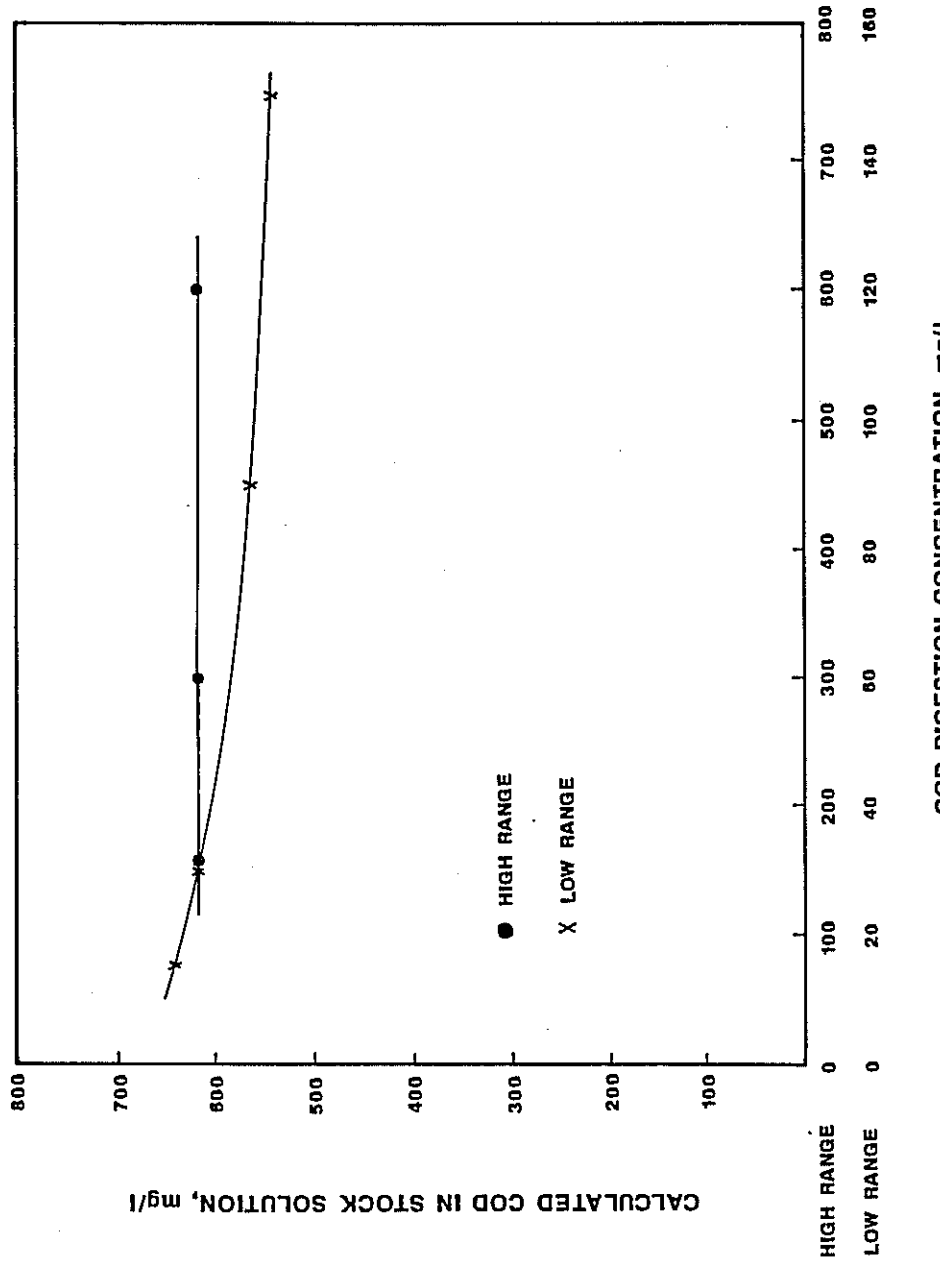


Figure 4.4. COD Concentration in a Stock Solution, as Calculated from Digestion at Different Dilutions.

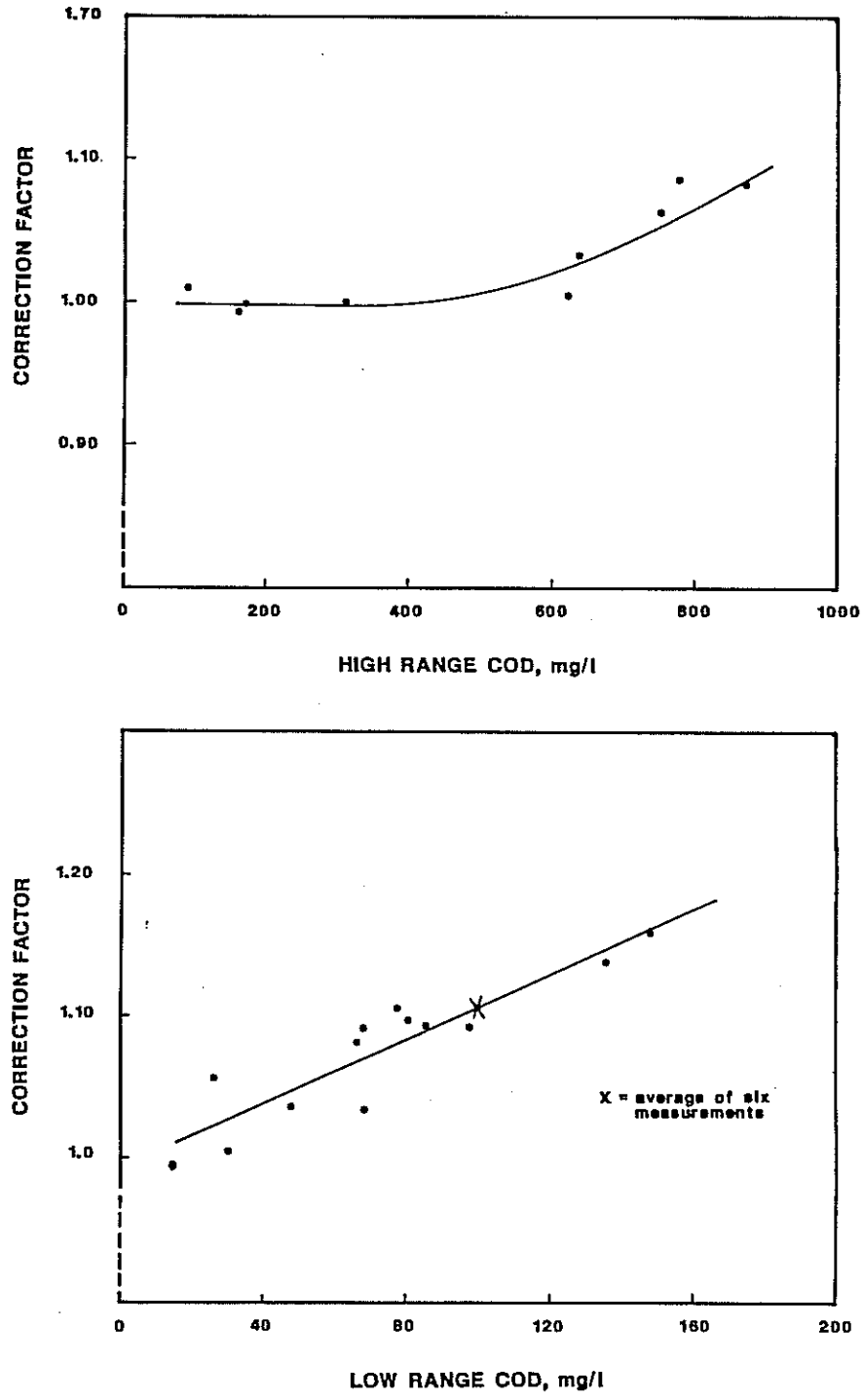


Figure 4.5. Correction Factors for COD Measurements.

observed over almost the entire range of the low COD test (20 to 160 mg/l), as well as in the upper end of the high COD test (above 500 mg/l).

The correction factors resulting from Figure 4.4 were used throughout the study in reporting COD values. It should be noted that these correction factors were substrate specific and should be revised when another type of organic material is analysed for COD. Furthermore, the serial dilutions were prepared with distilled water without correcting for salinity. Therefore, there could have been a varying degree of residual chlorine interference in different samples.

For determination of COD in the mixed liquor supernatant, a sludge aliquot was first centrifuged to provide a rapid separation of biomass and supernatant. In the batch type experiments, where time of the phase separation was critical for calculation of the kinetic relationships, the time of the centrifuge start-up was reported as the sampling time. After 1.5 min of centrifugation the supernatant was filtered to remove the residual biomass.

A series of experiments was performed in order to compare the effectiveness of two different filtering media. Two different filter media were used; Whatman AH-934 fiberglass filter with a nominal pore size of 1.4 µm and Gelman GA-6 membrane with pore size 0.45 µ m. The latter was considered for the application since most of the individual bacterial cells in the mixed liquor have a diameter of 0.5 to 1.5 µm and would have been retained on the membrane, while the Whatman AH-934 pore size of 1.4 µm was potentially too large for this purpose.

Results of this test indicated that Gelman membrane did not remove any additional, measurable COD from the samples as compared to Whatman filters. In fact, samples filtered through Gelman membrane had a slightly higher COD which indicated that these membranes might actually release some COD thus masking any enhanced removal of the bacterial cells or debris. Experiments carried out in the same laboratory with another batch of Gelman 0.45 µm membranes also gave evidence that these filters may release appreciable quantities of COD into the filtrate (Schneider, 1987). Consequently, only Whatman AH-934 filters were used for sample preparation throughout the study.

During filtration, the working surface of the filtering apparatus with a fresh filter in place was first rinsed with one portion of the sample. The initial filtrate was then discarded. Filtrate was then collected into a disposable, glass test tube with two drops of concentrated sulfuric acid. The tube was capped with parafilm and refrigerated until the analyses, usually no longer than two to three days.

Specific Oxygen Uptake Rate

The specific oxygen uptake rate (SOUR) was determined according to the Standard Methods (1981) procedure. A YSI dissolved oxygen (DO) probe and meter, with Hewlett-Packard model 680 strip chart recorder was used for these determinations. The calibration procedure was slightly modified compared to that recommended in Standard Methods. Firstly, the calibration was done on the recorder output, instead of on the DO meter scale, since the strip chart record was used for OUR

calculations. Secondly, zero on the recorder output was marked by allowing a complete leveling off of DO concentration in a sample of activated sludge which reliably indicated a complete removal of DO by biological demand. To complete the calibration, DO in saturated water was recorded on the strip chart with a portion of the sample analyzed for DO by the iodometric method (Standard Methods, 1981). The SOUR was calculated from the linear portion of the strip chart record. Occasionally, a changing rate of oxygen consumption was recorded, indicating a substrate concentration limitation on SOUR. In such cases, an initial SOUR was calculated and reported as an indication of the actual, in situ SOUR.

Sludge Settling Characteristics

Sludge settling characteristics were quantified by means of sludge volume index (SVI) and zone settling velocity (ZSV). Both parameters were measured simultaneously in a 1 l graduated cylinder, equipped with a slow speed (10 rph) mixer. SVI is defined as the volume 1 g of biomass occupies after 30 min of settling. ZSV is a rate of descent of the sludge blanket measured in a linear portion of the graph resulting from recording blanket interface position in the cylinder in time. In addition, in situ ZSV were measured during Phase III investigations. For this measurement, aeration in the reactor was stopped and the rate of sludge blanket descent in the reactor determined.

Filamentous Bacteria Length Count

One of the objectives of this study was to investigate the effect of reactor configuration and operating parameters on proliferation of filamentous microorganisms. In order to quantify filaments present in the sludge, a filament counting technique was adopted from Walker (1985). The method offers a compromise between several published methods of subjective assessment of relative abundance of filaments (Farquahr and Boyle 1971; Chudoba et al., 1973b; Raleigh et al., 1974; Rensink, 1974; Foster and Dallas-Newton, 1980) and very involved and time consuming methods of actual counting of total filament length described by Finstein and Heukelekian (1967) and Sezgin et al., (1978).

The method is based on counting the number of filaments present in selected squares of fine grid engraved on a microscopic slide. The square dimensions (50 by 50 μm) are small relative to the length of the filaments (usually in hundreds of μm range) and therefore, the number of filaments present in the selected squares is on average proportional to the total length of filaments present in the sample.

In this application, a Bright-Line counting chamber usually used for white and red blood cells counts, was used. The chamber's central, $1mm^2$ counting area was divided into 400 small squares (50 μm side). A flat cover glass covered the counting area 0.1 mm above the ruled surface. Thus, the sample volume contained over one small square is $2.5\ 10^{-4}\ mm^3$. During the sample examination the number of filaments crossing 20 squares located on one diagonal of the counting area was counted. Counting was performed at 100x magnification and it was

necessary to move the microscope's focus through the full cell depth (0.1 mm) in order to identify all filaments present. Only filaments which were within a square for at least 0.6 of the square width were counted. This minimum traverse width of 0.6 square was chosen (Walker, 1985) to balance those filaments which cross a square diagonally (i.e., 1.4 square width). Knowing the average number of filaments crossing 20 diagonal squares, the cell volume, and the MLSS of the sample, the average filament length per gram of dry solids was calculated using the following formula:

$$L = \frac{N \times W}{20 \cdot V \cdot MLSS}, \text{ cm/mg} \qquad (4.2)$$

where:

L = filament length, cm/gMLSS

N = average number of filaments crossing 20 diagonal cells

W = width of one cell (0.05 mm)

V = cell volume ($2.5\ 10^{-4}$ mm^3)

MLSS mixed liquor suspended solids, mg/l

After substitutions and units conversion, the final formula has the form:

$$L = \frac{N}{MLSS} \cdot 10^6, \quad \text{cm/mg} \qquad (4.3)$$

Before the sample examination, the test tube with the sludge was mixed for 30 sec in a vortex mixer at a high speed in order to break large sludge floc agglomerates and to provide a more uniform filament distribution. Despite this precaution it was found necessary to repeat the filament count 12 times (20 diagonal squares each count) in order to reduce coefficient of the variation between the measurements to 50 percent (average of 12 counts treated as an individual measurement).

Filamentous Bacteria Identification

Identification of filamentous bacteria was performed according to a procedure developed by Eikelboom and Van Buijsen (1981) and subsequently modified and supplemented by Strom and Jenkins (1984). The basis for the identification of filament type was a microscopic evaluation. The key morphological characteristic for identification was filament diameter. Other characteristics for identification included the presence of branching, attached unicells, crosswalls, inclusions; filament length, shape and location and shape of individual cells. Finally, two staining techniques, Gram and Neisser (after Eikelboom and Van Buijsen, 1981) were used to provide additional information.

An interference contrast microscope with magnifications from 100x to 1,000x was used for microscopic evaluations. Filament diameter was measured with the aid of anocular micrometer scale at 1,000x

magnification. Additionally, high resolution images of cell structure were obtained with a JOEL -100S electron microscope for several samples.

Other Analytical Methods

Monitoring of the substrate removal kinetics by a lumped parameter such as COD was aided on several occasions by analysis for a specific substrate, i.e., glucose. The analysis was performed using a Boehringer Mannheim Biochemicals test kit.

The remaining analytical work conducted during this study was performed according to Standard Methods (1981). This included: BOD_5, MLSS/MLVSS, TSS/VSS, NO_3-N, NH_4-N, TKN, PO_4-P. The only exception was the use of an NO^-_3 specific electrode for NO_3-N measurements performed in Phase III.

Experimental Procedures

Monitoring of PRZ Efficiency

During all phases of the experimental work, the PRZs' performance in terms of COD removal efficiency was routinely monitored. In order to obtain a representative sample of PRZ content, the feed flow and sludge recycle were stopped for several seconds and the contents of the PRZ stirred prior to sample withdrawal. The sample was then immediately centrifuged in order to separate liquid phase from the bulk of the biomass, and subsequently filtered.

Despite minimizing the volume of the samples withdrawn from the system, it was recognized that PRZ sampling was disrupting the systems' performance. Therefore, after each sampling event a minimum of three HRT's were allowed to elapse before the sampling was repeated.

Since different operating parameters of the PRZs was likely to lead to the development of biomasses with different characteristics, a separate experiment was designed to assess effect of the PRZs operating parameters on the substrate removal kinetics in the same reactor. In this experiment the operating parameters of the PRZ in the same reactor were changed at relatively short time intervals (1 to 2 hr). During the test, wide ranges of PRZ contact times and floc loads were tested in the same reactor. In this way, the possibility of a change in the sludge characteristic by physiological adaptation was minimized. During the experiment, the sludge recycle rate was initially kept constant, and HRT in the PRZ was changed, simply by changing the contactor volume. After each change, the system was allowed to reach a hydraulic equilibrium for a time longer than 3 HRT (in the contactor) before the samples were collected. Another three HRT were allowed to elapse before the duplicate sample was collected. After the process of changing HRT (PRZ volume) in the PRZ was completed, the sludge recycle rate was readjusted and process of changing PRZ volume repeated.

Track Studies

In order to document the dynamic behavior of the intermittently aerated reactors, their performance was monitored for several hours so

the data for all phases of the 3 hr cycle were collected. During such intensive studies, dubbed "track studies", COD and oxygen uptake rates were measured in the PRZs and aeration basins at short time intervals, along with flow rates, pH, temperature, DO and MLVSS.

Each reactor was subject to a track study at least once during the experimental work, after a complete acclimation of the reactor was achieved.

Batch Tests For Biosorption Capacity

One of the objectives of the study was to determine the effect of the PRZ design on the biosorption capacity of the activated sludge cultivated in the system. The biosorption capacity was defined as the ability of the sludge to remove the soluble substrate upon a short contact time in excess of a steady state utilization rate. One measure of the biosorption capacity of the sludge was biosorption efficiency as measured under continuous flow conditions in the PRZs. However, the operating parameters were different for each PRZ and a direct comparison of sludge characteristics was not possible from these data. In order to provide such a direct comparison, parallel batch tests were performed on sludge samples from all operating reactors. The tests were designed so to allow a quick, direct comparison of the initial substrate removal rate under a standard condition.

The tests were performed concurrently with sludge samples from all the reactors. Before a sludge sample was withdrawn for the test, the PRZ discharge was stopped for about 10 sec and aeration basin contents allowed to mix. In this way, the possibility of having

unrepresentatively high background COD in the sludge sample through short-circuiting from PRZ was minimized. Then the sludge sample (250 ml) was transferred to a beaker, dosed with 100 ml of reactor feed and aerated. Soluble COD concentrations in the mixed liquor were measured before the feed addition, and at 5 and 15 min into the aeration. MLVSS in each test beaker was determined at 30 min after the feed addition.

The batch tests were performed at approximately weekly intervals. In order to eliminate potential variability of sludge biosorption capacity during an aeration cycle, the sludge samples were always withdrawn at the end of the aeration phase.

For the calculation of the results in terms of mg COD sorbed per g of MLVSS, the initial MLVSS concentration was used. This was calculated by subtracting an estimated 40 mg/l increase in MLVSS due to the substrate biosorption/assimilation during the test from the MLVSS measured at 30 min into the aeration cycle.

Batch Studies on Kinetics of Substrate Removal

A series of experiments were performed on kinetics of substrate removal in batch tests under both aerobic and anaerobic conditions. The primary objective of the aerobic test was to develop a correlation between substrate removal kinetics in a batch-type test and in a PRZ in a continuously operating reactor.

During the preliminary batch tests, the procedure was as follows. A 2 l sludge aliquot was removed from the reactor and dosed immediately with a concentrated feed solution in order to provide a desired initial

COD concentration. COD concentrations, MLVSS and OUR were determined at short intervals for several hours during which the suspension was aerated. Considering the relatively short contact time in the PRZ (less than 30 min), of particular interest in batch test was substrate removal rate in the initial phase of an experiment, i.e., the first 30 min. In the subsequent experiments the procedure was modified in order to increase accuracy of data in this initial phase of the tests.

In a conventional batch test, a sludge aliquot is dosed with a relatively small volume of a concentrated substrate. As a consequence, the initial concentration of the substrate was calculated from a volumetric dilution ratio. In the experiment with multicomponent substrate where an unspecific, lumped parameter for measuring substrate concentration such as COD, TOC or BOD_5 was used, background organics concentration in the sludge sample had to also be taken into account. Such a procedure was burdened with several sources of potential error. Firstly, volumetric dilutions were involved, both during the test (dosing concentrated feed) and for determining substrate concentration in the stock feed solution. The initial concentration in the batch test was calculated from analytical data obtained in either a different range of an analytical test or through factoring in volumetric dilution ratio. Secondly, the background organics concentration in the sludge sample, presumably unavailable for assimilation, had to be substracted from the TOC or COD values measured during the test runs.

In order to eliminate such errors, the sludge sample withdrawn from the reactor was placed on a framed piece of filtering cloth,

dewatered, rinsed with distilled water, and transferred into a graduated cylinder. At the onset of the test, one hundred ml of concentrated sludge was then added to 650 ml of diluted feed, raising the volume only by a small factor. In this manner, the initial substrate concentration was measured directly, with only a minor correction factor for volume (0.87) involved. It was determined that under the test conditions, the supernatant COD in the concentrated sludge sample was in the range 3 to 5 mg/l, resulting in the negligible COD addition after dilution with the feed. The initial COD concentration as well as concentrations in the critical first minutes of the test run were than measured side by side in the same COD test range, providing maximum analytical accuracy.

A series of batch kinetic tests were performed under anaerobic conditions. In these tests, 250 ml of diluted feed was placed in an anaerobic test apparatus (Figure 4.6) and purged with N_2 to remove any dissolved oxygen. A sludge sample prepared as for an aerobic batch test was added into the reactor, the vessel sealed, and then mixed with N_2. The samples were withdrawn from the reactor by closing the gas escape tube thus forcing liquid out the sample tube.

Fed Batch Reactor Experiments

Results of the conventional batch tests and performance of the PRZ in continuous flow reactors provided only indirect data on the actual biosorption capacity of the sludge. Biosorption is defined in this case as the ability of the biomass to remove soluble substrate from the solution at a rate higher than the maximum steady state utilization

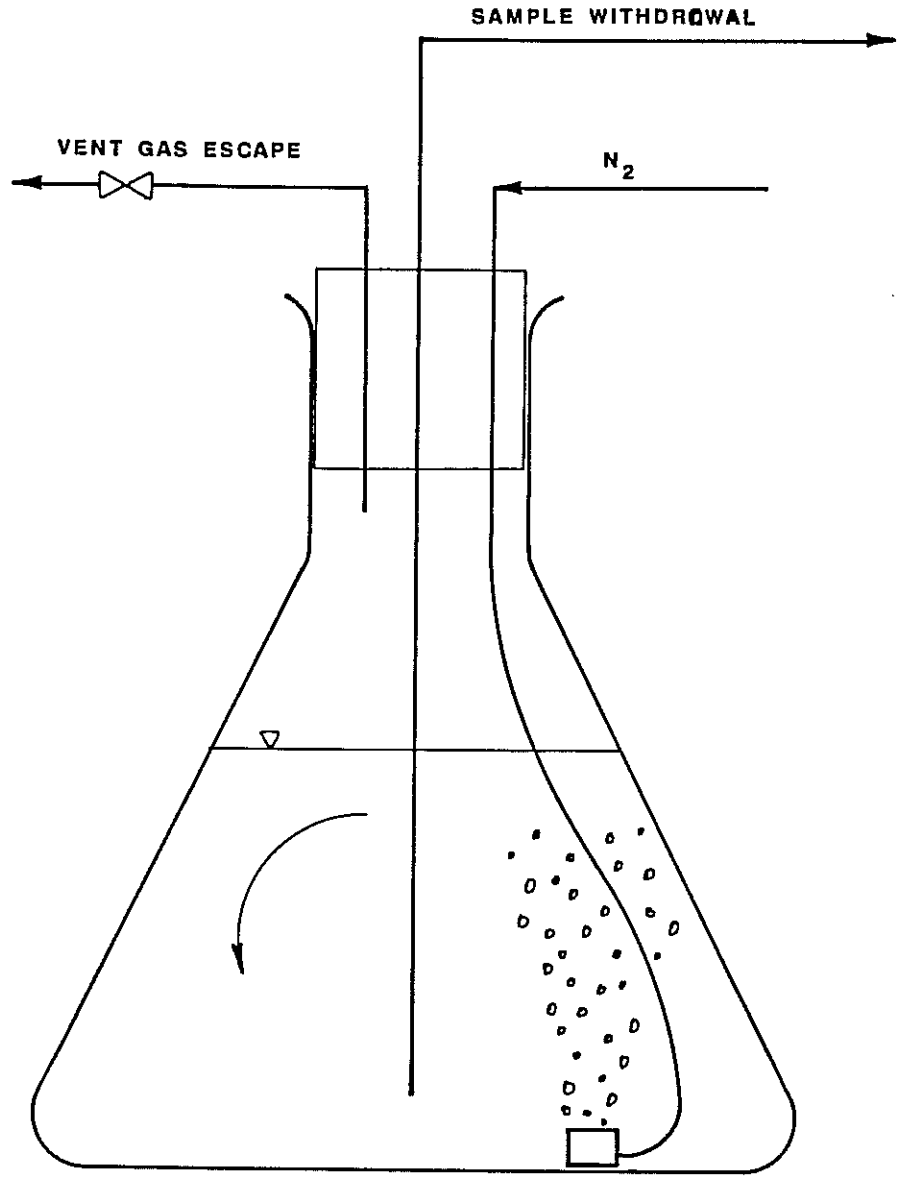


Figure 4.6. Apparatus for Anaerobic Batch Tests.

rate. To directly evaluate the actual biosorption capacity, fed batch reactor (FBR) tests were carried out.

An FBR reactor consists of an aerated vessel into which a concentrated feed is added continuously during the duration of the experiment. Initial FBR volume and feed flow rate are selected so that the feed volume added to the FBR during the test is small compared to the initial volume. Consequently, the sludge concentration in the vessel and the substrate addition rate (in terms of mg COD/l-hr) do not change appreciably during the duration of the test. Moreover, withdrawal of reactor contents for the sampling approximately compensates for the increase in volume due to the substrate addition. That also compensates for the build-up of biomass concentration due to growth, and for the decrease in substrate addition rate (per volume or per mass basis).

The FBR test was originally developed by Williamson and McCarty (1975) for rapid determination of nitrification kinetic coefficients. Philbrook and Grady (1985) used FBR techniques for evaluation of biodegradation kinetics for priority pollutants (2-chlorophenol). Watkin (1986) adapted the FBR procedure for determination of inhibitory effects of wastewaters on biomass. The FBR test was adopted in this study to identify biosorption capacity of the sludge since the procedure offers a unique way of directly measuring this sludge characteristics.

The theoretical response of an FBR test with sludge exhibiting biosorption is given in Figure 4.7. The biosorption effect is present if the linear portion of the measured build-up of the substrate concentration in the reactor intersects the ordinate below the origin of the abscissa. The biosorption capacity of the sludge in mg of substrate per g of biomass can then be read directly from the Y-axis.

FBR results from this study were processed using a simple computer program which calculated the expected substrate concentration (assuming no reaction) for each sampling interval. The program applied appropriate corrections due to the small changes in reactor volume and substrate withdrawal with the samples. The substrate uptake in each interval was calculated as a difference between expected and measured substrate concentration at the end of the sampling interval.

The substrate uptake was expressed in specific terms using a sludge concentration calculated for each sampling interval from a linear correlation of the experimental data on MLVSS concentration. The specific substrate uptake data were then used for construction of a cumulative plot from which the reaction rate and biosorption capacity were read directly (Figure 4.8).

FBR tests were performed in an open beaker with an initial volume of 2 l. A sludge sample for FBR testing was withdrawn from the continuous reactor and aerated for 10 minutes prior to the test start. Feed prepared from the feed concentrate was pumped at the rate of about 2 ml/min. During the testing period of two to three hours, COD, SOUR

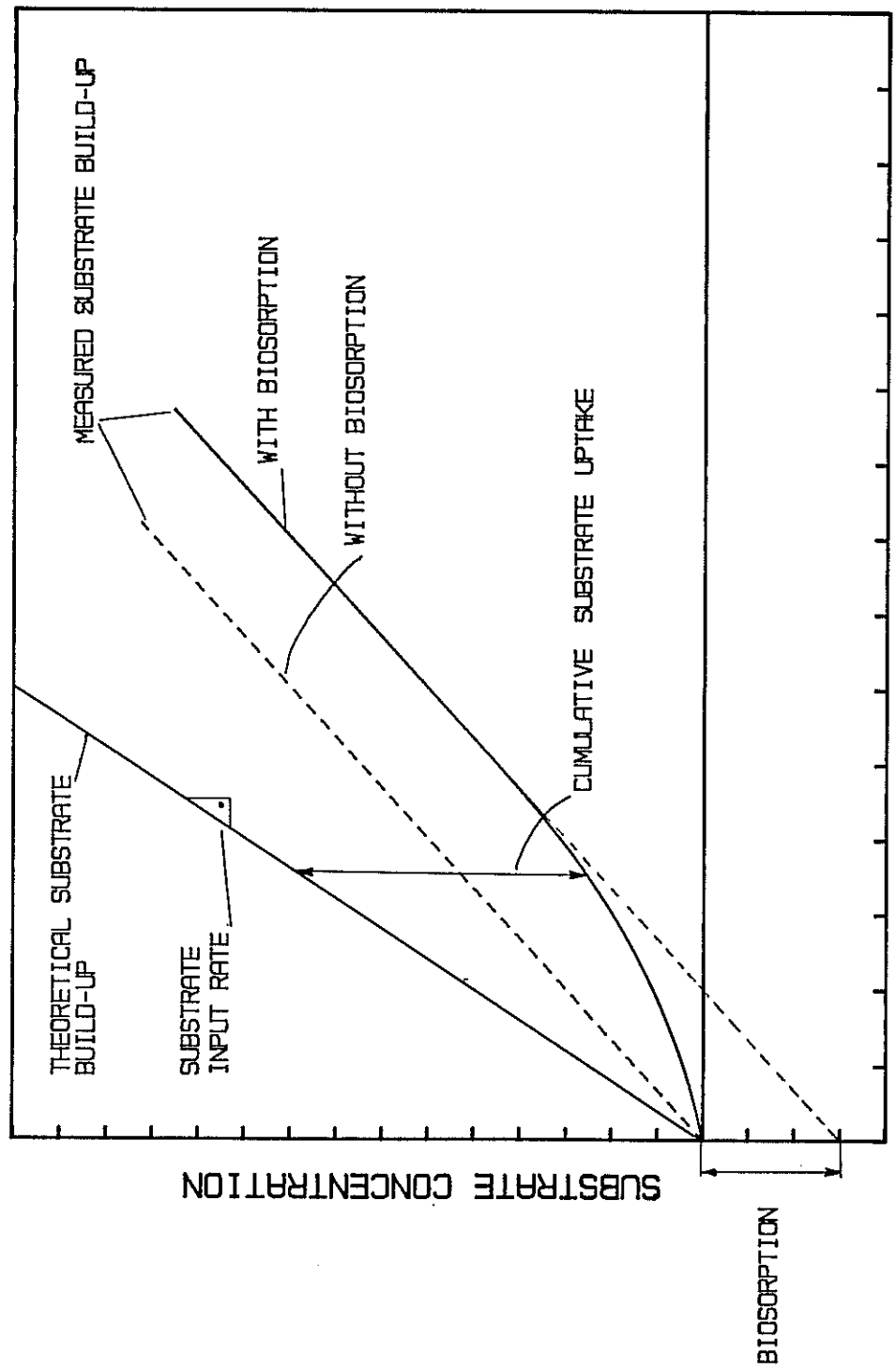


Figure 4.7. Theoretical Response of a Fed Batch Reactor (FBR).

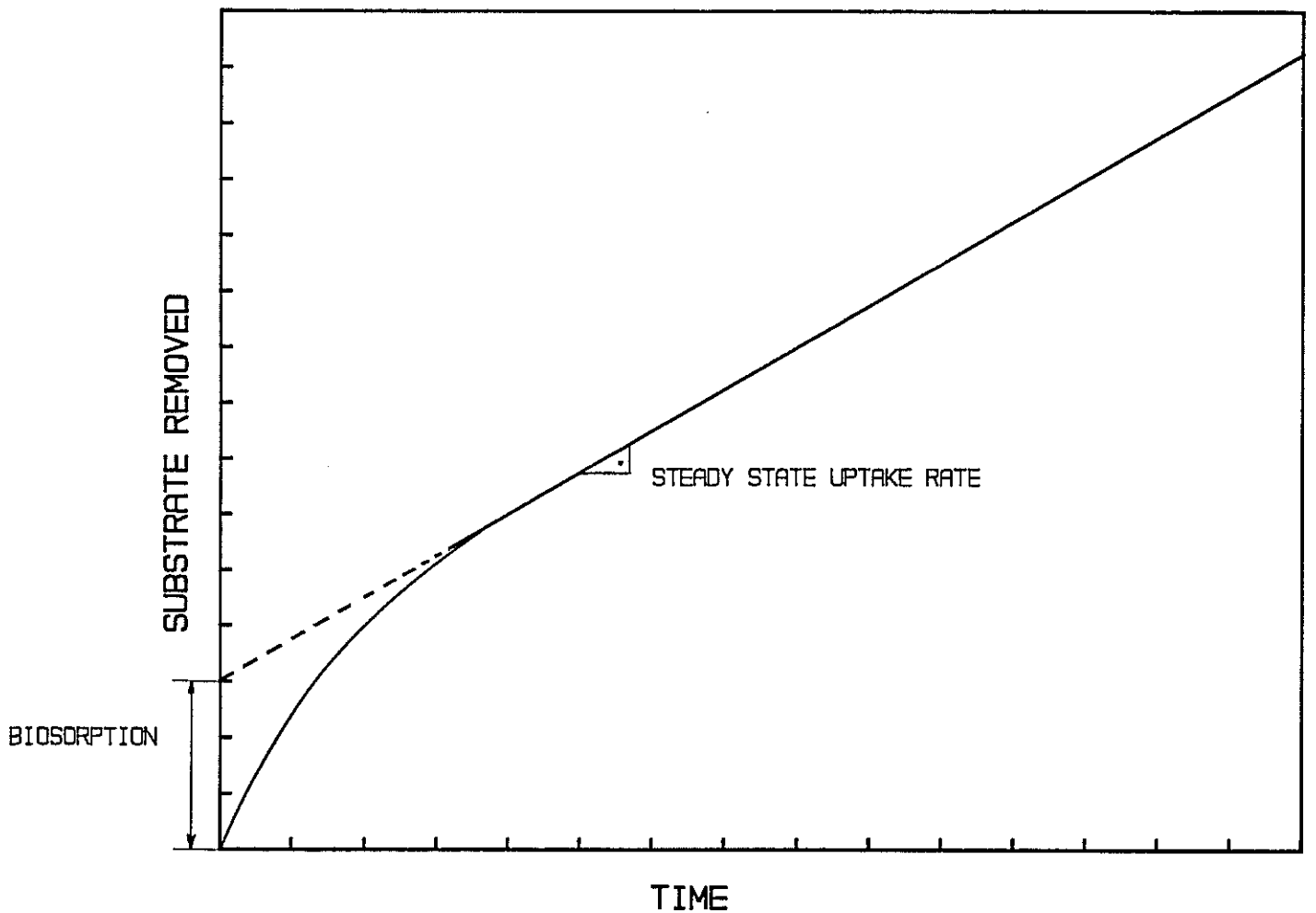


Figure 4.8. Cummulative Plot of Substrate Uptake in Fed Batch Reactor Test.

CHAPTER V

RESULTS AND DISCUSSION - PHASE I

PRELIMINARY WORK

Effluent Quality

During the preliminary work, four activated sludge reactors were operated with operating parameters provided previously in Table 4.3. Table 5.1 presents the average effluent quality data from that period. It is apparent that almost complete removal of the influent BOD_5 resulted, with residual effluent BOD_5's below 3.1 mg/l for all the reactors. In all intermittently aerated units almost complete nitrification was achieved, with effluent $N\text{-}NH_4$ generally below 1 mg/l (Table 5.1). Nitrification in the continuously aerated reactor (No. 4) was inhibited by daily addition of 1-Allyl 2-Thio-Urea (ATU). Effluents from the intermittent units averaged 7.0 mg $N\text{-}NO_3$/l, which yields an overall denitrification efficiency of about 75 percent. The mechanism of nitrification-denitrification is discussed in more detail later, along with OUR results obtained during the track studies.

COD Profiles from Track Studies

COD and OUR profiles obtained from track studies, the scope of which was detailed in Chapter IV, are presented in Figure 5.1 through 5.3. No significant pattern in COD concentrations are present in PRZs

TABLE 5.1

AVERAGE PERFORMANCE OF THE REACTORS-PHASE I

	Aeration Mode	Floc Load in PRZ (mg COD/g VSS)	COD (mg/l)	BOD_5 (mg/l)	TKN (mg/l)	NH_4-N (mg/l)	NO_3-N (mg/l)
Influent	--	--	182	118	29.1	5.9	--
Effluent							
Reactor 1	Intermittent	200	9.3	1.5	--	0.49	5.3
Reactor 2	Intermittent	20	10.5	2.0	--	0.66	8.0
Reactor 3	Intermittent	--	10.0	3.1	--	0.47	7.6
Reactor 4[a]	Continuous	--	12.6	2.4	--	24	3.2

[a]Nitrification in Reactor No. 4 was inhibited by daily addition of ATU.

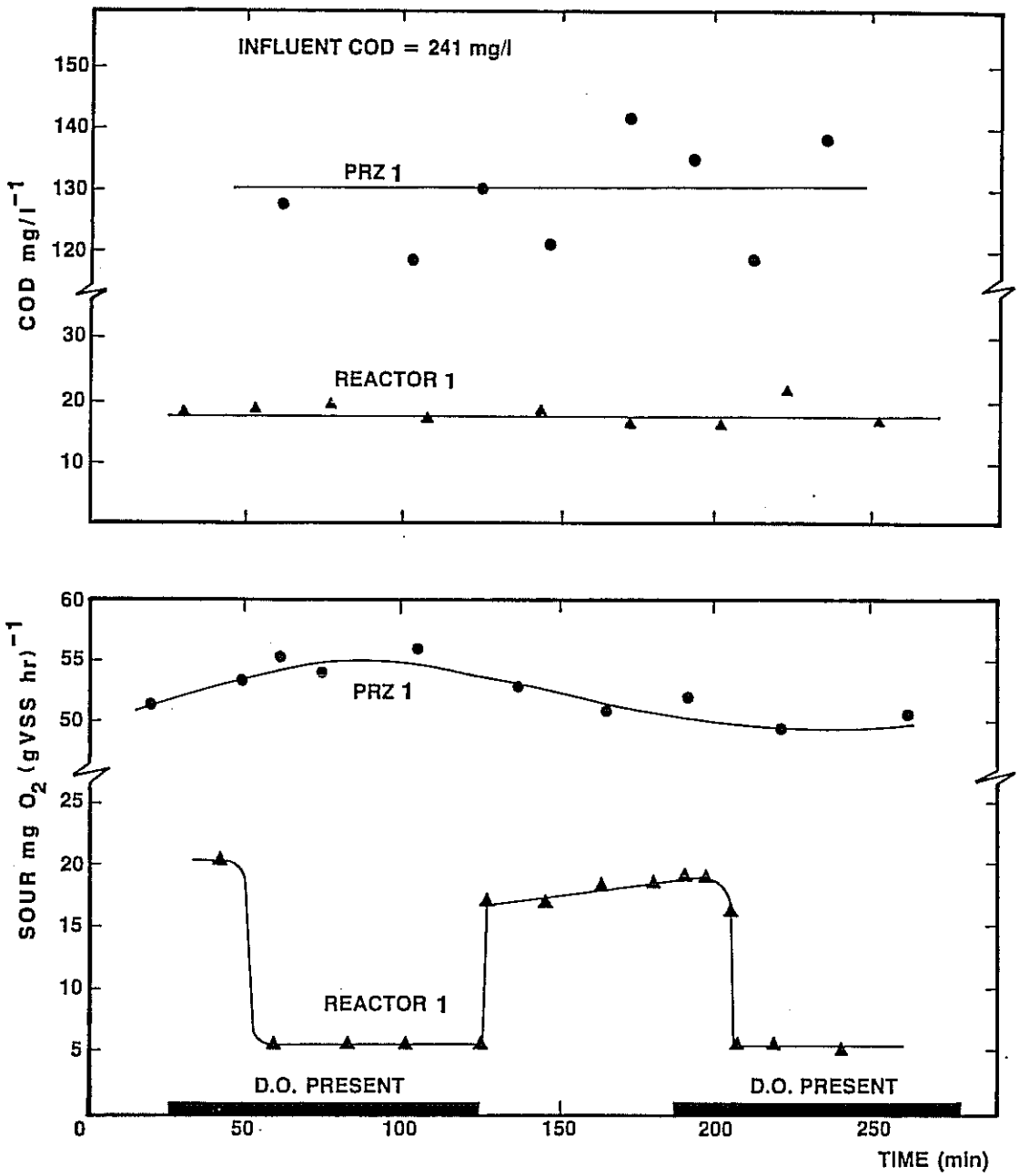


Figure 5.1. COD and Specific Oxygen Uptake Rate Profiles in an Intermittently Aerated Reactor No. 1.

