A NEW CONCEPT OF UPFLOW CLARIFICATION

IN

ACTIVATED SLUDGE SEPARATION

A Thesis Presented To
The Faculty of Graduate Studies
The University of Manitoba

In Partial Fulfillment
of the Requirement for the Degree
Master of Science
in Civil Engineering

by

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ABSTRACT

Continuous flow laboratory tests were conducted over a 135 day period to evaluate the efficiencies of a new upflow clarifier prototype in separating sludges with different settling characteristics. The continuous flux upflow clarifier (CFLUC) separated the different sludges produced, by an activated sludge system treating a synthetic dairy waste at four different F:M and/or SRT values. All sludges produced had different physical and biochemical characteristics. Other operating control parameters were selected to simulate actual operating conditions for a full-scale wastewater treatment facility at Souris, Manitoba.

It was concluded from the study that the CFLUC system could effectively separate sludge types produced over a wide range of organic loadings. At low F:M ratios of 0.10 to 0.24 g COD/g MLVSS.d the upflow clarifier removed 70% of the pinpoint floc that remained after quiescent batch settling tests in a 1000 mL cylinder. The sludges at these low organic loading rates were characterized as normal zoogleal sludges with considerable pinpoint floc. The good settling properties of these sludges were typical of extended aeration sludges. At high F:M ratios of 0.41 to 0.83 $days^{-1}$, the upflow clarifier was capable of separating filamentous bulking sludges. The sludges could be characterized as severe bulking sludges with very poor settling characteristics. The effluent quality (in terms of suspended solids) was similar to the results obtained under quiescent batch settling tests.

The study results indicate that the solids separation efficiencies of the CFLUC system exceed previous treatment efficiencies achieved in conventional clarifiers.

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TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

GLOSSARY

F:M	Food to microorganism ratio, $time^{-1}$
HRT	Hydraulic retention time, time
SRT	Solids retention time, time
SESS	Secondary effluent suspended solids, mass/volume
MLSS	Mixed liquor suspended solids, mass/volume
MLVSS (X)	Mixed liquor volatile suspended solids, mass/volume
SS	Suspended solids, mass/volume
SVI (30)	Sludge volume index after 30 minutes, volume/mass
ZSV	Zone settling velocity, distance/time
SPSS	Supernatant suspend solids, mass/volume
dx/dt	Net microbial growth, mass/time
So	Influent substrate concentration, mass/volume
S	Effluent substrate concentration, mass/volume
$\frac{ds}{dt}$	Substrate removal rate, mass/time
Y	Microbial growth yield, no dimensions
Y [ds/dt]	Specific oxygen utilization rate, $time^{-1}$
Kd	Organism decay coefficient, $time^{-1}$
E	Process efficiency, percent
Do	Dissolved oxygen, mass/volume
Q	Influent flow, volume/time
Q_R	Return sludge flow, volume/time
X_R	Concentration of return sludge solids, percent

CHAPTER 1

INTRODUCTION

1.1 NEED FOR INNOVATIVE TECHNOLOGY

The problem of treating wastewater efficiently and economically has always been an issue of major concern for small communities (Sparling, 1983). During the earlier period, design engineers placed emphasis on more sophisticated and complex treatment processes which were intended to improve overall treatment efficiencies. Today, small communities find these equipment-intensive plants very difficult and costly to operate and maintain (White, 1981). Financial constraints have compounded the problems as these communities strive to meet treatment objectives. Higher energy costs, inflation, lower tax support bases, and reduced government funding are the primary antecedents. For example, these communities can seldom compete for the services of skilled personnel required to operate such facilities. In general, there has been a decrease rather than a projected increase in treatment efficiencies.

There is a need for the use of simpler, lower-cost, and more efficient wastewater treatment systems (Saxon et al., 1981). The need was also recognized by the United States Environmental Protection Agency (herein referred to as E.P.A.). E.P.A. introduced the "Innovative and Alternative Program", which offered state funding as incentives for consulting firms to satisfy the aforementioned needs. Environmental engineering practitioners have now placed emphasis on designing processes which reduce capital and operating costs (Christopher et al., 1984).

In addition, there has been a trend towards improving treatment efficiencies. In general, there continues to be an improvement in the design of traditional elements which include: oxygen transfer, mixing and cell-separation performance.

The focus of this thesis is to present and evaluate a new concept of upflow clarification in activated sludge separation. Cowatt Systems of Manitoba, Canada have recently constructed a wastewater treatment facility of novel design. The wastewater treatment facility can be described as a continuous flux upflow clarifier system (herein referred to as the CFLUC system). As shown in the following case study, CFLUC offers the potential to improve cell-separation performance while minimizing capital and operating costs (Sparling, 1984).

1.2 CASE STUDY: TOWN OF SOURIS

In 1981, the Town of Souris, a small community located in southwestern Manitoba, was faced with the financial burden of constructing a new wastewater treatment facility. The Town had solicited design proposals and had two options, which included the construction of a conventional extended aeration plant, or an extended aeration facility of novel design. The primary reason the Town selected the latter option was that the capital cost was two-thirds the cost of the conventional facility. Other considerations included simple operation and lower projected operating costs.

A flow schematic of the proposed 920 m^3/day facility is shown in Figure 1.1. The CFLUC system consists of three basic components: extended aeration basin, newly conceived upflow clarifier, and an air supply.

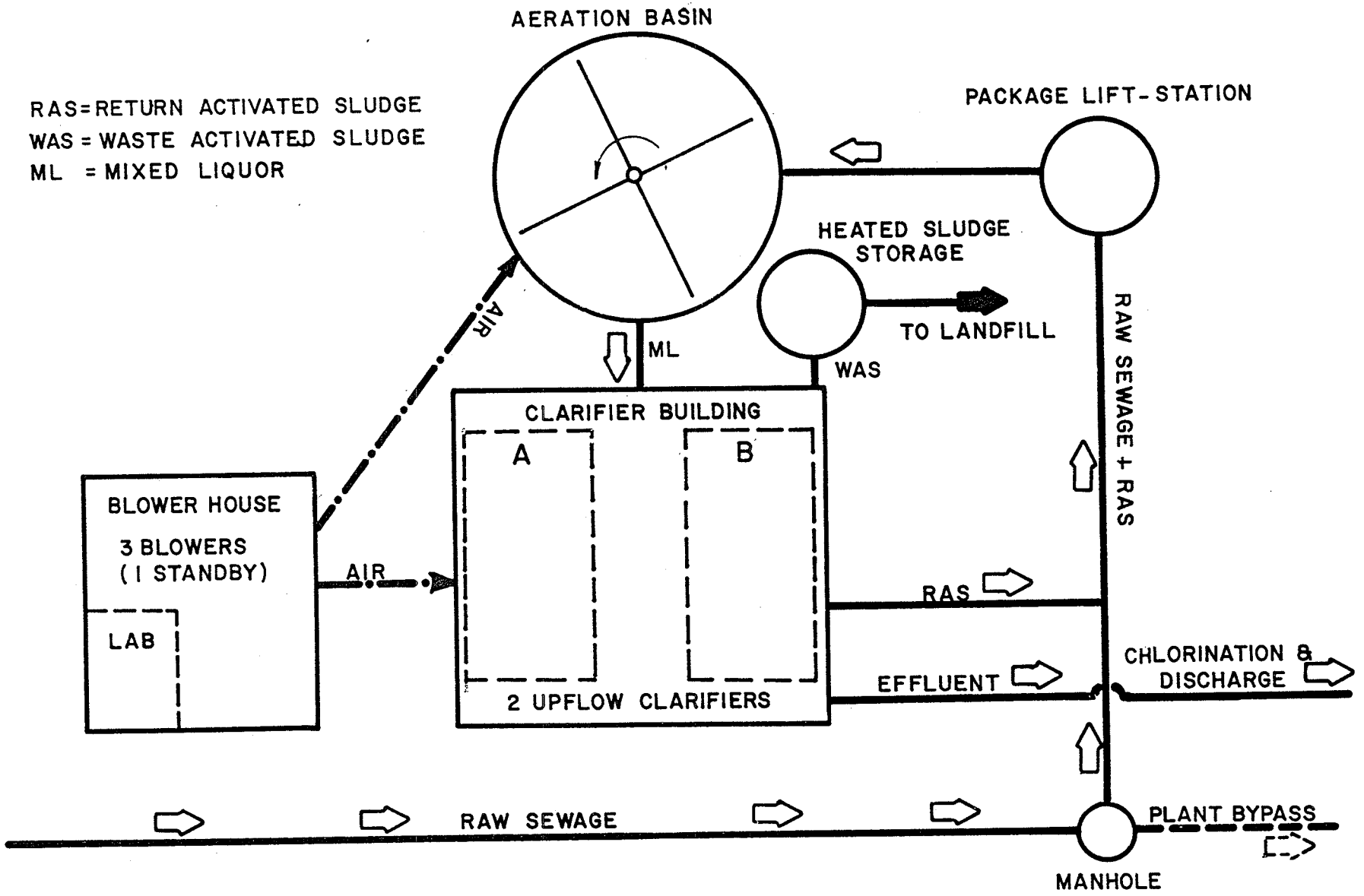


Figure 1.1 Schematic of Wastewater Treatment Plant at Souris, Manitoba (Sparling, 1984)