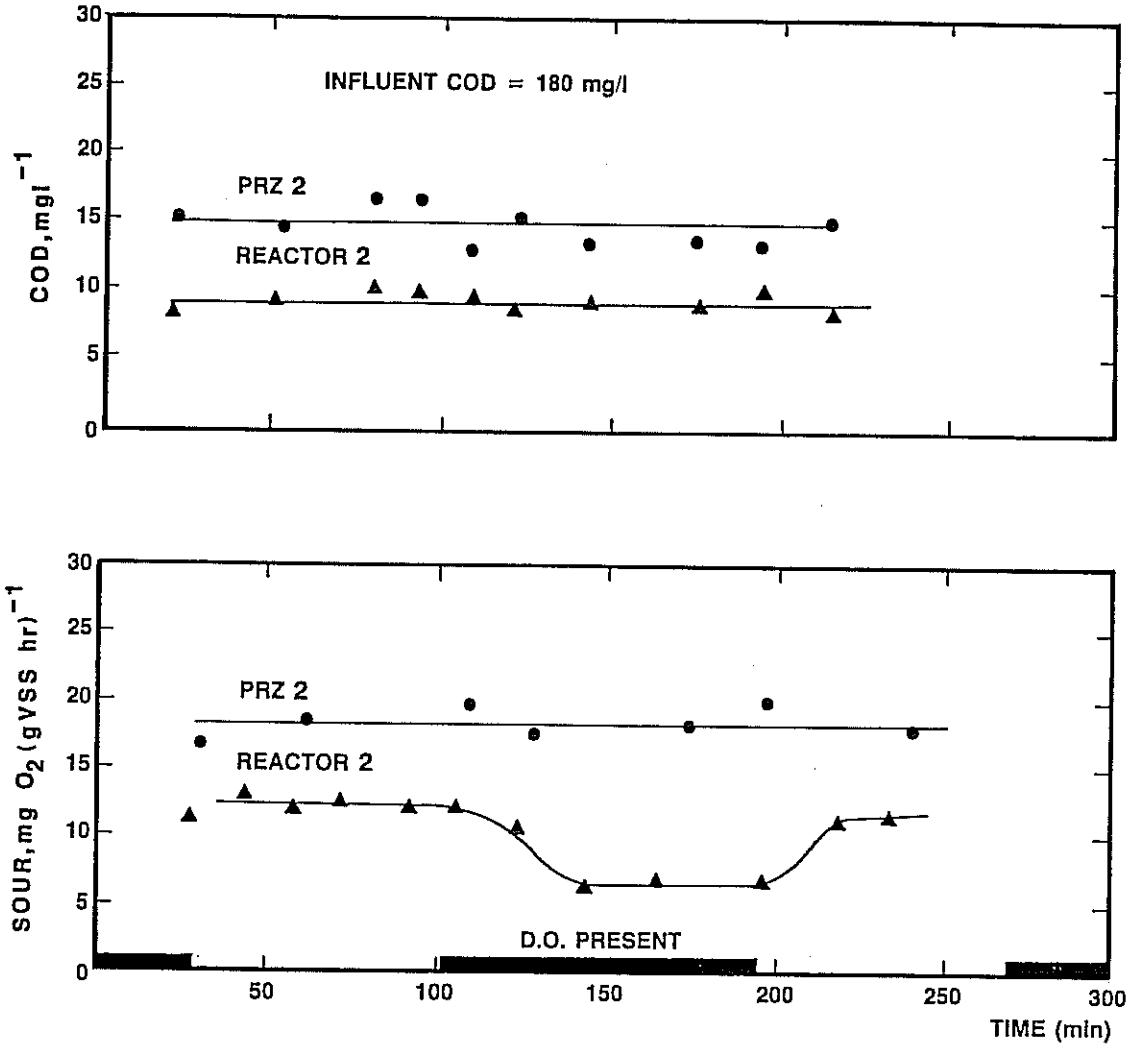


Figure 5.2. COD and Specific Oxygen Uptake Rate Profiles in an Intermittently Aerated Reactor No. 2.

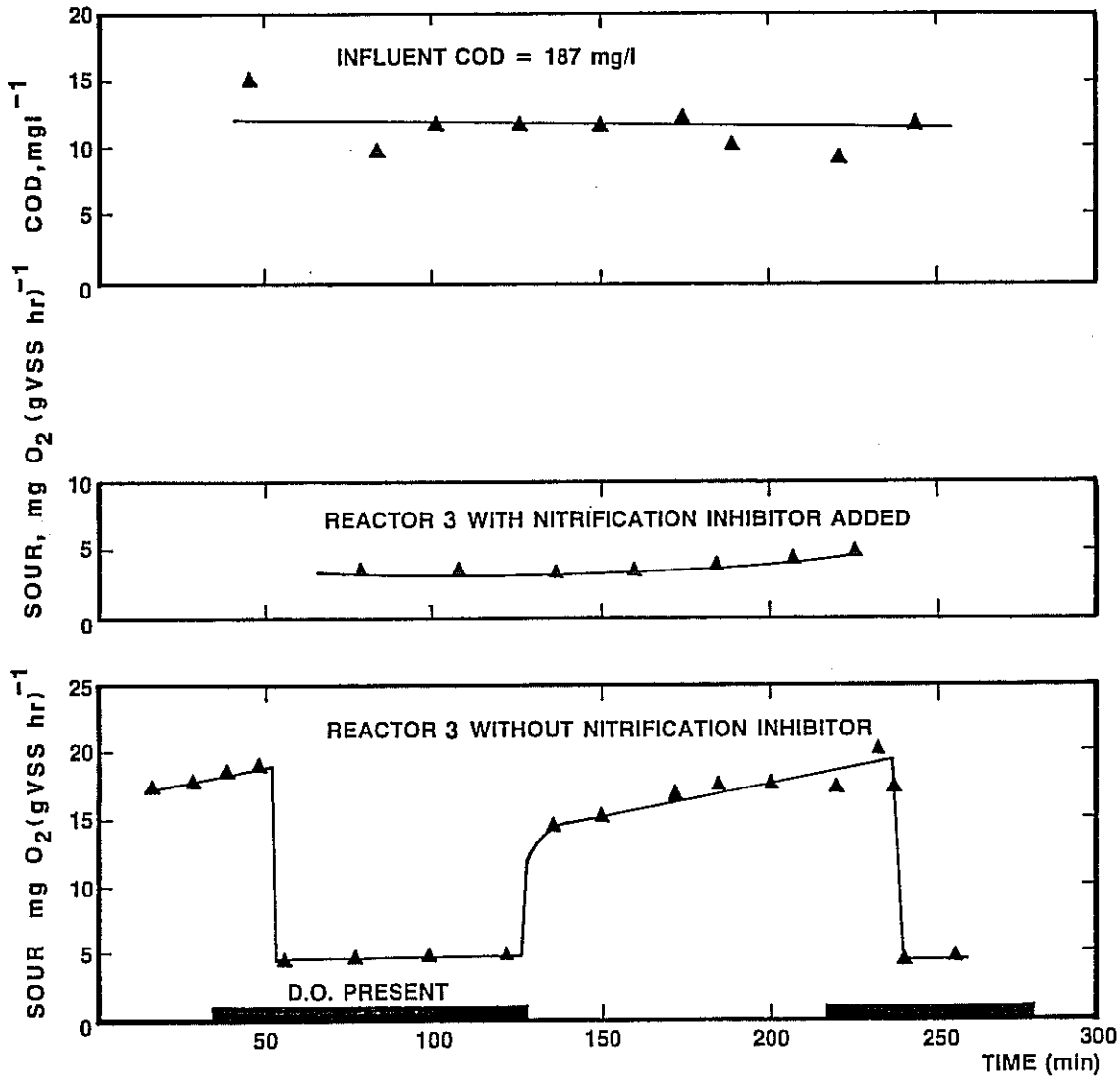


Figure 5.3. COD and Specific Oxygen Uptake Rate Profiles in an Intermittently Aerated Reactor No. 3.

or in the main tanks attributable to the different phases of a cycle. Lack of such a pattern can be explained by one or more of the following factors:

- Hydraulic buffering capacity of the reactor.
- Utilization of bound oxygen (nitrates) for carbon oxidation.
- Capability of sludge to remove soluble substrate without a terminal electron acceptor.
- Substrate uptake in the PRZ.
- Influx of DO from the continuously aerated PRZ or through surface diffusion.

Let us first examine the ability of a reactor to buffer the COD variations on a purely hydraulic basis. If any such variations were to be found, they should be most pronounced in the Reactor No. 3 with no PRZ. The mass of COD entering the reactor during 1.5 hr of the anoxic phase was 182 mg COD/l x 0.0187 l/min x 60 min/hr x 1.5 hr = 305 mg COD. If the sludge was not capable of removing soluble substrate during the air-off period, the accumulation of COD would have increased the mixed liquor (ML) COD by about 305 mg COD/20 l = 15 mg/l. The lack of such an increase (Figures 5.1 through 5.3) indicates that the biomass is able to remove substrate without the presence of DO.

Nitrate concentration at the end of the aeration period (indirectly measured) can be estimated from the average effluent data (Table 5.1). Since during the air-off phase NO^-_3 is biochemically reduced, the concentration of NO^-_3 at the end of the aeration phase was

no less than the 7.6 mg/l measured for the composite effluent. During denitrification, 1 mg of nitrate-nitrogen serves as equivalent of about 3.6 mg of oxygen, hence the total amount of O_2 available from denitrification was up to 28 mg O_2/l, a value which is sufficient for the 15 mg O_2/l COD which could theoretically accumulate during the air off phase.

In the reactors with PRZs, the potential for variations in ML COD is further reduced by substantial uptake of COD in the PRZ during all phases of operation.

It can be concluded, therefore, that in intermittently aerated systems with similar COD:TKN ratios, significant variations in effluent organic concentration are unlikely, unless caused by a direct short-circuiting.

Oxygen Uptake Rates

Specific oxygen uptake rates (SOUR) show extensive oscillation during the course of an aeration cycle (Figures Nos. 5.1 through 5.3). In all the reactors the uptake rates were high at the start of aeration and remained at that level for about 20 min into the aeration phase. At that moment, the uptake rates decreased abruptly to a new, low level and stayed relatively constant for the rest of the aeration period. After the end of the aeration period no changes in SOUR were noted for another 15 min, during which DO was still present in the reactors. As soon as the reactors became anoxic, the uptake rates rapidly increased when the sample was aerated. It should be noted that the uptake rates measured during the absence of DO in the reactor should be considered

as "potential" since they were exerted only in the SOUR vessel. In reactor No. 2 (large PRZ) the potential SOUR remained virtually constant throughout the remainder of the air-off phase and the initial period of aeration (Figure 5.2). In reactor No. 1 (small PRZ) potential SOUR showed tendency to increase slightly (from 17 to 19 mg O_2/g VSS-hr) during that time (Figure 5.1). In reactor No. 3 (no PRZ) this tendency is more pronounced and SOUR increased from 14 to 20 mg O_2/g VSS-hr (Figure 5.3).

Abrupt decreases in SOUR after the initial 20 min of aeration are also illustrated by changes in DO concentration in the reactor (Figure 5.4). After the start of aeration, the DO increased within a few minutes to 1.7 mg O_2/l and stabilized at that level for about a 20 min period, which corresponded with the period of high SOUR on Figure 5.2. After SOUR decreased abruptly, DO significantly increased and reached 5 mg/l. When the aeration stopped, the DO dropped linearly to undetectable levels within 15 min. DO profiles analogous to the one depicted on Figure 5.4 were obtained for all the intermittent reactors.

The abrupt changes in SOUR were found to be due to the occurrence of nitrification. When ATU (nitrification inhibitor) was added to the reactor, SOUR during the cycle showed no abrupt changes (Figure 5.3). Steady SOUR when nitrification was inhibited is illustrated also by a quick increase in DO to a steady state value (Figure 5.4), without the two plateaus characteristic of operation with uninhibited nitrification.

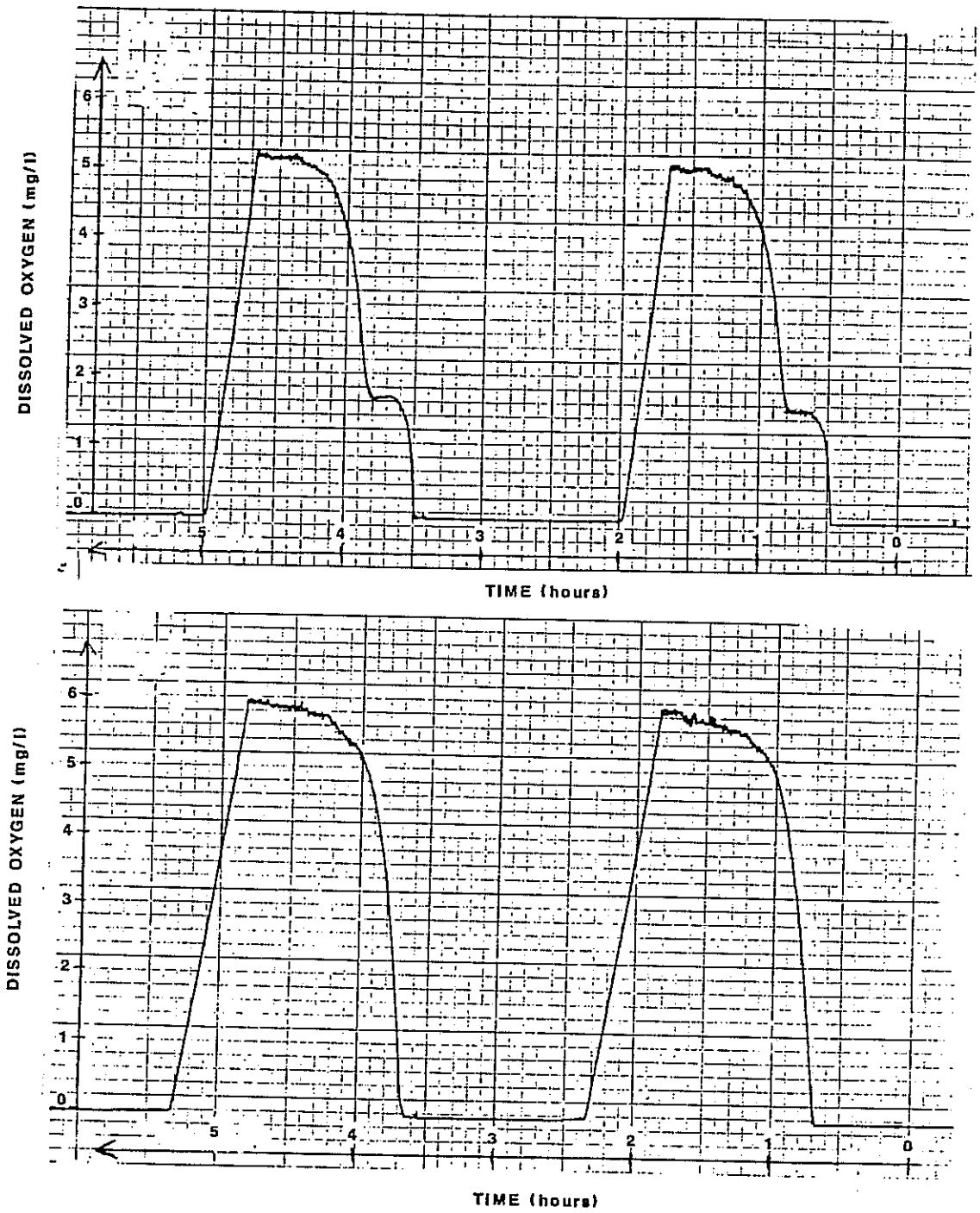


Figure 5.4. Dissolved Oxygen (DO) Profiles in Reactor No. 2 without (Top) and with (Bottom) Nitrification Inhibitor.

The ratio of the initial to final SOUR during the aeration period was affected by the PRZ presence and volume. For reactor No. 3, without a PRZ, that ratio was 3.8:1 (Figure 5.3). Presence of the continuously aerated PRZ decreased that ratio to 3.1:1 for reactor No. 1, with a small PRZ (Figure 5.1), and to 2:1 for the reactor No. 2, with a large PRZ (Figure 5.2). Since the amounts of bound oxygen available from denitrification should have been about the same in all the reactors, it appears that DO supplied to the PRZs throughout the cycle is used for nitrification, and further, that it decreases a high SOUR at the beginning of aeration.

Changes in SOUR from low to high and high to low levels were found to be very rapid and to occur within seconds. From a single sludge sample taken shortly before the change had occurred, two straight oxygen depletion curves were frequently recorded, with two slopes corresponding to a high and low OUR (Figure 5.5).

Swift changes in SOUR were characteristic of operation with uninhibited nitrification. It was found that the reaction proceeds as a zero order rate to a very low substrate (NH_4) concentrations (Huang and Hopson, 1974). Assuming that for nitrification the oxygen uptake parallels substrate removal, nitrification would continue at a high, constant rate with a constant nitrogenous oxygen uptake until all accumulated ammonia-nitrogen is converted to the nitrate. At that moment the nitrogenous oxygen uptake rate abruptly decreases to a value corresponding to the rate of TKN input into the reactor, or more precisely, to the sum of the ammonification rate and NH_4-N input.

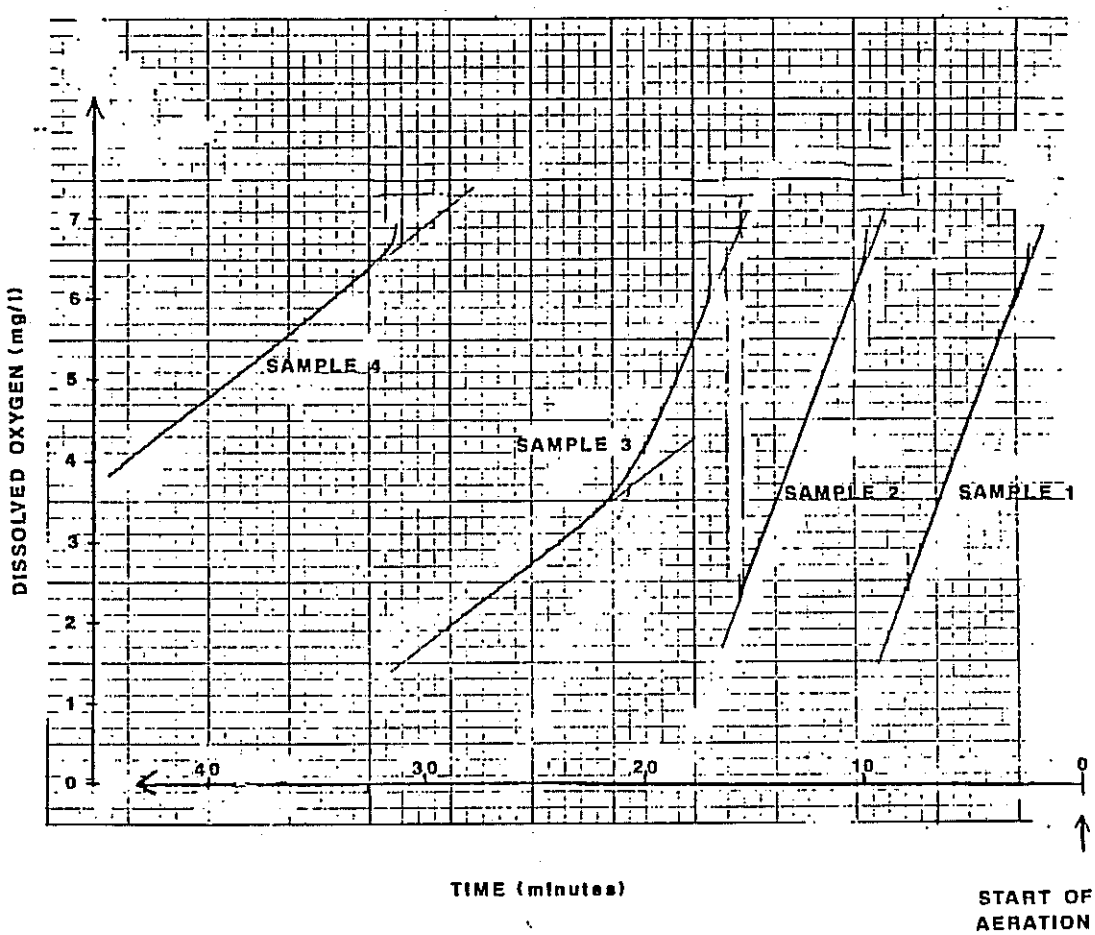


Figure 5.5. Illustration of Abrupt Changes in Oxygen Uptake Rates (OUR) in Reactor No. 1 (Slopes of the Oxygen Depletion Curves are Proportional to OUR. In Sample No. 3 Two Distinctly Different Slopes are Observed).

Performance of the PRZs

For the discussion of the PRZs' performance the following terms are defined after Eikelboom (1982):

1. Floc loading (F)- mass flow of available substrate into the PRZ per mass flow of VSS (mg COD/g MLVSS).
2. Floc uptake (B) - mass of substrate taken up by a unit mass of MLVSS in the PRZ (mg COD/g MLVSS).
3. Biosorption efficiency (E) - percent of available (biodegradable) substrate taken up in PRZ.

From the substrate mass balance around the PRZ (assuming no growth) the following formulas apply with nomenclature shown in Figure 5.6:

$$F = Q_o COD_o/(Q_R X_R) \quad (5.1)$$

$$B = [Q_o COD_o + Q_R COD_R - (Q_o+Q_R)COD_1]/(Q_R X_R) \quad (5.2)$$

$$E = (B/F)\ 100 \quad (5.3)$$

In the original definition of the floc loading (Eikelboom, 1982), concentration of the COD in the recycle stream was considered to be the nonbiodegradable part of influent COD and therefore it was subtracted from the influent concentration with the resulting formula:

$$F = \frac{Q_o(COD_o - COD_R)}{Q_R X_R} \quad (5.4)$$

In this study, a synthetic, completely biodegradable feed was used and therefore it was assumed that COD in the recycle stream consisted of metabolic byproducts and lysed material, rather than of

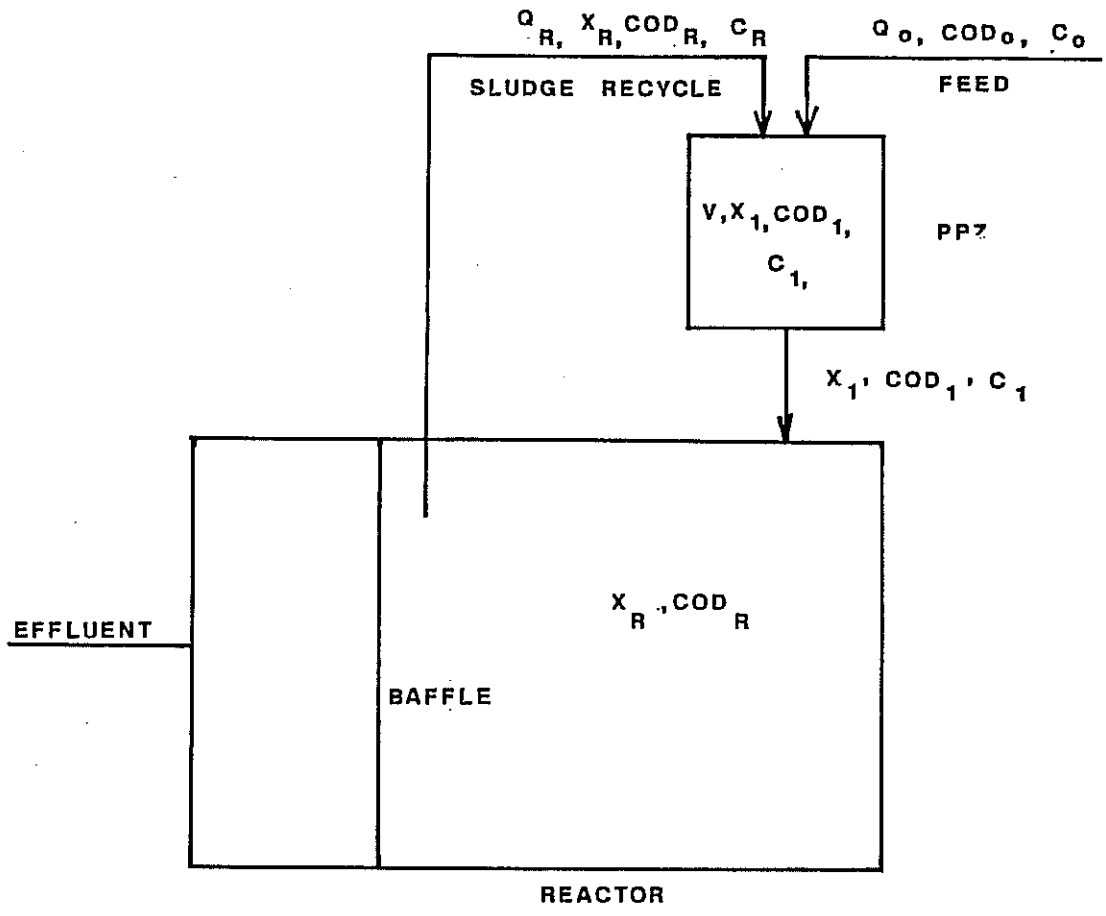


Figure 5.6. Graphical Presentation of the Nomenclature Used for PRZ.

nonbiodegradable part of the feed. Consequently, definition of floc loading was modified as shown in Formula (5.1).

Data collected on substrate uptakes in the PRZs are summarized in Tables 5.2 and 5.3. The tables include the average results from track studies and data from routine analyses performed during Phase I. The low loaded PRZ removed about 80 to 90 percent of the biodegradable COD, while the PRZ with a high floc loading removed 40 to 55 percent. The reaction rates in the PRZs can be calculated from the following mass balance (nomemclature as in Figure 5.6).

$$V(dCOD_1/dt) = -VX_V R_r + Q_o COD_o + Q_R COD_R - (Q_o+Q_R)COD_1 \quad (5.5)$$

Since the PRZ operates at a steady state, the term on the left side is equal to zero. The specific reaction rate, R_r, was then calculated from the final formula.

$$R_r = [Q_o COD_o + Q_R COD_R - (Q_o+Q_R)COD_1]/(VX_V) \quad (5.6)$$

The PRZs' performance data are summarized in Tables 5.2 and 5.3, and are included in the discussion on modeling of PRZ performance presented in Chapter VII.

Sludge Settling Characteristic

The history of ZSV in the four reactors operated during Phase I is presented in Figure 5.7. Interpretation of this data is obscured by the fact that the reactors did not operate in the same mode throughout the study. As was previously outlined in Chapter III, operation with intermittent aeration was preceded by a period of operation in a

TABLE 5.2

PERFORMANCE OF PRZ IN REACTOR NO. 1[a]
PHASE I

Date (1984)	C_l[b] (mg COD/l)	F[c] (mgCOD/gVSS)	B[d] (mgCOD/gVSS)	E[e] (%)	F/M[f] (gCOD/gVSS-day)	R_r[g] (gCOD/gVSS-day)
7/30	130	200	49.3	24.6	28.8	7.1
8/5	86.2	166	58.0	34.9	23.9	8.3
8/11	64.5	141	66.7	47.3	20.3	9.6
8/14	67.0	162	75.3	46.5	23.3	10.8

[a]For each test parameters such as feed and recycle flow, feed and recycle COD, and PRZ operating volume were measured (not provided here) and used for calculation of the PRZ performance parameters.

[b]Substrate concentration in the PRZ.

[c]Floc load.

[d]Biosorption.

[e]Substrate removal efficiency in PRZ.

[f]Organic loading in PRZ.

[g]Reaction rate in PRZ.

TABLE 5.3

PERFORMANCE OF PRZ IN REACTOR NO. 2[a]
PHASE I

Date (1984)	C_1[b] (mg COD/l)	F[c] (mgCOD/g VSS)	B[d] (mgCOD/gVSS)	E[e] (%)	F/M[f] (gCOD/gVSS-day)	R_r[g] (gCOD/gVSS-day)
7/22	14.5	15.1	12.5	82.8	2.17	1.80
7/30	20.6	20.3	16.4	80.7	2.92	2.35
8/5	11.9	16.5	15.4	93.1	2.38	2.21
8/11	12.7	13.4	11.3	84.6	1.93	1.63
8/14	10.1	15.5	14.3	92.3	2.23	2.05

[a]For each test parameters such as feed and recycle flow, feed and recycle COD, and PRZ operating volume were measured (not provided here) and used for calculation of the PRZ performance parameters.

[b]Substrate concentration in the PRZ.

[c]Floc load.

[d]Biosorption.

[e]Substrate removal efficiency in PRZ.

[f]Organic loading in PRZ.

[g]Reaction rate in PRZ.

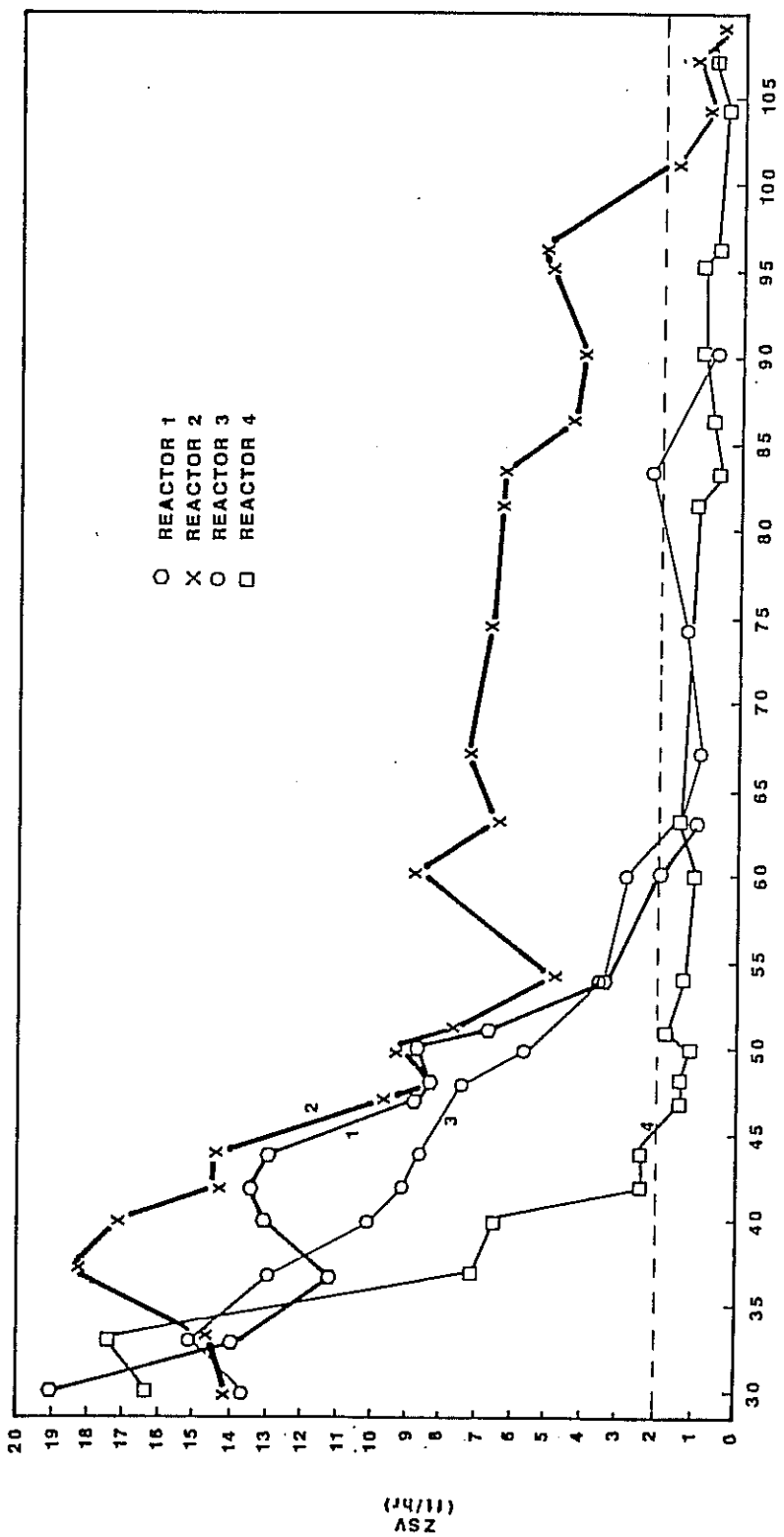


Figure 5.7. Chronological Plot of Stirred ZSV-Phase I.

continuous aeration mode. During that time, in addition to reactors No. 1 and 2, also reactor No. 3 was equipped with a PRZ of an intermediate size. Nevertheless, the control reactor, No. 4, which operated without a PRZ from the onset of the study, rapidly developed a bulking sludge. Reactor No. 3, which initially was equipped with a PRZ and later served as an intermittently aerated control, bulked two weeks later. Reactor No. 1, operating with a PRZ at high floc loading, bulked at the same time as reactor No. 3. Reactor No. 4, equipped with a low loaded PRZ, resisted bulking for 3.5 months.

CHAPTER VI

ANALYSIS OF DATA REPRODUCIBILITY - PHASE II

Parameter Selection

The phase II experiments were designed primarily to compare reproducibility of the parameters of the activated sludges grown in reactors operating in parallel under the same conditions. Parameters selected as the most consequential for the assessment of the significance of the study findings were: sludge settleability (ZSV, SVI), filamentous bacteria count and sludge biosorption characteristic.

Zone Settling Velocity

Zone settling velocities were measured using two protocols: stirred and in situ, as detailed in the methodology section.

The results of stirred ZSV measurements performed during the course of the 4-week testing period are summarized in Table 6.1 and are presented in Figures 6.1. A review of these data indicates that the sludge settling characteristic in all four reactors followed closely the same trend, and that the differences between the reactors were relatively small. Coefficients of variation between the stirred ZSV measurements performed on the same day on the different reactors (Table 6.1) were in the range 3 to 20 percent during the first three weeks of the reactors' operation and increased to 20 to 25 percent in

TABLE 6.1

SUMMARY OF STIRRED ZSV RESULTS (ft/hr)
PHASE II

Date 1985	Julian Day	Reactor No. 1	2	3	4	5	6	Average	Standard Deviation	Coefficient of Variation (%)
5/7	0	11.04	10.87	10.08	10.87	10.87	10.74	10.75	0.31	2.9
5/8	1	6.32	7.05	5.89	6.30	7.74	8.70	7.00	0.97	13.8
5/9	2	4.51	6.79	4.17	5.39	6.22	6.34	5.57	0.97	17.4
5/10	3	4.13	6.18	5.28	3.52	3.54	4.09	4.46	0.97	21.7
5/14	7	2.81	3.90	3.43	4.07	3.60	3.94	3.63	0.42	11.7
5/19	12	3.60	3.29	3.03	3.54	3.54	2.76	3.29	0.31	9.4
5/20	13	3.03	3.41	3.29	2.83	4.06	2.95	3.26	0.41	12.5
5/22	15	2.93	2.28	2.95	2.95	2.76	2.89	2.79	0.24	8.5
5/24	17	2.56	2.48	2.64	2.60	2.42	2.44	2.52	0.08	3.2
5/26	19	2.62	2.46	2.24	2.56	1.83	2.09	2.30	0.28	12.1
5/28	21	2.19	2.19	1.44	1.52	1.56	1.18	1.68	0.38	22.6
5/30	23	1.73	2.01	0.96	1.59	1.38	1.20	1.48	0.35	23.3
6/2	26	1.30	0.93	0.57	1.12	0.83	1.12	0.98	0.24	24.1

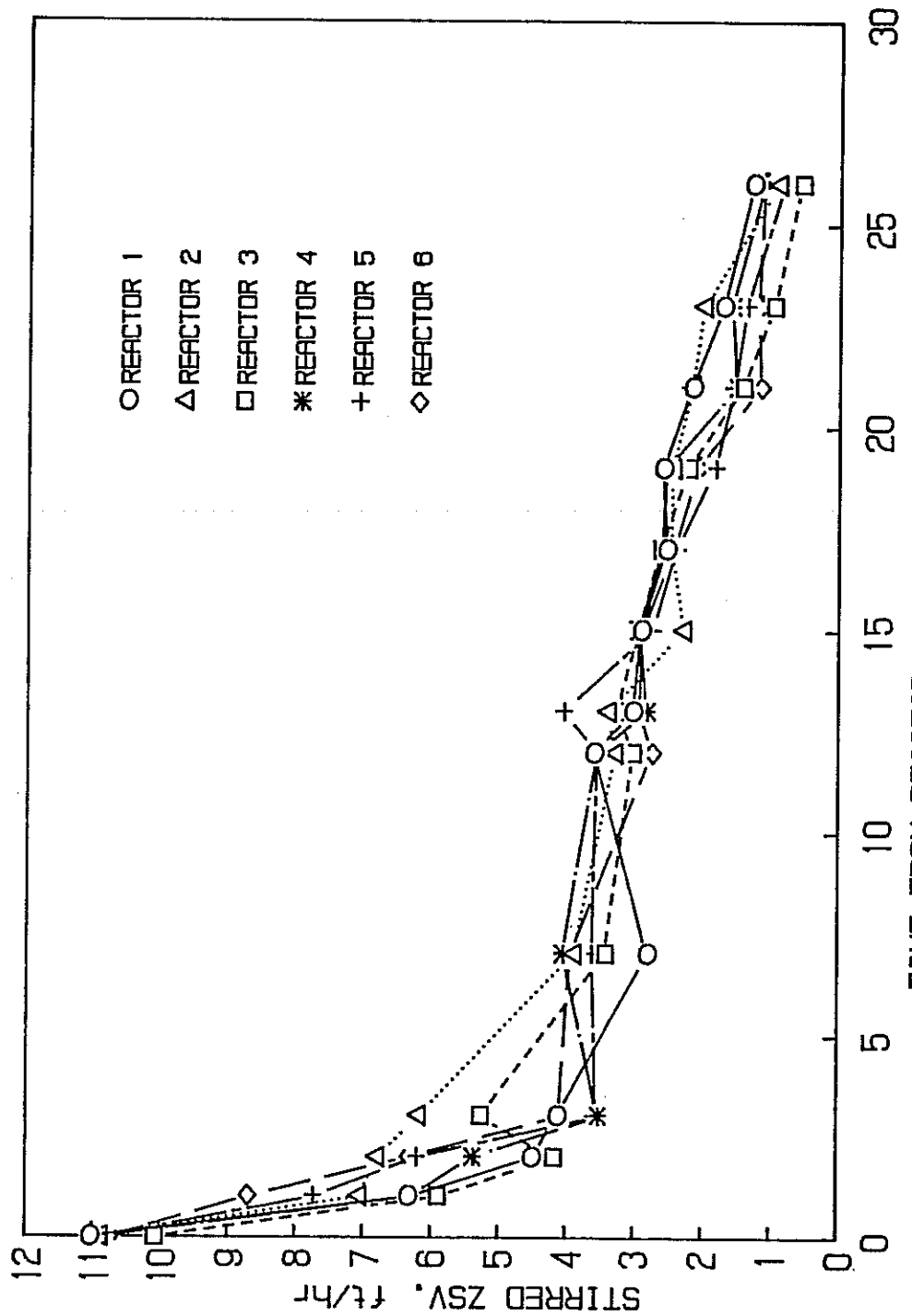


Figure 6.1. Chronological Plot of Stirred ZSV-Phase II.

the last week of the study. At that time ZSV's in most of the reactors dropped below 2 ft/hr, an indication of a bulking condition. The average variation between the stirred ZSV measurements performed on the same day on different reactors was 14 percent.