The facility had been designed on the basis of preliminary pilot plant studies conducted on site at Souris, Manitoba. The pilot plant consisted of a pilot scale aeration basin (volume: 10 m^3) and an upflow clarifier (diameter: 750 mm) for final sedimentation. The design flow, overflow rate and HRT were 20 m^3/day, 22 m^3/m^2.day, and 12 hours, respectively. Similar to the full-scale treatment facility, the pilot plant received domestic sewage and industrial wastewaters from local cheese and plastic factories. An evaluation of the wastewater characteristics indicated significant daily variations in the strength of the wastewater and the quantity of flow.

It was concluded from the pilot plant studies that, (Sparling, 1982):

1. "The treatment provided by the pilot plant is capable of treating domestic sewage to a degree that complies with the discharge requirements."
2. "Care must be taken by the Town to ensure shock loadings from the cheese plant to be minimized."
3. "Every effort should be taken by the Town to eliminate storm-runoff and infiltration to the distribution system."

In summary, the pilot plant studies had shown the CFLUC system to be workable, and answered many questions related to the wastewater characteristics. Other operating conditions of the system such as the performance of the clarifier at various F:M and/or SRT ratios, however, remain unanswered. Other limitations of the pilot plant studies include conducting tests for operating conditions which were different than those used in design. For example, all pilot testing was conducted at an HRT of 12 hours, however, the design of the full-scale plant was based on an HRT of 24 hours. Unfortunately, the discrepancy in operating parameters was a result of financial and time constraints

imposed on the designer during the selection of the pilot-scale equipment. There is evidence to show that the operating conditions in the biological reactor can have a marked effect on clarifier performance (Bosogni and Lawrence, 1970).

In order to evaluate and project the performance of the CFLUC system at Souris, there was a need to conduct further laboratory studies.

1.3 OBJECTIVES

The general purpose of this study was to present and evaluate a new concept of upflow clarification in activated sludge separation. The specific objective of the test program was to evaluate the efficiency of the upflow clarifier in separating sludges with different settling characteristics. An activated sludge system was used to treat a synthetic dairy waste at four different F:M ratios and/or SRT values. All sludges produced had different physical or biochemical characteristics. Operating control parameters such as HRT, overflow rate and dissolved oxygen levels were selected to simulate operating conditions at Souris, Manitoba.

CHAPTER 2

BACKGROUND

2.1 GENERAL DISCUSSION

The activated sludge process has been defined as "a continuous or semi-continuous (fill-and-draw) aerobic method for biological wastewater treatment including carbonaceous oxidation and nitrification" (Ganczarczyk, 1983.) In simple terms, the process can be thought of as a conversion process whereby soluble organics are converted to biological growth in an aerobic environment (i.e. aeration basin). The biological growth is then separated from the treated wastewater in the secondary clarifiers. A portion of the biological growth is wasted from the system and the remainder is returned to the aeration basin. A flow diagram of the conventional activated sludge process is shown in Figure 2.1.

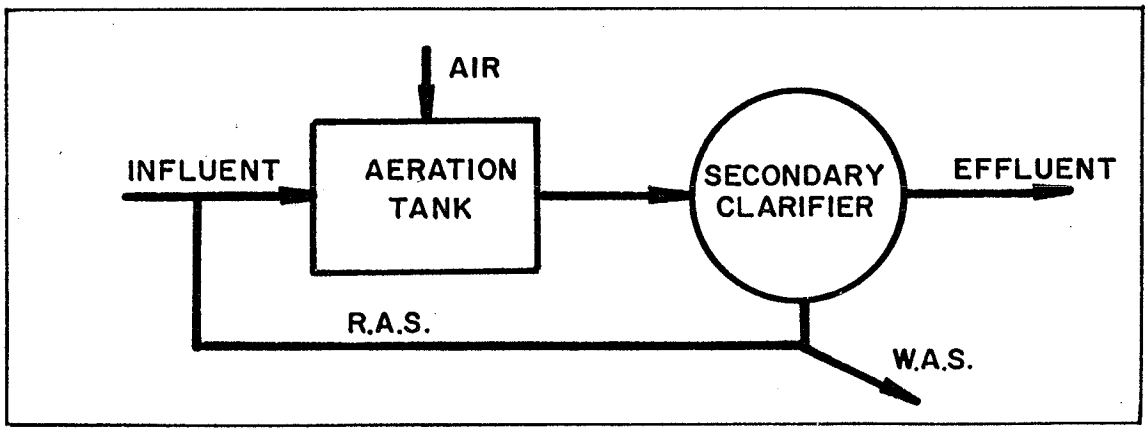


Figure 2.1 Flow diagram of the conventional activated sludge process

In the operation of the activated sludge process, the efficient performance of the secondary clarifiers is as important as the operating

control parameters in the aeration basin. This is because the clarifier has to provide solids-separation with the dual purpose of firstly providing a clarified effluent low in SESSS and, secondly, sludge thickening so that activated sludge can be returned to the aeration basin (Pitman, 1980). Inadequate solids-separation can result in high SESS values and also a failure of the process caused by insufficient return of the biomass to the aeration basin. Sedimentation efficiencies can be optimized by controlling the operating variables of the biological process and by improving the hydraulic design of the secondary clarifier to enhance flocculation and sedimentation. Both approaches are addressed in the following sections.

The following section, entitled "Activated Sludge Process Concepts" includes a historical review and a discussion of biological and environmental concepts that influence process efficiencies. In Section 2.3, an assessment of past and current secondary clarification design concepts are presented, which include hydraulic, operating and bioflocculation considerations. A conceptual model of the CFLUC system is presented based on new applications of established concepts. The final section of the chapter includes a description of the new upflow clarifier, and its operating characteristics in relation to conventional clarifiers.

2.2 ACTIVATED SLUDGE PROCESS CONCEPTS

2.2.1 Historical Perspective

The activated sludge process was initially developed by Fowler, Ardern, Mumford and Lockett in 1914. Historically, one of the most remarkable facts subsequent to the discovery, was that rudimentary

tests conducted by the aforementioned researchers would be transformed into full-scale applications in a matter of years (Sawyer, 1965; Alleman, 1983). However, despite the monumental value of the discovery, activated sludge did not find widespread application until the 1940's and 1950's. The main reasons for this were patent litigations and a poor understanding of the biological factors of the process.

The years of legal controversy were also highlighted by two opposing theories. The debate centered around the fundamental behavior of the process, more specifically, in terms of whether it was a physical-chemical or a biological process (Alleman and Prakasam, 1983). By 1930, the process was considered by many researchers (Baly, Parker, Theriault, Mohlman) to be primarily a physical process. The physical theory of activated sludge treatment during this period may be summarized as follows, (Metcalf and Eddy, 1930):

> "The first and perhaps most noticeable function of the process is that of coagulating or flocculating suspended and colloidal matters in the sewage. The floc is a sponge-like mass, or as expressed by Stein, an open mesh network which, in the process of formation may envelop, entrap, or entrain colloidal matter and bacteria. The sponge-like structure of the floc offers a very large surface area for contact. Buswell has estimated that the sludge surfaces present an area of 500 square feet per cubic foot of tank volume. When the floc is driven about in the liquid, it has a sweeping action, or, as stated by Parker, the "process may be regarded as passing a filter through the water in place of passing the water through a filter.""

By 1935, the arguments in favor of the biological theory of activated sludge were well established. It would not be until 1982, however, that physical mechanisms such as conditioning (opportunity to provide flocculation/solids contact) would be utilized integrally with biological mechanisms in the design of a full-scale treatment facility (Parker, 1983).

New concepts in the application of the activated sludge process were introduced in the 1940's, and resulted primarily from improvements in plant operation (McKinney, 1962). These improvements led to tapered aeration, step aeration, and modified aeration systems. Further innovations resulted in the high rate, Kraus process, and contact stabilization systems.

One disadvantage of the conventional activated sludge process is the large quantities of sludge produced, which resulted in disposal problems. This factor was significant in that it eliminated the practical application of the activated sludge process in many small installations. This led to the development of the extended aeration system, where sludge volume was reduced. Today, the extended aeration process has made the activated sludge process a practical system of treatment for many small communities.

In all of the aforementioned modifications of the activated sludge process, initial design methods were empirical in nature, and based primarily on hydraulic detention times and developed loading criteria. It has only been within the last thirty years that design concepts have been based on the fundamental and scientific concepts of mass balance and microbial growth kinetics. With this new design approach, it is now possible to design a process on biological parameters.

2.2.2 Process Microbiology

To operate and design an activated sludge system efficiently, it is necessary to understand the role of microbiology in the process. Activated sludge is a mixed biological population of bacteria, viruses,

protozoa, fungi, and other multicellular organisms. Entrapped in the sludge floc are dead cells, waste bi-products, and inert and/or organic solids.

The species of microorganisms will depend on several factors which include: wastewater characteristics, environmental considerations, operating conditions, and process design. The microbial community is dominated by heterotrophic bacteria and, to a lesser extent, by autotrophic bacteria. Heterotrophic bacteria utilize organic wastes as a source of both carbon and energy. Autotrophic bacteria utilize carbon dioxide as a carbon source and derive their energy from either photosynthesis or the oxidation of mineral compounds as an energy source (Ganczarczyk, 1983). A schematic presentation of the heterotrophic and autotrophic metabolisms is shown in Figure 2.2.

In the aeration basin, a portion of the organic waste (herein referred to as a substrate) is removed and assimilated by the heterotrophic bacteria to obtain energy for further synthesis of substrate into cell tissue (Metcalf and Eddy, 1979). The cell tissue will be converted to gaseous end products and energy through endogenous respiration. The aerobic conversion of organic matter is stoichiometrically represented in Table 2.1.

Activated sludges are generally classified into two main categories: normal and/or bulking sludges. There is a significant number of different bacterial species for each category. Eikelboom (1975) identified a variety of over thirty filamentous organisms which are attributed to bulking sludges.

Early investigators noticed that zoogleal bacteria predominated in normal activated sludge and contributed significantly to the

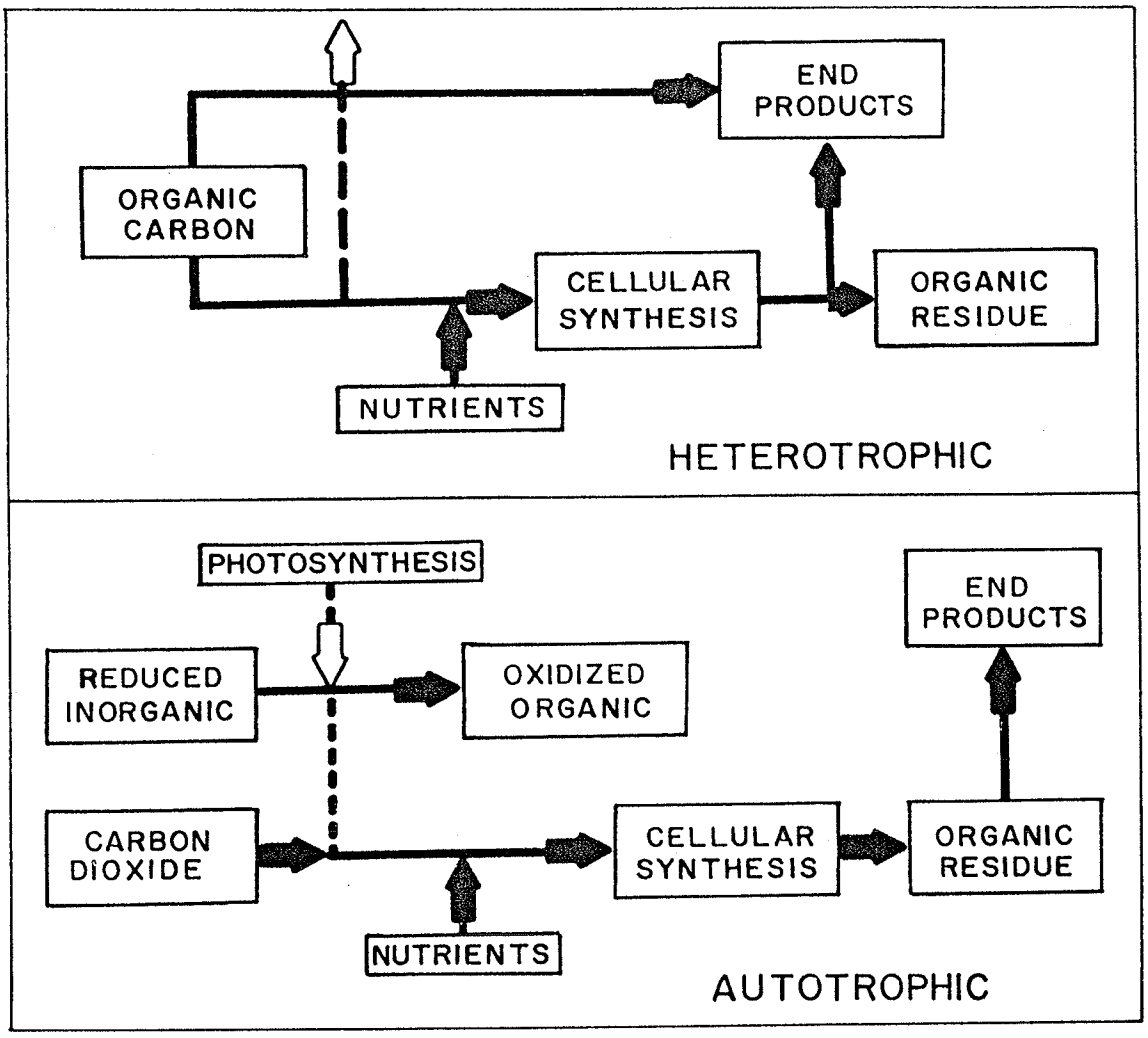


Figure 2.2 Schematic representation of heterotrophic and autotrophic bacterial metabolisms (Metcalf and Eddy, 1979)

TABLE 2.1

AEROBIC CONVERSION OF ORGANIC SUBSTRATE

1.	Oxidation (dissimilatory process)	$COHNS + O_2 + bact \rightarrow CO_2 + NH_3 +$ energy + end products
2.	Synthesis (assimilatory process)	$COHNS + O_2 + bact + energy \rightarrow C_5H_7NO_2$ (new cells)
3.	Endogenous Respiration	$C_5H_7NO_2$ (cell tissue) $+5O_2 \rightarrow 5CO_2 + NH_3 + 2H_2O$ + energy

development of a stable floc (McKinney, 1952). Other bacteria have been shown capable of performing similar functions, and include members of the genera *Pseudomonas*, *Achromobacter*, *Flavobacterium*, *Alcaligenes*, *Bdellovribrio*, *Mycobacterium*, and the two autotrophic bacteria, *Nitrosomonas*, and *Nitrobacter*.

The term "bulking sludge" has been used in the past to describe a variety of problems associated with the separation of activated sludge. Chambers et al. (1982) defined bulking as the phenomenon which occurs in activated-sludge plants when the sludge occupies an excessive volume and does not settle rapidly, so that in extreme cases, the effluent contains excessive concentrations of SESS. One cause of sludge bulking is the excessive growth of filamentous organisms resulting from adverse changes in the aerobic environment. Many different organisms have been identified which may cause sludge bulking. Several of these organisms include: *Sphaerotilus*, *Bacillus*, *Beggiatoa*, *Thiothrix*, *Geotrichum* and *Nocardia*. (Eikelboom, 1975; Gerardi, 1983).

While the bacteria are the microorganisms that actually assimilate the organic substrate, other organisms such as protozoa and rotifers take part indirectly in the stabilization of the organic waste. The relative growth of micro-organisms stabilizing an organic substrate is shown in Figure 2.3.