The in situ ZSV results followed the same trend (Table 6.2 and Figure 6.2), with an average variation of 21 percent between the measurements performed on different reactors on the same day.

Additional experiments were performed in order to determine how the variations in the ZSV data were distributed between measurement errors and true differences in the sludge characteristic between the reactors. Stirred ZSVs were measured on the same day six times in all the reactors. The data are presented in Table 6.3 and results of analysis of variance performed on this data are summarized in Table 6.4. The computed F value for this statistic was F=22.1 and is much greater than the critical value for F distribution (F=2.53), resulting in the conclusion that at least two of the reactors did not have the same ZSV at the test date. The 95 percent confidence intervals for the mean presented in Table 6.3 further illustrate this conclusion.

Another factor contributing to the variations in ZSV measurement in a reactor between different days, as well as in the different reactors on a given day was MLSS concentration. The sludge was wasted daily from all the reactors in order to keep sludge concentration constant at the design MLVSS value of 2,000 mg/l. However, due to the

TABLE 6.2

SUMMARY OF IN SITU ZSV RESULTS (ft/hr)
PHASE III

Date 1985	Julian Day	Reactor No. 1	2	3	4	5	6	Average	Standard Deviation	Coefficient of Variation (%)
5/10	3	10.50	10.00	10.00	10.00	10.50	11.25	10.38	0.45	4.3
5/11	4	7.50	7.75	6.50	7.25	7.88	8.75	7.60	0.68	8.9
5/13	6	5.50	6.25	5.25	5.38	5.75	6.25	5.73	0.40	6.9
5/15	8	5.05	2.18	3.93	3.35	4.05	3.13	3.61	0.89	24.6
5/17	10	1.25	2.23	2.28	1.58	1.80	1.72	1.81	0.36	19.7
5/19	12	1.70	2.40	2.07	1.78	1.85	1.80	1.93	0.24	12.4
5/20	13	1.75	2.13	2.28	2.13	1.65	1.75	1.95	0.24	12.2
5/22	15	1.28	2.03	1.63	1.75	1.13	1.05	1.48	0.35	23.9
5/24	17	0.88	1.18	1.35	1.25	0.75	0.68	1.01	0.26	25.5
5/26	19	0.70	0.70	0.78	0.60	0.28	0.28	0.55	0.20	36.8
5/28	21	0.33	0.50	0.43	0.15	0.18	0.40	0.33	0.13	39.1
5/30	23	0.30	0.25	0.25	0.23	0.13	0.13	0.21	0.07	30.9
6/2	26	0.16	0.18	0.15	0.16	0.25	0.13	0.17	0.04	22.6

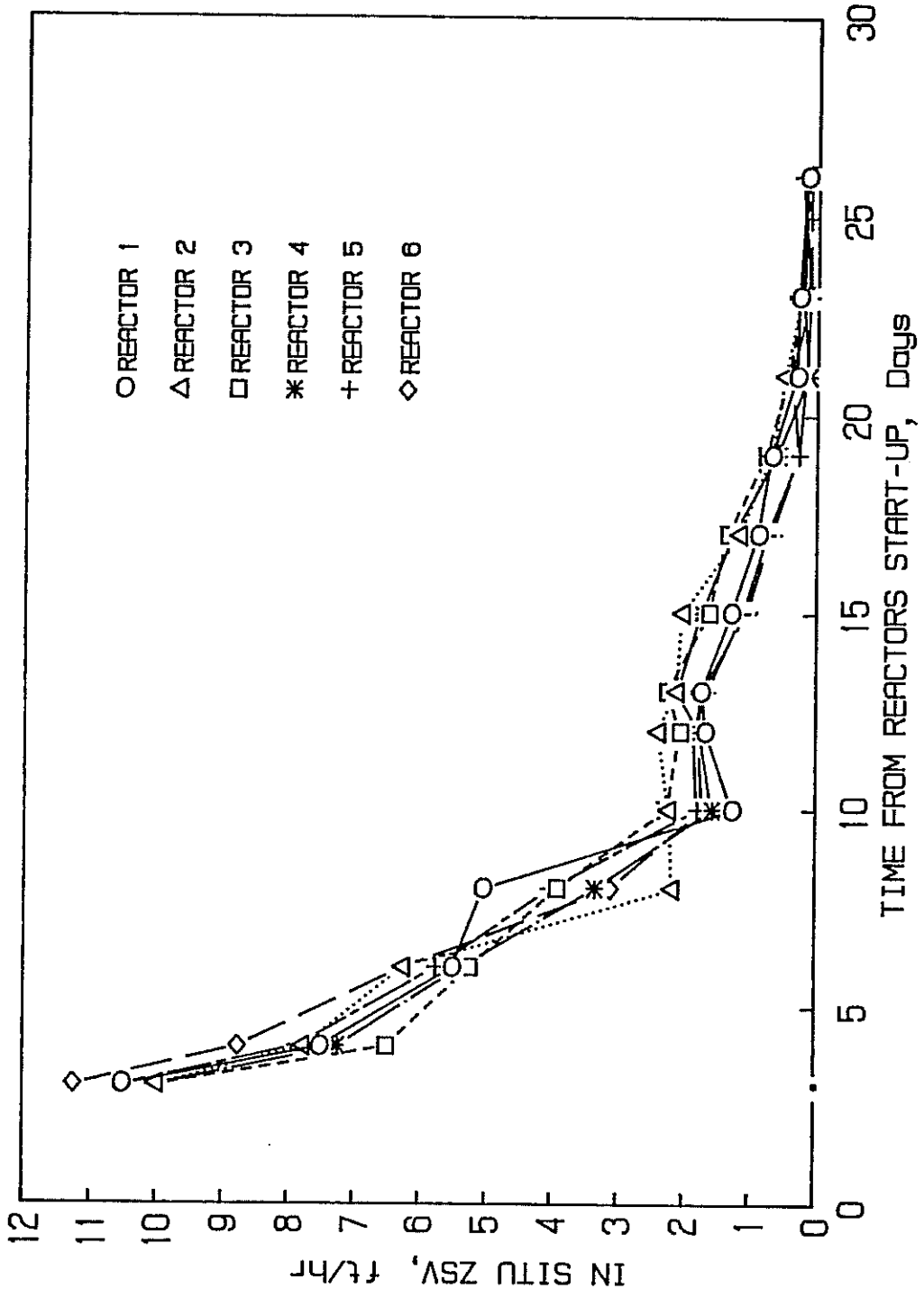


Figure 6.2. Chronological Plot of In-Situ ZSV-Phase II.

TABLE 6.3

RESULTS OF STIRRED ZSV TESTS (ft/hr)
MAY 14, 1985

Parameter	Reactor 1	2	3	4	5	6
Test No.						
1	3.07	4.11	3.67	4.23	3.34	4.19
2	2.55	4.26	3.71	4.09	3.61	4.02
3	2.81	3.56	2.98	3.90	3.45	3.62
4	2.63	4.01	3.35	4.33	3.77	4.25
5	2.92	3.61	3.63	3.80	3.50	3.85
6	2.88	3.86	3.22	4.09	3.94	3.73
Mean	2.81	3.90	3.43	4.07	3.60	3.94
Standard Deviation	0.175	0.254	0.267	0.181	0.202	0.231
95 percent Confidence Interval for Mean	2.68-3.01	3.57-4.19	2.91-3.73	3.88-4.28	3.71-3.83	3.46-4.21

TABLE 6.4

SUMMARY OF ANALYSIS OF VARIANCE
ON STIRRED ZSV RESULTS
MAY 14, 1985

Source of Variation	Sum of Squares	Degrees of Freedom	Mean Square	Critical F Region for $\alpha = 0.05$	Computed F Value
Reactor	6.498	5	1.300	2.53	22.1
Error	1.761	30	0.0587		
Total	8.258	35			

error involved in MLSS/MLVSS measurement and due to the discrete wasting procedure, variations in MLSS/MLVSS concentrations were unavoidable.

In order to determine the effect of MLSS variations on ZSV results, the ZSV was tested in a series of dilutions of sludge from the reactor No. 1. The relationship between ZSV and MLSS is shown on Figure 6.3 and has the form of the following exponential function:

$$ZSV = 20.67 \cdot \exp(-0.45\ MLSS) \qquad (6.1)$$

(Correlation coefficient = 0.975)

Where:

ZSV = zone settling velocity, ft/hr

MLSS= mixed liquor suspended solids, g/l

The error in MLSS measurement is estimated at 0.10 g/l, however, a larger variation in actual MLSS in the reactors might have been common, due to the previously discussed discrete nature of the sludge wasting. Assuming that variations of MLSS in a reactor were less than 0.2 g/l, and that the average MLSS in the reactors was 3.12 g/l Formula 6.1 indicates that at the test conditions the ZSV adjusted to the actual MLVSS in the reactor could have varied between 4.6 and 5.8, depending on the actual MLSS.

The discussion presented above on the reliability of ZSV measurements as indices of sludge settling characteristic addresses only a part of the problem. The main question to be answered was: is the sludge settling characteristic a function of the operating

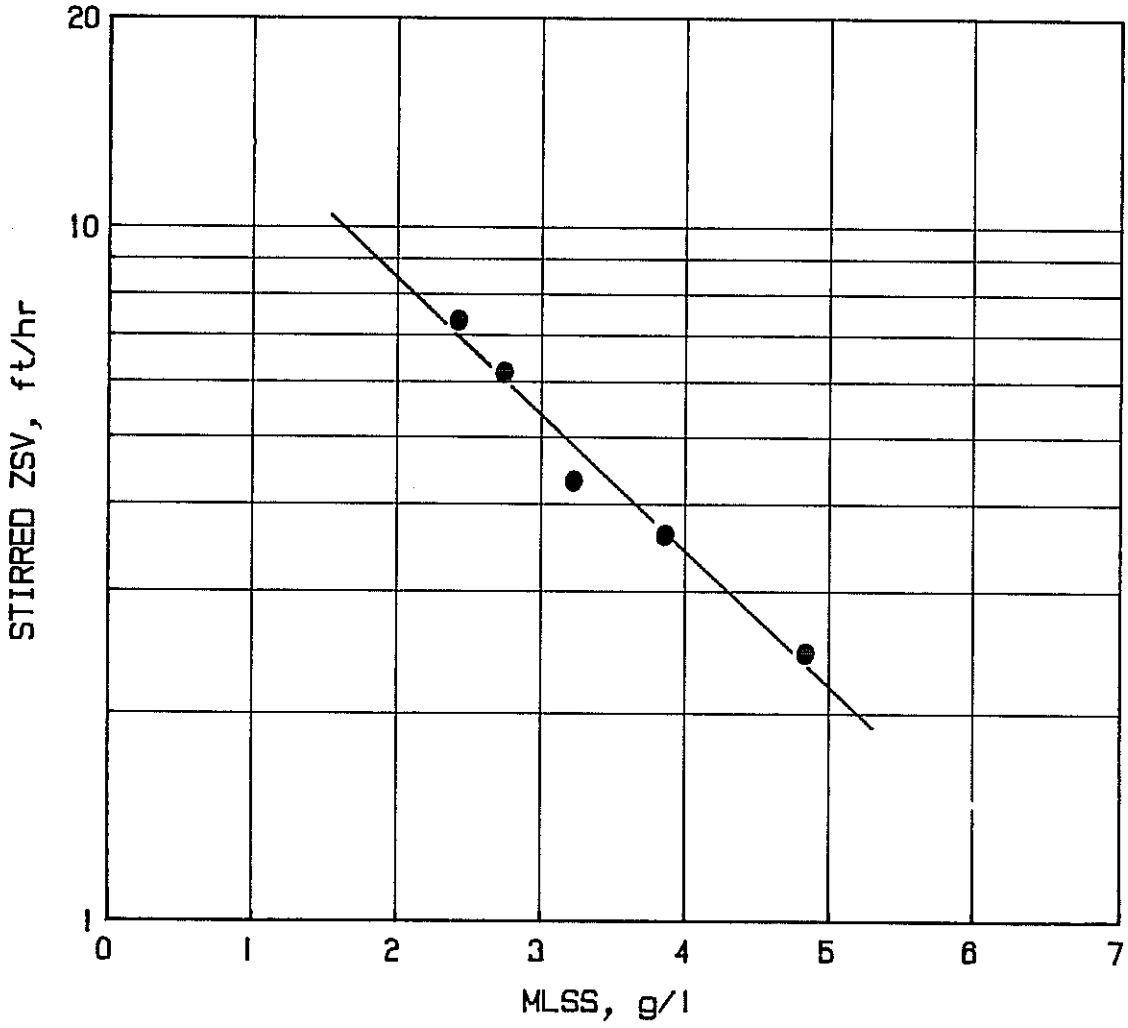


Figure 6.3. Effect of MLSS on ZSV.

parameters of the reactor? There is no clear answer. On one hand, the statistical analysis performed on one set of data demonstrated that there were true differences between the ZSVs of the tested reactors. However, even if there were statistical differences in ZSVs of the reactors throughout the study, review of Figures 6.1 and 6.2 convinces us that the differences were small. Both stirred and in situ ZSV data indicated that sludge settling characteristic in the studied biological environment were reproducible and all the reactors followed the same trend leading to their bulking at about the same time.

Unfortunately, these observations are of limited value only. Reproducible performance of one set of CSTRs gives only a qualified confidence that other reactors, particularly with more complex configuration (PRZ) would behave similarly. A definite answer would be only possible by operating multiple reactors for each reactor configuration and set of operating parameters.

Biosorption Characteristics

Three times during the course of the phase II study, biosorption tests were performed on the six parallel reactors in order to experimentally define variations in biosorption characteristic of the reactors operating at identical conditions. The tests were performed according to the procedure outlined in Chapter IV.

The results of the tests are summarized in Table 6.5. The average biosorption capacity of sludges from the reactors showed a small but

TABLE 6.5

SUMMARY OF THE BIOSORPTION DATA
PHASE II

Parameter	Reactor 1	2	3	4	5	6	Average	Coefficient of Variation (%)
5 min. Biosorption, mg COD/g VSS								
5/15/85	28.0	27.3	25.5	25.1	24.4	22.7	25.5	7.5
5/24/85	26.9	26.9	25.5	24.3	28.3	27.5	26.6	5.3
6/3/85	32.8	27.9	29.9	31.5	24.6	26.4	28.9	10.7
15 min. Biosorption, mg COD/g VSS								
5/15/85	39.0	42.3	38.2	40.5	39.9	41.1	40.2	3.7
5/24/85	41.6	40.6	39.6	38.2	41.5	39.7	40.2	3.2
6/3/85	47.5	44.6	38.3	46.4	39.6	42.2	43.1	8.6

measurable increase during the four week testing period, for both the 5 min and the 15 min measurements. Coefficients of variation for the measurements performed on the same day on different reactors varied from 3.2 to 10.7 percent, with the variations resulting from both measurement error and, potentially,actual differences in sludge characteristic between the different reactors.

To assess the systematic error involved in biosorption capacity measurements a series of six parallel tests were performed on sludge samples withdrawn at the same time from reactor No. 1. The results are presented in Table 6.6. The coefficient of variation was about 5 percent for both the 5 min and 15 min biosorption measurements. Comparison with the data presented in Table 6.5 indicates that testing of different reactors resulted in coefficients of variation comparable or smaller to the ones expected from the systematic error, with the exception of the data from June 3, 1985.

Filamentous Bacteria Count

Estimation of the proliferation of filamentous bacteria was performed throughout the Phase II investigations using the technique outlined in the Methodology Section. Figure 6.4 presents a chronological plot of the results for the six reactors.

As was discussed in Chapter IV, each result is an average of 12 individual counts from the same sludge sample. This number of repetitions assured that the obtained value is within 33 percent of the

TABLE 6.6

BIOSORPTION TEST ON REACTOR NO. 1
MAY 7, 1985

Parameter	Test No. 1	2	3	4	5	6	Average	Coefficient of Variation (%)
5 min. Bisorption, mg/l	29.3	28.0	26.0	28.0	30.8	27.6	28.3	5.7
15 min. Bisorption, mg/l	42.5	44.9	41.8	41.6	38.9	40.1	41.6	5.0

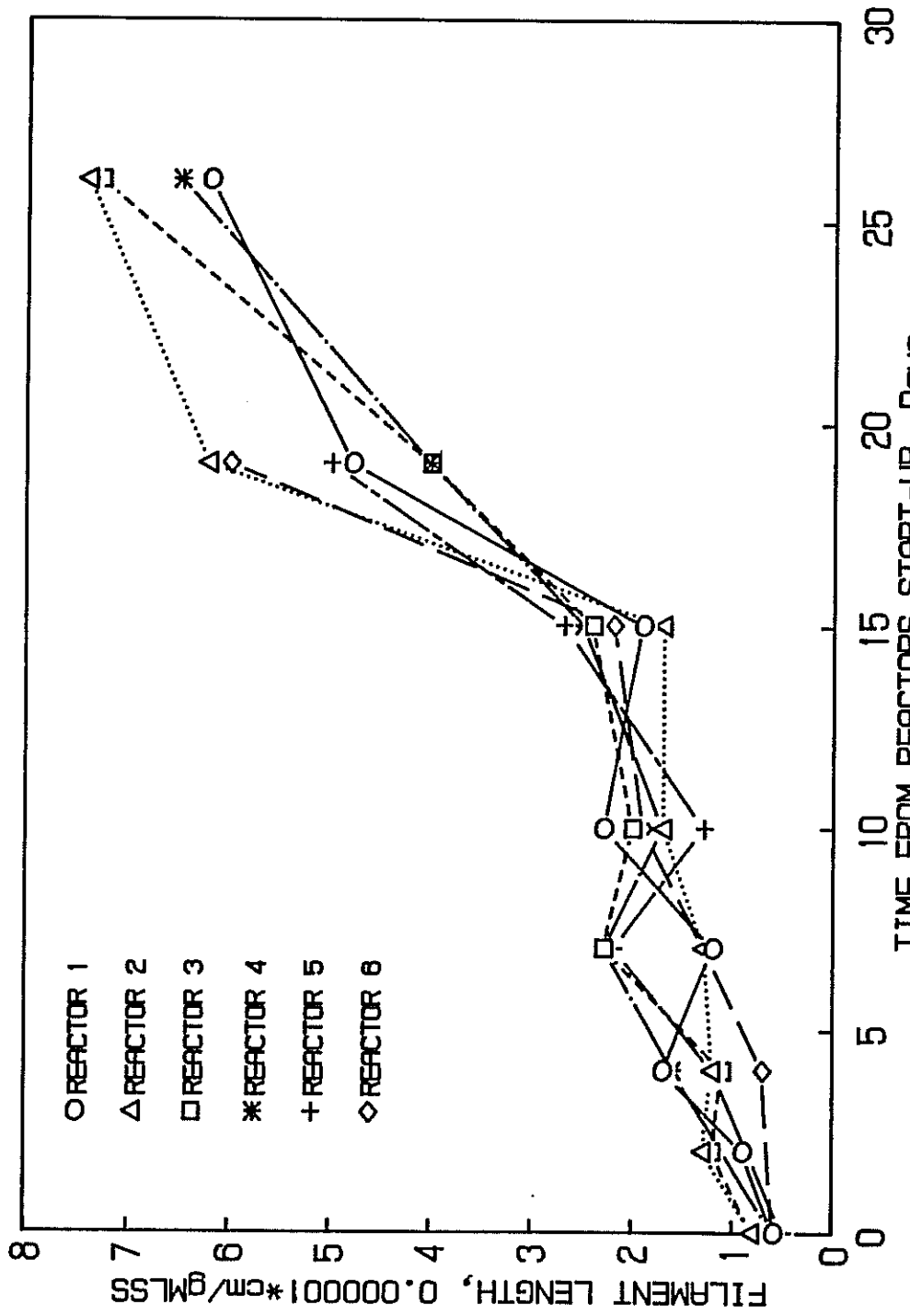


Figure 6.4. Chronological Plot of Filament Length Measurements.

true value (with 90 percent confidence). The relatively lenient criteria accepted for the accuracy of the filament length measurement are reflected in Figure 6.4, where a considerable scatter of the data is apparent. Considering the time involved in each individual count (about 1.5 min.) and the fact that the number of repetitions required is inversely proportional to the second power of the error (fraction) desired, a substantial increase in precision was not feasible. For example, a reduction of error to 10 percent would have required about 150 counts, or more than 3 hrs per sample.

Taking advantage of the fact that each data point is an average of multiple observations, analysis of variance is possible for data from each day. In Table 6.7, a summary of such analysis performed on three sets of data (three different days) is presented. In all cases, filamentous bacteria length measurements did not reveal any differences between the reactors, at the 95 percent confidence level. In conjunction with a rather substantial variation in the average measurements seen on Figure 6.4, it is apparent that the filament counting technique utilized in this study is burdened with a high random error, arising from the highly inhomogeneous microstructure of the activated sludge.

In order to illustrate the relationship between sludge settling indices and filament length count, values of these parameters obtained from all six reactors at any given day were averaged, and these averages are presented in Figure 6.5. It is recognized that the averaging utilized for construction of Figure 6.5 is not quite

TABLE 6.7

SUMMARY OF ANALYSIS OF VARIANCE IN FILAMENT LENGTH MEASUREMENT

Date	Source of Variation	Sum of[a] Squares	Degrees of Freedom	Mean Square	Critical F Region (95%)	Computed F
May 11, 985	Reactor	65	5	13.1	2.35	1.82
	Error	475	66	7.2		
	Total	540				
May 19, 1985	Reactor	51	5	10.2	2.35	0.62
	Error	1,084	66	16.4		
	Total	1,135				
May 26, 1985	Reactor	123	5	24.6	2.35	0.68
	Error	2,325	66	35.2		
	Total	2,448				

[a]Analysis was performed on unprocessed count numbers (N), i.e., before conversion to the filament length (Formula 4.2).

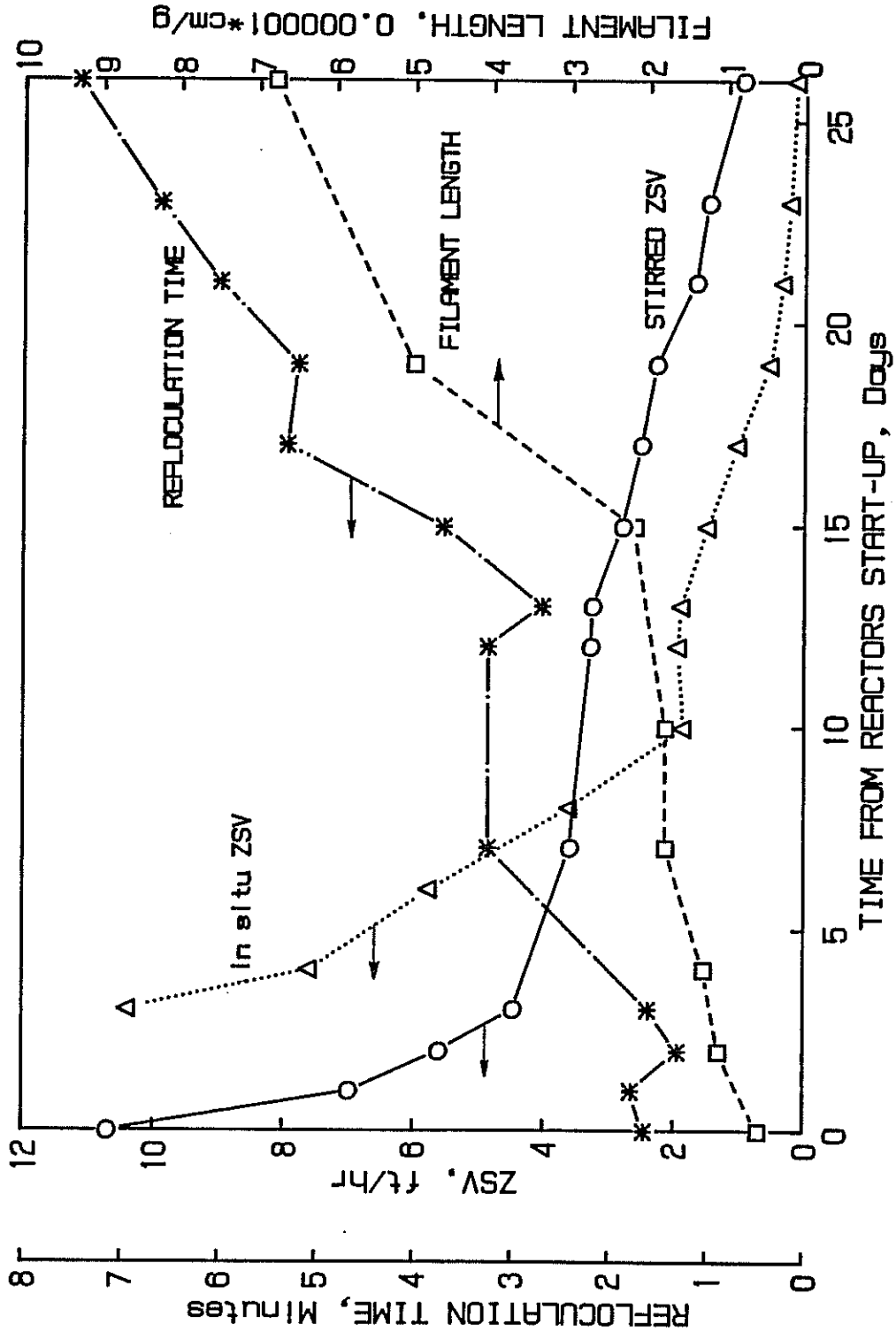


Figure 6.5. Comparison of Different Sludge Settling Indices.

justified. Analysis of stirred ZSV data, for example, demonstrated that there were true differences between the reactors. Nevertheless, it is believed that such averaging is acceptable for illustrative purposes.

Figure 6.5 reveals an expected relationship between ZSVs and filament length. Increase in the total filament length is accompanied by a decline of both stirred and in situ ZSVs. Reflocculation time closely follows the increase in the filament length, illustrating the detrimental effect of protruding filaments on the formation of the floc agglomerates.

It is interesting to note that at relatively high ZSV (above 3.5 ft/hr), the in situ ZSV gave a much higher value than the stirred ZSV, while for low ZSVs the opposite was true. The most likely explanation for this phenomoneon is that for well flocculating and settling sludge, the walls of the narrow cylinder and stir rods present an obstacle (wall effect), rather than aid in flocculating and breaking bridging (stir rods). At low ZSVs the relative importance of the wall effect (friction) diminishes, while stirring is effective in breaking filament-induced bridging between the agglomerates.

When the stirred ZSV's are plotted versus inverse of filaments length a well defined, linear relationship is obtained (Figure 6.6). The correlation is very satisfactory for stirred ZSV values above 2 ft/hr. For lower stirred ZSV values the relationship is undefined for the lack of sufficient data in that range.

The relationship presented in Figure 6.6 demonstrated that for practical purposes measurement of ZSV and filament length count are equivalent indices of the sludge settling characteristic. It had been expected that perhaps a direct observation of filamentous bacteria proliferation might be useful for anticipating onset of bulking. However, filament length count did not appear to provide any additional information regarding the sludge conditions. Consequently, use of this time-consuming and tedious analytical method was discontinued and only stirred ZSV measurements were used for sludge evaluations in the subsequent phase of the study.

Predominant filamentous bacteria present in the reactors during Phase II (as well as in the subsequent Phase III) were identified as type 0041. This filament type has been frequently associated with bulking under low organic loading conditions, as well as in plants treating readily biodegradable substrates such as food processing wastewater (Strom and Jenkins, 1984). Secondary filamentous species present in some samples were 021N and H. hydrossis.

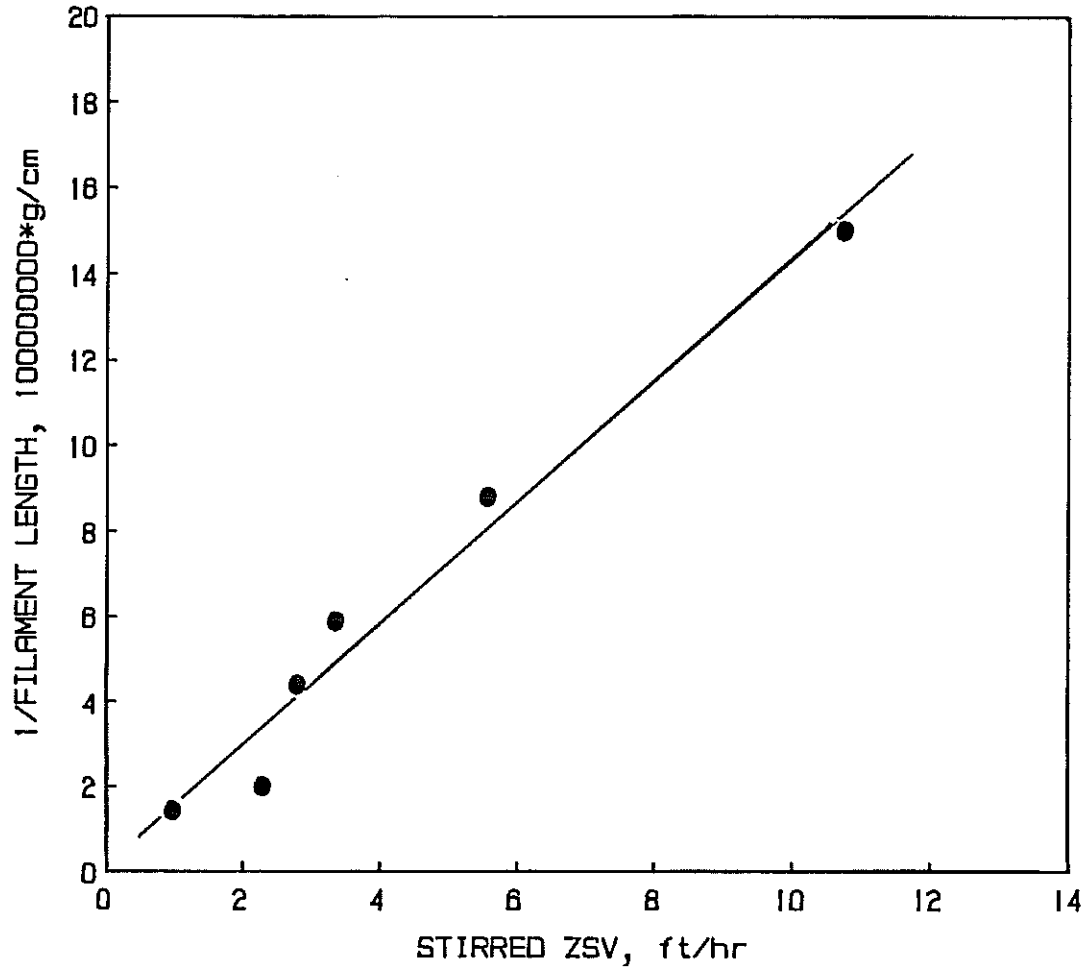


Figure 6.6. Correlation Between Stirred ZSV and Filament Length Measurement.

CHAPTER VII

RESULTS AND DISCUSSION-PHASE III

Sludge Settling Characteristic and Reactors Configuration

Sludge Settling Characteristics

The chronological chart of the sludge settling characteristics from the seven reactors operated during phase III is presented in Figure 7.1. The pattern of ZSV changes is rather surprising, showing a substantial cyclic variation for all reactors, except No. 2 (PRZ with intermittent mixing) and No. 7 (batch fed).

The completely mixed reactor No. 1 after some mild, initial ZSV oscillations, bulked after 16 days from the start-up. After 10 days, it rebounded for a brief period, only to severely bulk for the reminder of its operation (ZSV less than 0.05 ft/hr). The recovery of reactor No. 1 in days 27 to 31 (Figure 7.1) can be attributed to a small change in its operation. On day 26 during a track study on reactor No. 1, the pulse mixing of the reactor's content with N_2 during the air-off phase was temporarily discontinued. As a result, the feed delivered to the reactor during the air-off time collected at the reactor's bottom below the point of addition (temperature and density gradient), and during that time was in contact with only a small fraction of the sludge. Only after the aeration was resumed, was the reactor's stratification

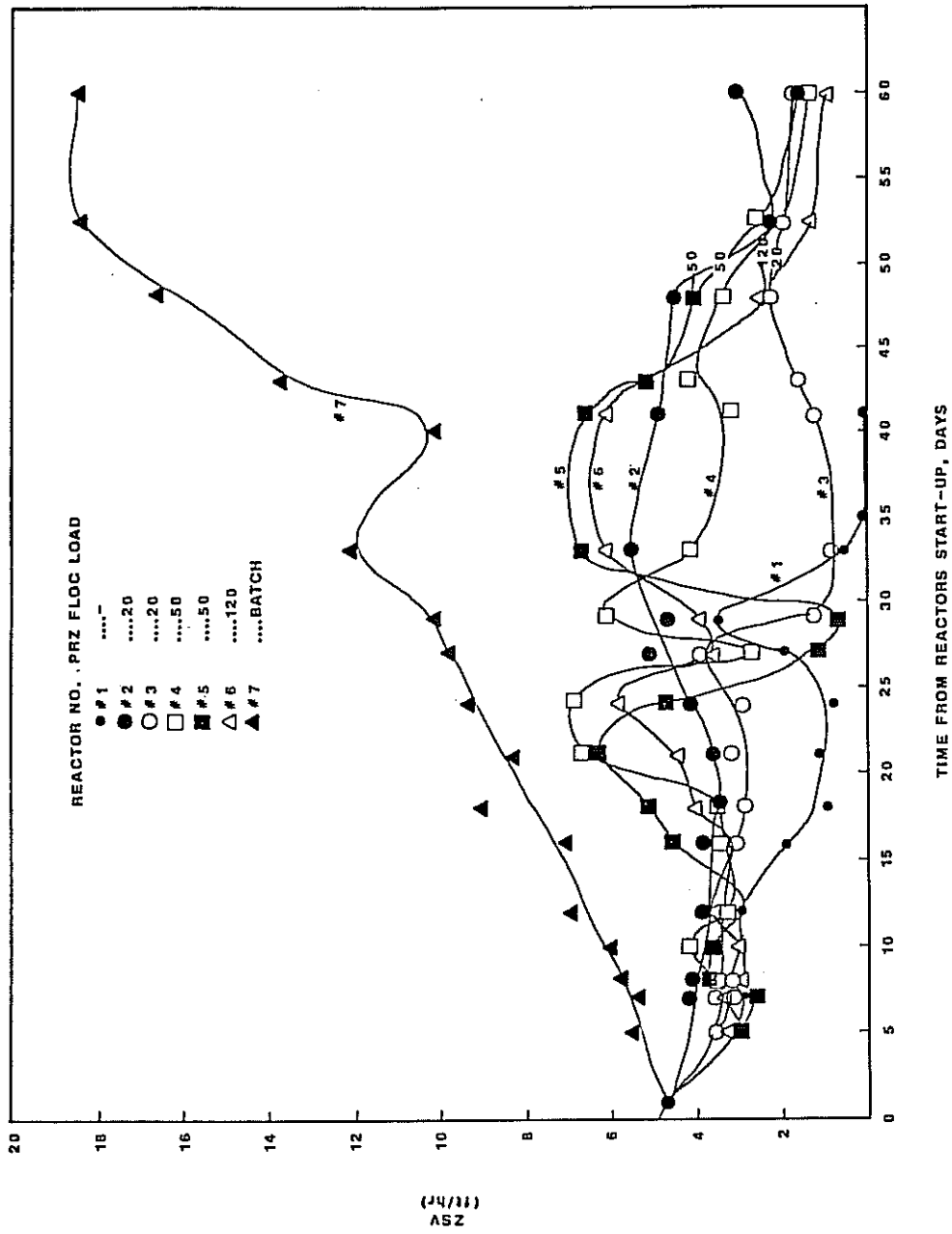


Figure 7.1. Chronological Plot of Stirred ZSV-Phase III

broken and the feed distributed over the whole aeration basin's content. The normal reactor's operation mode with continuous agitation during the air-off phase was restored on day 28 and shortly thereafter ZSV dropped to almost zero. The connection between the change in operating mode and the rapid but short lived decrease of bulking is not conclusive but highly probable.

Reactor No. 2 with the PRZ operating at a floc load of 20 mgCOD/gVSS, with intermittent PRZ aeration, demonstrated a stable operation till the very end of the study. ZSV values in this reactor varied between 3.5 and 5.5 ft/hr for 50 days, only to deteriorate to a bulking condition at the end of the study. The relative stability of this reactor is most likely associated with its unique operational mode. In this reactor, the PRZ was not mixed during the air-off time and the sludge was not recycled during that time. This allowed most of the feed which was delivered continuously to the PRZ to accumulate there since the volume of this PRZ was 970 ml versus a total of 1100 ml of feed pumped into PRZ during 105 min of the air-off phase. From the data on Figure 7.2 the amount of the substrate which was contained in the PRZ during the air-off phase was calculated by a graphical integration method to be 55 percent of the total feed delivered to the reactor during that period. This substrate was partially taken up by the sludge entrapped in the PRZ during the air-off phase and by the fresh sludge recycled to the PRZ in a period of several minutes immediately following the resumption of aeration. This substrate uptake was occurring at much higher substrate concentration than in any

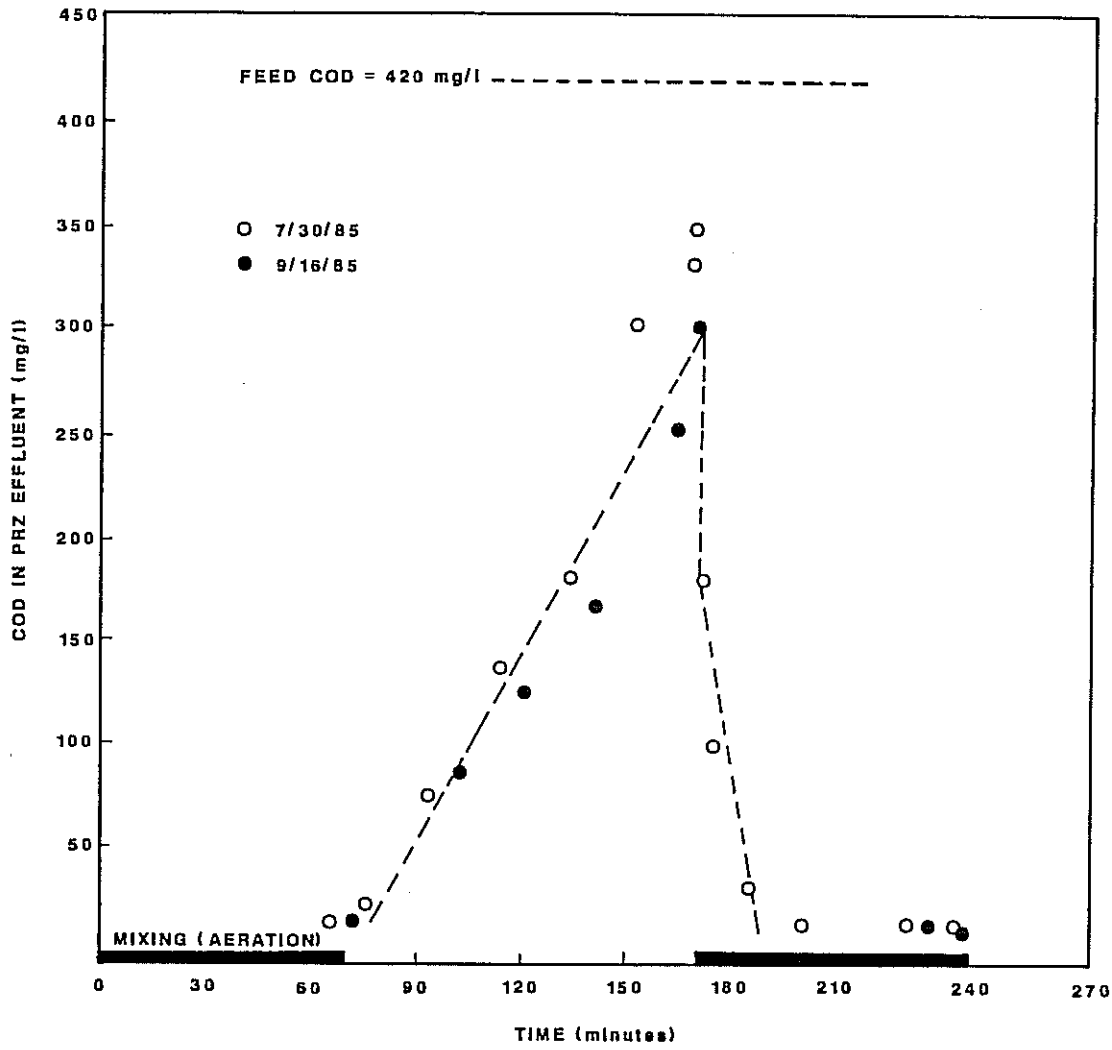


Figure 7.2. COD Profile in Reactor No. 2.

of the remaining, continuously aerated PRZs with a continuous sludge recycle. This process appears to account for a relatively stable, non-bulking condition of the reactor No. 2.

The sludge settling characteristics in the remaining reactors with a PRZ were unstable throughout the study, with difficult to explain oscillations in ZSV values (Figure 7.1). The MLSS concentrations in these reactors was kept nearly constant by daily, routine sludge wasting. MLSS was measured twice a week and the sludge concentration corrected to the design value and wasting rate readjusted, when necessary. Solids concentration variations were therefore kept to a minimum and cannot account for ZSV oscillations; moreover, the oscillations are documented in each case (except No. 4 ZSV decline on day 27) by several consecutive measurements, dismissing an accidental (i.e., measurement error) nature of these variations. The oscillations were noticeable from the very beginning of the study and they tended to increase in the amplitude and period as the study progressed.

Reactor No. 3 with a PRZ floc load of 20 mgCOD/gVSS exhibited relatively stable sludge settling characteristic for the first four weeks of operation, only to rapidly deteriorate from ZSV of 4 ft/hr to 1 ft/hr between the 27th and 29th day of operation. No changes in the operating parameters or reactor's performance were noticed prior to or during that period. The sludge biosorption efficiency (Figure 7.3) was rising steadily from the onset of the study until day 60, and apparently there was no relation between the ZSV decline after day 27

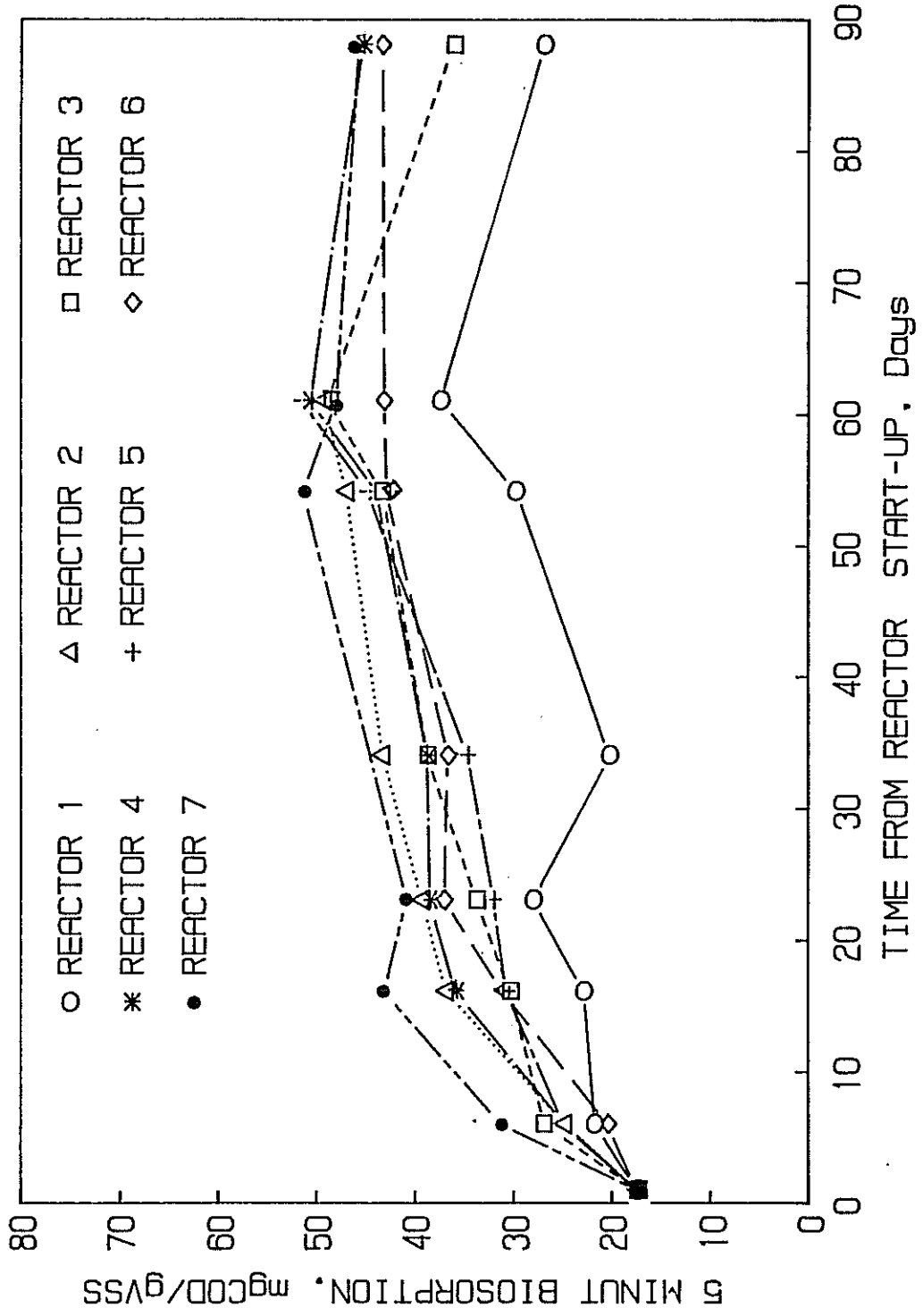


Figure 7.3. Chronological Plot of Five Minute Biosorption Efficiences.

and the sludge biosorption capacity. After day 40 the ZSV slowly improved to reach about 2 ft/hr and then started to deteriorate again until the end of the study.

Reactor No. 4 with a PRZ operating at floc load of 50 mgCOD/gVSS and HRT of 8 min and reactor No. 5 with the same PRZ floc load and HRT of 20 min both developed extensive oscillations in their ZSV (Figure 7.1). No change in reactor operation or in performance parameters accompanied these oscillations. Reactor No. 4 bulked eventually at day 53 and remained in this condition until the end of the study. Reactor No. 5 during an extensive ZSV oscillation had a brief bulking episode between days 26 and 30 and bulked again around day 55 to remain in this condition until its shutdown on day 62.

Reactor No. 6 operating with PRZ load of 120 mgCOD/gVSS (HRT = 8 min) developed an oscillation pattern similar to that observed in reactor No. 4 (Figure 7.1). The amplitude of the oscillations were smaller, however. The reactor eventually bulked on the 50th day of the study.

Reactor No. 7, batch fed once a day, and without a PRZ, exhibited a steady increase in ZSV from an initial value of 4.7 ft/hr to about 18.5 ft/hr which was reached on day 53. The ZSV remained at this level until the reactor's shutdown on day 62.

Changes in Sludge Biosorption Capacity

The history of the 5 min and 15 min biosorption capacities of the reactor sludges are presented in Figures 7.3 and 7.4, respectively.

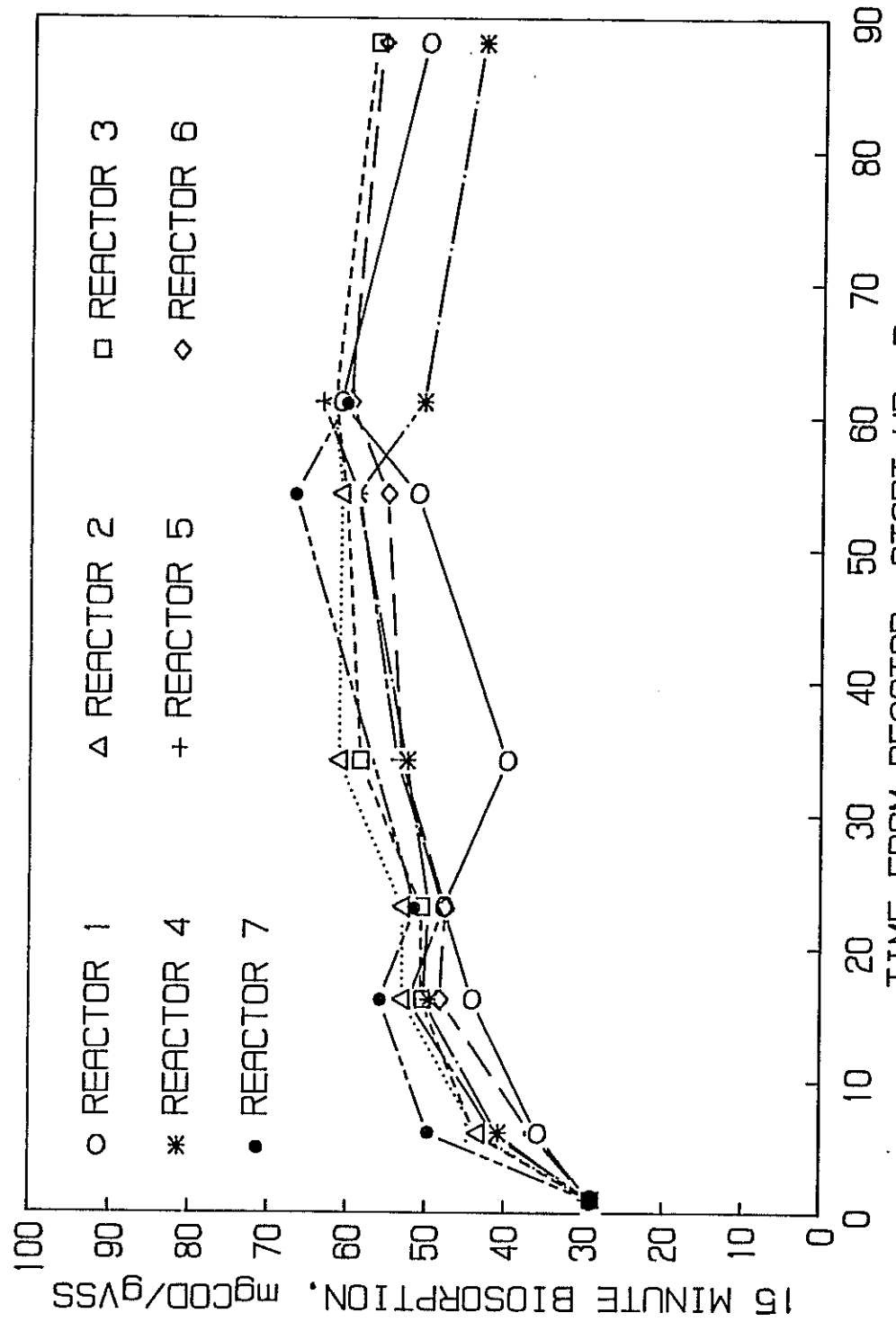


Figure 7.4. Chronological Plot of Fifteen Minute Biosorption Efficiencies.

In all the reactors (except No. 1) the biosorption capacity was rising steadily from an initial value of 19 mgCOD/gVSS (at 5 min contact time) to 45 to 50 mgCOD/gVSS on day 61. This increase, however, did not progress at the same pace in all the reactors. Inspection of Figure 7.4 reveals that the reactors which generated sludges with a nonbulking and stable characteristic had, on average, a higher biosorption capacity than the reactors with oscillating and/or bulking sludge.

Batch reactor No. 7 which had an excellent and continuously increasing ZSV demonstrated the highest biosorption capacity measured at the end of the daily cycle, 22 hr after the feed addition. Particularly noteworthy is the rapid increase in the biosorption capacity of this reactor at the beginning of the study, when it more than doubled in 16 days from start-up. The sludge from reactor No. 2, had a consistently higher biosorption capacity than the other reactors with PRZs, which corresponds to a similarly favorable settling characteristic of this sludge. The completely mixed reactor No. 1 developed sludge with both the lowest biosorption capacity and the worst settling characteristic of all the reactors.

Based on the data presented, it can not be conclusively demonstrated that the sludge biosorption characteristic is directly related to the ZSV. The filamentous bacteria content in the sludge, directly related to the ZSV, is rather independent of the biosorption characteristic of the whole biomass. Comparison of Figures 7.1 and 7.4 provides numerous examples of this. For example, on days 33 to 34 ZSV

(and therefore filament content) in reactors No. 1 and No. 3 were almost identical, while the biosorption capacity of reactor No. 3 was almost twice that of No. 1.

Therefore, while association of a high sludge biosorption capacity with a good settling characteristic is, on average, certainly more than accidental, these two events are not necessarily linked by a cause-effect relationship. Instead, it can be argued that the same environmental pressures which favor growth of non-filamentous organisms (Chapter II), namely relatively high substrate uptake (PRZ), cause physiological adaptation of the biomass to a starvation-feast cycle with a resulting improved substrate uptake capacity during the short period of sludge contact with the substrate in the PRZ. While the presence of any PRZ logically selects and/or adapts microbial populations for a rapid substrate uptake, it does not guarantee prevention of filamentous bacteria proliferation.

Effect of PRZ Performance on Sludge Settling Characteristic

Figure 7.5 presents the relationship between time to the first incidence of bulking and the weighted average concentration at which the substrate was taken up by the biomass. The concentration was calculated as a weighted average, assuming that, in the aeration basin, concentration of the available substrate was negligible. The return sludge COD contribution was subtracted from the measured COD in the PRZ under the assumption that the aeration basin COD reflected the

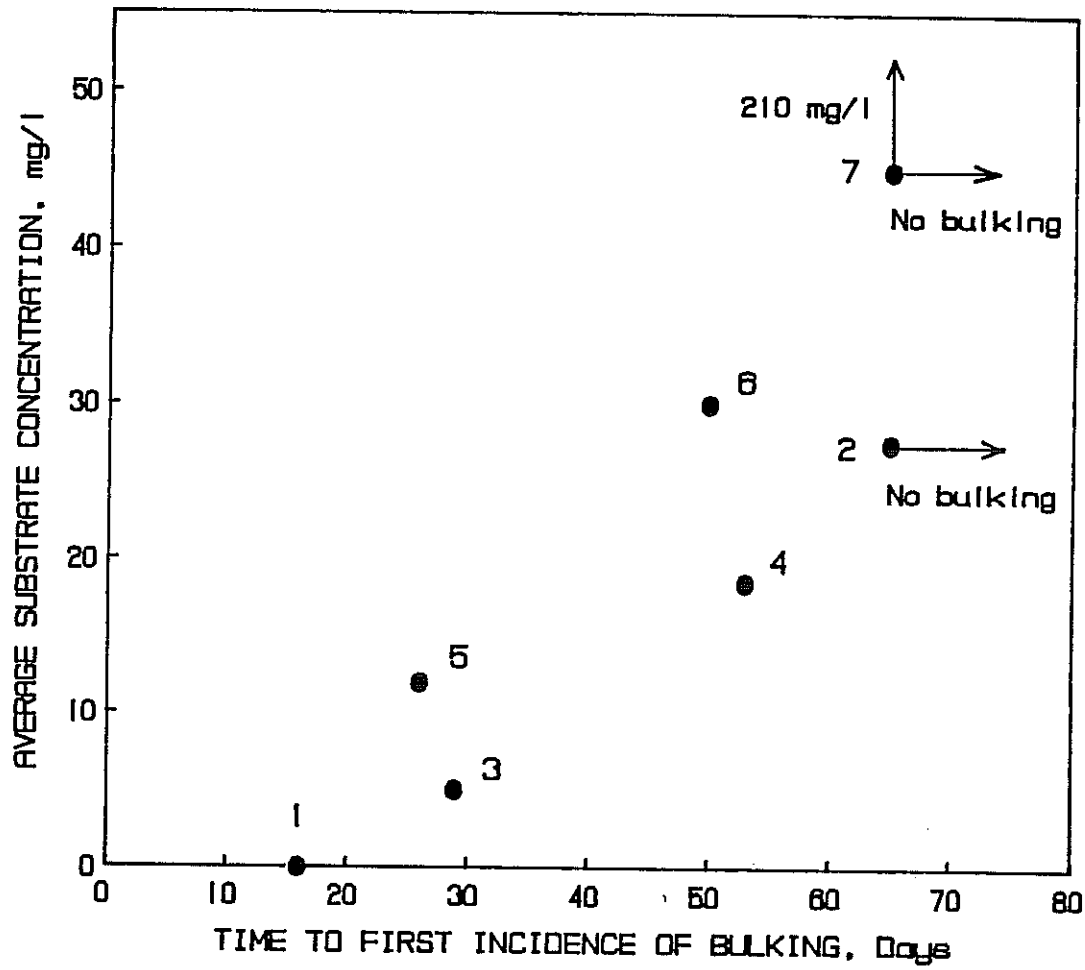


Figure 7.5. Relationship Between Weighted Average Biosorption Concentration (C_B) and Reactors' Resistance to Bulking. Reactor No. is Indicated on the Graph.

biodegradation byproducts, rather than the residual feed COD. The formula for calculation of the weighed average biosorption concentration (C_B) has the form:

$$C_B = EC_1 \qquad (7.1)$$

Where:

E = fraction of substrate removed in PRZ

$C_1 = COD_1\ COD_R Q_R/(Q_o+Q_R)$ = available substrate concentration in the PRZ (Refer to Figure 5.6 for nomenclature).

It should be noted that the fraction of the available substrate which leaked into the main aeration tank (1-E) is not reflected in the Formula 6.1. This is the consequence of the high dilution rate of the substrate in the aeration tank which results in a negligible concentration of the available substrate.

Figure 7.5 gives a clear indication that the increasing average biosorption concentration (C_B) increases the reactor's resistance to bulking. The weighted average approach used in the construction of Figure 7.5 provides a simple approximation of the prevailing conditions at which the substrate is being removed in the activated sludge system with a PRZ. Maximization of this average uptake concentration should provide optimum conditions for the growth of the floc forming bacteria.

In order to optimize C_B, a method of predicting the performance of a PRZ is necessary. Such a predictive model is proposed in the subsequent section.

Modeling of PRZ Performance

The performance of a PRZ in all the reactors was routinely monitored throughout the study as described in Chapter IV. The results of these measurements are summarized for the individual reactors in Tables 7.1 through 7.4. A series of additional tests were performed on reactor No. 4 in which a wide range of PRZ operating parameters was tested for their effect on PRZ performance, as outlined in Chapter IV. Results of these experiments are summarized in Table 7.5.

Efficiencies of substrate removal in the PRZ presented in Table 7.1 through 7.5 were calculated from Formulas 5.1 through 5.3. Figure 7.6 gives a chronological plot of PRZ performance for the reactors and comparison with Figure 7.4 indicates that the sludge biosorption capacity and performance of the PRZ demonstrated an expected trend, increasing as the study progressed.

The average performance of each reactor's PRZ is shown in Figure 7.7 as a function of the average floc loading and HRT in the PRZ, indicating, as expected, that increased floc loading at the constant HRT results in a lower substrate removal efficiency. A similar graph constructed from data collected during the intensive study on reactor No. 4 (Figure 7.8) gives a more complete picture of the relationship between the PRZ operating parameters and its performance. Figure 7.8 was constructed by reading biosorption efficiencies at several HRT's from a smooth line drawn through the experimental points presented on Figure 7.9.

While the results presented in Figure 7.8 provide a well defined and consistent series of curves, they are valid only for the specific

TABLE 7.1

PERFORMANCE OF PRZ IN REACTOR NO. 3[a]

Test No.	Date (1985)	C_1[b] (mg COD/l)	F[c] (COD/g VSS)	B[d] (mg COD/g VSS)	E[e] (%)	F/M[f] (g COD/g VSS-day)	R_r[g] (g COD/g VSS-day)
1	7/30	20.3	19.1	13.7	71.6	3.63	2.54
2		14.8	19.1	16.6	86.8	3.63	3.07
9	8/9	18.1	17.8	15.6	87.8	3.26	2.76
10		17.6	17.8	15.9	89.3	3.26	2.81
17	8/16	20.2	19.6	17.2	87.5	3.73	3.14
18		19.9	19.6	17.3	88.4	3.73	3.17
19		22.8	19.6	15.7	80.1	3.73	2.88
29	8/23	19.4	19.0	14.0	73.6	3.64	2.61
30		15.3	19.0	16.2	85.3	3.64	3.03
37	8/25	15.8	19.3	16.6	86.1	3.70	3.11
41	9/16	14.1	20.1	17.2	85.5	3.88	3.24
45	9/23	13.2	19.6	17.9	91.4	3.74	3.33
50	10/20	23.0	21.1	18.6	88.4	4.17	3.52
53		18.4	19.0	16.7	87.6	3.68	3.11

[a]For each test parameters such as feed and recycle flow, feed and recycle COD, and PRZ operating volume were measured (not provided here) and used for calculation of the PRZ performance parameters.

[b]Substrate concentration in the PRZ.

[c]Floc load.

[d]Biosorption.

[e]Substrate removal efficiency in PRZ.

[f]Organic loading in PRZ.

[g]Reaction rate in PRZ.

TABLE 7.2

PERFORMANCE OF PRZ IN REACTOR NO. 4[a]

Test No.	Date (1985)	C_1[b] (mg/l COD)	F[c] (mg COD/g VSS)	B[d] (mg COD/g VSS)	E[e] (%)	F/M[f] (g COD/g VSS-day)	R_r[g] (g COD/g VSS-day)
3	7/30	48.0	48.8	24.9	51.1	9.16	4.57
4		42.9	46.9	27.0	57.5	8.80	4.95
11	8/9	39.1	44.2	27.1	61.4	8.35	4.99
12		34.7	44.2	29.8	67.4	8.35	4.38
20	8/16	45.8	46.9	27.8	59.3	8.88	5.06
21		44.5	46.9	28.6	61.0	8.88	5.21
22		47.3	46.9	26.8	57.2	8.88	4.88
31	8/23	4.39	49.1	28.1	57.1	8.96	5.00
32		39.6	49.1	30.7	62.5	8.96	5.47
39	8/29	31.6	47.8	34.2	71.6	9.02	6.31
42	9/16	32.6	49.4	35.3	75.7	8.55	6.32
46	10/20	35.6	47.2	33.0	69.9	8.73	5.92
54	10/20	3.56	47.2	33.0	69.9	8.73	5.92

[a]For each test parameters such as feed and recycle flow, feed and recycle COD, and PRZ operating volume were measured (not provided here) and used for calculation of the PRZ performance parameters.

[b]Substrate concentration in the PRZ.

[c]Floc load.

[d]Biosorption.

[e]Substrate removal efficiency in PRZ.

[f]Organic loading in PRZ.

[g]Reaction rate in PRZ.

TABLE 7.3

PERFORMANCE OF PRZ IN REACTOR NO. 5[a]

Test No.	Date (1985)	C_1[b] (mg COD/l)	F[c] (mg COD/g VSS)	B[d] (mg COD/g VSS)	E[e] (%)	F/M[f] (g COD/g VSS-day)	R[g] (g COD/g VSS-day)
5	7/30	29.1	46.0	34.3	74.5	3.34	2.43
6		25.2	46.0	36.6	79.5	3.34	2.59
13	8/9	26.2	48.3	39.4	81.6	3.18	2.52
14		25.6	48.3	39.8	82.4	3.18	2.54
23	8/16	33.4	47.5	36.5	76.9	3.35	2.47
24		33.1	47.5	36.7	77.3	3.35	2.48
25		34.7	47.5	35.7	75.1	3.35	2.41
33	8/23	28.4	48.8	37.9	77.7	3.22	2.43
34		27.6	48.8	38.4	78.8	3.22	2.47
38	8/29	21.9	48.0	40.3	83.8	3.21	2.63
43	9/16	19.4	49.2	42.7	86.8	3.54	2.84
47	9/23	16.9	50.7	45.3	89.4	3.26	2.86

[a]For each test parameters such as feed and recycle flow, feed and recycle COD, and PRZ operating volume were measured (not provided here) and used for calculation of the PRZ performance parameters.

[b]Substrate concentration in the PRZ.

[c]Floc load.

[d]Biosorption.

[e]Substrate removal efficiency in PRZ.

[f]Organic loading in PRZ.

[g]Reaction rate in PRZ.

TABLE 7.4

PERFORMANCE OF PRZ IN REACTOR NO. 6[a]

Test No.	Date (1985)	C_1[b] (mg COD/l)	F[c] (mg COD/g VSS)	B[d] (mg COD/g VSS)	E[e] (%)	F/M[f] (g COD/g VSS-day)	R_r[g] (g COD/g VSS-day)
7	7/30	80.6	80.1	31.7	39.6	15.1	5.84
8		79.1	80.1	32.7	40.9	15.1	6.03
15	8/9	79.4	76.6	29.6	38.7	15.5	5.84
16		73.5	76.6	33.7	44.0	15.5	6.64
26	8/16	74.8	75.9	34.0	44.8	15.8	6.84
27		73.3	75.9	35.1	46.2	15.8	7.05
28		75.2	75.9	33.8	44.5	15.8	6.78
35	8/23	64.9	79.0	39.7	50.3	16.2	7.96
36		64.9	79.0	39.7	50.3	16.2	7.96
40	9/16	51.4	81.0	52.0	64.3	16.7	10.5
48	9/23	50.2	76.4	50.0	65.5	16.0	10.2
49		40.4	76.4	56.7	74.3	16.0	11.6
52	10/20	60.3	69.5	37.1	53.4	14.9	7.71
55		60.5	70.2	37.6	53.6	15.1	7.83

[a]For each test parameters such as feed and recycle flow, feed and recycle COD, and PRZ operating volume were measured (not provided here) and used for calculation of the PRZ performance parameters.

[b]Substrate concentration in the PRZ.

[c]Floc load.

[d]Biosorption.

[e]Substrate removal efficiency in PRZ.

[f]Organic loading in PRZ.

[g]Reaction rate in PRZ.

TABLE 7.5

PERFORMANCE OF PRZ IN REACTOR NO. 4
DURING INTENSIVE STUDY
October 26 through 28, 1985

Test No.	V[a] (ml)	t[b] (min)	C_l[c] (mg/l)	F[d] (mg COD/ g VSS)	B[e] (mg COD/ g VSS)	E[f] (%)	F/M[g] (g/g-day)	Rr[h] (g/g-day)
70	410	7.4	41.7	47.3	29.8	63.1	9.51	5.78
71	273	4.9	47.5	47.0	25.9	55.1	14.21	7.55
72	575	10.4	36.9	47.8	33.3	69.7	6.85	4.60
73	187	3.4	49.1	46.5	24.6	52.9	20.52	10.47
74	1,550	28.1	25.2	51.1	43.8	85.7	2.72	2.24
75	755	13.7	31.5	48.6	37.4	77.1	5.30	3.94
76	653	6.1	23.9	22.7	18.8	82.7	5.61	4.44
77	1,025	9.5	23.9	23.4	19.4	82.8	3.69	2.93
78	1,600	14.9	21.4	24.2	21.7	89.3	2.44	2.09
79	325	3.0	30.1	22.4	15.0	66.8	11.14	7.13
80	693	21.7	48.8	104.0	78.7	75.7	7.19	5.22
81	315	9.8	66.8	100.8	62.2	61.7	15.38	9.10
82	136	4.2	82.9	99.7	49.1	49.3	35.43	16.72
83	113	5.4	142.5	204.8	74.1	36.2	57.16	19.89
84	113	5.6	155.4	219.2	70.2	32.0	59.17	18.21
85	315	14.6	1,115.9	197.6	95.7	48.4	20.26	9.43
86	410	18.9	91.8	196.9	119.6	60.7	15.58	9.09
87	214	9.9	122.3	195.1	87.1	44.7	29.53	12.69

[a] PRZ volume

[b] Hydraulic retention time in PRZ.

[c] Substrate concentration in the PRZ.

[d] Floc load.

[e] Biosorption.

[f] Substrate removal efficiency in PRZ.

[g] Organic loading in PRZ.

[h] Reaction rate in PRZ.

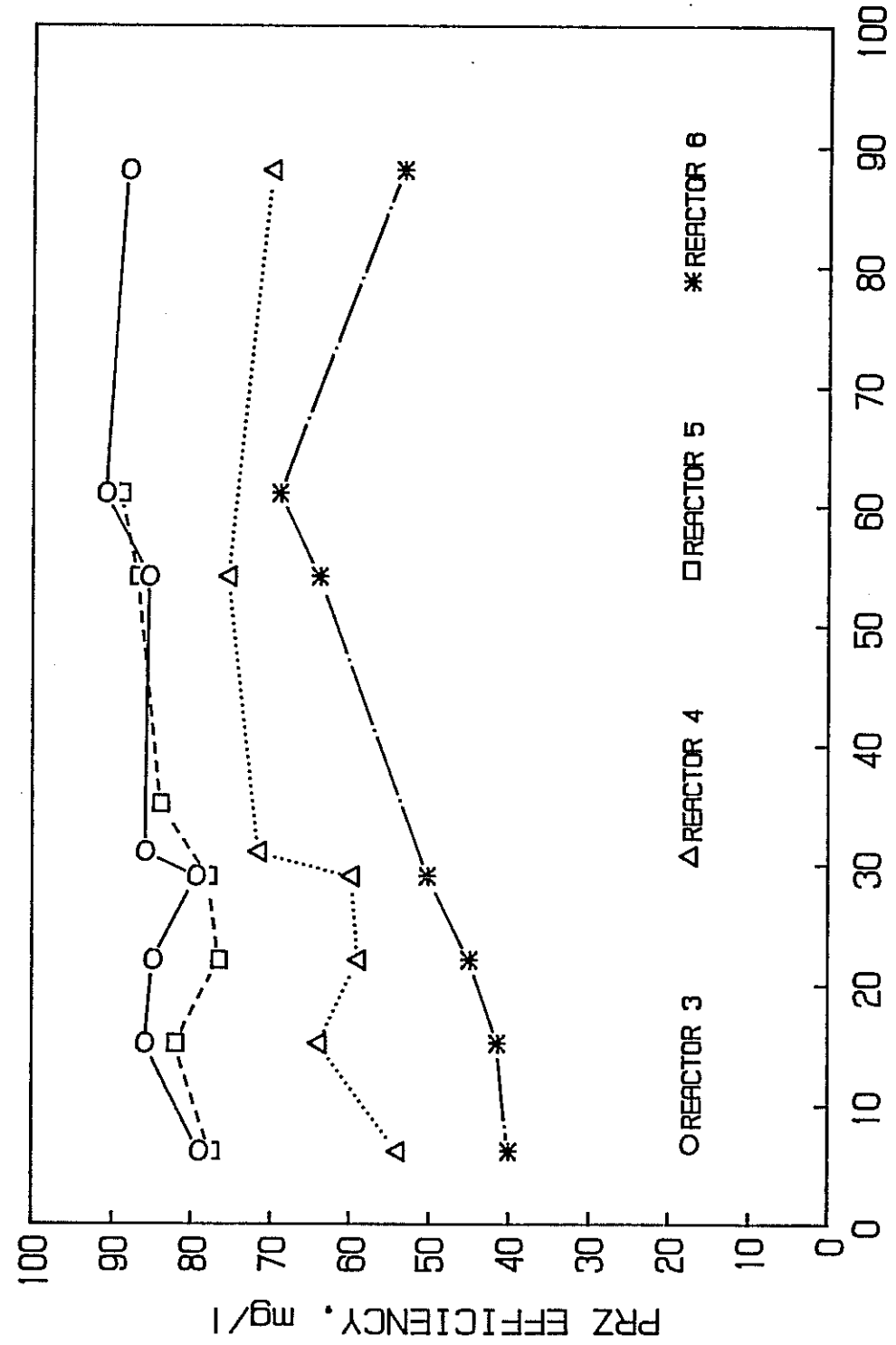


Figure 7.6. Chronological Plot of PRZs Substrate Removal Efficiencies (E).

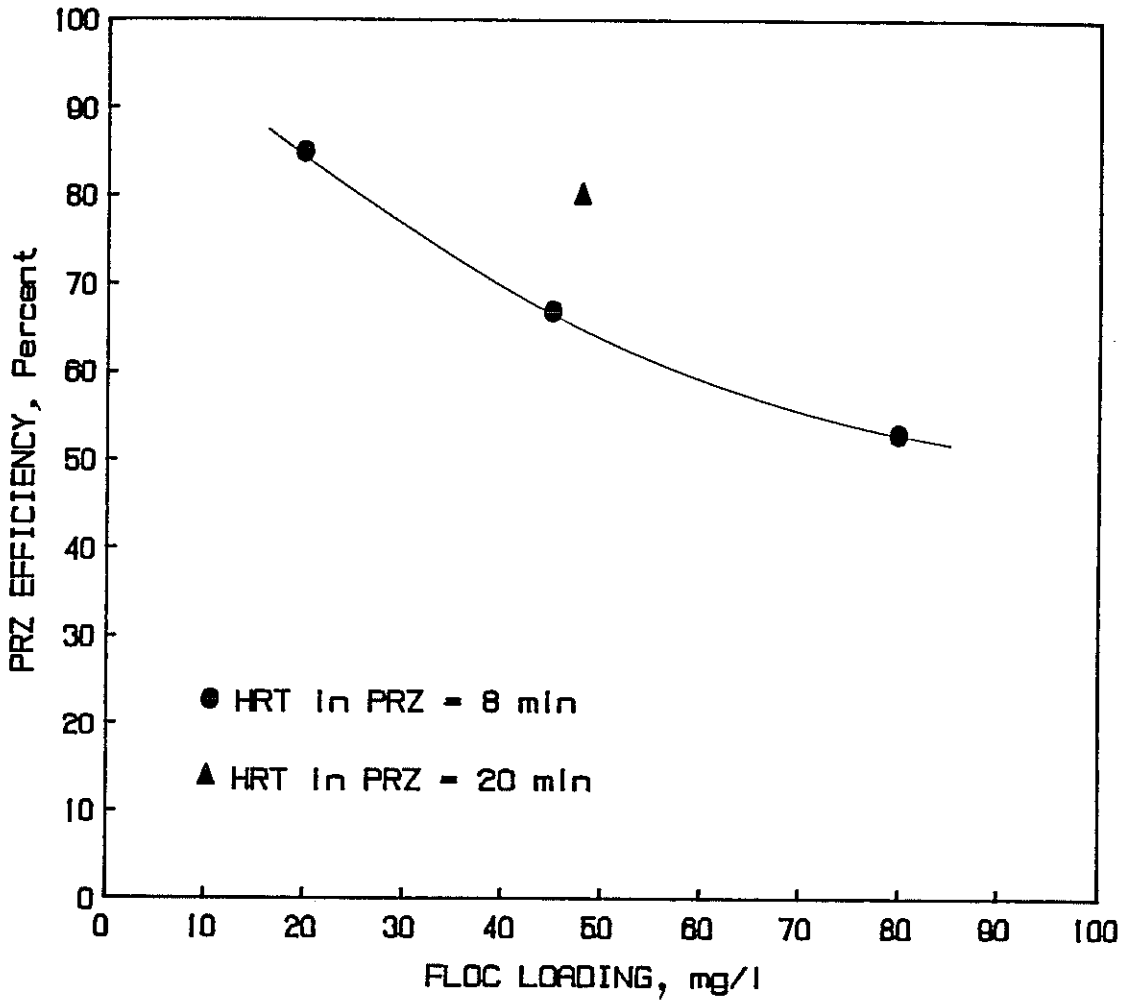


Figure 7.7. Correlation Between Average PRZ Efficiency and Floc Loading.

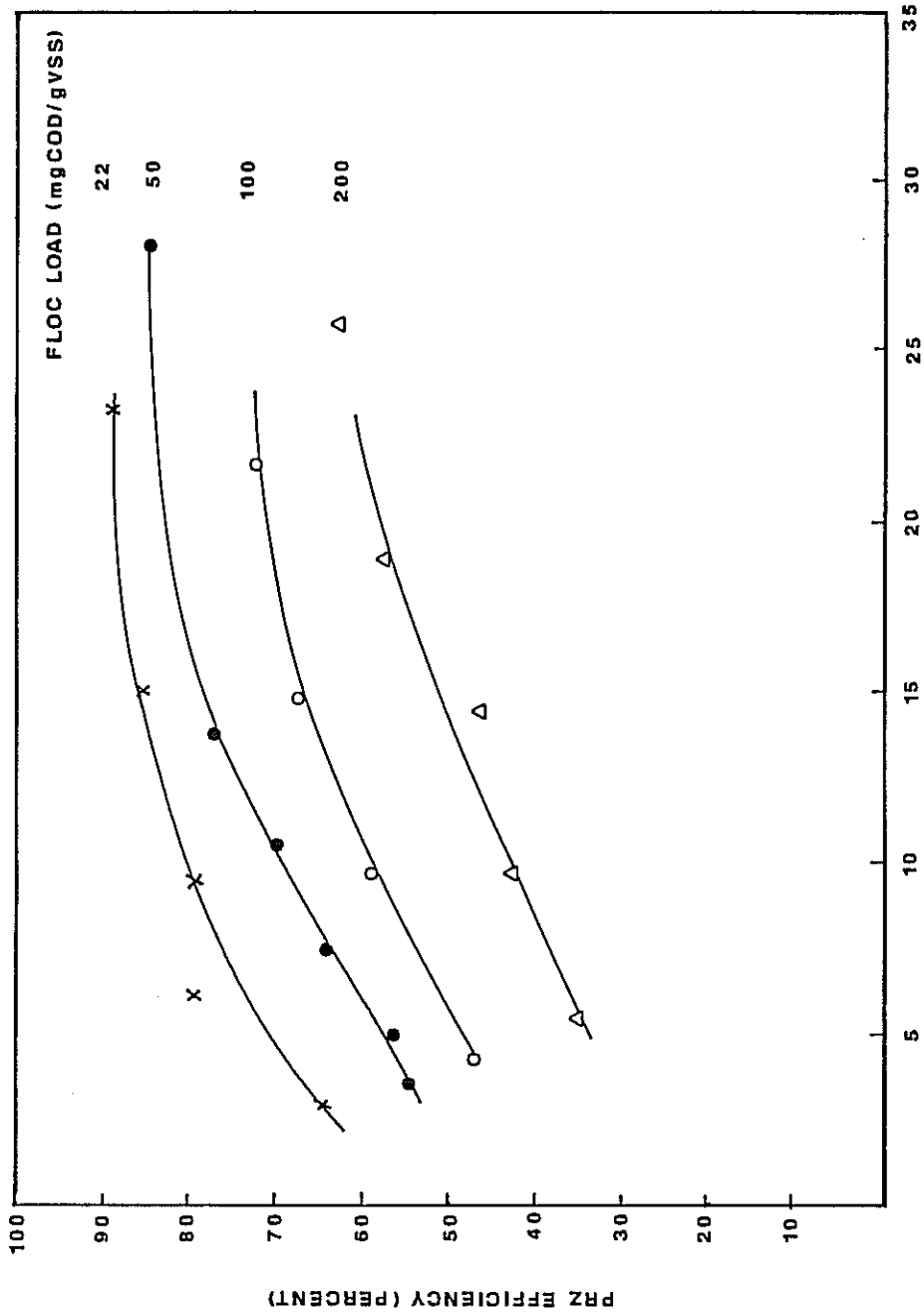


Figure 7.8. Correlation Between PRZ Efficiency and Hydraulic Retention Time in PRZ (Reactor No. 4).