During the early stages of the growth cycle (i.e. at high substrate loadings), flagellata and/or free swimming ciliates predominate. These microorganisms consume dispersed bacterium and thus improve the quality of the effluent. As the substrate load decreases with time, stalked ciliates become the predominant organism. This protozoa form is normally a sign of a well-operating process

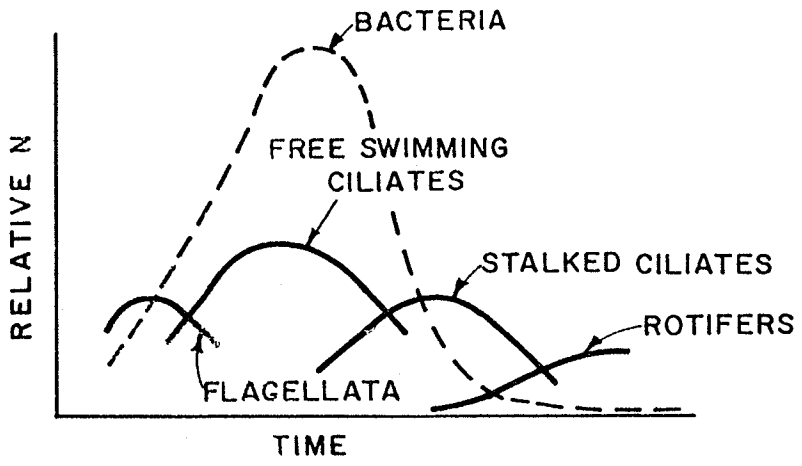


Figure 2.3 Relative growth of microorganisms in a start-up phase of the activated sludge process (after Ganczarczyk, J.J., 1983)

(W.P.C.F., 1976). Rotifers are present in over-aerated and under-loaded systems. The organism is common in extended aeration systems. Rotifers consume small biological floc particles and are common indicators of a stable biological system.

For many years, the visual examination of mixed liquor has been used as a guide in controlling the operation of the activated sludge process.

In the past, attention has been given to the growth pattern of bacteria, since this is the major microorganism responsible for the removal of pollutants in wastewater. The growth pattern of bacteria is an integral part of understanding the fundamental concepts of substrate utilization and microbial growth kinetics. The bacterial growth pattern for a normal activated sludge is shown in Figure 2.4. The ideal growth curve is based on a batch culture application.

Various positions on the curve are associated with changes in the physical or biochemical characteristics of the normal zoogleal populations. The relationships between the fundamental characteristics

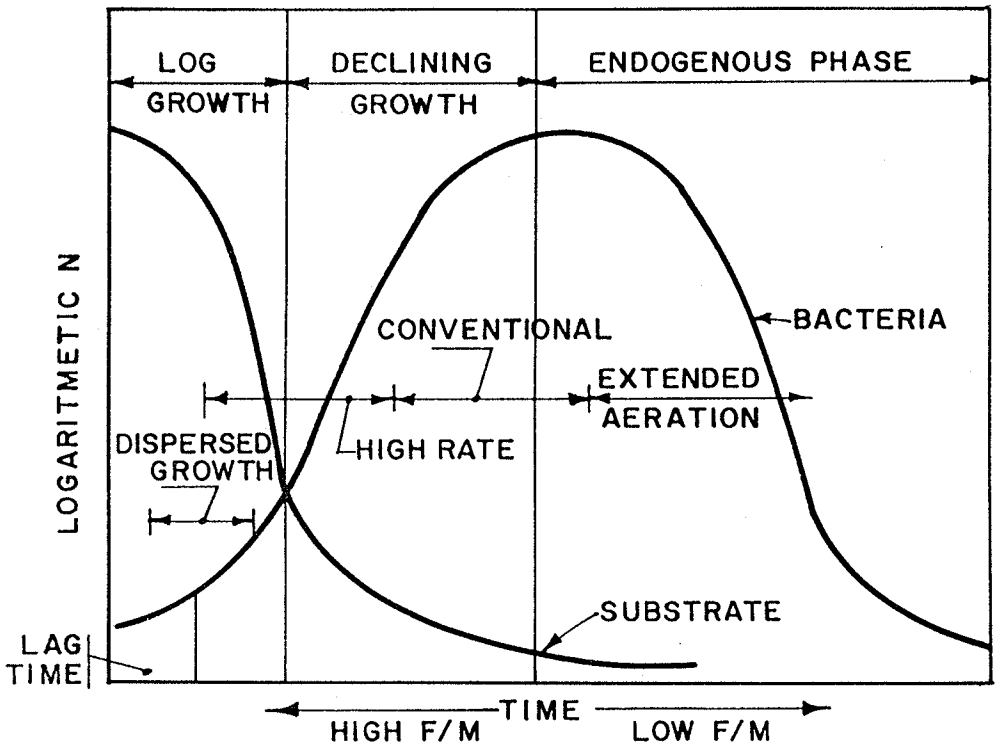


Figure 2.4 Ideal growth curve for a batch type activated sludge unit (after WPCF, 1976)

of the curve and settleability have been established. At the beginning of the growth cycle there is a lag phase or a period of time when the microorganisms are acclimated to the substrate. The lag phase is followed by a rapid increase in growth, often termed "logarithmic rate of growth". In this growth phase there is a high oxygen demand and the substrate is removed at a maximum rate of conversion into new cells. In the logarithmic growth phase the substrate is non-limiting and is often characterized by a high F:M ratio. It was theorized by McKinney (1962) that high energy levels kept the microorganisms dispersed, resulting in poor settleability. As the substrate is utilized and the F:M ratio decreases, the bacteria show a declining rate of growth. The substrate is said to be a limiting factor for further growth. Most conventional activated sludge plants operate in the declining growth

phase.

As the substrate is further depleted, the growth rate will eventually equal the death rate, hence, the beginning of the endogenous phase. In the endogenous phase, the bacteria metabolize the remaining substrate at a very slow rate. Old cells are lysed to produce metabolic substrate for new cells, resulting in a net reduction in energy levels. It has been hypothesized by several researchers (McKinney, 1962; Bosogni and Lawrence, 1970) that in the endogenous phase the bacteria start to produce extracellular biopolymers which promote the formation of floc particles. This phase is often characterized by light, good settling sludges, with the formation of pinpoint floc in the supernatant.

The previous discussion shows that the operating conditions in the aeration basin can have a marked effect on the physical or biochemical character of normal zoogleal populations. Biological parameters which are used in operational control and design include MLSS, SRT, and HRT. These parameters are addressed in the following section.

2.2.3 Factors Affecting the Activated Sludge Process

(i) Wastewater Characteristics

One of the major complaints concerning the activated sludge process is its lack of stability which is mainly attributed to fluctuating wastewater strengths and flows. The activated sludge process is continuous rather than the batch type process represented in Figure 2.4. Daily variations in the wastewater characteristics result in constant oscillations along the growth curve, and often result in a continuous imbalance of the microorganisms. The application of flow

equalization basins offer a practical solution to improve the stability of the process.

(ii) Hydraulic Detention Time

One important process parameter in the activated sludge process is the concept of HRT. The HRT is the time duration required to aerate wastewater with activated sludge. It is usually expressed in hours, and can be determined from the following relationship:

$$\frac{\forall}{Q} = HRT \qquad (1)$$

where $\forall$ = Volume of aeration basin, m^3

Q = Wastewater flow, m^3/h

The recommended values of HRT varies for each modification of the activated sludge process. Recommended design values are summarized in Table 2.2 (Metcalf and Eddy, 1979).

Longer HRT values, such as that comparable to the extended aeration process, are recommended in wastewater treatment plants experiencing shock loads. Ganczarczyk (1976) reported that there is an increase in "buffering capacity" for such systems.

(iii) Food to Microorganism Ratio

As indicated in Table 2.2 the F:M ratio is an important design concept often used to characterize a specified activated sludge process modification. The F:M ratio is measured in terms of kg of BOD_5 or kg of COD per kg MLVSS.d. The F:M ratio is determined from the following relationship (Metcalf and Eddy, 1979):

$$F:M = \frac{So}{(HRT)(X)} \qquad (2)$$

TABLE 2.2

RECOMMENDED DESIGN PARAMETERS FOR THE ACTIVATED SLUDGE PROCESS

	Process Modification	HRT (h)	F:M (d^{-1})	SRT (d)	MLSS (mg/L)
1.	Conventional	4-8	0.2-0.4	5-15	1,500-3,000
2.	Tapered Aeration	4-8	0.2-0.4	5-15	1,500-3,000
3.	C.F.S.T.R.	3-5	0.2-0.6	5-15	3,000-6,000
4.	Contact Stabilization	0.5-1.0[a] 0.3-6.0[b]	0.2-0.6	5-15	1,000-3,000[a] 4,000-10,000[b]
5.	High Rate	0.5-2.0	0.4-1.5	5-10	4,000-10,000
6.	Extended Aeration	18-36	0.5-0.15	20-30	3,000-6,000

a Contact Unit
b Solids Stabilization Unit

The success in operating an activated sludge plant is to maintain an optimum F:M ratio. The ultimate control of the system is based on the wasting of solids (i.e. new growth). The wasting of solids based on F:M ratios is difficult and seldom practiced. The evaluation of So and X is usually what makes the use of the F:M ratio impractical as a control parameter.