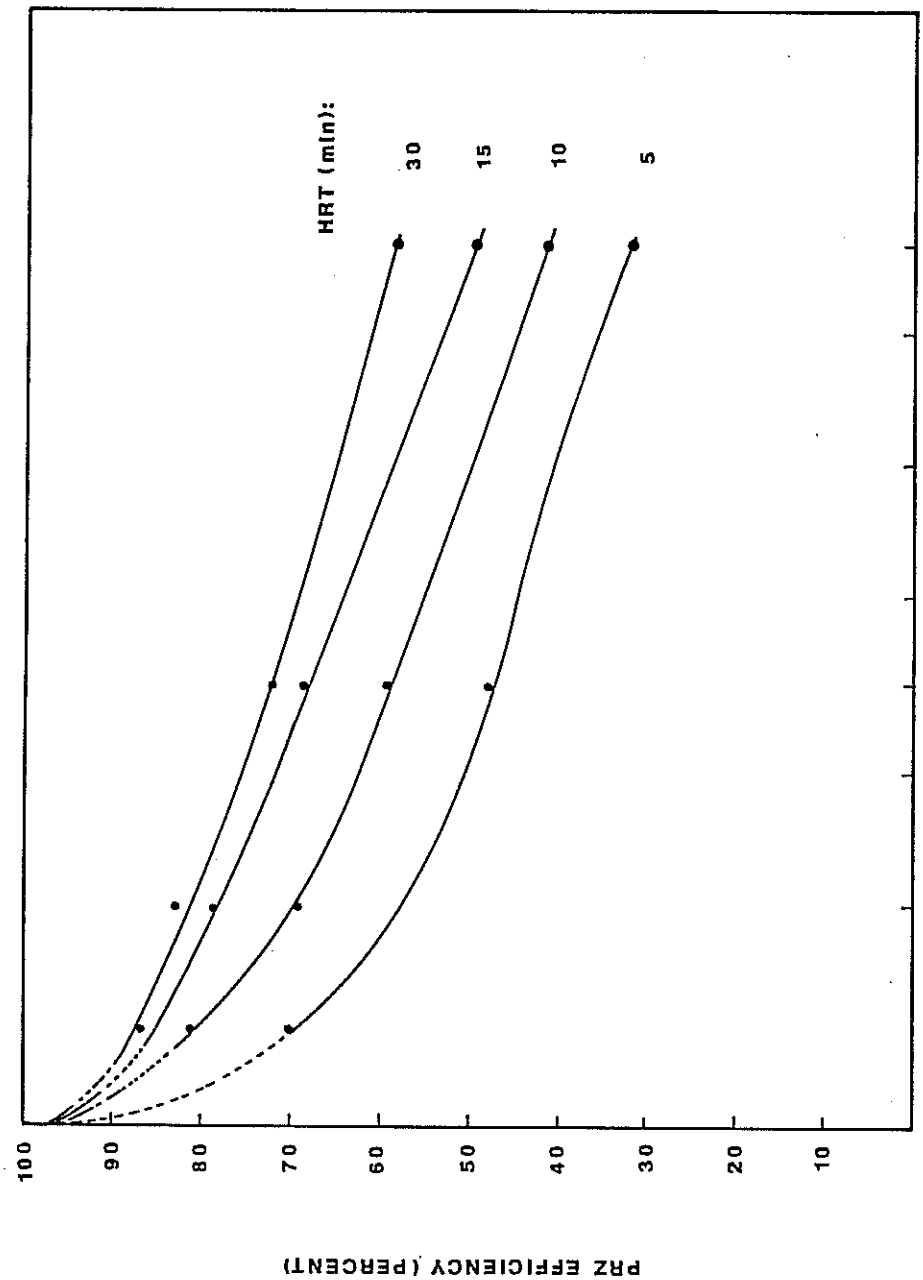


Figure 7.9. Correlation Between PRZ Efficiency and Floc Loading (Reactor No. 4).

substrate and other parameters used during the operation of the tested reactors.

In order to develop a more general relationship the PRZ performance data were correlated using a semi-empirical function similar to the one proposed by Suschka (1980) for a completely mixed activated sludge system (refer to Chapter IV). The proposed model has the following form:

$$R_r = R_m(F/M)/(K_w+F/M) \qquad (7.2)$$

Where:

Si Rr = reaction rate in PRZ, gCOD removed/gVSS-day

F/M = organic loading in PRZ, gCOD/gVSS-day

R_m = maximum reaction rate, gCOD/gVSS-day

K_w = half-velocity loading, gCOD/gVSS-day

Transformation of Equation 7.2 gives its linearized form:

$$1/R_r = (K_w/R_m)(1/F/M) + 1/R_m \qquad (7.3)$$

Correlation of the performance data from the intensive study in reactor No. 4 using equation 7.3 is shown in Figure 7.10. The correlation obtained can be considered extremely good with a correlation coefficient of 0.995. The resulting equation is:

$$1/R_m = 1.089 \cdot 1/F/M + 0.0428 \qquad (7.4)$$

Recalculation of the constants gives the final formula for modeling the PRZ efficiency in Reactor No. 4.

$$R_r = \frac{23.4\text{g/g-day} \cdot F/M}{25.5\text{g/g-day} + F/M}, \quad \text{gCOD removed/gVSS-day} \qquad (7.5)$$

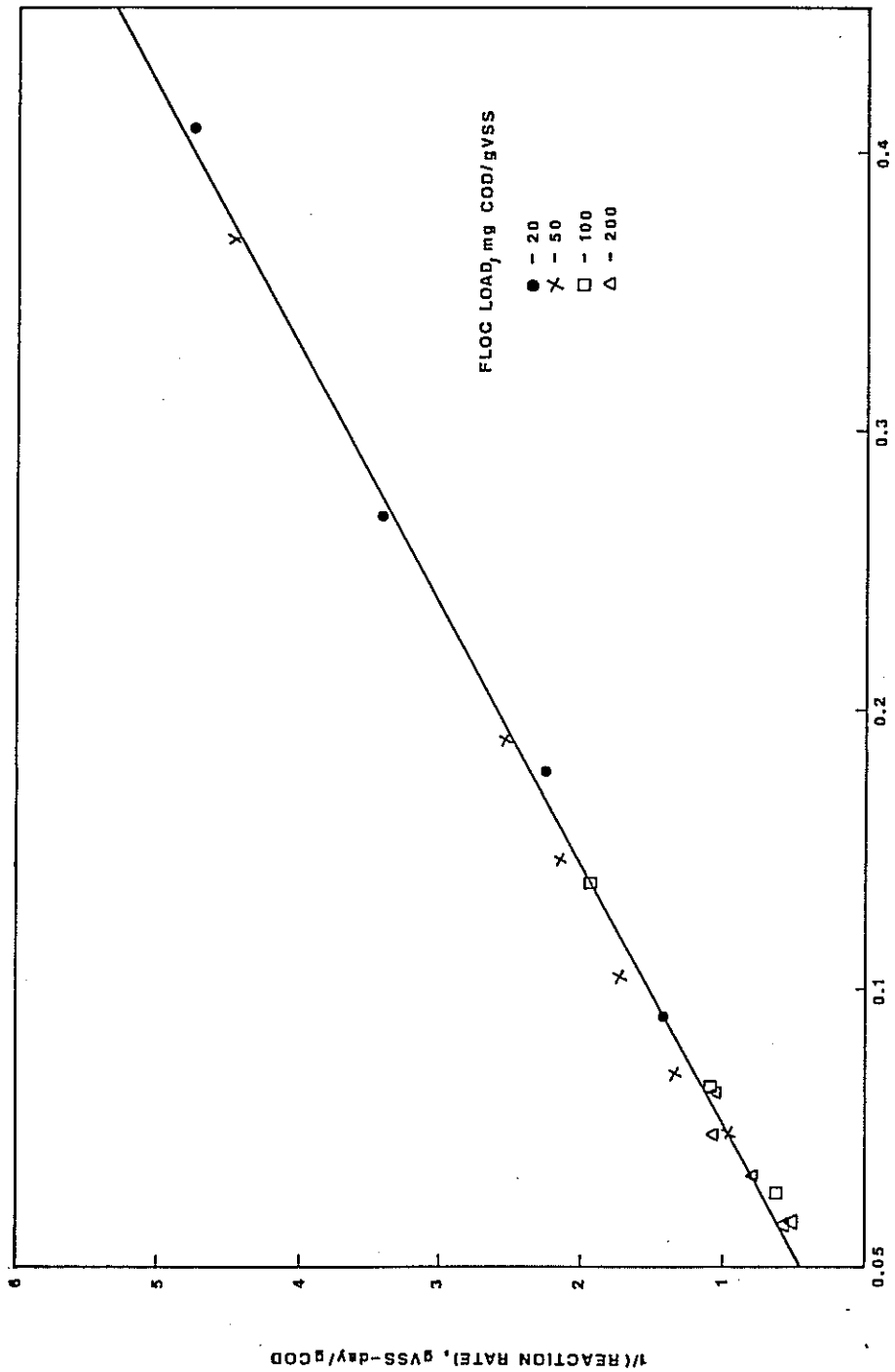


Figure 7.10. Correlation of PRZ Performance Data (Reactor No. 4).

Use of Formula 7.2 can be supported by the following theoretical considerations applicable to any biological system.

The reaction rate in any bacterial culture can not exceed a limiting, maximum rate, corresponding to the rate of the limiting reaction or transport step. In this case the most probable limiting factor is a finite amount of enzymes available and/or limited bacteria surface (active sites) which control the substrate's transport through the cell's membrane. Both phenomena are characteristic of catalytic and heterogenous reactions. Obviously, the activated sludge process fits into both categories, considering that enzymes are in fact biocatalysts. The maximum reaction rate is specific for a given microbial culture and substrate type.

Continuous (or intermittent) exposure to a high substrate concentration can likely result in a physiological adaption of the culture and/or in a change in the bacterial species distribution. This will result in a somewhat higher maximum reaction rate for the culture, though there is still a limiting maximum specific reaction rate attainable for the biological assimilation of a given substrate mix.

At the other end of the spectrum, at very low substrate concentrations (loadings) the system's performance is bound to approach complete substrate removal, and the observed reaction rate will correspond to the applied loading. This is particularly true when the loading rate is expressed in terms of biodegradable substrate such as

TABLE 7.6

RESULTS OF CORRELATION OF LITERATURE DATA ON PERFORMANCE OF SINGLE-TANK COMPLETELY MIXED ACTIVATED SLUDGE SYSTEMS

Source	Wastewater Type	Parameter	R_m (g/gVSS-day)	K_w (g/gVSS-day)	Correlation Coefficient	K_w/R_m
Suschka, 1980	Phenolo-formaldehyde	COD	46.1	51.6	0.9916	1.12
Wood and Sheldon, 1975	Mixed industrial and municipal	BOD_5	19.0	20.3	0.9998	1.07
Guarino et al., 1974	Municipal	BOD_5	16.1	17.4	0.9968	1.09
Krishnan and Gaudy, 1966	Glucose	COD	40.3	39.9	0.9968	0.99
This study	Synthetic, soluble, readily biodegradable	COD	23.4	25.5	0.9950	1.09

simple and reliable way of correlating and predicting the performance of a completely mixed activated sludge system, as the literature data presented above indicates.

Experimental data from this study, collected on the PRZs under a wide range of operating conditions, demonstrate that Formula 7.2 can successfully model performance of an initial contact chamber (selector). In view of a rapidly growing interest in application of selectors for control of filamentous bulking, the proposed model can potentially be a valuable research and design tool.

Validity of the formula for prediction of the performance of selectors was verified only for the systems studied in this work. Despite a large body of literature on selectors application to sludge bulking control, no complete set of selector performance data was found to confirm or reject the proposed model.

From Formula 7.2, expressions for the calculation of the substrate concentration in a PRZ and the PRZ's substrate removal efficiency can be readily derived.

$$C_1 = C_A - R_r X_1 t \qquad (7.7)$$

Where:

C_1 = substrate concentration in PRZ

C_A = weighted average substrate concentration in feed and recycle = $(C_o Q_o + C_R Q_R)/(Q_o + Q_R)$

X_1 = $X_R Q_R/(Q_R + Q_o)$ = sludge concentration in the PRZ (neglecting growth)

t = hydraulic retention time in the contactor

and, assuming that $K_w = R_m$

$$E = R_r/(F/M) = R_m(F/M)/(F/M + R_m)/(F/M) = R_m/(F/M + R_m) \qquad (7.8)$$

Where:

E = Fraction of the substrate removed in PRZ

An alternative formula for E has a form:

$$E = (C_A - C_1)/C_A \qquad (7.9)$$

The only parameter which requires an experimental determination is R_m, the maximum reaction rate. In the subsequent sections it will be demonstrated that the R_m value which was obtained from the PRZ data correlation is indeed a maximum reaction rate for a given sludge-substrate system, as measured in batch-type experiments.

Model for Optimization of the PRZ Design

Theoretical Considerations

As was previously discussed in the literature review section, it was postulated that filamentous bacteria tend to have a lower saturation constant and a lower maximum reaction rate than the floc forming bacteria. This would result in a higher growth rate of filamentous bacteria at low substrate concentrations with their resulting predominance under such conditions. An opposite situation would exist at high substrate concentrations, where the floc forming bacteria would grow faster and eventually prevail.

The theory was directly verified for two species of filamentous bacteria (Van den Eynde et al., 1982 and Van Veen et al., 1982) and indirectly for bulking and non-bulking heterogeneous cultures for a

variety of different specific substrates (Chudoba et al., 1985). Application of this theory in practice is not a simple matter. Qualitatively, creation of a zone of high substrate concentration, either in space or time, certainly works toward maximizing growth of the floc formers. In particular, an initial contact chamber was studied by numerous workers with mostly positive results (refer to Chapter II).

Based on the experimental data, different operational parameters of a selector were proposed as a controlling factor in optimization of the selector performance for bulking control. These included: dispersion number (Chudoba et al., 1973a), floc load (Eikelboom, 1982), F/M (Boyle, 1982), and performance number (Lime and Chiesa, 1985). Only the last concept offered a quantitative criterion of system "goodness" for bulking control. It was based on the measured performance data from the actual systems (substrate profile, oxygen uptake rates), however, and therefore is not useful for a predictive design.

Van Niekerk et al. (1986) presented the only quantitative model of the selector effectiveness in bulking control. The model utilizes a standard mass balance and microbial growth kinetic equations for two competing filamentous and floc-forming species. The resulting matrix of six differential equations was solved numerically for a set of about two dozen experimentally determined or assumed parameters to arrive at the optimum size (hydraulic retention time) of a selector, or rather at a range of HRT's in which the net specific growth of the floc former

was greater than that of the filamentous species. Such process parameters as sludge recycle rate, reactor hydraulic retention time, sludge age and substrate concentration were fixed. Specific, individual kinetic constants for the two bacterial species, (maximum growth rate, saturation constant, yield coefficient and endogenous decay rate) different for balanced (main tank) and unbalanced (selector) growth were derived from the experimental data and used in the model. Obviously, such an approach is not a practical design tool, because of the number of assumptions involved and the model's complexity.

In an attempt to optimize the design of the selector two important factors need to be considered. One of them is the substrate concentration in the selector. From the already verified theory, the higher the substrate concentration the more favorable is the growth rate (or substrate uptake) of the floc formers in their competition with the filamentous bacteria. From the published data (Chudoba, et. al., 1985), it appears that for most of the individual substrates a cutoff concentration, at which floc formers have higher growth rate than filaments, is below 5 mg/l. In practical application, this concentration is substantially exceeded in almost any PRZ in terms of total substrate concentration expressed in terms of collective parameters such as TOC, COD or BOD_5. It is quite obvious, though, that for a heterogeneous substrate, a multitude of different filamentous species would compete with different floculant species,each of them showing a different affinity for the different components of

the feed. In any case, the assumption can safely be made that the higher the substrate concentration in the selector the better chance the floc formers will have to overgrow the filamentous species.

The second factor involved is the fraction of the substrate removed in the contactor. Obviously, if the selector efficiency is low, the majority of the substrate passes to the main aeration tank. In such circumstances the selector will not serve its purpose, even if the substrate concentration in it is high, since the bulk of the substrate will be removed at the prevailing low concentrations of the main aeration basin favoring the filaments. At the same time, if the selectors' removal efficiency is high, the substrate concentration in the selector will be necessarily lowered, working against the high substrate concentration requirement.

Model Development

In order to balance these two conflicting requirements, the weighted, average substrate concentration at which the substrate was removed is proposed to be an optimization parameter. For a system consisting of two completely mixed tanks (selector and aeration basin) the formula for the weighted average reactant concentration has the following form (as previously used for construction of Figure 7.5).

$$C_B = C_1E + C_2(1-E) \qquad (7.10)$$

Where:

C_1 = biodegradable substrate concentration in the contactor

C_2 = biodegradable substrate concentration in the aeration basis

E = fraction of the substrate removed in the contactor.

For practical purposes, the biodegradable substrate concentration in the aeration basin is negligible, and Formula 7.10 reduces to the previously used equation:

$$C_B = C_1E \tag{7.1}$$

It is proposed here that the criterion for optimization of the PRZ design is to maximize the value of C_B. For a solution of the problem a relationship between the PRZ operating parameters and its removal efficiency (and consequently C_1) must be known. Formula 7.2, developed from the experimental data, provides such a relationship. The formula defines the reaction rate in a PRZ as a function of the PRZ's organic loading and has a form:

$$R_r = R_m(F/M)(K_W+F/M) \tag{7.2}$$

The following relationships hold (refer to Figure 5.1 for the nomenclature).

$$F/M = C_oQ_o/(Q_RX_Rt) \tag{7.11}$$

$$X_1 = X_RQ_R/(Q_o+Q_R) \tag{7.12}$$

Let us define D, the dilution rate as:

$$D = Q_o/(Q_o+Q_R) \tag{7.13}$$

From a steady state mass balance, the available substrate concentration in the PRZ (C_1) is equal to:

$$C_1 = C_oD - R_rX_1t \qquad (7.14)$$

Incorporation of Formulas 7.2 and 7.11 through 7.13 into Formula 7.14 then yields equation for biodegradable substrate concentration in the PRZ:

$$C_1 = [D^2C_o^2/X_R + DC_oK_W(1-D)t - R_mC_oD(1-D)t]/[Dc_o/X_R + K_s(1-D)t] \qquad (7\text{-}15)$$

The value of K_W can be related to R_m:

$$K_W = aR_m \qquad (7.16)$$

Where a is a dimensionless constant, found previously to be close to unity.

Rearrangement of Formula 7.15 yields:

$$C_1 = C_o[D^2+AD(1-D)(a-1)t]/[D+a(1-D)At] \qquad (7.17)$$

Where:

A = system constant = R_mX_R/C_o

The efficiency of substrate removal in the PRZ is related to C_1 as follows:

$$E = (D(C_o-C_1)/(DC_o) \qquad (7.18)$$

Substitutions yield

$$E = A(1-D)t/[D + a(1-D)At] \qquad (7.19)$$

Incorporation of Formulas 7.17 and 7.19 into the expression for the average biosorption concentration (Formula 7.1) leads to the final formula:

$$C_B = C_oA(1-D)t[D^2+AD(1-D)(a-1)t]/[D+a(1-D)At]^2 \qquad (7.20)$$

Model Analysis

Examination of Formula 7.20 indicates that C_B is directly proportional to C_o, and therefore the PRZ operational parameters (D and t) for which C_B achieves a maximum value is not a function of C_o (barring involvement of C_o in the calculation of the system constant A). C_B is a function of two independent variables, the PRZ dilution rate, D, and the PRZ hydraulic retention-time, t. All other parameters can be assumed constant for a given system and are lumped into the system number, A, and the constant, a.

In order to determine the maximum of C_B, partial derivatives of the function should be found and the set of two equations with two unknowns solved. Initially, the partial derivatives were obtained, and set to zero for a special case when a=1. The resulting set of equations is as follows:

$$\partial C_B/\partial t = D-A(1-D)t = O \qquad (7.21)$$

$$\partial C_B/\partial D = D^2(At-1)-3ADt+2At=O \qquad (7.22)$$

Analysis of the equations indicates that the function does not have a local extremum in respect to t, and its value at optimum D continuously increases with increasing t.

Figure 7.11 illustrates the behavior of the function expressed in Formula 7.20 for several values of t and with A set to 10. From Figure 7.20 it is apparent that when value of A is fixed, the dilution rate corresponding to the maximum value of C_B approaches unity with increasing values of t.

The physical situation corresponding to this case is a PRZ with a large detention time and a very low sludge recycle rate. Such a system is in some sense an equivalent of batch treatment for a continuous flow situation. It can be readily shown that for batch treatment, C_B is equal to $C_o/2$. This is the highest C_B achievable for any system configuration. It is certainly encouraging to note that, indeed, batch treatment was frequently reported to be the most effective process configuration for filamentous bacteria control. This is evident from both this study and from others, discussed in Chapter II. A plug flow reactor performance with respect to C_B is theoretically equivalent to a batch reactor. However, a considerable degree of axial dispersion is present in any practical system, which diminishes the C_B value practically achievable in such systems.

It is apparent that a high value of C_B (i.e., $C_B = \frac{1}{2}C_o$) is not achievable in a continuous flow, completely mixed reactor with a PRZ since this would require removal of 50 percent of the substrate at its full (influent) concentration, or removal of 100 percent of substrate at half its initial concentration. Both of these situations and any combination in between is self-contradictory.

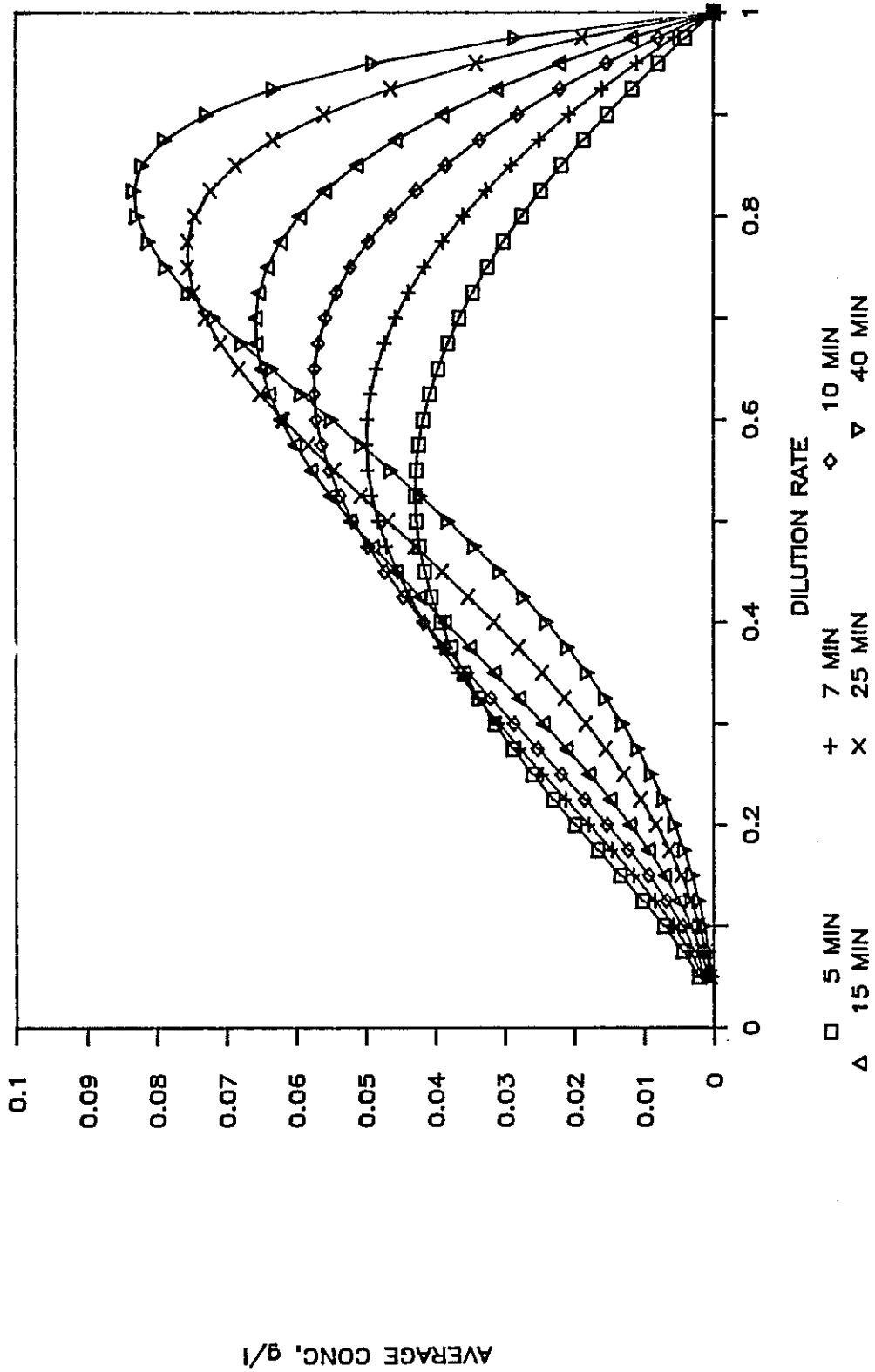


Figure 7.11. Effect of Dilution Rate on the Weighted Average Biosorption Concentration (C_B) for PRZ with Different Contact Times, (A=10, C_O=0.42 g/l, X_R=2 g/l, R_m=24 g/g-day).

In practical applications, the hydraulic retention time in the selector is limited by the physical and/or economic constraints and in most cases is between 10 to 30 minutes. With the selector size limited by the physical constrains of the existing facility, or set by a designer, the problem of finding the optimum selector operating parameters is reduced to finding the maximum of a function of one variable - dilution rate.

In this application (PRZ volume constant), the hydraulic retention time in the PRZ will depend on the dilution rate. It is therefore convenient to use a nominal hydraulic retention time in the PRZ (T). The following formula holds:

$$T = V_1/Q_o = t/D \tag{7.23}$$

Substitution into Equation 7.20 yields:

$$C_B = C_o(1-D)LD[1+L(1-D)(a-1)]/[1+a(1-D)\ L]^2 \tag{7.24}$$

Where:

$L = TA$ = system constant = $V_1R_mX_R/Q_oC_o$

For a special case, when a=1, the local maximum can be readily found and has the following form.

$$D_o = (L+1)/(L+2) \tag{7.25}$$

Where:

D_o = dilution rate at which the maximum C_B is obtained

Expressing the dilution rate in the more convenient form of the sludge recycle rate provides the following formula:

$$R_o = 100/(L+1), \text{ percent} \quad (7.26)$$

Where: R_o is the optimum sludge recycle rate (as percent of the influent flow).

When the coefficient a does not equal 1, differentation of the Function 7.24 with respect to D yields a third order function (in the numerator). Analytical solution of the resulting equation, while straightforward, is extremely tedious and it was found more practical to find the local maximum numerically for a matrix of system constants (L) and coefficients (a).

The results are presented in the form of a nomogram (Figure 7.12) from which the optimum sludge recycle rate can be determined for given values of L and a. The nonogram shows that the optimum sludge recycle rate to the selector is always less than 100 percent, and approaches this value for the systems characterized by a low value of the system constant, L. The optimum sludge recycle rate is for practical purposes independent of a, particularly after considering that for the most systems, a is very close to unity.

When for the special case a=1, the expression for the optimum dilution rate are substituted into Formulas 7.14, 7.18 and 7.20, and the following relations result:

$$C_1 = C_o/2 \quad (7.27)$$

$$E = L/2(L+1) \quad (7.28)$$

$$C_B = C_o L/4(L+1) \quad (7.29)$$

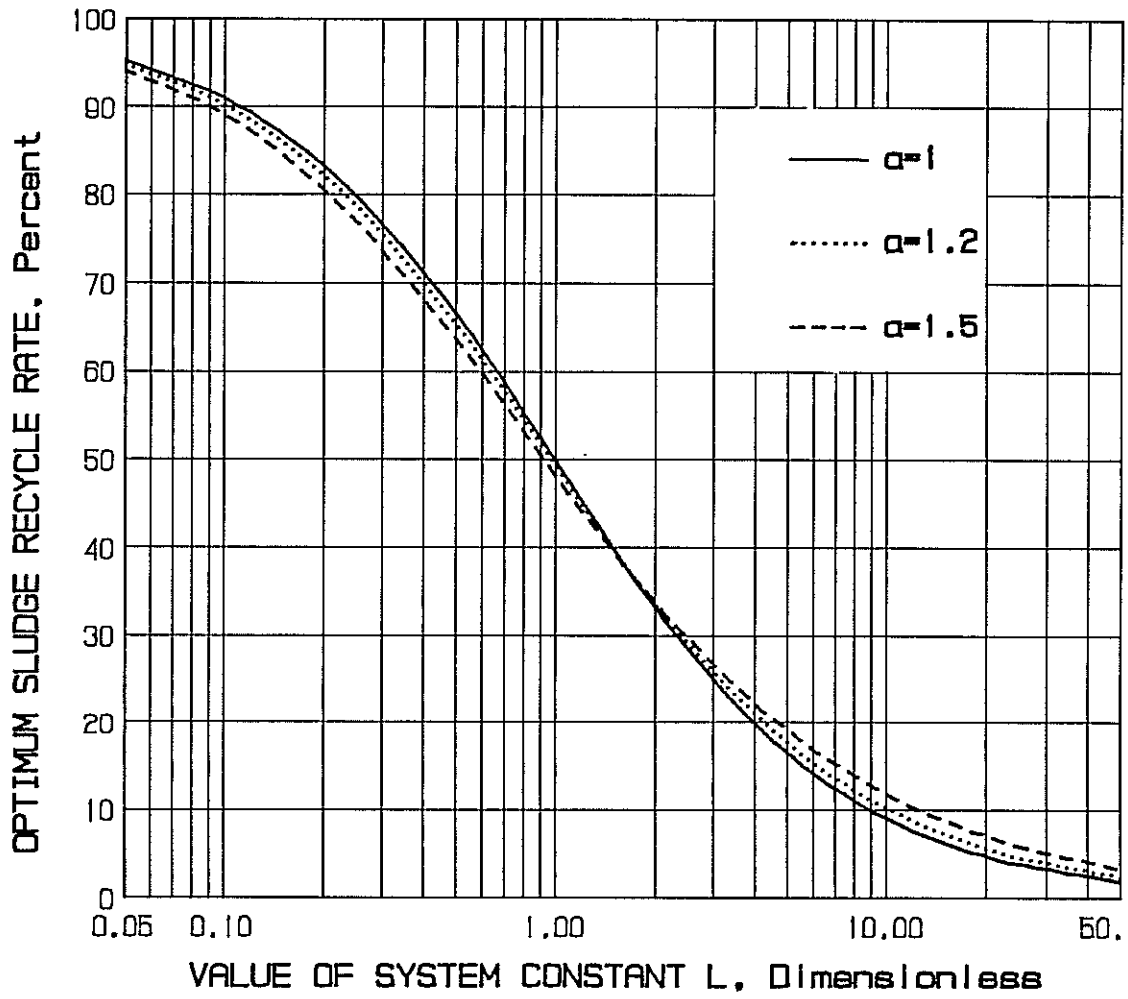


Figure 7.12. Nomogram for Determination of Optimum Sludge Recycle Rate.

It is then apparent that the selector operates at a maximum C_B when the substrate concentration in the selector is half that of the influent concentration. The efficiency of the substrate removal at the optimum sludge recycle rate is a function of L and increases asymptotically to 0.5 with an increasing system number. The relationships between selector efficiency, C_B, and L at the optimum recycle rate (R) are presented graphically in Figure 7.13 (For a = 1).

On Figure 7.14 the values of C_B are plotted as a function of the sludge recycle rate for several values of L. Figure 7.14 illustrates that little premium can be gained by increasing the system number above about 10, since at this point the maximum C_B reaches 23 percent of the feed concentration (C_o), while the maximum obtainable C_B for the continuous flow contactor is 25 percent of the C_o.

Model Limitations

It should be realized that the proposed model has been developed under several assumptions which limit its applicability.

Formula 7.2 describing the reaction rate in the PRZ as a function of F/M is felt to be applicable for any soluble substrate and for a wide range of F/M's in a PRZ. However, the equation describing the reaction rate in the PRZ was derived under an assumption that the biomass concentration in the PRZ is constant. In this and other studies the PRZs were operated with floc load as a design parameter and were usually in the range 20 to 100 mg substrate/gVSS. Under such loads (and assuming 50 percent removal efficiency), an increase in the

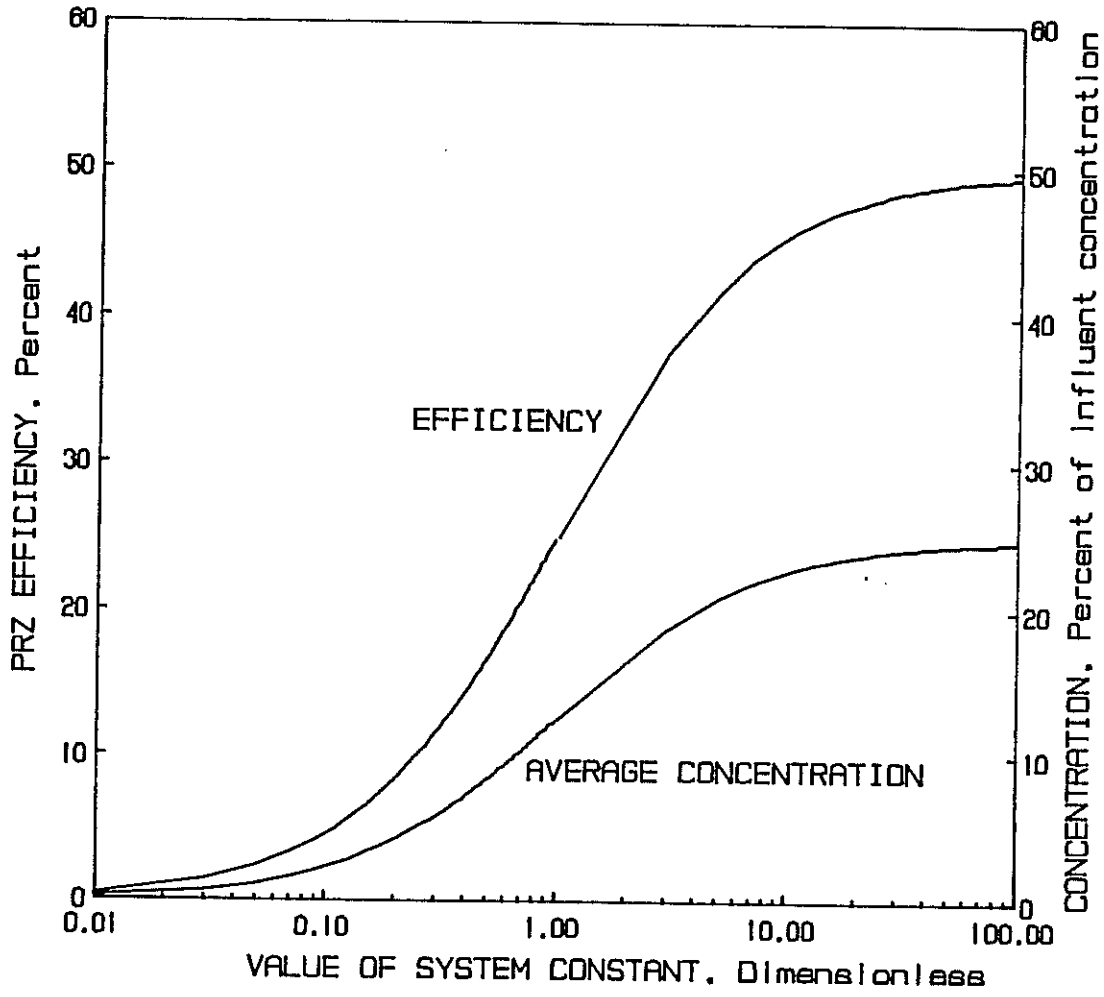


Figure 7.13. Expected PRZ Efficiency and Average Biosorption Concentration (C_B) at the Optimum Sludge Recycle Rate as a Function of the System Constant.

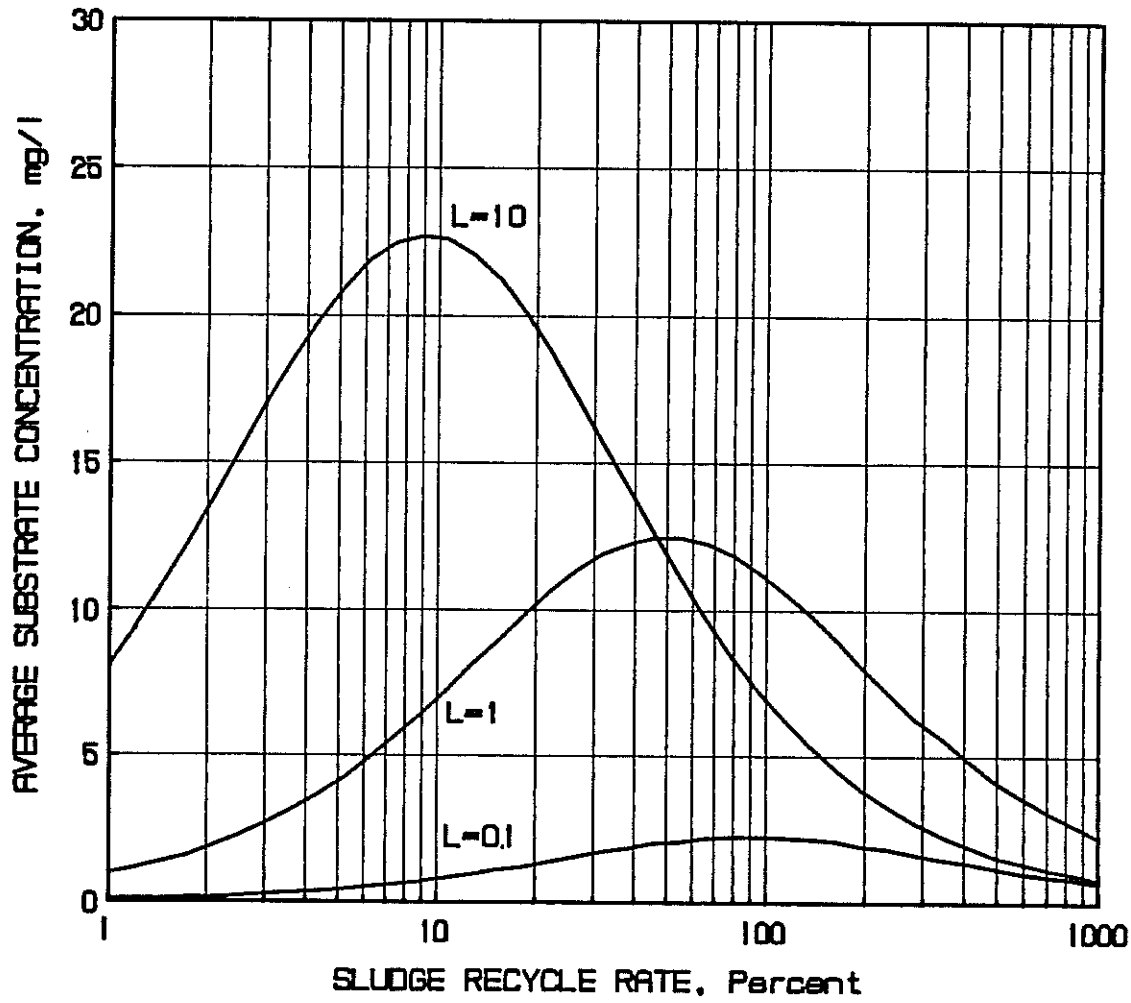


Figure 7.14. Average Substrate Concentration as a Function of Sludge Recycle Rate. (Initial Substrate Concentration = 100 mg/1.)

sludge concentration in the PRZ would be negligible (max about 5 percent). From the proposed model, much higher floc loads may be desirable under some conditions. For example, with system numbers greater than 5 it would be common to operate at floc loads around 1000 mg/g or more. Under such floc loads, the increase in the biomass concentration in the PRZ will be appreciable, and it could decrease the optimum sludge recycle rate for a given system.

Under extreme conditions, (high floc load, long HRT), equation 7.2 may not describe accurately the reaction kinetics in PRZ, since it would be out of the experimentally verified range. It is felt, however, that even in such a case the discrepancies from Formula 7.2 should be minor. This confidence is based on the previously discussed inherent validity of Formula 7.2 for the two boundary conditions (i.e., for low F/M, the R_r approaches F/M and for high F/M, R_r approaches R_m), and on the exceptionally good fit of the experimental data presented in this study for the proposed formula. If a modification of Formula 7.2 is deemed necessary a possibility of expressing R_r as a sum of two (or more) terms, each of them having the form of a Monod-type function with properly adjusted constants, should be explored. Preliminary work with such two-term functions demonstrated that it can model almost any realistic, continuous curve which meets the above delineated boundary conditions.

A more critical area of concern is the applicability of Formula 7.2 for influents where a part of the substrate is in colloidal or particulate form. In this study a completely soluble substrate was

used. However, frequently up to 30 percent of available substrate is in a colloidal form, particularly in municipal wastewaters.

Apart from the considerations regarding the adequacy of Formula 7.2, more questions might arise concerning the selection of Function 7.1 as the criterion of the PRZ optimization for bulking control. Equation 7.1 should be considered as a first generation approach, the simplest formula which satisfies theoretical requirements of maximizing removal concentration and removal efficiency in the PRZ at the same time. It is possible, and even quite likely, that some other formula embracing these two contradictory requirements is more appropriate.

The most obvious limitation of Formula 7.1 is the presence, at least in the theory, of a minimum concentration at which the floc formers have a higher growth rate than the filamentous species. If conditions in PRZ resulting from maximization of C_B are such that the substrate concentration C_1 is lower than the threshold concentration, proliferation of floc formers will not occur. It seems, however, unlikely that such a concentration can be defined for any real system, considering the multitude of species and substrates involved. It perhaps may be demonstrated that a more adequate optimization criterion can be formulated by giving more weight to the substrate concentration C_1, (e.g., by rising it to the second power in the Formula 7.1). Experimental work required for verification of any such formula is very difficult, considering the instability of the sludge settling characteristic in the reactors operated in this study.

At the same time if the selector efficiency resulting from application of Figure 7.12 is much less than 50 percent, most of the substrate would leak into the aeration basin. In such cases it may be advisable to increase recycle rate (efficiency), particularly if the resulting, lower value of C_1 is still above the critical concentration for the given system. Unfortunately, such adjustments are possible only if the specific values of the growth parameters for floc-formers and filamentous bacteria are available.

Kinetics of Substrate Removal

Batch Tests under Aerobic Conditions

A series of standard batch tests for evaluation of reaction kinetics was performed on sludge from reactor No. 4, using the procedure detailed in Chapter IV. Figures 7.15 through 7.19 present the data obtained in these tests under both aerobic and anaerobic conditions. Anaerobic data will be analyzed in the subsequent section, and the ensuing discussion concerns the aerobic results only.

The form of the substrate depletion curves in Figures 7.15 through 7.19 suggests a reaction with kinetics of higher than zero order, as expected for a heterogeneuous substrate. The substrate (COD) concentration levels-off after an initial decrease, approaching the "ultimate" COD concentration. Value of this "ultimate" COD was determined from Figures 7.15 through 7.19 and was subsequently used as a measure of the "non-biodegradable COD". This non-biodegradable COD is considered to be mostly metabolic byproducts released by the

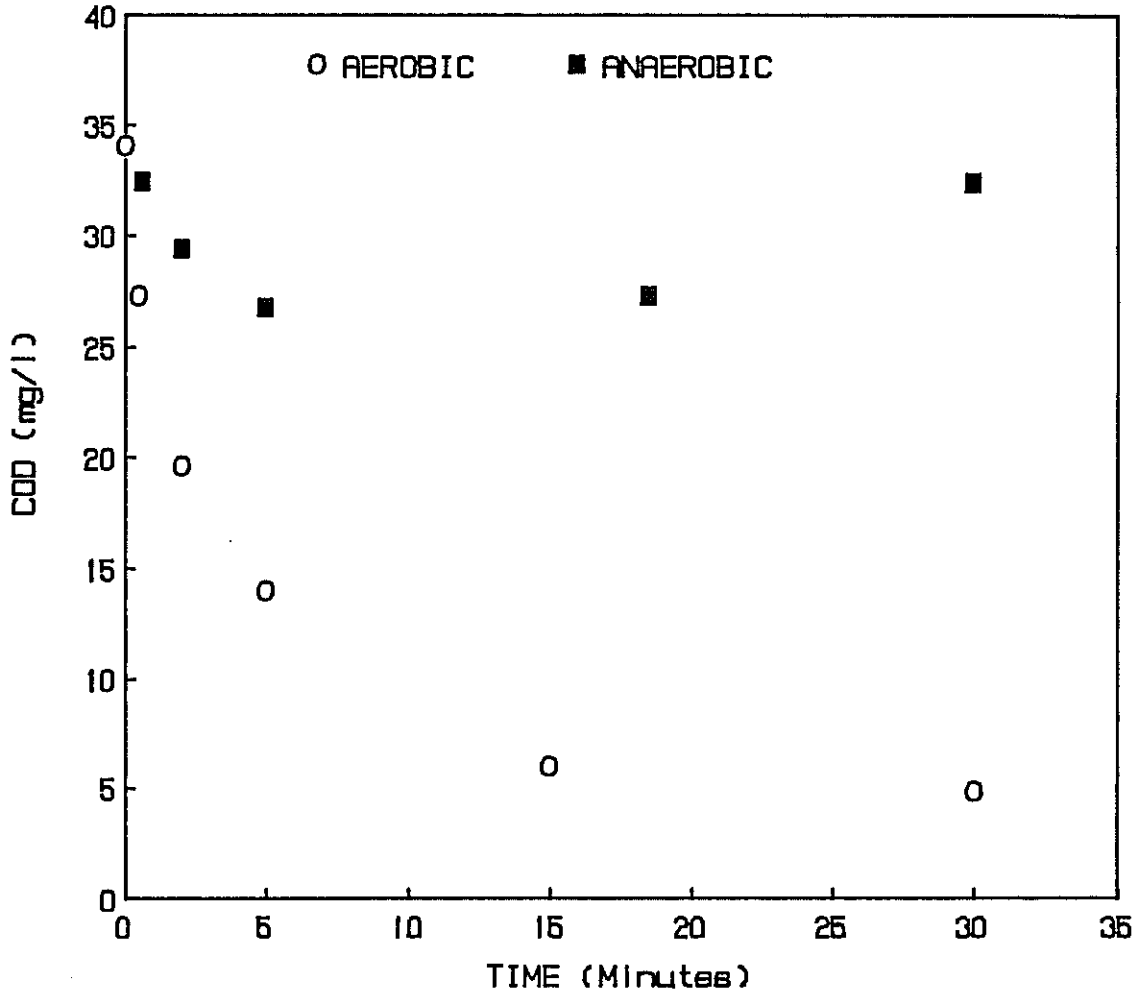


Figure 7.15. Results of Aerobic and Anaerobic Batch Tests on Reactor No. 4 - Run 1.

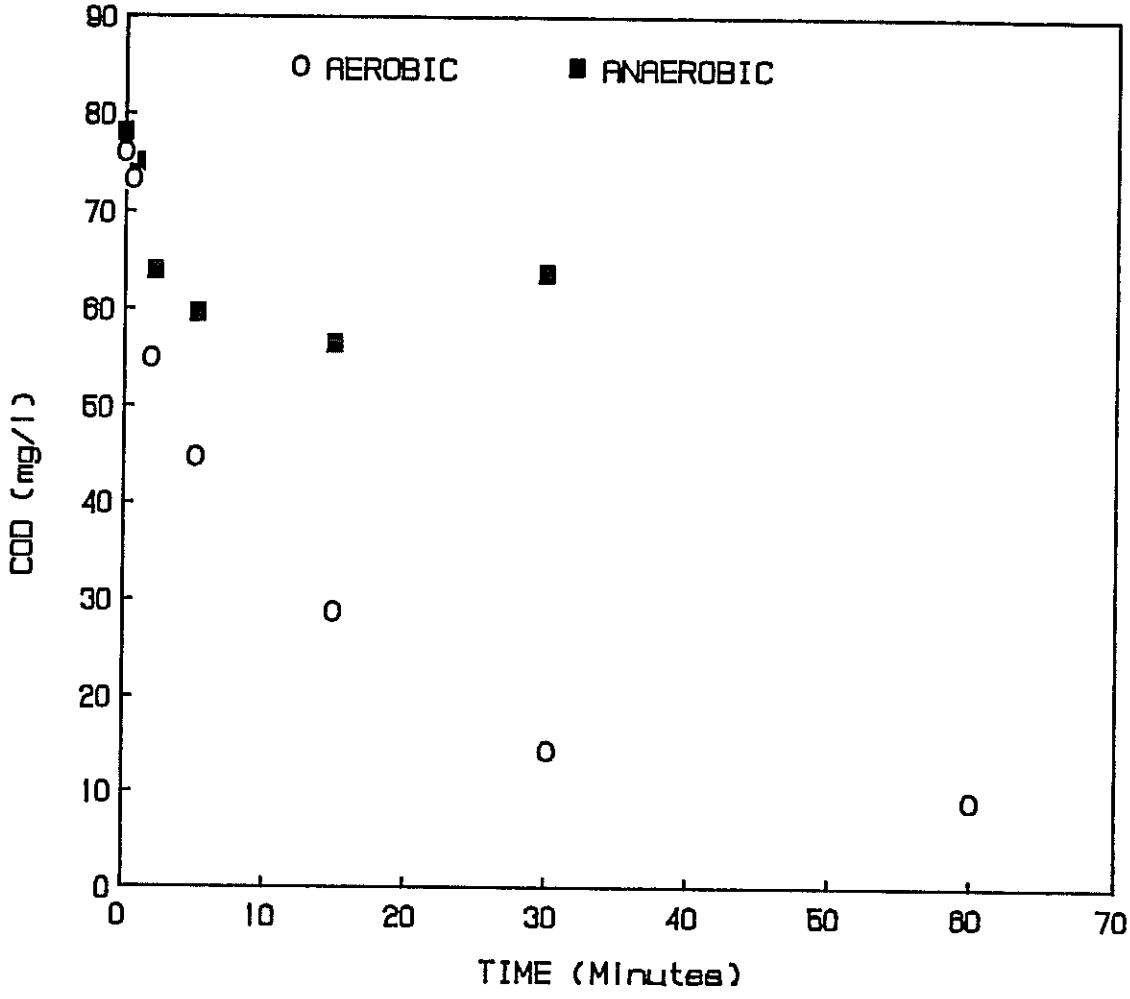


Figure 7.16. Results of Aerobic and Anaerobic Batch Tests on Reactor No. 4 - Run 2.

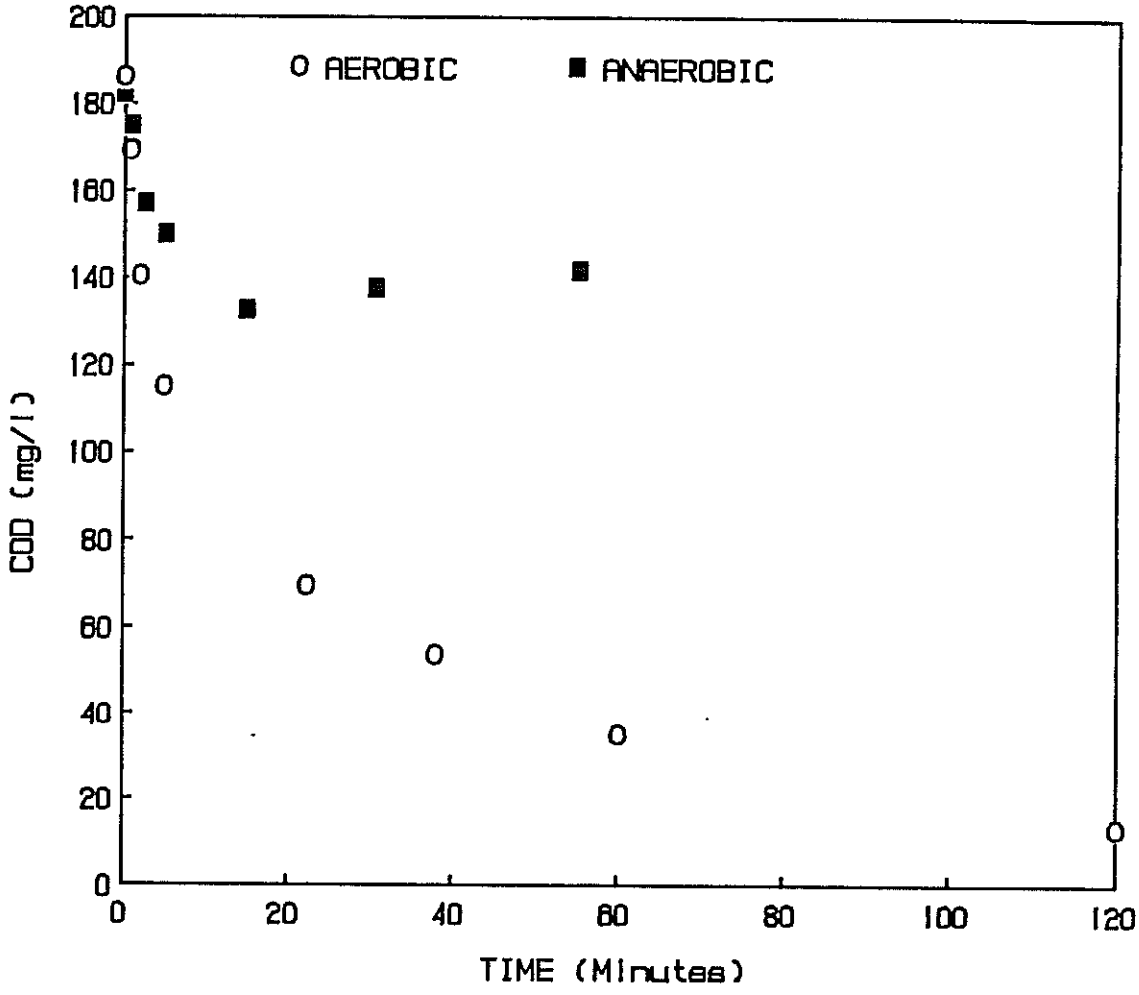


Figure 7.17. Results of Aerobic and Anaerobic Batch Tests on Reactor No. 4 - Run 3.

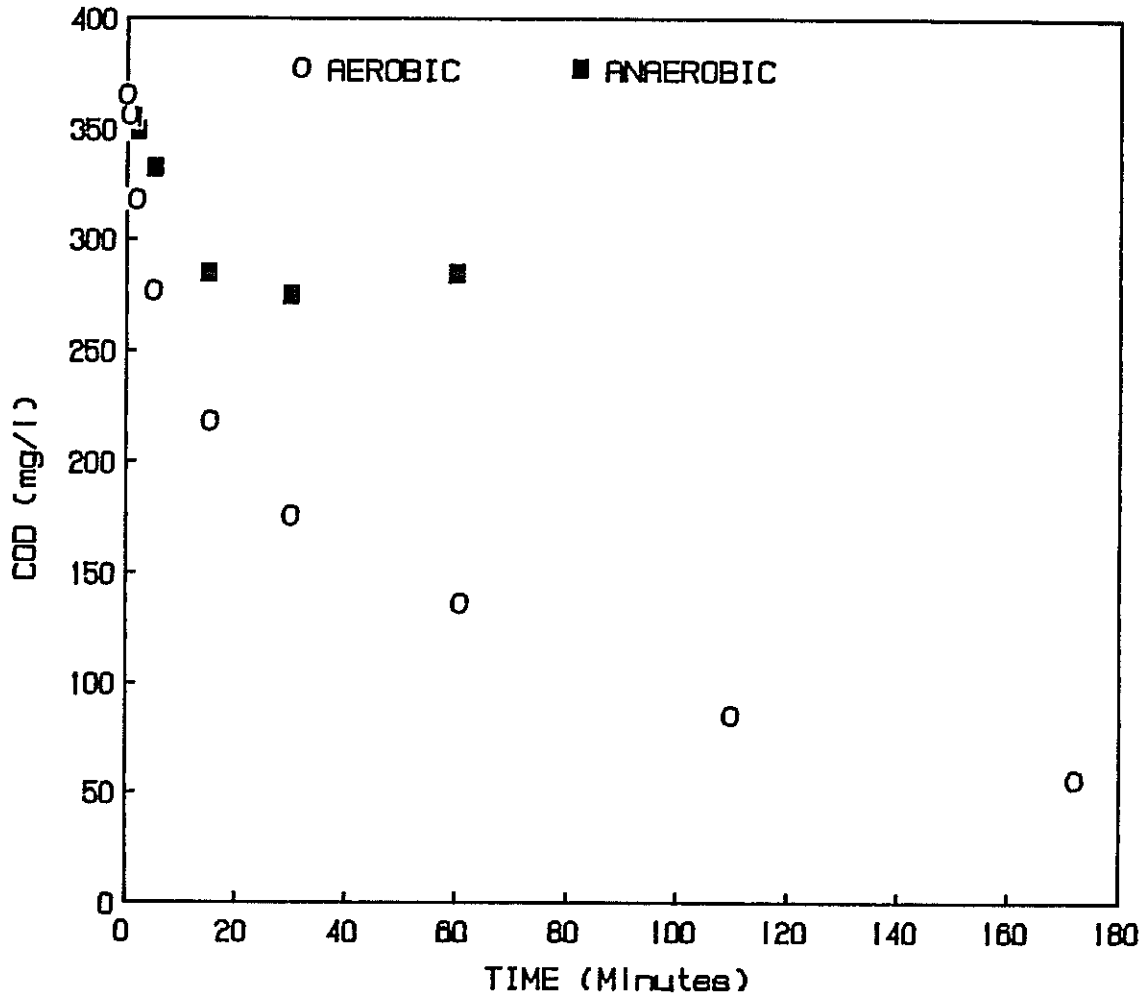


Figure 7.18. Results of Aerobic and Anaerobic Batch Tests on Reactor No. 4 - Run 4.

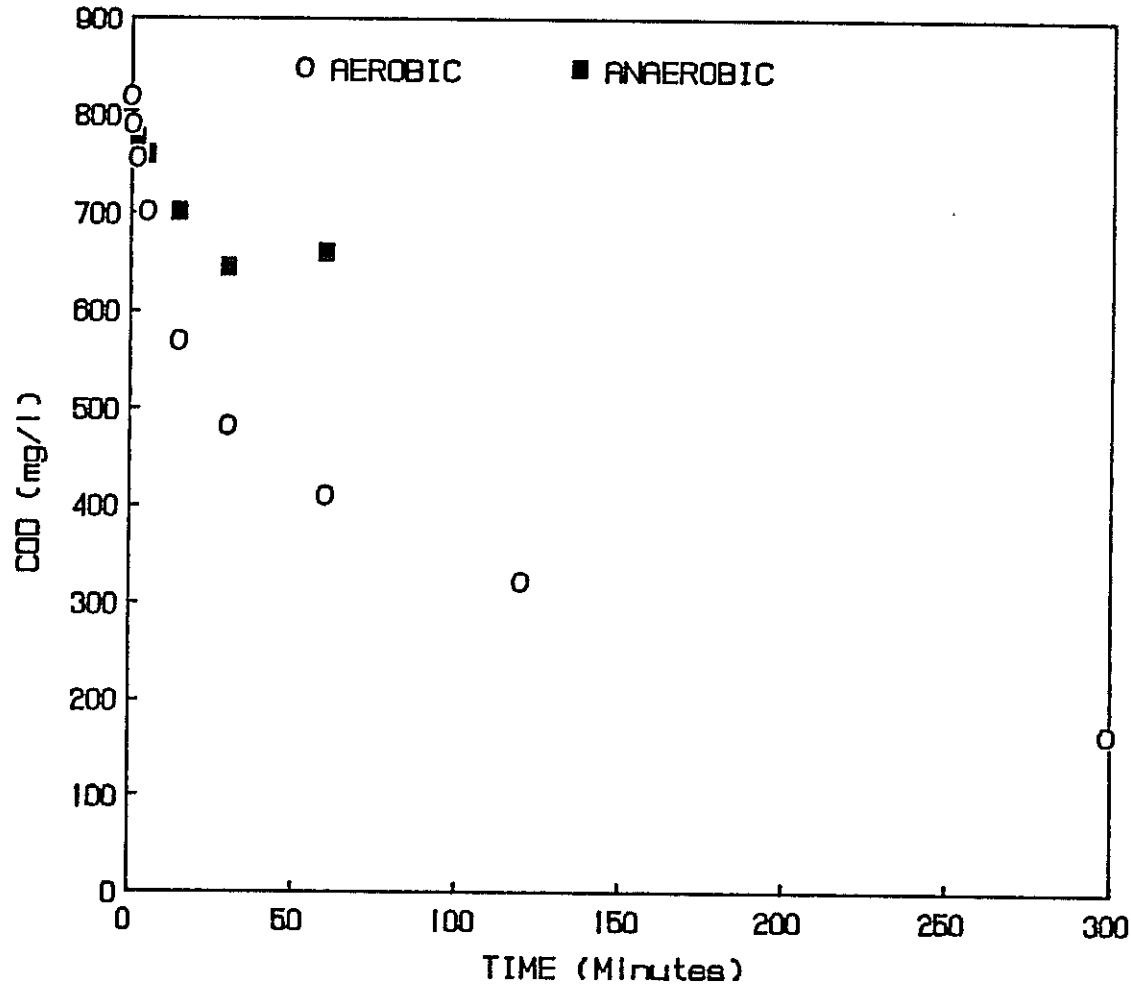


Figure 7.19. Results of Aerobic and Anaerobic Batch Tests on Reactor No. 4 - Run 5.

biomass, since the synthetic substrate used is thought to be completely biodegradable.

In order to determine if the reactions followed first order kinetics, the results are plotted on a semi-logarithmic scale on Figures 7.20 through 7.22. Values of the COD used on these graphs were adjusted to account for the build-up of the "byproduct" COD in proportion to the substrate removed. Review of these figures suggests that the reaction proceeds in two phases, both of which approximated first order kinetics with the first phase rate constant much higher than for the second phase. The first order rate constants determined from the presented data are summarized in Table 7.7, together with the other test parameters.

Two mechanisms can potentially account for the observed shift in the rate constants: concurrent substrate removal, and biosorption capacity. Under the first scenario, the individual substrate components are removed from the solution concurrently as was discussed in the literature review section. The substrate constituent with the highest individual zero-order rate constant (or, more correctly, with the lowest ratio of its initial concentration to the rate constant) will be removed from the solution first, with the resulting decrease in the overall reaction rate. The most likely candidate for such behavior is glucose, which constituted about one-third of the feed in terms of COD. The shifts in rate constants were indeed observed after the substrate concentration decreased by about one-third of its initial value. The at which the shift occurred was read from Figures 7.20

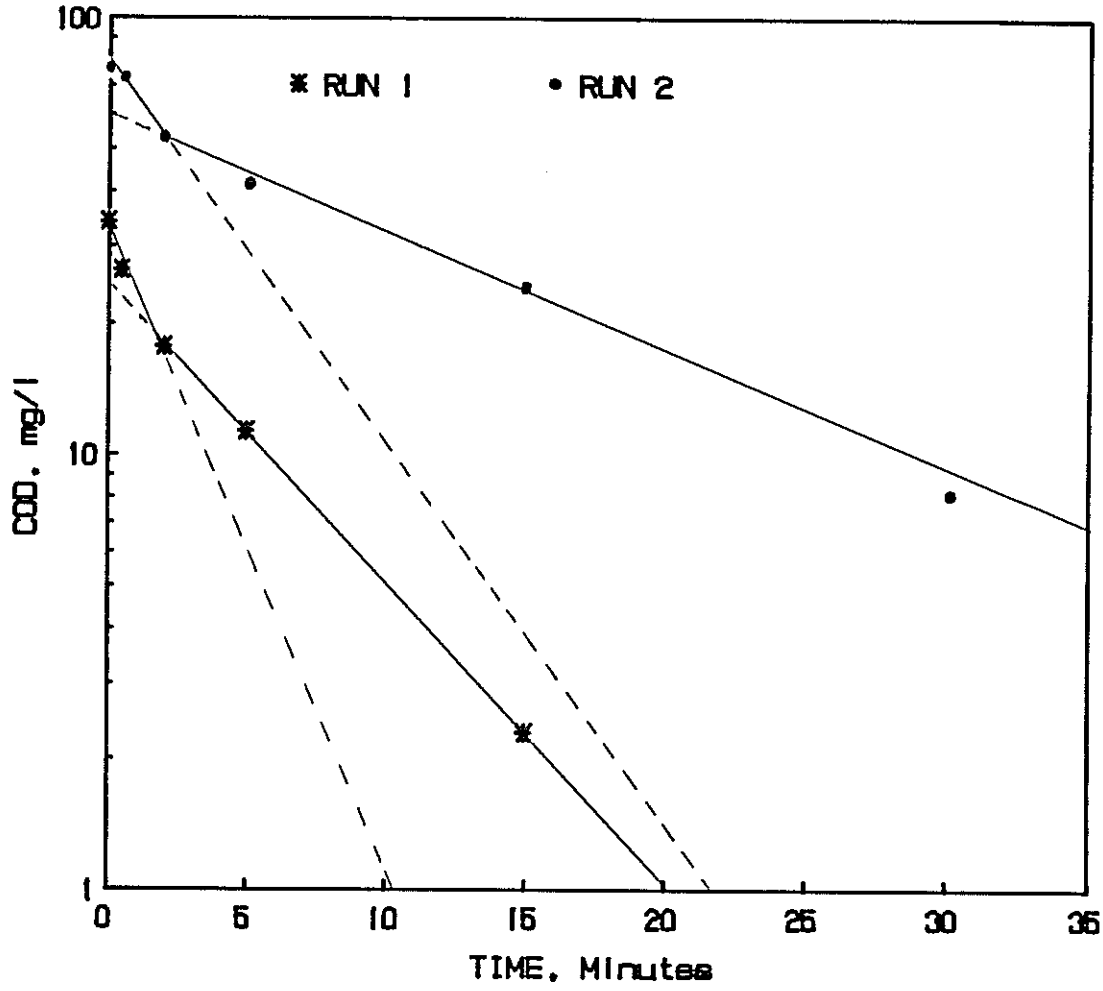


Figure 7.20. Determination of Aerobic Reaction Rates - Run 1 and 2.

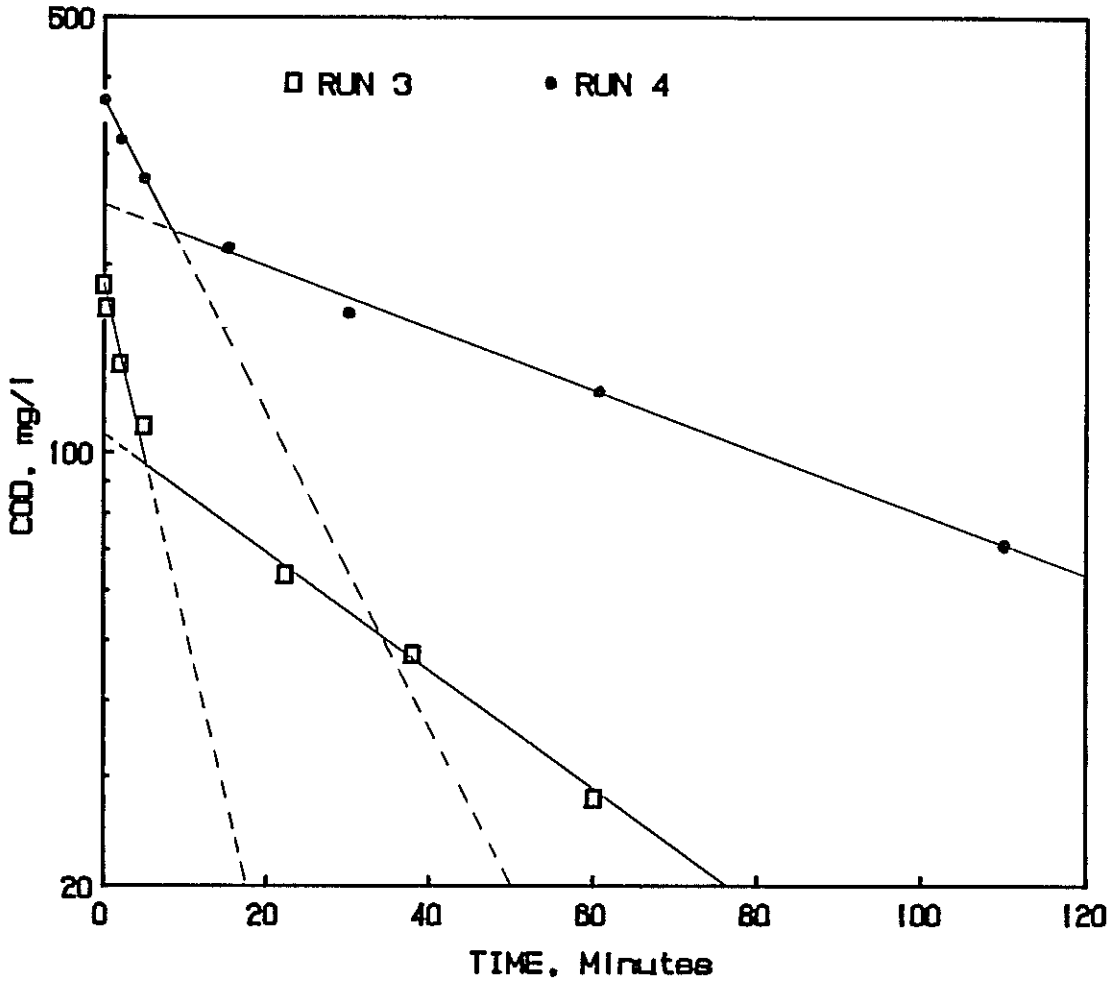


Figure 7.21. Determination of Aerobic Reaction Rates - Run 3 and 4.

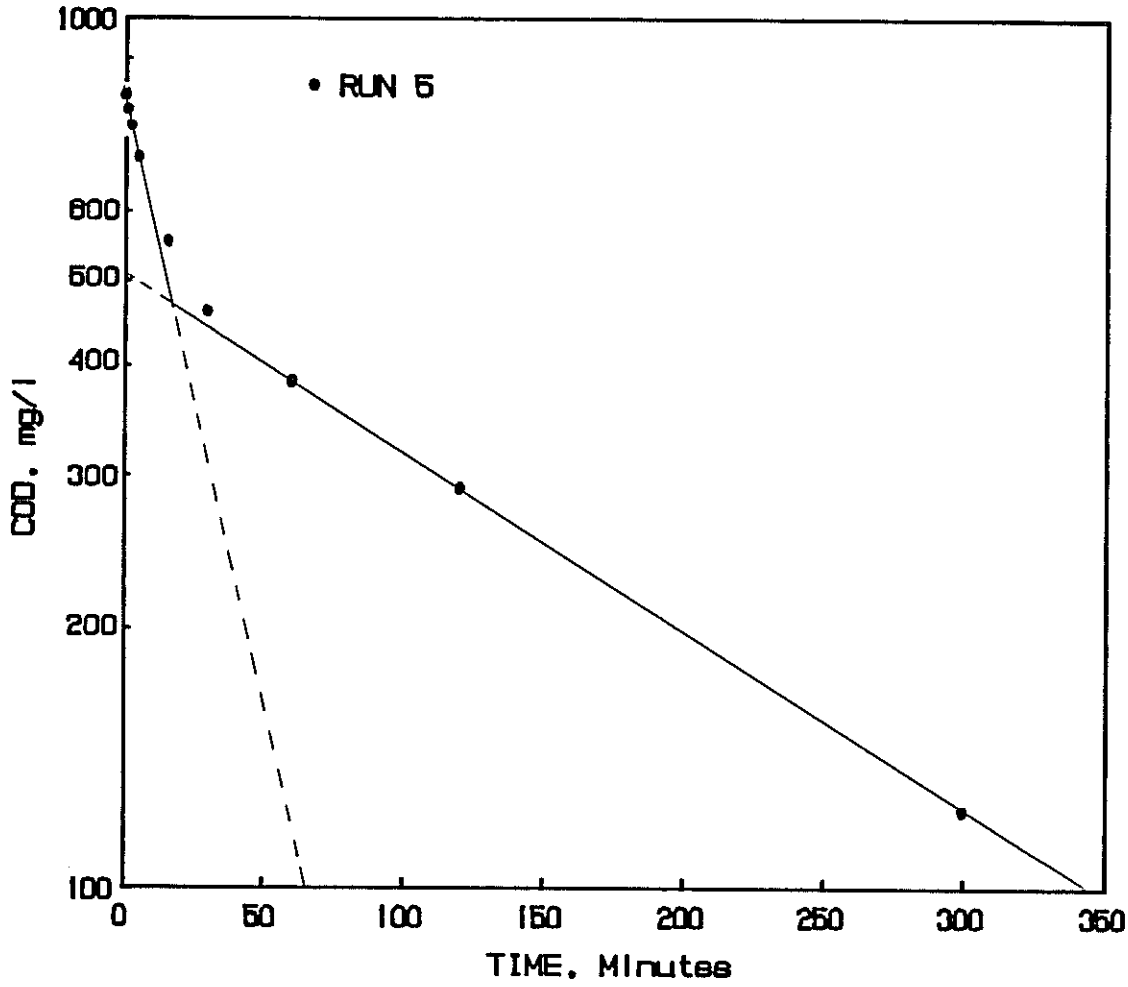


Figure 7.22. Determination of Aerobic Reaction Rates - Run 5.

TABLE 7.7

SUMMARY OF BATCH TESTS ON
SLUDGE FROM REACTOR NO. 4

SEPTEMBER 5, 1985

Parameter	Test No.				
	1	2	3	4	5
Initial COD concentration, mg/l	34	76	186	365	819
Initial MLVSS, g/l	1.23	1.16	1.24	1.22	1.25
Floc load, mg COD/g VSS	27	64	149	298	654
Biosorption, mg COD/l	10	24	77	133	308
Biosorption, mg COD/g VSS	8	21	62	109	264
Time of shift, min	1.8	2.6	4.1	7.7	15
Shift concentration, mg COD/l	18	48	98	215	470
Phase I rate constant, 1/g VSS-l-day	413	245	178	80	42
Phase II rate constant, 1/g VSS-l-day	182	74	26	12	5.4
Phase I initial reaction rate g COD/g VSS-day	14.0	18.6	33.2	29.1	34.0
Phase II initial reaction rate g COD/g VSS-day	4.3	4.1	2.8	2.7	2.8

BOD_5. In such a case, for a CSTR operating at low F/M, a virtually complete removal of BOD_5 is observed in practical applications and the residual effluent BOD_5 is usually a few mg/l.

Let us consider a typical CSTR operating at 2000 mg/l MLVSS and an influent strength of 100 mg/l BOD_5. Hydraulic retention time corresponding to a loading of F/M of 0.2 is ½ day. On the average, then, the substrate was in contact with a high (2000 mg/l) concentration of active, acclimated biomass for 12 hr. During the BOD_5 test, the concentration of biomass provided by innoculation is about 3 to 4 orders of magnitude lower (below 1 mg/l) with a contact time (5 days) only several times longer than in the reactor. Consequently, the fraction of the substrate which did not degrade in the reactor and which would subsequently degrade under a much milder conditions in BOD_5 bottle is bound to be marginal. Therefore, the observed reaction rate in the CSTR at low organic loadings is for all practical purposes equal to the rate at which the substrate is provided.

The simplest mathematical expression for the reaction rate embracing the two above discussed bounder conditions has the following form:

$$R_r = R_m(F/M)/[R_m+(F/M)] \qquad (7.6)$$

The maximum reaction rate (R_m) appears in both numerator and denominator to fulfill the condition that for loads approaching zero the reaction rate is equal to the load.

Correlation of four sets of data published in the literature on completely mixed activated sludge systems using Equation (7.2), where

the half-loading constant K_W is not equal to the maximum reaction rate R_m, provided the results presented in Table .6. It is interesting to note that the half-loading constant in all cases presented in Table 7.6 has a value slightly higher or equal to the maximum reaction rate. This supports the above arguments regarding the reaction rate at low loadings, that is that K_W should be equal (or close) to R_m. On the average, the half-loading constant (excluding data by Krishana and Gaudy) is higher by about 8 percent than the maximum reaction rate.

One possible explanation of the observed higher value of the half-velocity constant is the release of metabolic byproducts by the activated sludge. Such byproducts are usually included in the collective effluent concentration measurements used in wastewater treatment practice, such as COD or TOC. If at low loadings such byproducts are produced at a higher rate with respect to the substrate removed than at high loadings, which will result in observed deviation from the ideal relationship expressed in equation 7.6.

Most likely, however, this inequality relates to the definition of the half-loading constant in the sense of a Monod-type equation. In this sense it is simply the organic load at which the reaction rate is half of the maximum reaction rate. The fact that its value is apparently close to the maximum reaction rate is a consequence of mathematical compromise between the boundary conditions requiring equality of these values and the actual value of a true half-loading constant in a Monod sense; a compromise forced by the manner of data correlation. This notwithstanding, Formula 7.2 appears to provide a

through 7.22 at the point corresponding to the intersection of two straight segments of the substrate depletion curves. The time of shift appears to be roughly a linear function of the initial concentration (Figure 7.23) which would be consistent with the postulated zero order glucose removal rate during the first phase. At the same time, the shape of a semi-logarithmic plot resulting from a superimposition of the general, first-order reaction for the remaining feed constituents and the zero order glucose removal in the first phase should have a shape as illustrated in Figure 7.24, which is not consistent with the experimental curves.

The second potential mechanism is the biosorption capacity of the sludge--an initial, rapid substrate accumulation at a rate higher than the maximum, steady state reaction rate. The experimental data would then suggest that biosorption is a first order reaction, with the maximum biosorption capacity increasing with the initial substrate concentration (Figure 7.25). If biosorption capacity, defined as a difference between the initial substrate concentration and the intersection of the second phase line, is plotted versus the initial concentration, an apparently linear function results (Figure 7.25). While it was postulated that the biosorption (accumulation) capacity is a function of the initial concentration (Rickenberg et al., 1956), it is highly unlikely that the biosorption capacity increases linearly with the initial concentration up to, and above 250 mg/g as Figure 7.25 would indicate. The more conclusive results of the FBR tests discussed

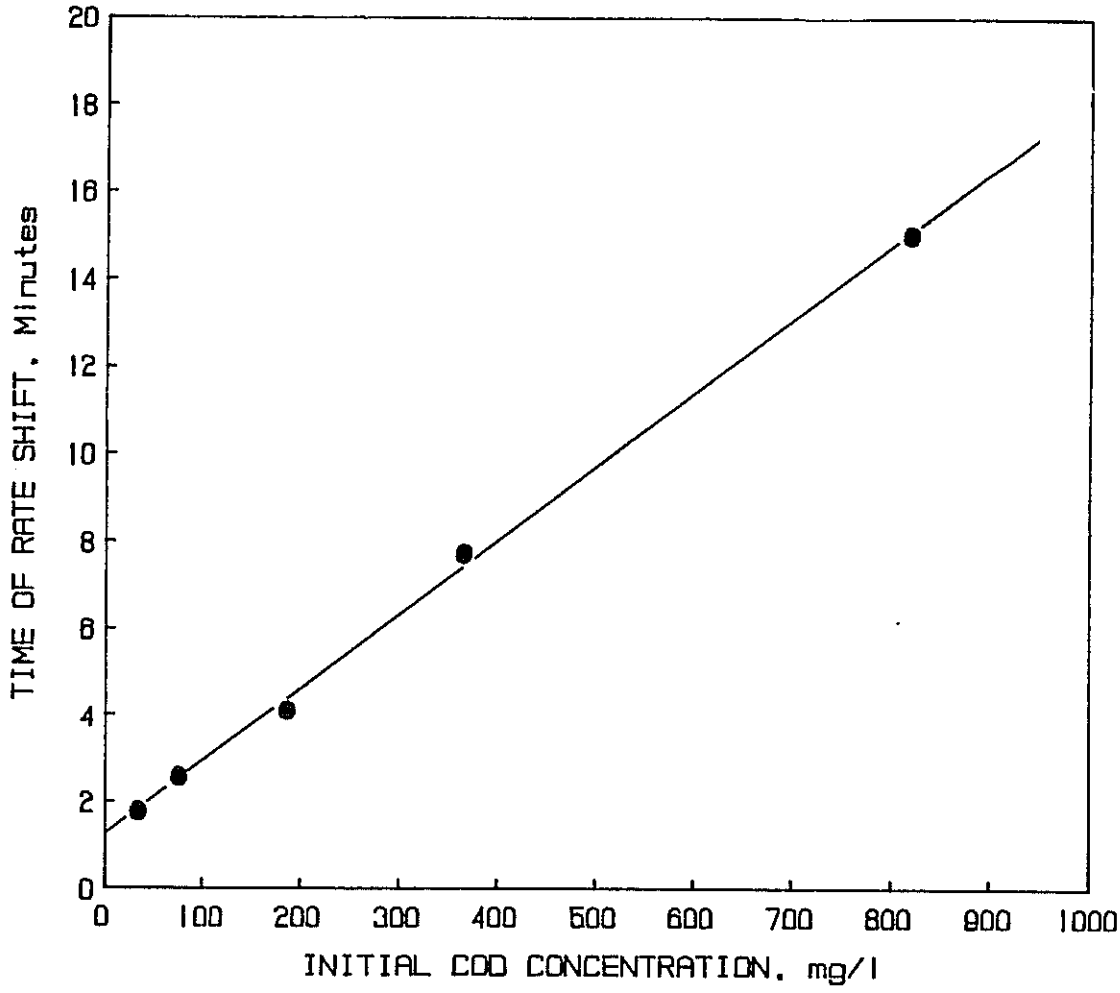


Figure 7.23. Correlation Between Time of Shift in the Reaction Rate and Initial Concentration.