(iv) Mixed Liquor Suspended Solids

The concentration of MLSS in treated wastewater is only an estimate of the density of microbial solids in solution. The MLSS in a treatment unit consists of active microorganisms, non-active microorganisms, non-biodegradable organics and inorganic mass (Ganczarczyk, 1983). The most common parameter used as a measure of biological solids is the volatile suspended solids (herein referred to as MLVSS). This parameter is also inaccurate since it does not measure the active cellular material. Ganczarczyk (1983) reported that the active microbial mass represents only 30% or less of the MLSS.

The evaluation of the active mass of micro-organisms is usually what makes the MLSS impractical as a control parameter. In addition, this parameter ignores the concept of the F:M ratio. For example, significant increase in organic load could result in poor settleability (i.e. a shift in the growth curve) if the operator was to control the process by maintaining a constant MLSS level.

(v) Solids Retention Time

The value of SRT is based on the concept that to control the growth rate of micro-organisms, a percentage of the cell mass (i.e. new growth) must be wasted from the system daily. In literature, SRT is often referred to as mean cell residence time and/or the sludge age. The net microbial growth rate can be expressed mathematically as shown in Equation 3.

$$\frac{dx}{dt} = \frac{Yds}{dt} - KdX \qquad (3)$$

The SRT is related to the reciprocal of the net microbial growth rate (i.e. $\frac{dx}{dt}$). Equation 3 can be rearranged to yield:

$$\frac{1}{SRT} = Y \left[\frac{F:M}{100} \right] E - Kd \qquad (4)$$

The value of SRT may also be determined by calculating the total mass of micro-organisms in the system and dividing it by the rate at which micro-organisms are wasted from the system.

As shown in Equations (3) and (4), the SRT is inversely proportional to the F:M ratio and the rate of microbial growth. One is indirectly controlling the F:M ratio and the microbial growth by controlling the SRT of the system. Presently, maintaining a constant SRT in the aeration basin is considered the best control method that can be utilized for the activated sludge process (Ganczarczyk, 1983; Metcalf and Eddy, 1979). For example, if an SRT of 20 days is required for a desired treatment efficiency then 5% of the total cell mass is wasted from the system daily. Wasting solids in this manner is a simple method of controlling and measuring SRT.

The performance of the activated sludge process is influenced by the SRT value selected. It is possible, in most applications, to determine the optimum SRT value required to maximize effluent equality.

(vi) Sludge Volume Index

The SVI index, introduced by Mohlman in 1934, characterizes how well a sludge settles and compacts. Pipes (1979) defined the SVI index as the volume occupied by the sludge after 30 minutes settling divided by the MLSS concentration. The SVI index is defined mathematically as follows:

$$SVI = \frac{SV_{30}}{MLSS} \qquad (5)$$

The SVI index for a particular test can vary depending on

factors such as: initial MLSS concentration, temperature, test cylinder diameter, initial depth and the effect of stirring (Dick and Vesiland, 1969).

The index, although limited in its application, represents the combined influence of the physical properties of the sludge. It is not a scientific parameter but can be used as an operational tool for in-plant control.

The SVI test also gives the operator an indication of the return sludge rate. The return sludge rate to the effluent flow rate can be calculated as the sludge volume divided by the supernatant volume after settling for 30 minutes in a 1000 mL graduated cylinder.

As a general rule, it is accepted that SVI values greater than 200 mL/g indicate severe filamentous bulking. SVI values less than 100 mL/g are considered typical for a normal sludge. Values between the aforementioned extremes indicate moderate bulking (Pipes, 1979).

Variations in the SVI index can be mainly attributed to two biological responses. The first response, discussed in Section 2.2.2 is associated with variations in the physical and biochemical character of normal zoogleal sludges. The second response involves the shift in normal zoogleal populations to the excessive growth of filamentous type organisms. The latter response is more complex and is still poorly understood.

With regard to the first response, changes in the F:M ratio may result in shifts in the biological growth curve and poor settling performance. These settling problems include the formation of pinpoint floc and deflocculation. Pinpoint floc consists of small yet visible flocs and usually occurs in the extended aeration mode (i.e. F:M $<$

0.2 d^{-1}). Although the SVI index values are low (typical SVI < 100 mL/g), indicating a good settling sludge, the effluent is turbid. The term deflocculation is also referred to as dispersed growth. In laboratory studies, Bosogni and Lawrence (1970) reported predominantly dispersed growth at high F:M or low SRT values (i.e. SRT = 0.25-2.00 d). These results were supported by Pipes (1979) who concluded, after analyzing sludges from 32 plants throughout the United states, that sludges have a tendency to deflocculate when F:M ratios exceeded 0.4 d^{-1}.

Investigations into the phenomenon of bulking sludges have been undertaken for the past 60 years to identify the conditions responsible for bulking sludges. In 1978, Sezgin et al. presented a unified theory describing those factors which appear to be responsible for filamentous bulking sludge. The authors concluded that filaments form backbones on which zoogleal bacteria attach and form strong flocs. An excess of filamentous growth presents agglomeration and a lack of filaments led to the formation of pinpoint floc. It was concluded that a filament length of 10^7 u/mL was a good dividing point between a filamentous bulking sludge and a non-bulking sludge.

Similarly, Magar et al. (1976) concluded that the density of the floc particle decreased with filament length and this was the most significant factor repsonsible for high SVI values.

Chao and Keinath (1979) in treating a synthetic substrate derived a relationship between SVI and various SRT values for the system. The authors concluded that the lower settling rates and poor compaction could be attributed to the condition of the floc surface. Normal sludge flocs have relatively smooth surfaces. The rough surface of filamentous

type flocs results in increased frictional drag forces, hence, reduced settling rates.

The effects of bulking on final effluent quality will depend on the seriousness of the problem. One characteristic of a high SVI index is that low SESS concentrations are produced. Keefer (1963) and Pipes (1979) have observed that unusually clear effluents result when settleability is poor but not excessive.

Sezgin et al. (1978) and Pipes (1979) postulated that an abundance of filaments protruding from the floc produce a network and absorb and entrap small floc particles, resulting in a clear supernatant.

Factors associated with the occurrence of bulking have included: overloading, underloading, over-aeration, under-aeration, short-circuiting, nutrient imbalance, high pH value, low pH value, high temperature, sulphides, and many other identifiable conditions (Barnard, 1978; Chambers and Tomlinson, 1982). It is apparent that, with the many factors responsible for sludge bulking, it becomes difficult to predict the respective trends of SVI on a universal basis. This is indicated on Figure 2.5 by two prominent researchers in the field who attempted to correlate SVI with SRT values (Barnard, 1978) of microorganisms which remain dispersed after quiescent settling.

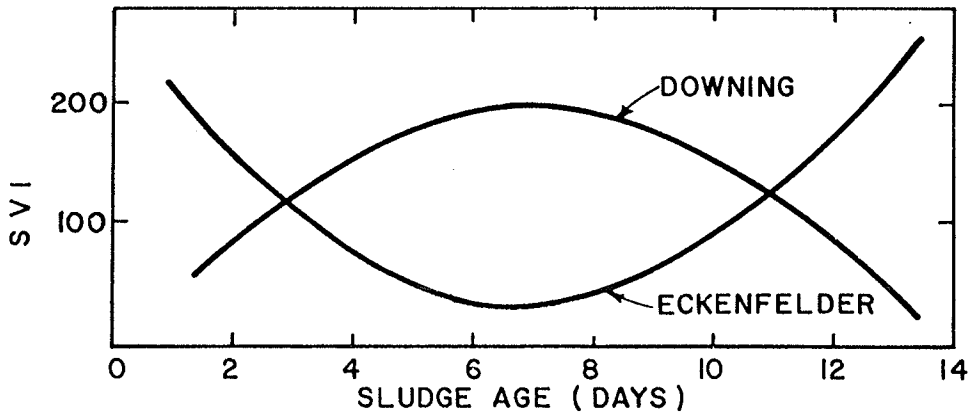


Figure 2.5 Relationships between SVI and sludge age (after Barbard, 1978)

Although it is presently recognized that there is no universal relationship between SVI and SRT, many researchers (Ford and Eckenfelder, 1967; Ganczarczyk, 1983; Chao and Keinath, 1979; Segzin et al., 1978) show the tendency for activated sludges to bulk at very low loadings and at increased loadings. Palm et al. (1980) verified Eckenfelder's curve and concluded that it would be possible to develop unique curves for each DO concentration in the aeration basin. The authors postulated that the rising limb, at high F:M ratios, was caused by low DO values in the floc interior. The research also verified that by increasing the DO concentration in the aeration basin, the range of optimum SVI values was extended. It was postulated that the rising limb may be attributed to nutrient deficiencies which favored the growth of filamentous type organisms. The result of this study is schematically illustrated in Figure 2.6.