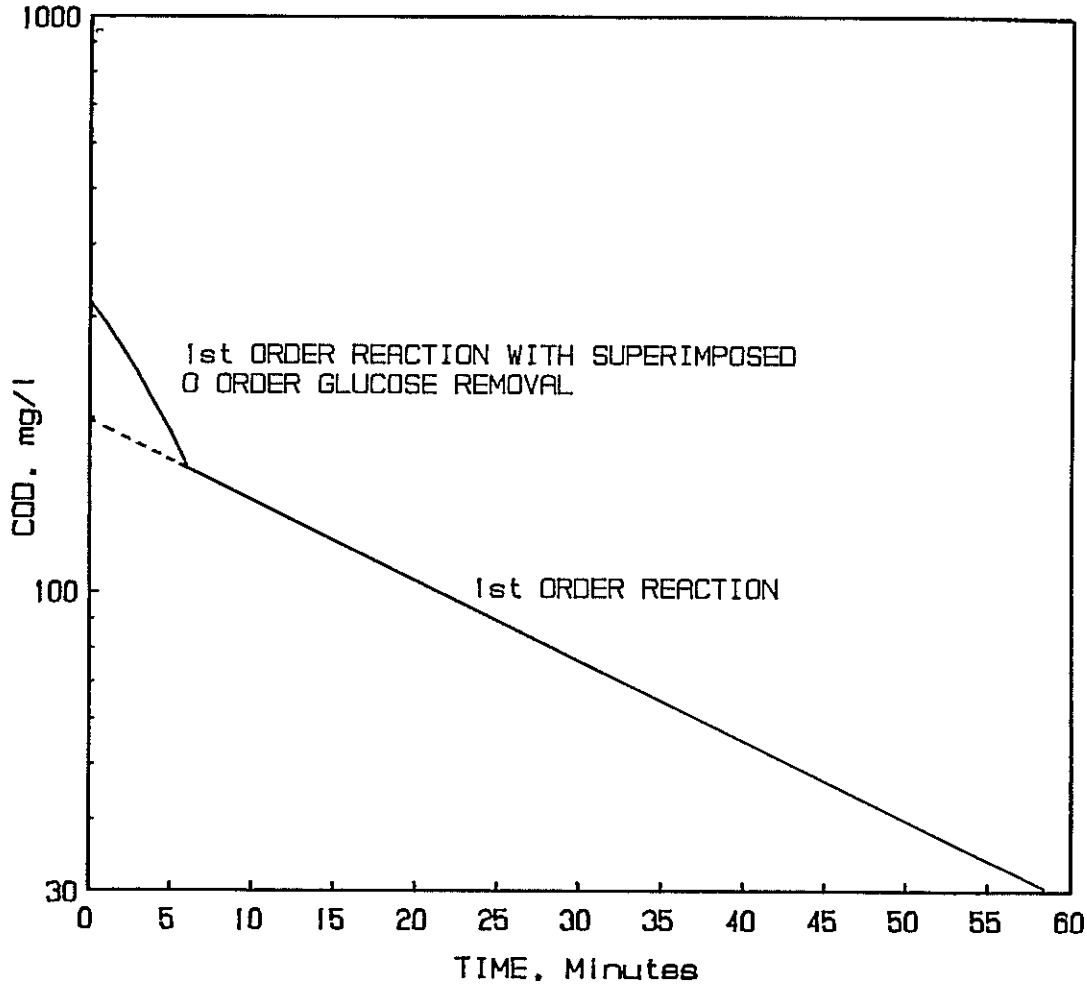


Figure 7.24. Theoretical Substrate Depletion with Superimposed 0-Order Glucose Removal Rate and First Order General Reaction Rate.

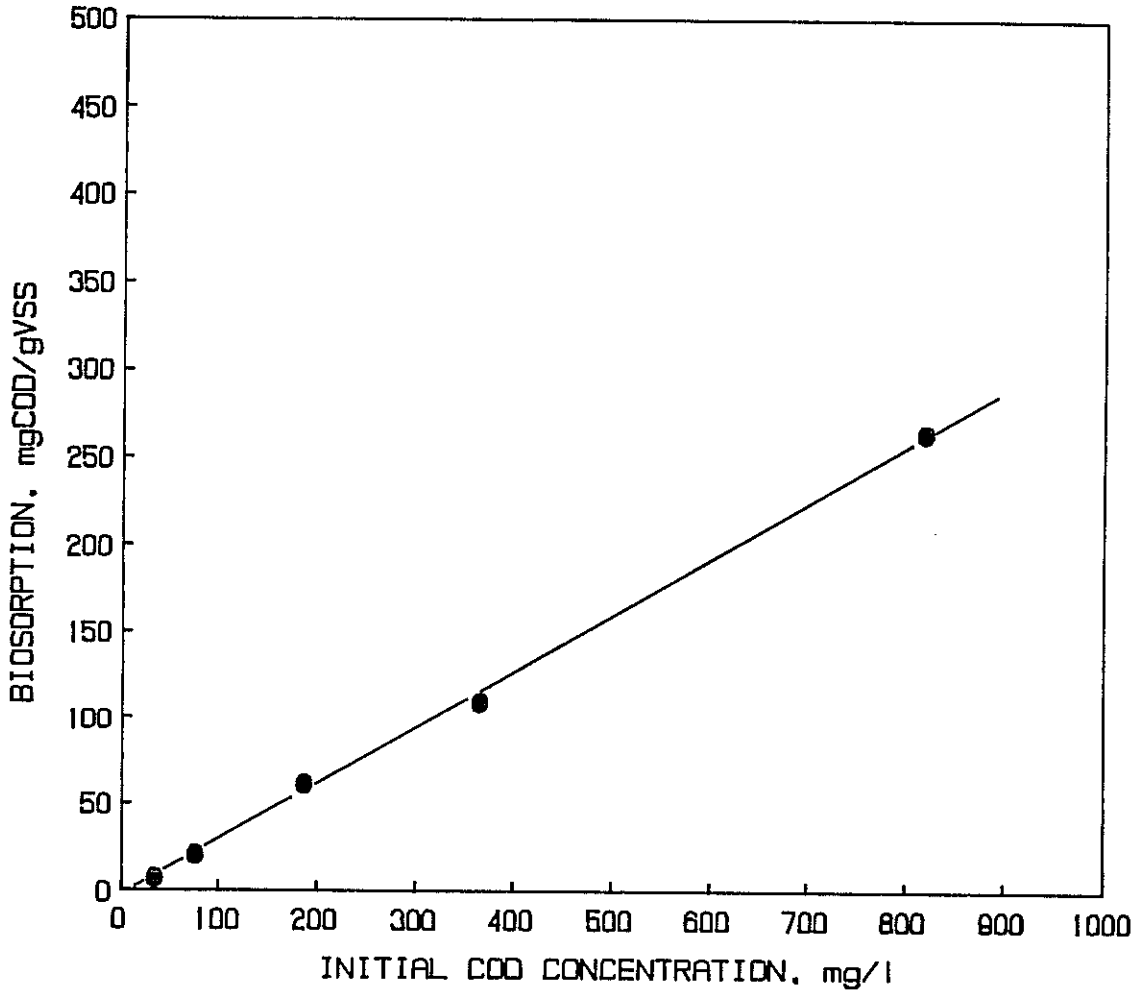


Figure 7.25. Correlation of the Biosorption Capacity with the Initial Substrate Concentration.

in the following section indicate that maximum biosorption capacity of the tested sludge-substrate systems did not exceed 150 mg/l.

In summary, it is most likely that the kinetics of the substrate removal in the initial phase of a batch test is a complex phenomenon, involving concurrent removal of substrate components with the utilization of the sludge biosorption capacity.

From the specific, first-order rate constants for both the first and second phase of the reaction, the maximum reaction rates at the initial substrate concentrations can be calculated. The results are summarized in Table 7.7.

The first phase maximum reaction rates initially increase with the initial substrate concentration, and then stabilize at about 33 gCOD/VSS-day (Figure 7.26). When these rates are correlated as a Monod-type function with the initial concentration, the resulting "absolute" maximum reaction rate is calculated to be 33 g/g-day. This calculated first phase maximum reaction rate is 9 g/g-day higher than the maximum reaction rate calculated for the proposed model of PRZ performance, as was presented in the preceding section.

Fed Batch Reactor (FBR) Tests

A series of FBR tests was performed on sludges from all seven reactors operated in the Phase III of the study. The FBR testing protocol and method of data processing are detailed in Chapter IV. The final product of the FBR tests, cumulative plots of substrate uptake, are presented in Figures 7.27 through 7.35. From these Figures the steady state reaction rate, sludge biosorption capacity and maximum

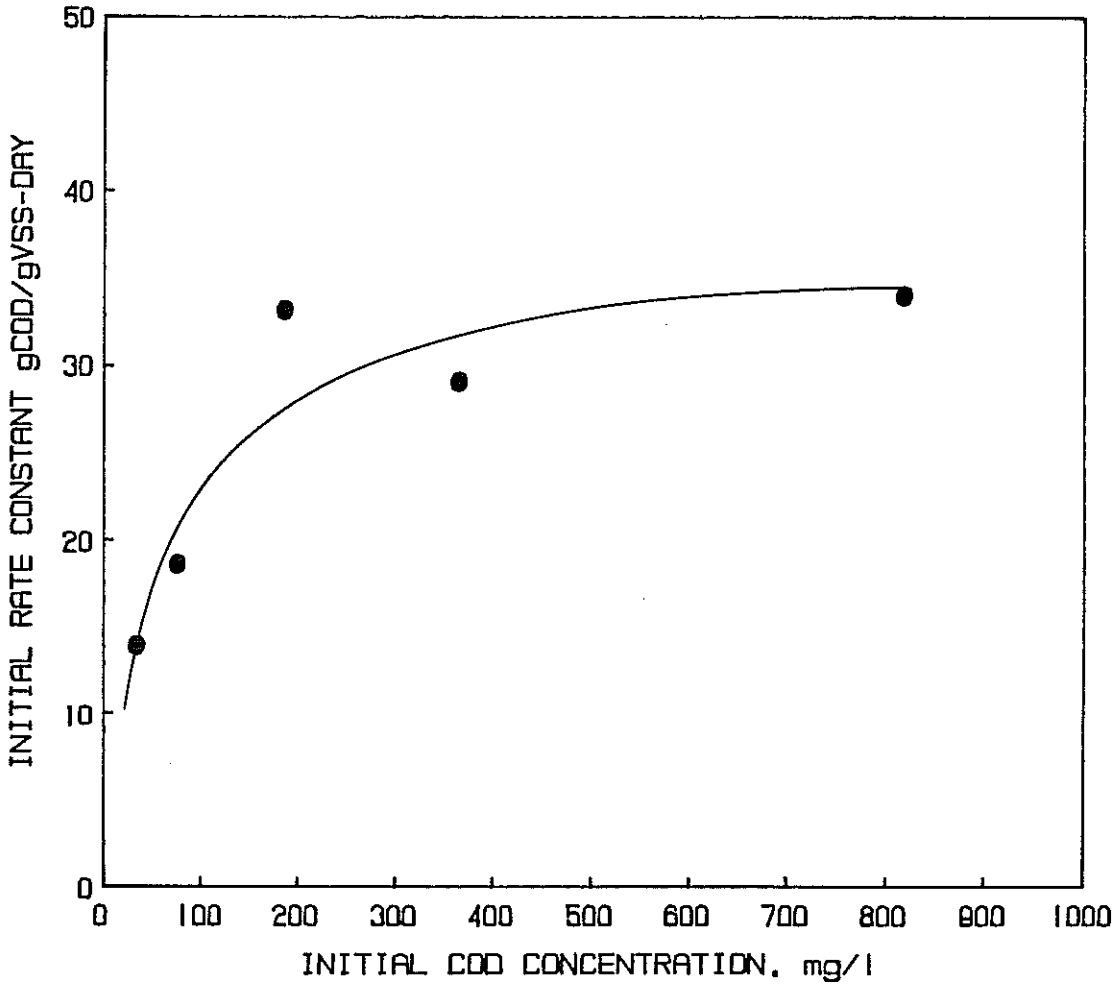


Figure 7.26. Correlation of the Batch Initial Reaction Rates with Initial Concentration.

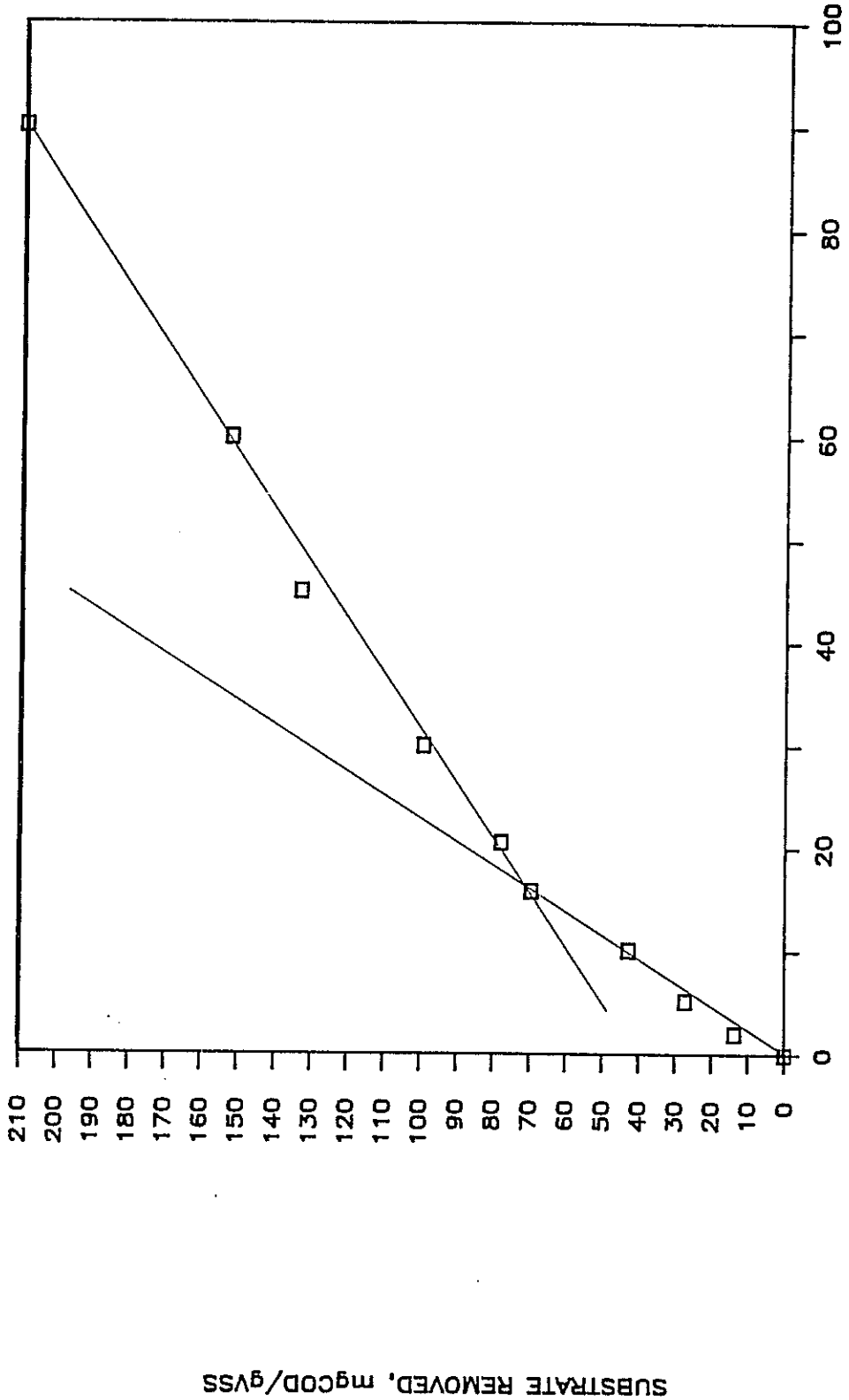


Figure 7.27. Cumulative Substrate Removal in FBR Test, Reactor No. 1.

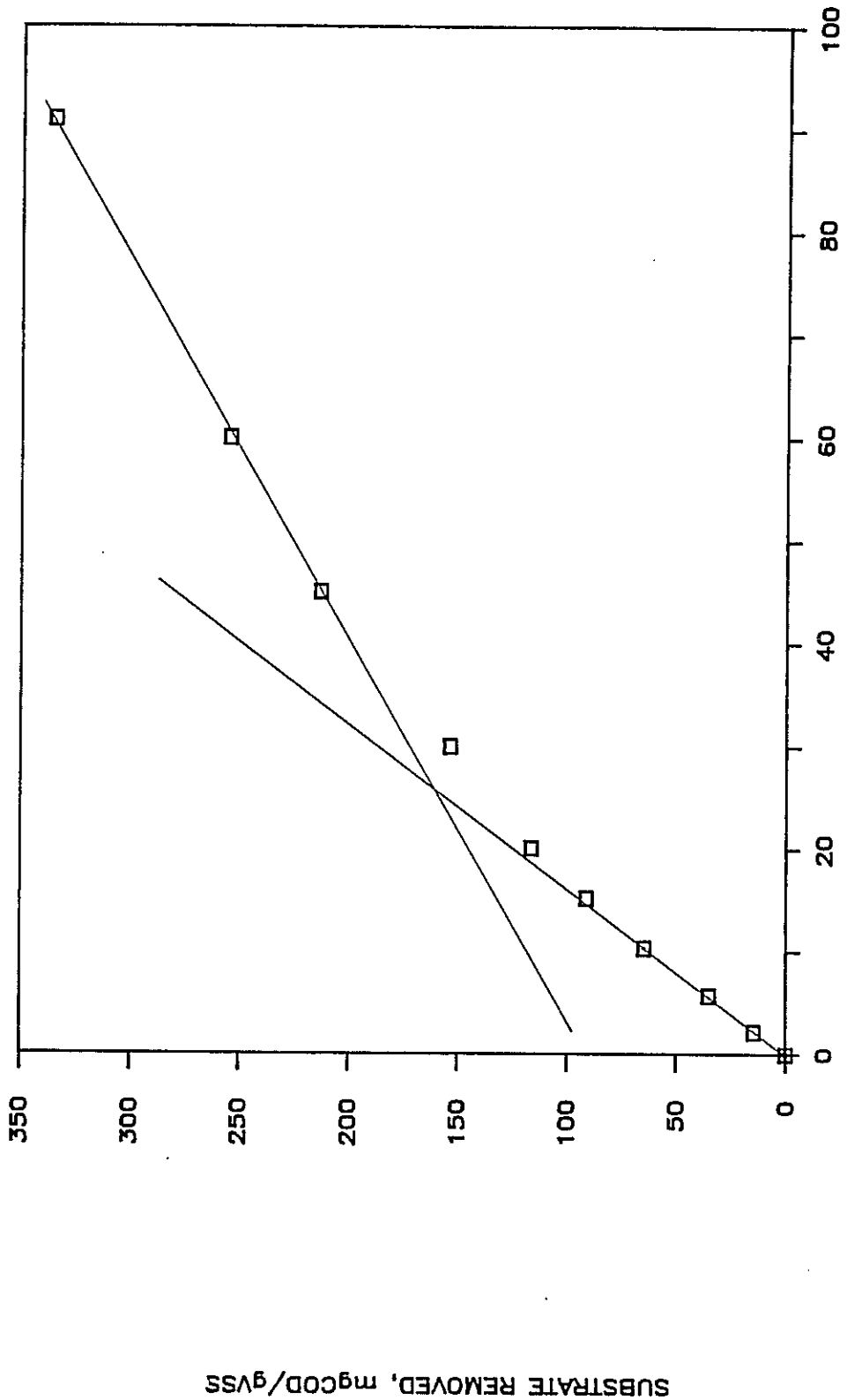


Figure 7.28. Cummulative Substrate Removal in FBR Test, Reactor No. 2.

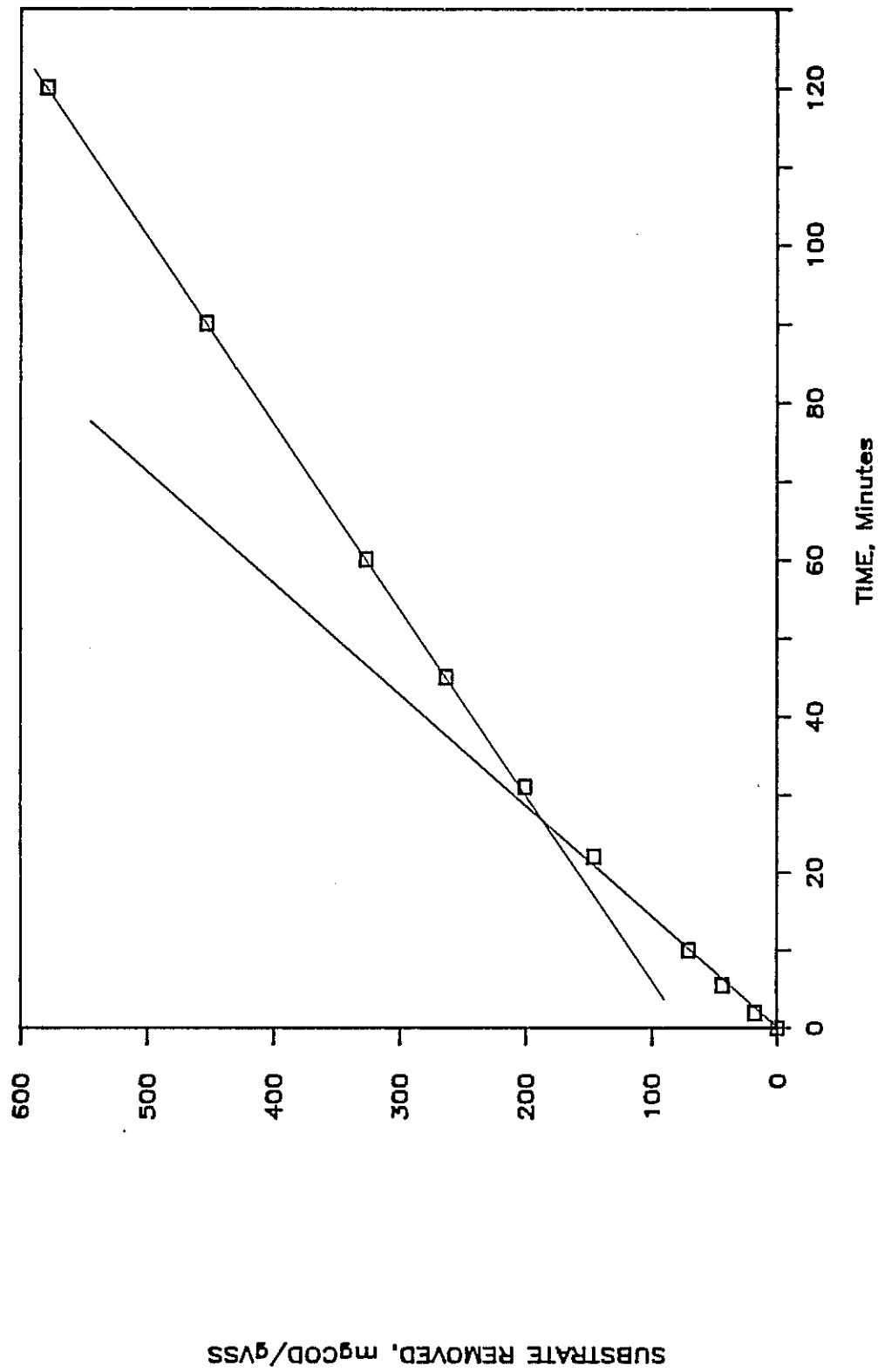


Figure 7.29. Cummulative Substrate Removal in FBR Test, Reactor No. 3.

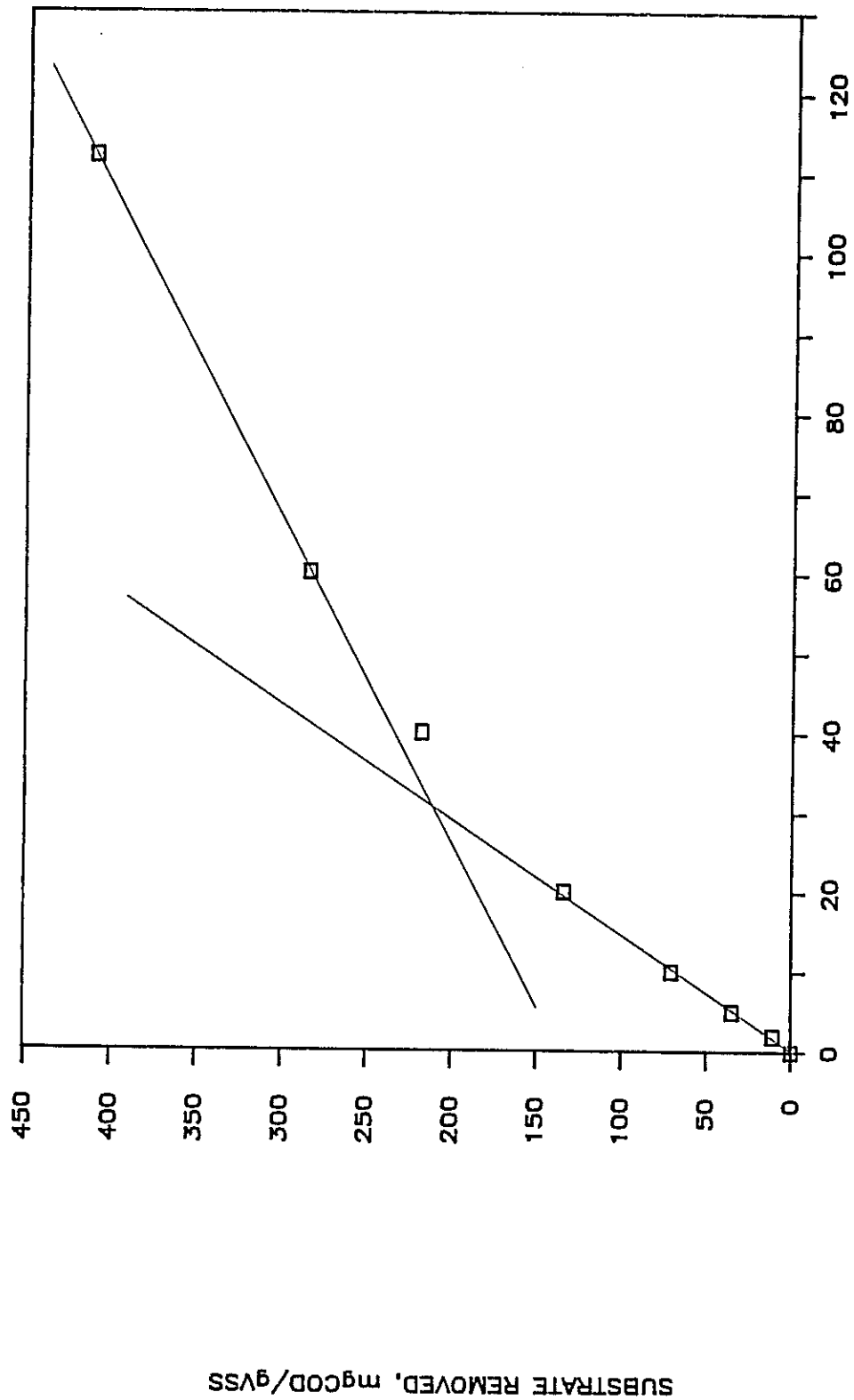


Figure 7.30. Cumulative Substrate Removal in FBR Test, Reactor No. 4.

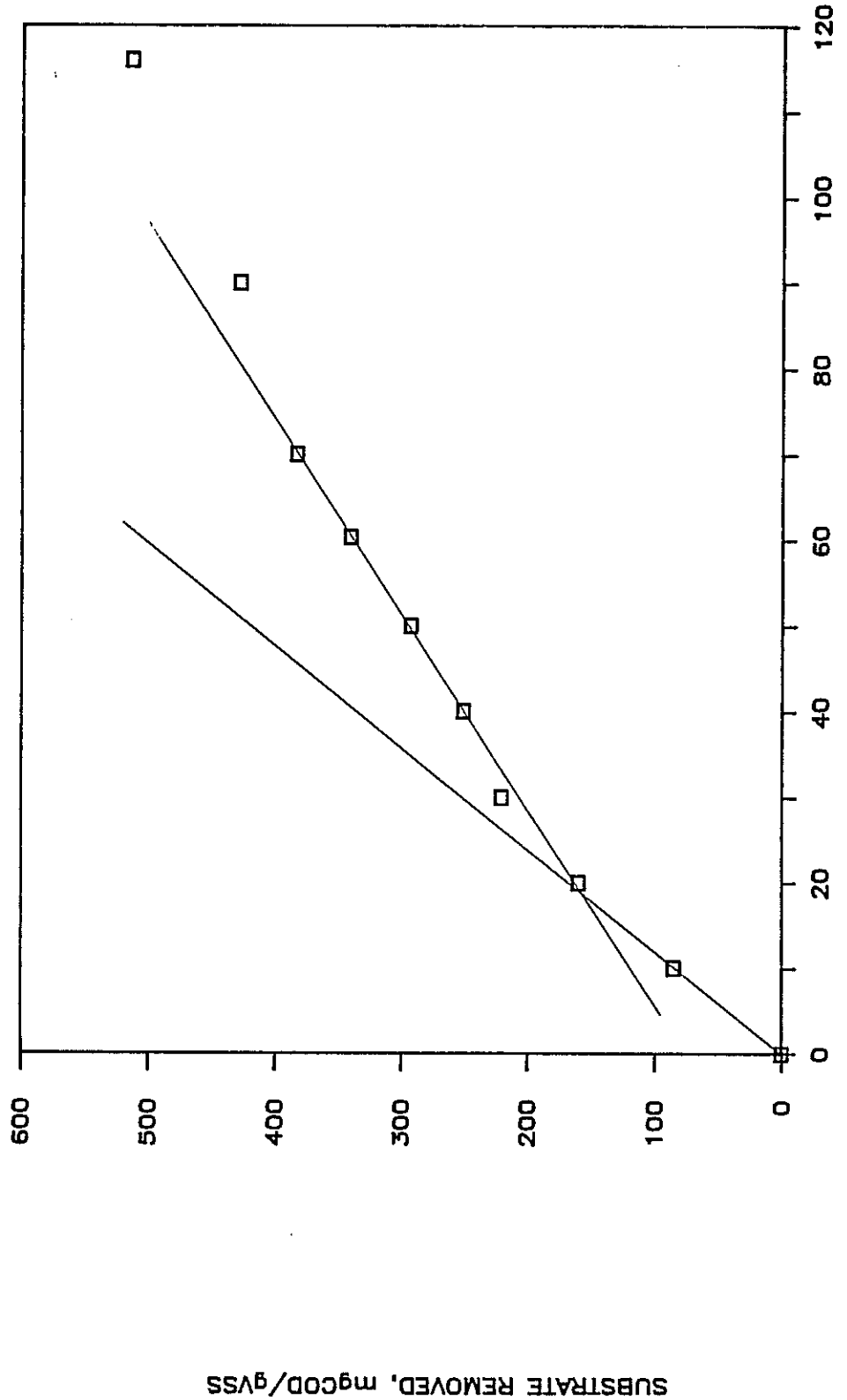


Figure 7.31. Cumulative Substrate Removal in FBR Test, Reactor No. 5, Run 1.

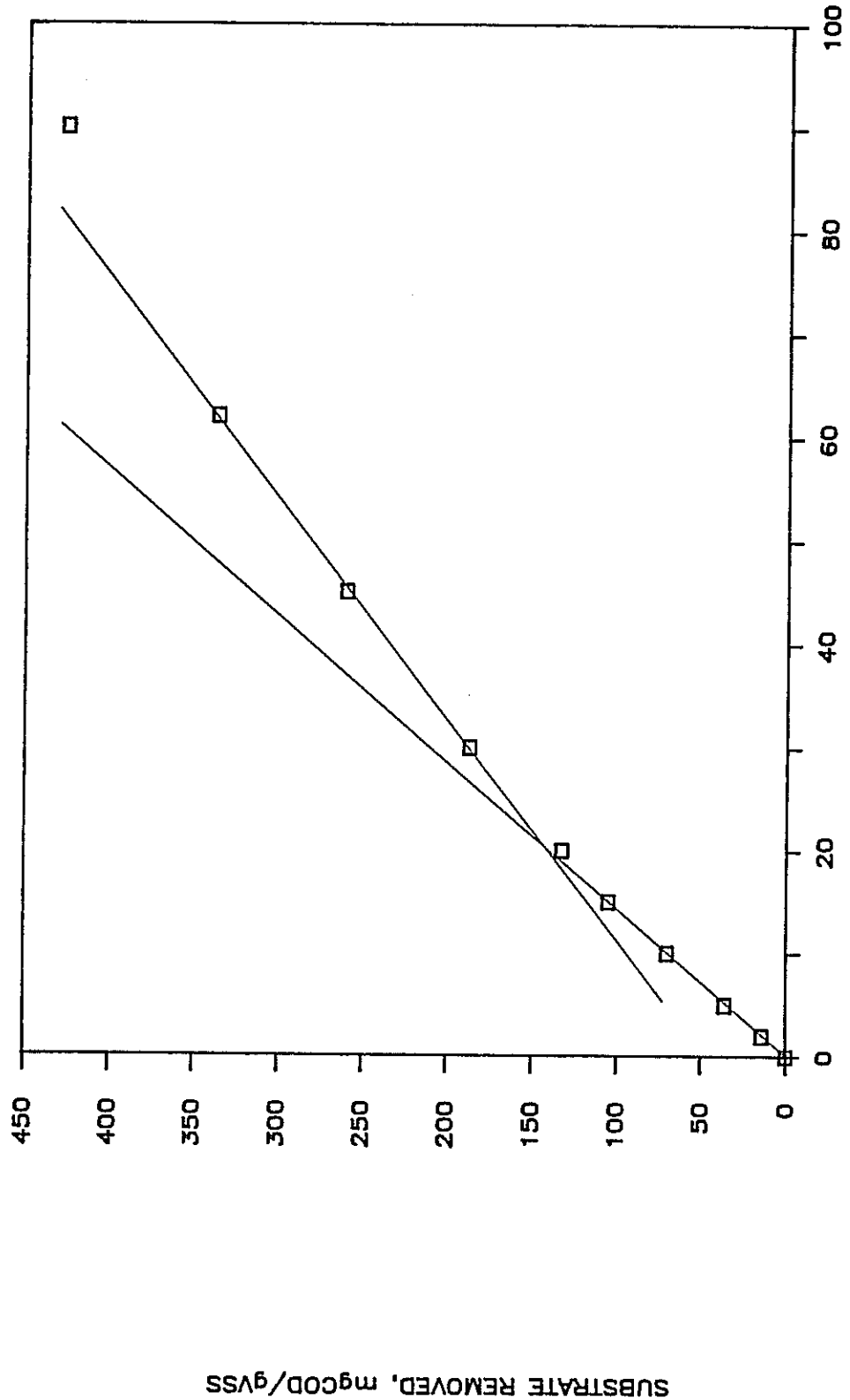


Figure 7.32. Cumulative Substrate Removal in FBR Test, Reactor No. 5, Run 2.

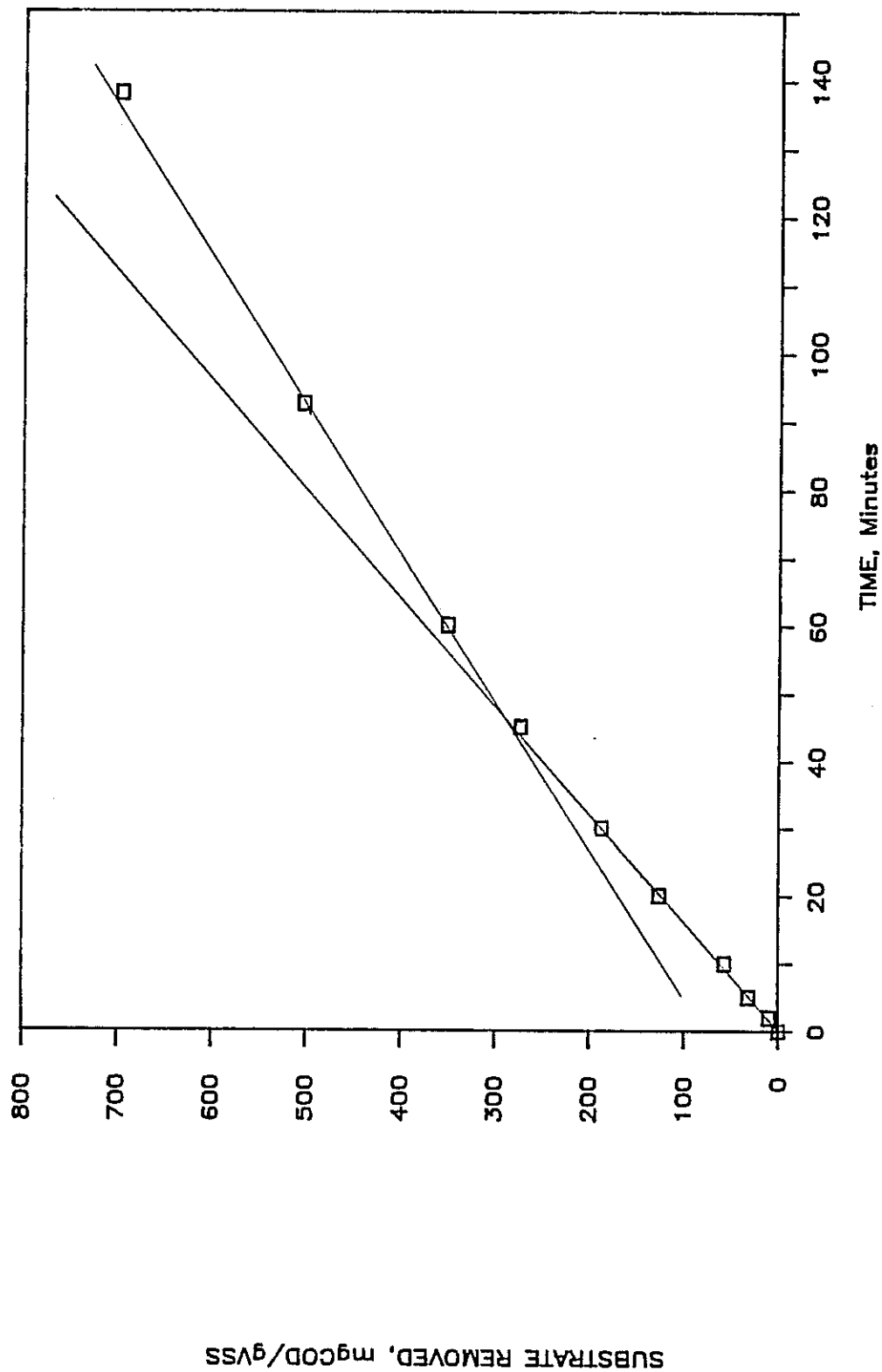


Figure 7.33. Cumulative Substrate Removal in FBR Test, Reactor No. 6.