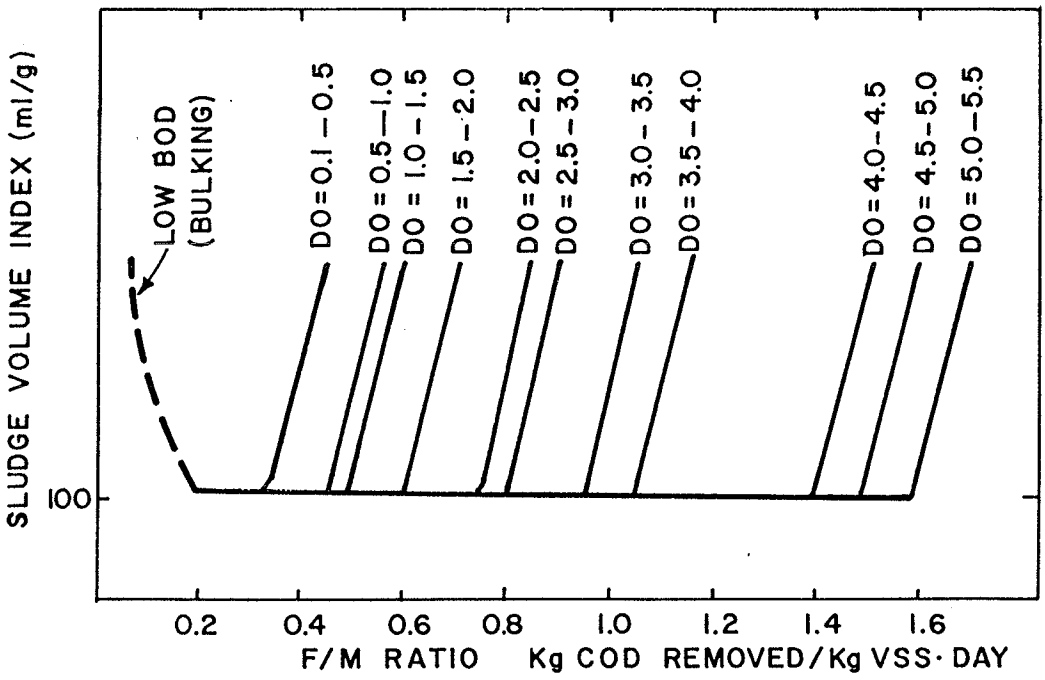


Figure 2.6 Proposed relationships between SVI and F:M at various dissolved oxygen levels

Chao and Keinath (1979) concurred with the above relationship, but their results indicated that two optimum F:M ranges existed for low SVI values. These ranges compared with the extended aeration and high rate modes of operation. Further investigations are required in this area to substantiate the author's results.

(vii) Zone Settling Velocity and Percent Dispersion

The settling characteristics of activated sludge can be defined by two additional parameters: zone settling velocities, and the percent of microorganisms which remain dispersed after quiescent settling.

The use of ZSV to characterize sludge settling performance is generally considered superior to SVI (Dick and Vesiland, 1969). The values of ZSV for a particular sludge can be determined by conducting batch settling tests. The solids-interface height is plotted versus time, enabling the computation of interface settling velocity in the linear range for each suspension. A typical normal activated sludge settling curve is shown in Figure 2.7.

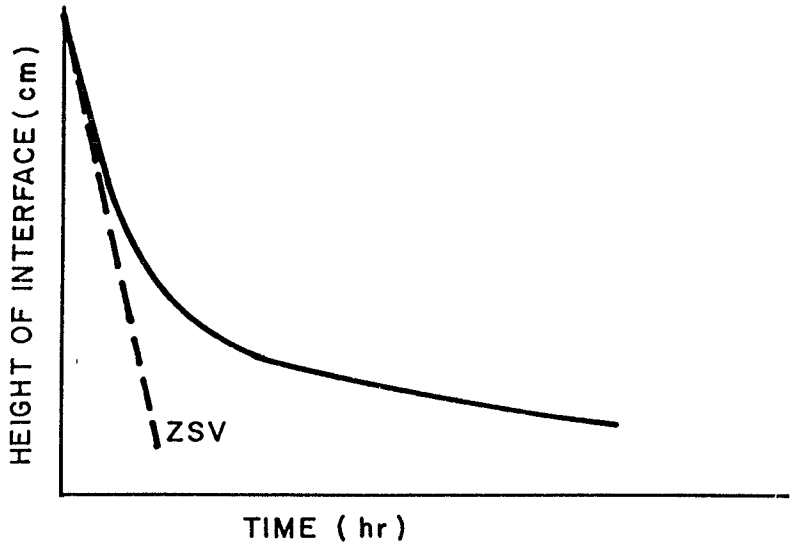


Figure 2.7 Typical normal activated sludge settling curve

Bosogni and Lawrence (1970) reported ZSV values of "2-6 ft per hour and 15 ft per hour" for well formed floc and pinpoint floc, respectively. Sezgin et al. (1978) reported that ZSV values decreased linearly with high filament lengths in the range 10^6-10^8 u/mL.

The second method used to define the settling characteristic of floc particles is "percent dispersion" or simply the percent of solids which do not settle. The test is performed by allowing the SVI sample to settle for an additional ½ hour (i.e. a total time of 1 hour). The solids remaining in the test cyclinder (SPSS) only gives an idea of the percent of MLSS which remain dispersed under quiescent settling conditions. It is not to be construed to represent the SS concentration that will be dispersed in the clarifier (SESS).

The SPSS obtained under batch conditions is a useful parameter for comparing the performance of existing clarifiers. Parker (1983) reported that when SESS exceeds SPSS the floc carryover is indicative of some operating problem within the clarifier. These problems may include: density currents, short circuiting, high sludge blanket, or some other factor causing significant deviations from the ideal settling batch conditions. To the contrary, when the SESS concentration is less than the SPSS level, this condition is indicative of improved efficiencies in the clarifier; most likely resulting from enhanced bioflocculation. The following section addresses the concept of bacterial flocculation.

2.2.4 Bacterial Flocculation

There is no universally accepted theory to explain why activated sludge flocculates. To further complicate matters, there is no accepted method of measuring the extent of the flocculation. Pipes (1979) attempted to describe flocculation in terms of SESS and SPSS. If the effluent contained low concentrations of SPSS and SESS, it is described as pinpoint floc. If the effluent is turbid and also contains high concentrations of SESS and SPSS, it is termed deflocculated. Although this description is not scientific, it can be concluded that concentrations of SS that do not settle are somehow related to the lack of bioflocculation.

Ganczarczyk (1983) presented two models which partially explain the flocculation of microorganisms. These included the physical-chemical and biopolymer models.

(i) Physical-chemical Model

The theory of floc formation based on the physical and chemical properties of the bacterial surfaces was proposed by McKinney (1952). McKinney concluded that because of the chemical structure of the bacteria, which he postulated involved ionized carboxyl and amino groups, there was the high possibility for chemical exchanges to occur. McKinney suggested that cations from salt solutions were adsorbed on the bacterial surface, reducing the net surface charge. Through physical means, such as gentle agitation, McKinney hypothesized that cells (which have a large surface area to volume ratio) approach one another, forming chemical bonds such as salt and ester linkages.

(ii) Biopolymer Models

The theory of biopolymers is receiving increased attention, especially in the food industry. The present observations concerning biopolymers suggest that floc forming bacteria produce an exocellular polysaccharide biopolymer during the early declining growth stage. It is assumed that these polyelectrolyte-type compounds are responsible for bacterial flocculation (Chao and Keinath, 1979). This observation was also in accordance with Pavoni et al. (1972).

Monosaccharides and disaccharides in combination with their derivatives are the predominant chemical components of biopolymers extracted from sludges (Forster, 1968). Some researchers have found that, with the presence of excessive nutrients, the concentrations of these biopolymers are very low (Ganczarczyk, 1983).

Parker at al. (1971) recognized the significance of physical agitation as a means of improving bioflocculation. The authors demonstrated the efficiency of a flocculation step to provide an environment more conducive to encourage aggregation of dispersed floc into well-formed floc particles.

2.3 APPLICATION OF CONCEPTUAL THEORY TO DESIGN

2.3.1 Conventional Clarifiers

This section pertains to the design of conventional clarifiers for gravity sedimentation of biological secondary treatment, excluding separation techniques such as filtration, tube and lamella separators, flotation and microstrainers. In many cases, the latter separation techniques are costly and can be avoided by optimizing the design of secondary clarifiers and controlling the operating variables of the

biological process (Parker, 1983).

Among the general types of conventional clarifiers available for gravity sedimentation are the following:

(i) Horizontal flow clarifiers

(ii) Upflow clarifiers (example, Dortmund tanks)

(iii) Solids contact unit

(i) Horizontal Flow Clarifiers

Circular clarifiers (radial or square radial) are the most common clarification units utilized in activated sludge treatment. In general, the hydraulic principles of circular clarifiers apply to rectangular horizontal flow basins as well. The theory of solids-separation in a clarifier basin being a function of overflow rate or HRT was first proposed by Allen Hazen in 1904. Camp (1953) further developed the theory, in that he hypothesized that, if one could obtain plug flow in a clarifier, its performance would be identical to that predicted by quiescent batch settling tests. As shown in Figure 2.8, Camp (1953) proposed to divide the clarifier volume into four zones: inlet zone, clarification zone, outlet zone, and sludge zone.

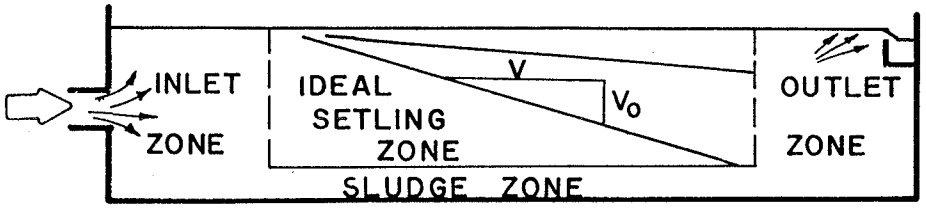


Figure 2.8 Idealized model for a horizontal rectangular continuous flow clarifier (after Camp, 1953)

Camp (1953) suggested that in actual basin hydraulics, short-circuiting cannot be prevented, and as a result, there will be

a decrease in the actual time of detention. This model has led to the use of inlet and outlet baffles to insure uniform flow distribution across the clarifier, thus reducing short circuiting and dead spaces in the clarifier.

In the early stages, horizontal flow clarifiers were designed on arbitrary values of overflow rate and HRT. More recently, the importance of solids loading has been researched, leading to the limiting solids flux theory proposed by Dick (1970).

According to this theory, there is a limiting solids flux concentration for a given sludge. It is this limiting solids flux or critical solids loading condition that could govern the design of secondary clarifiers. The limiting solids flux can be determined by conducting batch settling column tests. The data should then be analyzed using simple graphical methods (Metcalf and Eddy, 1979).

Although the design of secondary clarifiers without settling tests is not recommended by Metcalf and Eddy (1979), there may be situations where engineers must rely on arbitrary overflow and solids loading rates. The overflow rates given in Table 2.3 are typical values used in the design of activated sludge systems (Metcalf and Eddy, 1979).

(ii) Upflow Clarifiers

The first upflow clarifiers were constructed in Germany and were referred to as Dortmund Tanks. As shown in Figure 2.9, the Dortmund Tank resembles a funnel-shaped tank with vertical upflow. The influent pipe extends below the sludge blanket. The mixed liquor then passes through the fluidized sludge bed to take advantage of bioflocculation enhancement and the removal of very fine suspended solids (Resch, 1980).

The fluidized bed is located between the thickening zone

TABLE 2.3

TYPICAL DESIGN INFORMATION FOR SECONDARY CLARIFIERS

	Overflow Rate $m^3/m^2.d$		Loading $kg/m^2.h$		Depth m
Type of Treatment	Average	Peak	Average	Peak	
Settling following air-activated sludge (excluding extended aeration)	16-32	40-48	3.0-6.0	9.0	3.5-5.0

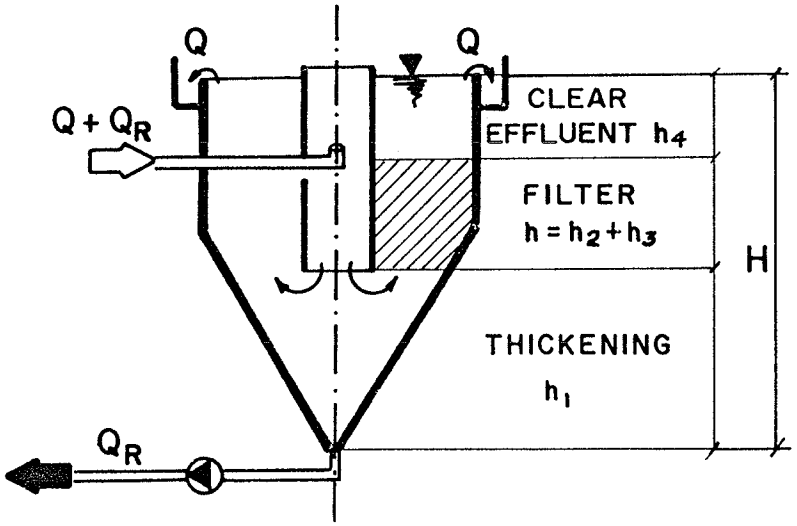


Figure 2.9 Dortmund Tanks with Vertical upflow (after Resch, 1980)

(h_1) and the clear water zone (h_4). The solids-separation phase (h_2) and the storage of activated sludge (h_3) occur in the fluidized bed region (h_2+h_3). The fluidized bed is characterized by a constant

distribution of S.S. and by a horizontal sludge blanket level. The fluidization of a sludge layer in an upflow mode is equal to the falling velocity of the suspension at the same concentration (Resch, 1980).

The required depth for secondary clarification in a Dortmund Tank may be determined by calculating the different layered heights:

h_1 = thickening zone

h_2 = separation zone

h_3 = storage zone

h_4 = clear water zone

Kalbskopf (1974) indicated the following formulas are recommended by the ATV Committee in the Federal Repulic of Germany.

TABLE 2.4

ATV COMMITTEE DESIGN GUIDELINES FOR UPFLOW CLARIFIERS

h_1 = MLSS x SV1 /1,000 . (m)
h_2 = 0.80 - 1.0 m (h_2 < .5 m if h_4 > 1.0 m)
h_3 = ΔMLSS x $\forall$ x SV1 / (500 x A) where: ΔMLSS = the difference between the solids content during dry and wet weather and must not exceed 1.3 g/l $\forall$ = volume of aeration basin (m^3) A = surface area of clarifier (m^2) h_4 = 0.5 m (a minimum height of 0.5 m is recommended)

As shown in the above equations, SVI is equally as important as solids loading. Pflanz (1969) indicated that both of these parameters

determine the required storage volume in a secondary clarifier. It is, therefore, important in designing Dortmund Tanks, to consider depth in conjunction with the surface area.

Pflanz (1969) introduced the comparative sludge volume, VS_V, which is the product of SVI and the MLSS in the aeration basin. In Germany, the ATV Committee on guidelines for "Settling Methods" recommended that horizontal flow rates should be a function of VS_V according to Figure 2.10.