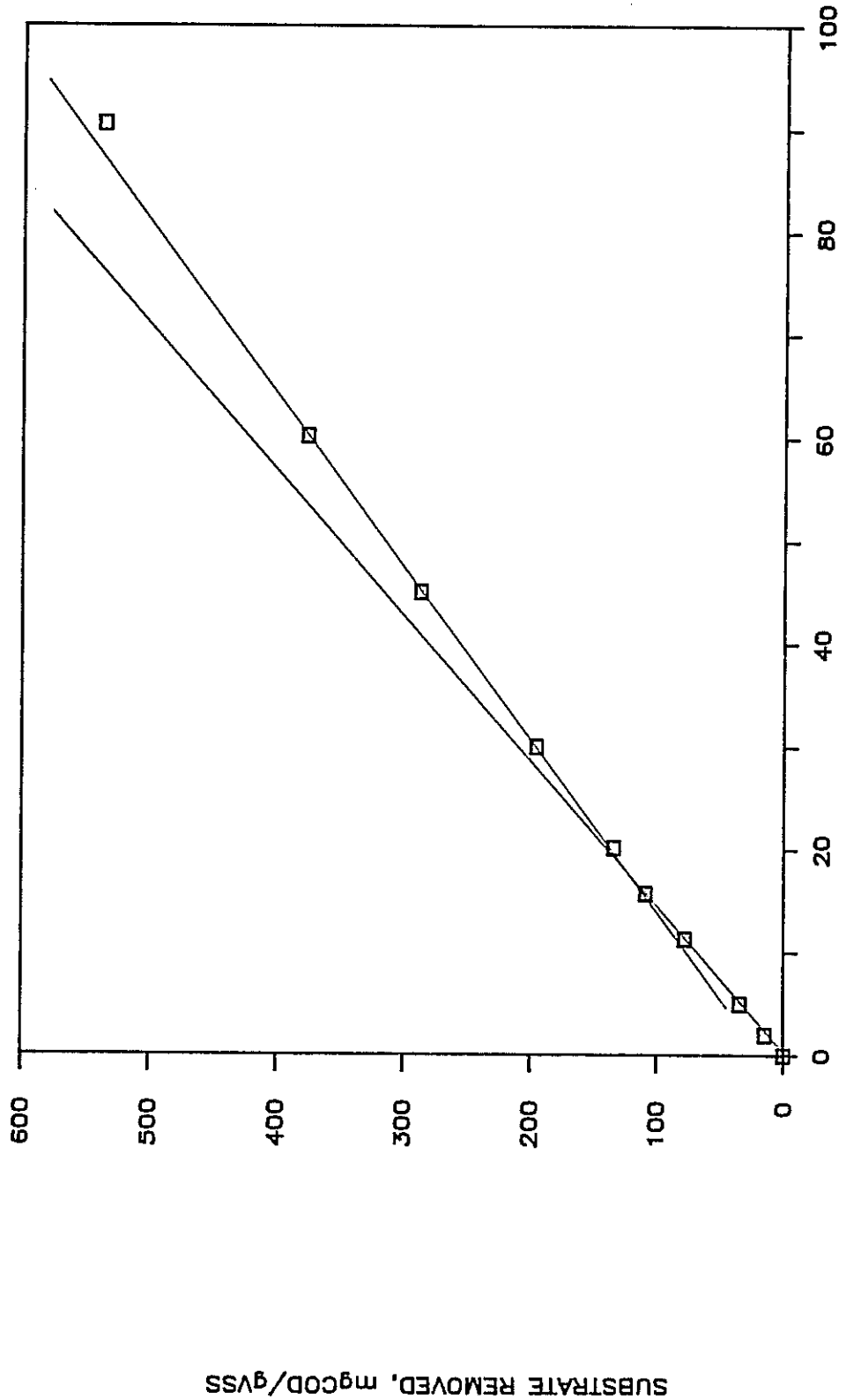


Figure 7.34. Cumulative Substrate Removal in FBR Test, Reactor No. 7.

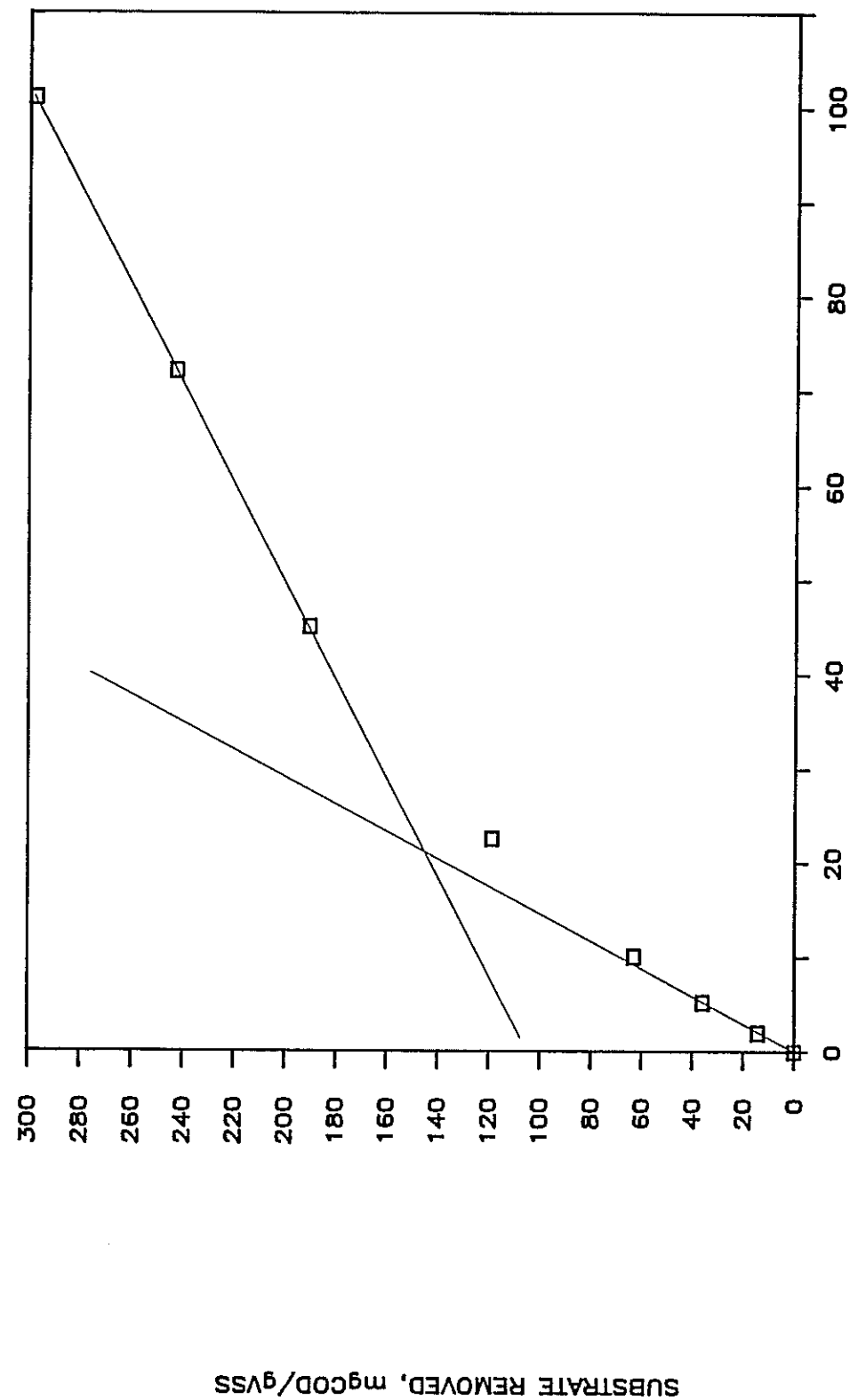


Figure 7.35. Cumulative Substrate Removal in FBR Test, Reactor No. 4, Glucose Feed Only.

(initial) reaction rate can be directly determined, as was previously shown on Figure 4.8.

The intersection of the steady state line with the ordinate axes represents the amount of substrate removed from the solution in the initial phase of the test in excess of what would have been expected from the steady state rate. The FBR test allows then, a direct measurement of the biosorption capacity of the sludge as defined earlier in this work. The values of the biosorption capacity obtained from the FBR tests are summarized in Table 7.8.

The reaction rates recorded during the FBR tests demonstrate a characteristic pattern. After an initial period of high substrate uptake rate the reaction proceeds at a relatively steady state rate until the end of the test (1 to 2 hr). The initial phase of the reaction, lasting 5 to 15 minutes into the test, proceeds at a fairly constant rate and for comparative purposes was calculated as an average slope between 0 and 10 minutes. The resulting values range from 7 to 12 g COD/g-day and are summarized in Table 7.8. Considering that results of the standard batch tests and correlation of PRZ performance demonstrated that the reaction rates were a function of the organic loading (or floc load) at which the system operated, it was felt appropriate to apply a similar approach to the interpretation of FBR tests results. Therefore, the initial reaction rate should be a function of system loading in the form given in Formula 7.2. In order to calculate the two constants in Formula 7.2, at least two FBR test results for each reactor are needed. Such data were not available.

TABLE 7.8

SUMMARY OF FED BATCH REACTOR TEST RESULTS[a]

Parameter	Reactor No.								
	1	2	3	4	5	5	6	7	4 (glucose only)
Date	9/24/85	9/24/85	10/20/85	11/1/86	9/20/85	9/24/85	10/20/85	9/24/85	11/2/85
Steady State Reactor Rate (R_s), gCOD/gVSS-day	2.69	3.90	6.05	3.51	6.47	6.81	6.50	8.73	2.74
Initial Reaction Rate (R_r), gCOD/gVSS-day	7.2	8.9	10.3	9.8	12.2	10.1	9.4	10.2	10.0
Biosorption Capacity, mg COD/gVSS	40	90	75	135	70	50	75	15	110
Applied, initial F/M, gCOD/gVSS-day	16.9	14.9	20.5	17.1	29.5	15.6	17.4	15.5	15.9
$R_m = \frac{R_r(F/M)}{(F/M) - R_r}$ g COD/gVSS-day	12.5	22.1	20.7	23.0	20.8	26.4	20.4	29.8	26.9

[a] In all tests, except as noted, diluted feed concentrate was used as the substrate.

However, as was previously discussed, the half-loading constant, K_w, is expected to have a value very close to that of the maximum reaction rate, R_m. Assuming that these two constants are equal, the maximum initial reaction rate can be approximated from a single FBR result. The R_m values calculated in this manner are given in Table 7.8. It is interesting to note that for all the reactors operating with a PRZ, R_m is in a narrow range from 20 to 25 g/g-day. The control reactor's sludge (without PRZ) yielded a much lower R_m at 12.5 g/g-day. The highest R_m was obtained for batch reactor No. 7 at 30 g/g-day.

Perhaps the most interesting observation from this data is that R_m for reactor No. 4 (23.0 g/g-day) is close to the R_m obtained from the batch tests (33 g/g-day), and almost equal to the R_m obtained from correlation of PRZ data (R_m = 23.4 g/g-day).

The data presented indicate that R_m is a parameter characteristic for a given sludge-substrate system, as can be expected from theoretical considerations. The value of R_m is not simply a standard, maximum, steady state reaction rate as classically determined from the first order correlation of batch test data. The steady state reaction rates with excess substrate available were obtained in this study from FBR tests and were substantially lower than the initial (maximum) reaction rates obtained in the same tests.

The initial, maximum reaction rate is an experimental quantity which encompasses a steady state reaction rate with superimposed initial substrate removal capacity, most likely in a form of storage. It was shown to be a Monod-type function of the organic loading (or

concentration) in the system, either in a standard batch test, or in a continuous contactor. A similiar value for the initial reaction rate was obtained from the FBR test.

Kinetics of Substrate Removal Under Anaerobic Conditions

During the course of the study a number of batch tests were performed under anaerobic and anoxic conditions. The objective of these experiments was to gain some insight into the effect of anaerobic (anoxic) conditions on the ability of the sludge to remove soluble substrate from solution. Such information is desirable in order to better understand the mechanisms of biosorption under aerobic conditions and to assess the behavior of the reactor during air-off cycle.

As a preparatory step, an experiment was designed to assess the ability of the biomass to remove the substrate by physical adsorption on the bacterial surface. Metabolic activity of the sludge was impaired in this test by Hg^{+2} added in a form of a $HgCl_2$ solution. Figure 7.36 presents the results obtained with two different Hg^{+2} concentrations. The small, decreases in COD between time 0 and 0.5 minutes are probably artifacts arising from the procedure of calculating the initial COD concentration from the concentration of the feed (100 ml at 419 mgCOD/l) and sludge sample (300 ml at 15 mgCOD/l). (Subsequently, the batch test procedure was modified to limit this type of error, as was discussed in Chapter IV.)

In any case, no appreciable substrate uptake was noted, indicating that physical adsorption was insignificant if not entirely absent.

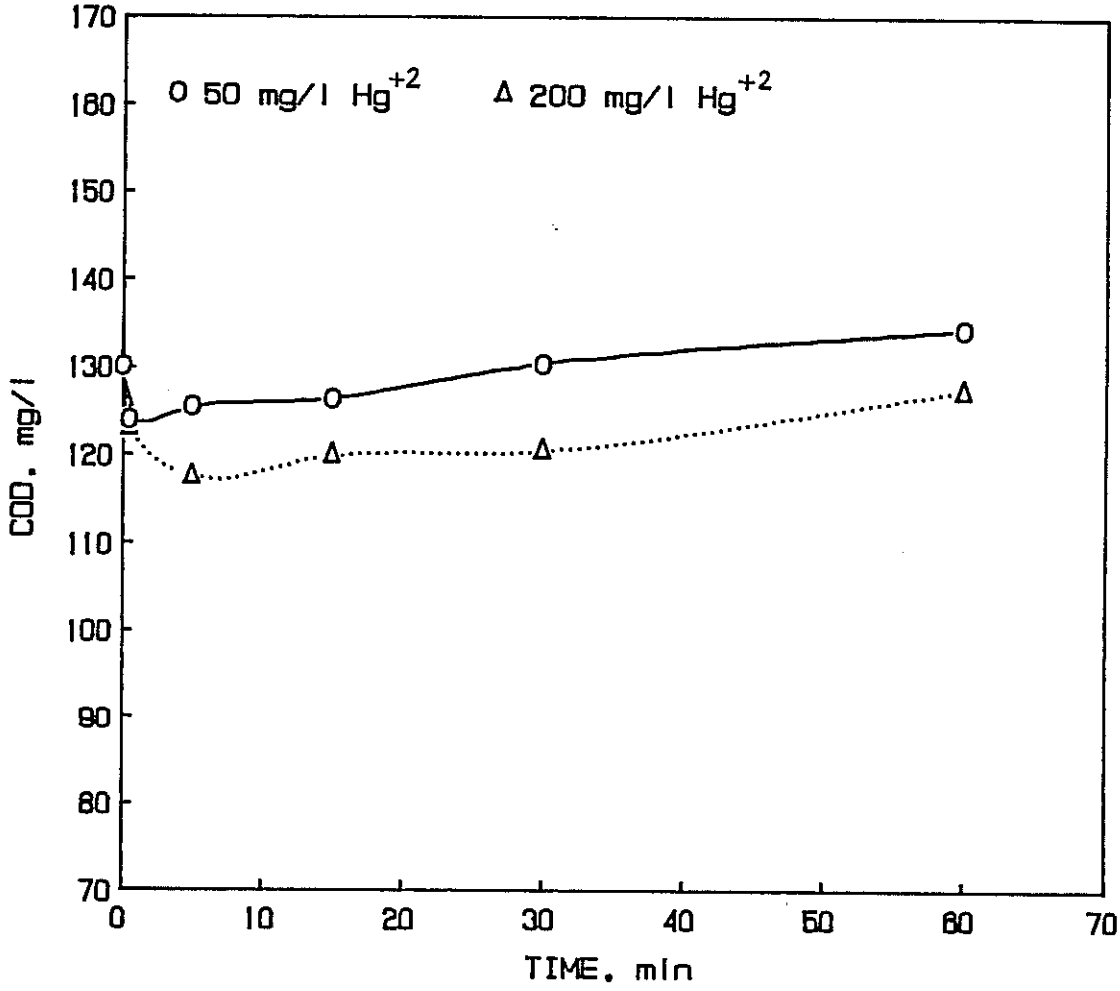


Figure 7.36. Results of Batch Tests with Hg^{+2} Poisoned Sludge.

Moreover, the COD concentrations tended to slightly increase with time, reflecting release of lysis products of the desactivated bacteria. The data are too limited to comment on the apparently higher COD concentrations in the sample dosed with lower Hg^{+2} concentration.

Subsequently, an experiment was performed to compare the release of the organic matter from the sludge in the absence of substrate under aerobic and anaerobic conditions. A sample of activated sludge was withdrawn from the reactor, rinsed twice with distilled water to remove background COD, split into two aliquots and then mixed with air or nitrogen in the batch reactor described in the Methodology Section. During the initial 30 minutes, a small, (few mg/l) increase of COD was observed in both samples (Figure 7.37), probably reflecting adjustment of the osmotic pressure by the bacteria as proposed by Parkins and McCarty (1981). During the following 2.5 hr only a very small (about 1 mg/l-hr) increase in COD was noted in the aerated sample. In the sample mixed with N_2, the COD concentration substantially increased between 30 and 60 minutes into the test, and then stabilized, reaching a total of about 20 mg/l of COD increase from the test start. It is difficult to assess if this increase is a result of lysis of the organisms, or if it originates from metabolic byproducts of the organisms fighting for survival under the stressed conditions. In any case, a substantial amount of organic matter is being released from the biomass under anaerobic conditions, while under aerobic conditions the release was small.

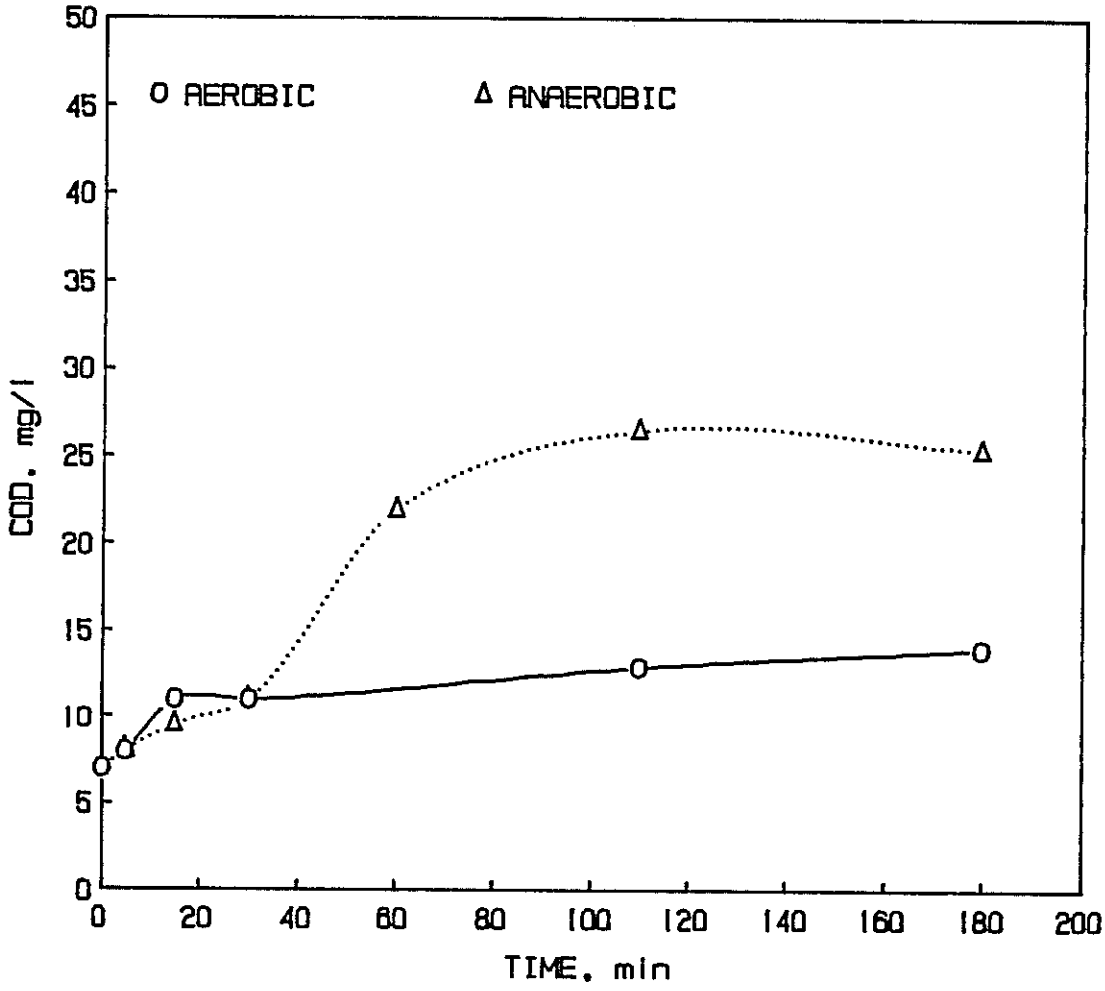


Figure 7.37. Release of COD by Activated Sludge without Substrate Addition.

An initial experiment for determination of the extent of substrate uptake under anaerobic (or anoxic) conditions was performed with an unrinsed sludge sample from reactor No. 1. A 625 ml aliquot of sludge containing 20 mg/l of NO_3-N was combined with 250 ml of the reactor's feed in the anaerobic, batch set-up. The results of COD and NO_3-N measurements are presented in Figure 7.38. The COD concentration decreased by 25 mg/l in the first 5 minutes of the test and then continued to decline at a relatively steady rate for 1 hr. At 1 to 1.5 hr into the test the measured COD started to increase. The NO_3-N concentration was decreasing continuously at a slightly retarded zero order rate (Figure 7.38). The onset of the increase in COD concentration coincided with the depletion of NO_3-N. When the average rates of COD and NO_3-N depletion between 15 and 60 minutes are compared (Figure 7.38) their ratio of 3.8 is only slightly higher than the one expected from the stichiometry of the denitrification reaction (3.6).

Following these initial experiments a series of anaerobic tests with different initial substrate concentrations was performed. In order to eliminate anoxic COD uptake, the sludge samples used in these experiments were concentrated and rinsed with distilled water. The nitrate concentration in the feed was negligible, as was confirmed in the previous anoxic test by a good agreement between the calculated initial NO_3-N concentration and the one measured 0.5 minutes after the addition of feed to the sludge (Figure 7.38). In the calculation of the initial concentrations shown on Figure 7.38 it was assumed that no nitrates were present in the feed. Additionally, spot tests performed

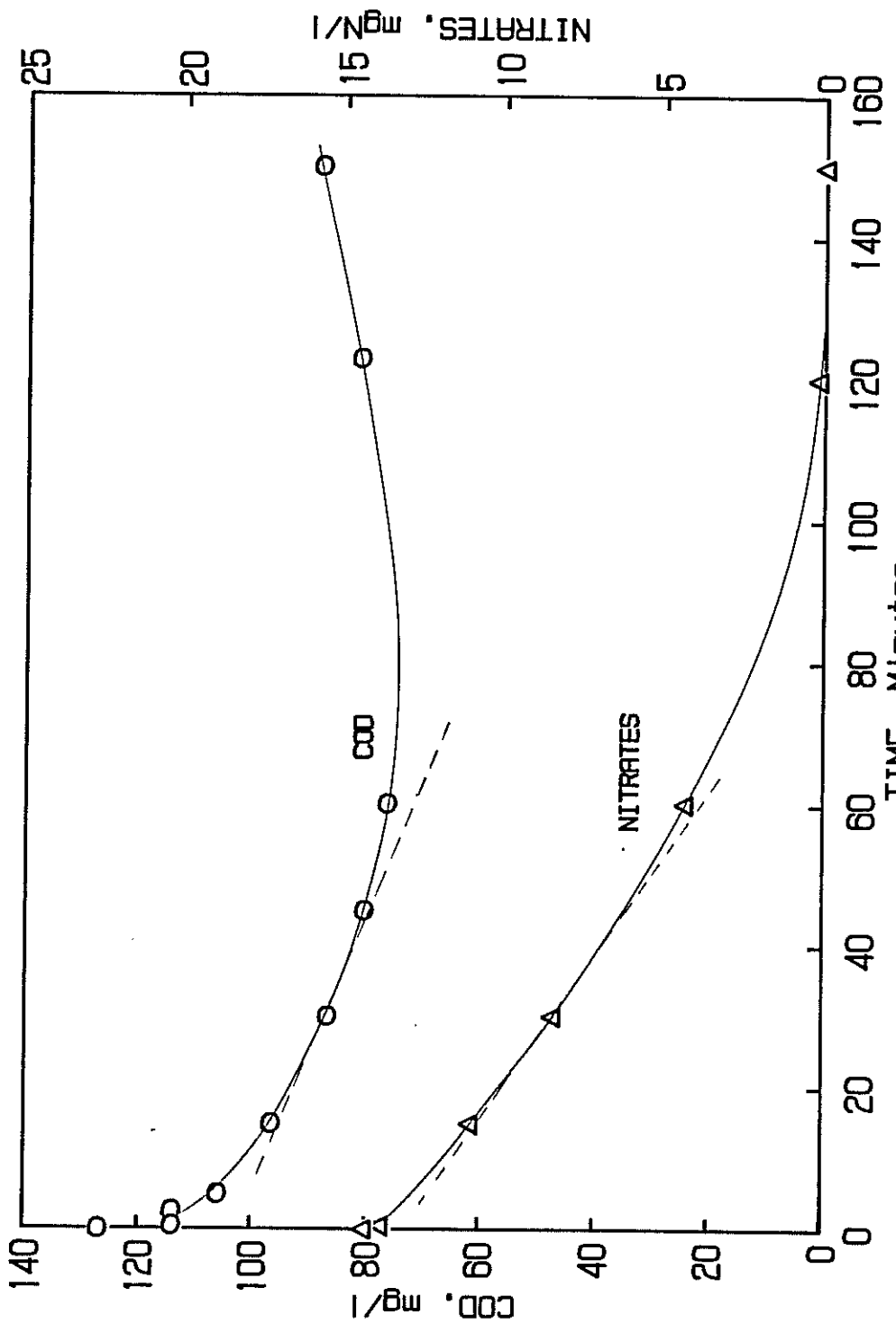


Figure 7.38. Results of Batch Test Under Anoxic Conditions.

during the anaerobic tests discussed below showed that only traces of NO_3 were present in the solution 0.5 minutes after the feed addition.

The results of the anaerobic tests were presented previously in Figures 7.15 through 7.19 (together with the results from the parallel aerobic tests). In all the anaerobic runs, COD was initially decreasing at a high rate, then abruptly stabilized and subsequently started to increase. The minimum recorded COD concentration, the approximate rate of COD removal and other parameters of the anaerobic runs are summarized in Table 7.9. It is interesting to note that for lower floc loads the COD uptake was proportional to the initial substrate concentration. Only at the highest initial concentration tested, was a deviation from this trend observed (Table 7.9).

These results indicate that some form of internal storage without external energy supply is taking place under anaerobic conditions, with the sludge capable of accumulating at least 100 mgCOD/gVSS under these conditions. At the same time, a significant amount of COD is released from the sludge, which eventually leads to an increase in the COD concentration (with respect to the minimum COD measured).

The initial anaerobic substrate removal rates given in Table 7.9 follow a trend similar to the aerobic reaction rates (Figure 7.26). The maximum initial anaerobic substrate removal rate has a value of 12 g/g-day; one-half of the rate determined for the aerobic conditions. Incidentally, all rate constants determined for the tests performed at 385 mg/l initial COD concentration are apparent outliers (aerobic Phase I and II - Figure 7.25, and aerobic - Table 7.9). Considering

TABLE 7.9

SUMMARY OF BATCH ANAEROBIC TESTS

Parameter	Test No. 1	2	3	4	5
Initial COD, mg/l	31	77	180	365	755
Minimum COD, mg/l	24	56	130	275	620
Uptake, mg/l[a] Substrate	7	21	50	90	135
Percent of Substrate Uptake[b]	30	37	38	33	22
Initial MLVSS, g/l	1.23	1.16	1.24	1.22	1.25
Substrate uptake, mg COD/g VSS	6	18	40	74	108
Approximate Reaction Rate, g COD/gVSS-day	2.6	7.3	10.6	6.9	11.7

[a]Difference between initial and minimum COD recorded during the test.

[b]In respect to the initial COD concentration.

that the aerobic and anaerobic tests were performed in parallel, with the sludge originating from a split sludge sample previously dewatered and rinsed it would seem that this particular sludge aliquot might have been inhibited or contaminated during handling.

The subsequent tests were designed to separate the two parallel mechanisms of organic transfer through the cell membrane: substrate uptake and release of metabolic byproducts and/or lysed material. For that purpose glucose was used as a sole substrate, and its concentration measured by a specific analytical method. An aerobic control test was performed in parallel. The sludge for these experiments was prepared as usual, i.e. rinsed with distilled water, so no nitrates were present. The results are presented in Figures 7.39 and 7.40 where the glucose concentration is expressed in COD equivalents. From Figures 7.39 and 7.40 it can be calculated that the glucose uptake rate is about three times higher under aerobic than anaerobic conditions. Glucose removal proceeds as an approximately zero order reaction under both conditions, with retardation effect observed below 40 mg COD/l residual glucose concentration.

As expected, under anaerobic conditions much more organic material was released by the biomass than in the aerobic tests. Comparison of Figure 7.37 with Figure 7.40 reveal that under aerobic conditions almost the same concentration of COD was released with and without food addition. Under anaerobic conditions a much higher release was observed in the presence of the substrate (compare Figure 7.37 with 7.39). This substantiates the conclusion that at least part of the

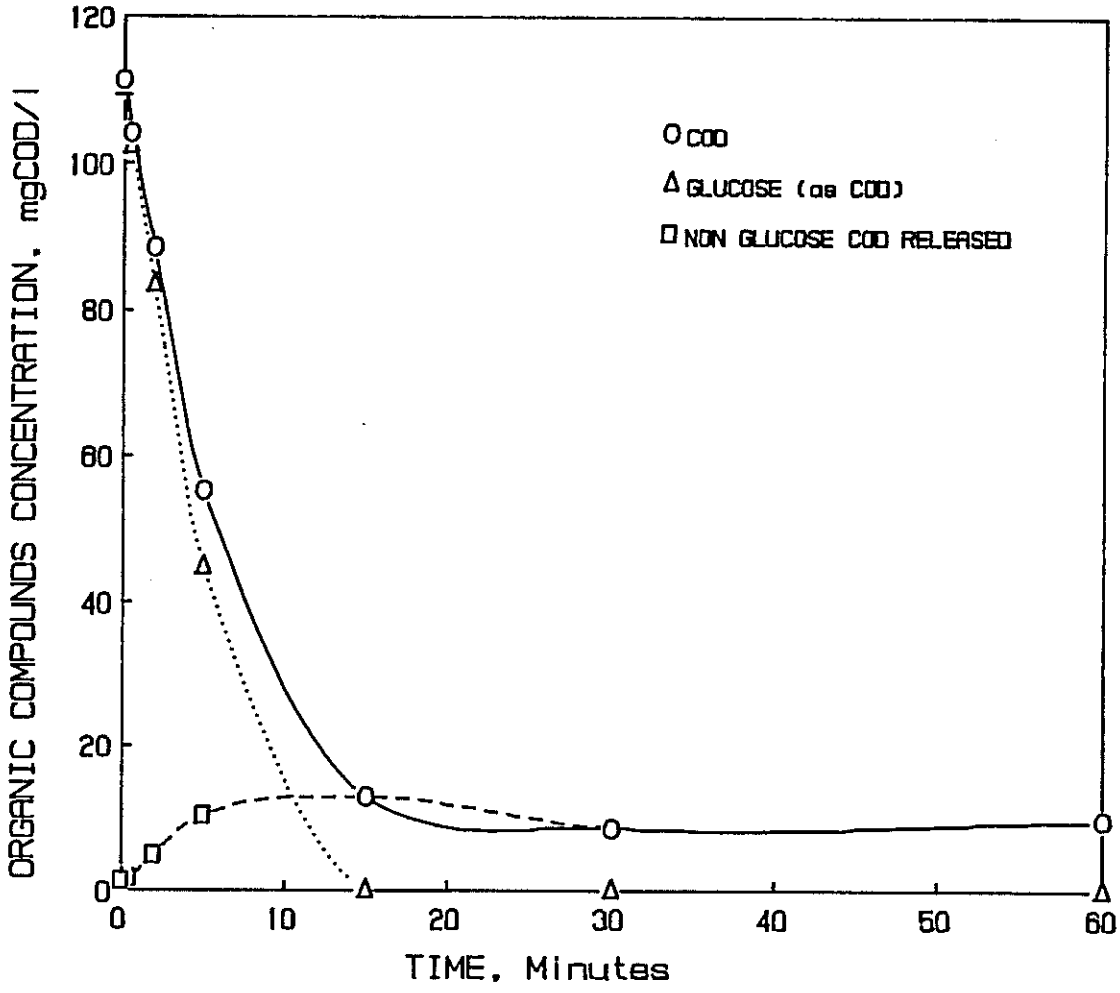


Figure 7.39. Results of Batch Test with Glucose Under Aerobic Conditions.

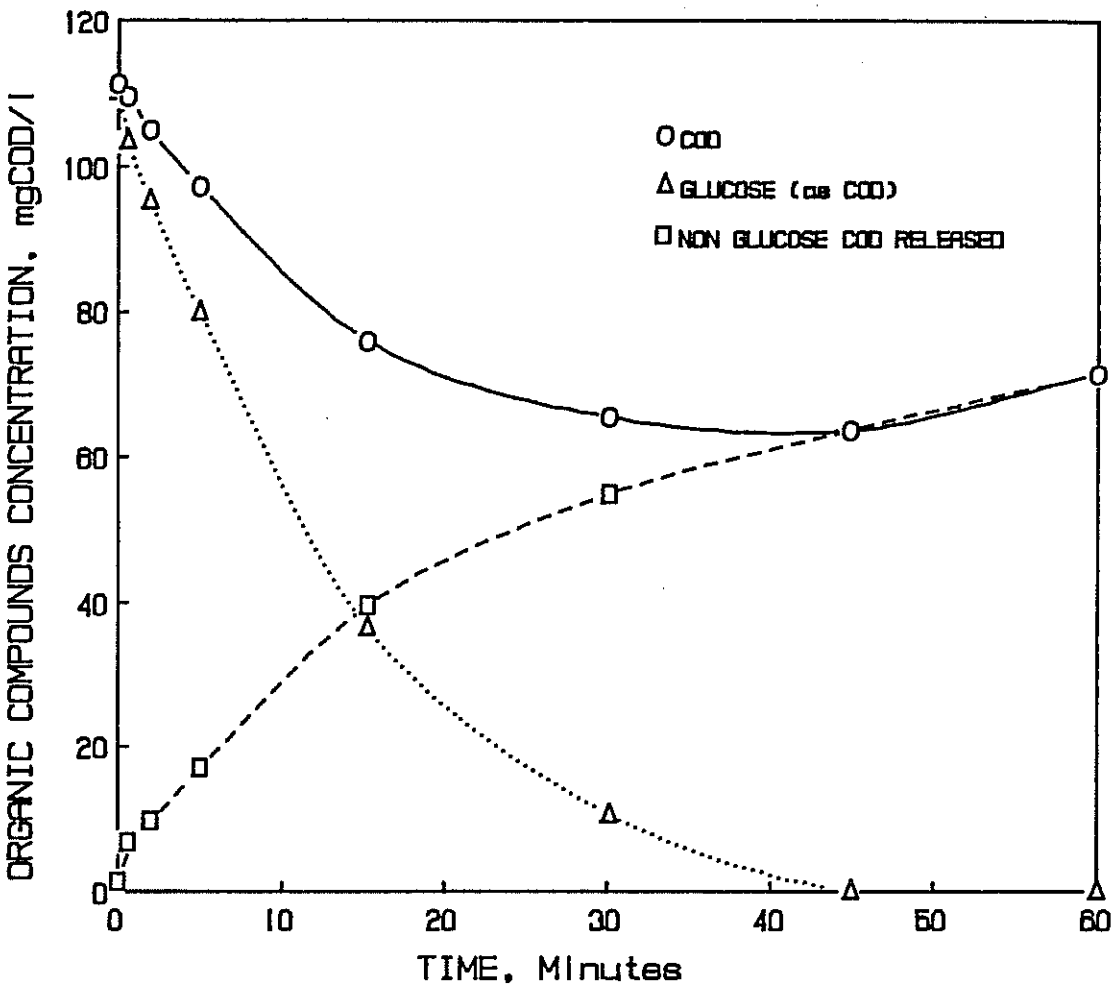


Figure 7.40. Results of Batch Test with Glucose Under Aerobic Conditions.

organic matter released under anaerobic conditions is byproducts arising from the expenditure of internal energy sources for storage or accumulation of the substrate.

CHAPTER VIII

CONCLUSIONS

The following conclusions were drawn from the study on intermittently aerated, continuous flow reactors with PRZs treating a readily biodegradable, soluble substrate:

1. Sludge settling charcteristics from the intermittently aerated reactors operated during the study were influenced by the reactors' configuration. The continuous-flow, control reactor without a PRZ experienced early and severe bulking problems. The presence of the continuously aerated PRZ delayed onset of bulking; and periodical releases of this condition were observed in these reactors. The continuous flow reactor with a PRZ operating in an intermittent (aeration and sludge recycle) mode produced a more stable sludge. The sludge settling characteristic, however in this reactor deteriorated toward the end of the study. Only the batch-fed reactor consistently produced a sludge with excellent settling characteristics. The reactors' resistance to bulking was shown to increase with an increase in the average concentration at which the substrate was removed from the liquid phase in the system.

2. The reaction rate in the PRZ was shown to be a simple function of the PRZ organic loading. The semi-empirical correlation derived from the experimental data has a form similar to the Monod-type equation with a maximum reaction rate and half-velocity loading being the functional parameters. The formula allows prediction of the performance of any PRZ for the tested sludge-substrate system.
3. The maximum reaction rate appearing in the equation for reaction rate in the PRZ was shown to be close to the true initial maximum reaction rate for a given sludge-substrate system, as independently determined from the batch and FBR tests.
4. Based on the formula for prediction of the reaction rate in the PRZ, and on the postulated relationship between the PRZ performance and resistance to bulking, a model for optimization of selector design was developed. A resulting nomogram allows the determination of the optimum sludge recycle rate for any set of the system's parameters lumped into a single system constant. The optimum sludge recycle rate results in a substrate concentration in the PRZ equal to one-half of the influent concentration. The optimum recycle rate is always less than 100 percent and approaches this value for small values of the system constant.

5. For the sludge-substrate system studied, the reaction rate in batch tests proceeded in two distinct phases. A high, initial reaction rate was attributed to both parallel substrate removal and biosorption.
6. Fed-batch reactor tests demonstrated that sludges grown in the continuous reactors were capable of removing substrate up to 135 mgCOD/gVSS above the uptake expected from a steady state reaction rate. The phenomenon, referred to as biosorption, was completed in 10 to 15 minutes after the test start.
7. Activated sludge grown in an intermittently aerated system (aerobic-anoxic) was capable of removing up to 300 mgCOD/gVSS of substrate in batch tests under anaerobic conditions.
8. The COD released from the biomass under anaerobic conditions (with and without substrate present) was several times greater than in the parallel aerobic tests.
9. The substrate concentration in the intermittently aerated reactors studied did not change significantly during a cycle. "Bound oxygen" available from denitrification and/or the sludge capability to uptake substrate without an external electron acceptor prevented accumulation of soluble substrate during the air-off phase.

10. SOUR showed significant and rapid changes during the cycle attributable to nitrification. In the reactor without a PRZ, SOUR during the initial part of the aeration cycle was about four times higher than in the remainder of the aeration phase.
11. The stirred ZSV were found to be inversely proportional to the total length of the filamentous bacteria per unit mass of activated sludge.

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