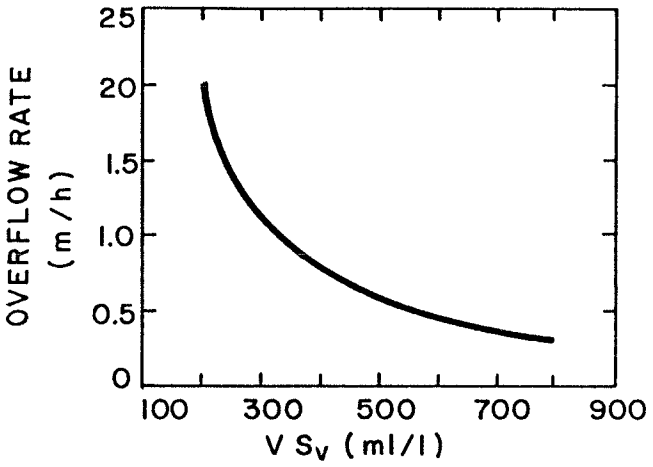


Figure 2.10 Overflow rate as a function of comparative sludge volume (after Kalbskopf, 1974)

It is interesting to note that in the case of predominantly vertical flow secondary clarifiers with fluidized blankets, the overflow rate can be set at 30% higher for the same VS_V value. Upon evaluating several full-scale upflow clarifiers in Germany, Resch (1980) concluded that even higher overflow rates could be achieved in upflow clarifiers (that is, greater than 30%).

Pflanz (1969) also introduced the concept of "solids surface feed", which is defined as the volume of activated sludge which can

be introduced to the surface of secondary clarifiers for a specified HRT value. The parameter VS_f is mathematically defined as follows:

$$VS_f = \frac{\Psi}{HRT} \times VS_v \ (L/m^2.h) \qquad (6)$$

= overflow rate x comparative sludge volume

Pflanz (1969) and Resch (1980) have shown in their studies that, with upflow clarifiers incorporating a sludge blanket, there was no deterioration in effluent quality as a result of increasing VS_f and/or the sludge recycle rate. This was the case as long as there was sufficient depth in the clarifier to prevent the sludge blanket level from reaching the outlet weirs. This was contrary to observations for horizontal flow clarifiers. Table 2.5 presents the recommended solids loadings to achieve a SESS of 30 mg/L at various SVI values (Ganczarczyk, 1983).

TABLE 2.5

RECOMMENDED SOLIDS LOADINGS AT VARIOUS SVI VALUES

(after Pflanz, 1969)

SVI, ml/g	Solids Loadings per hr	
	lbs per sq. ft	kg per sq. m
100	0.7	3.5
200	0.2-0.3	1.1-1.3
300	0.1-0.2	0.8-1.1

(iii) Solids Contact Clarifier

Upflow solids contact clarifiers have been extensively applied in the treatment of potable water since the early 1930's (Metcalf and

Eddy, 1930). The solids contact clarifier provides chemical coagulation, flocculation, and sedimentation in a single tank. In the process, treated water flows upward through a sludge blanket which is comprised of flocculated material.

Recently, solids contact units have been used in advanced wastewater treatment for the chemical treatment of industrial and municipal wastewaters. There have been, however, reported difficulties in maintaining the sludge blanket (Culp et al., 1978). As a result, a good quality effluent could not be achieved with any consistency; thus limiting the degree of reliability associated with their use.

Fisher and Hillman (1940) and Camp (1953) had recognized and emphasized the need for a preflocculation step prior to sedimentation. Camp (1953) recommended that the flocculation chambers be constructed integrally with the clarifier to minimize the possibility of floc destruction. Further work by Parker et al. (1971) supported the concept of physical conditioning of sludge prior to sedimentation. The author concluded that the opportunities for floc formation in conventional clarifiers were limited.

Recently, Norris et al. (1982) used a modified concept of a solids-contact clarifier (herein referred to as SCC) to treat a trickling filter effluent. The SCC unit was a modified version of a circular clarifier used extensively in the activated sludge process. The SCC unit, however, included an added center well which provided a mildly stirred environment for intimate contact between the flocculated particles. In contrast to conventional upflow solids contact units, the SCC units were operated at minimal sludge blanket levels. Solids contact was practised in the flocculator center well rather than in

the sludge blanket. Norris et al. (1982) concluded from the studies that it was possible to achieve consistent advanced wastewater treatment efficiencies on the SCC units (i.e. BOD_5 and SESS concentrations less than 10 mg/L). A schematic of the SCC unit is shown in Figure 2.11.

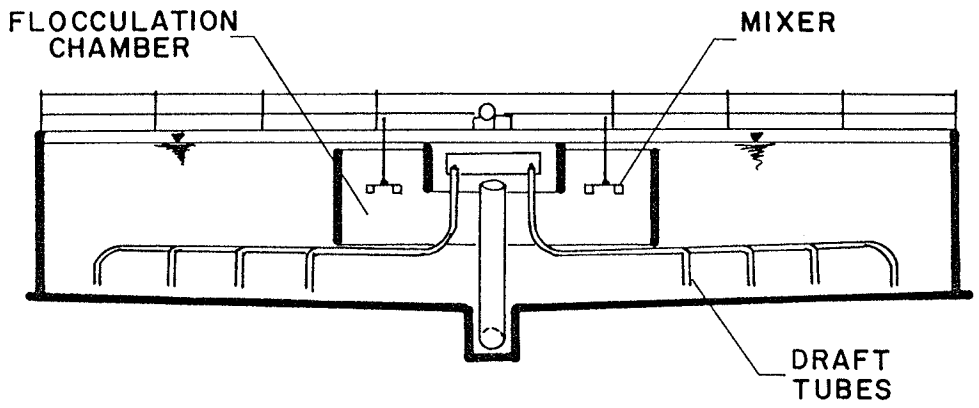


Figure 2.11 Schematic of a modified solids-contact clarifier (after Parker, 1983)

The SCC units were 35 m in diameter, with a 5.5 m sidewater depth and featured a center well with a HRT of 20 minutes.

2.3.2 The Development of a New Design Model

The development of a new design model involves a two-step approach. Firstly, existing theories and established concepts are reviewed. Secondly, the concepts which have resulted in substantial improvements are incorporated into the new design model. It is hypothesized that, through the iteration of optimizing established concepts and existing theories, one may approach an ideal design model such as proposed by Camp (1953) for an "ideal sedimentation basin".

In secondary clarification, the following concepts are sginificant in their application to the design:

(i) Hydraulic Conditions

(ii) Operation Conditions

(iii) Bioflocculation Conditions

An assessment of the aforementioned concepts is presented in the following section. The conditions that will result in substantial improvement of effluent quality are summarized for each concept. It is these conditions which have been used in the development of a new concept of upflow clarification in activated sludge separation.

(i) Hydraulic Conditions

The importance of overflow rate, clarifier depth, uniform flow distribution, and weir placement are endemic, if not ubiquitous, in the literature. It is evident that the modern approach to secondary clarifier design is dominated by the consideration of overflow rate (W.P.C.F., 1976). In horizontal flow clarifier design theory, SESS has been shown to be directly related to the overflow rate (Parker, 1983; Crosby and Bender, 1980). There was no deterioration in effluent quality with a corresponding increase in overflow rates, however, for upflow clarifiers (Resch, 1980). As a result, higher overflow rates are permitted in the Federal Republic of Germany for the design of upflow clarifiers. This observation, however, was only valid as long as the sludge blanket did not reach the outlet weirs. Consequently, clarifier depth is an important consideration in the design of upflow clarifiers.

Parker (1983) has indicated that there was also a trend towards utilizing deeper tanks in horizontal flow clarifiers. This result is based on the realization by designers that the distance from the sludge blanket level to the outlet is directly related to effluent

quality.

Camp (1953) had rationalized that if one could obtain plug flow in a clarifier, there would be little need for the use of inlet and outlet baffles. The phenomena of density currents, short-circuiting and turbulence have been well documented, which explains the difficulty of implementing ideal conceptual theory in practical design.

Contemporary clarifier design suggests that inlet velocities should be minimized, flows distributed equally in the outlet zone, and short-circuiting prevented. These guidelines have resulted in the use of solid skirted baffles in most designs to direct the influent flows towards the bottom of the clarifier. It has now been observed that the high density mixed liquor, when introduced in this manner, flows along the sludge blanket until it strikes the far wall and flows toward the effluent weir (Katz et al., 1962; Parker, 1983; Crosby and Bender, 1980). Typical results also indicate that dead zones are developed by the baffle and good vertical distribution is prevented. To reduce this problem, the authors recommenced placing effluent weirs, 2/3 to 3/4 of the radial distance from the center of a typical center feed circular clarifier. It should be noted, however, that this recommendation does not resolve the problem. By any reasonable estimate, horizontal flow clarifiers are turbulent, not laminar. With conventional design, one is confronted with eddys which move particles upward while horizontal transport is occurring. Rebhun and Argaman (1965) conducted experiments using tracers in a horizontal clarifier with inserted baffles at the inlet and outlet, and concluded that the effective flow consists of 70% mixing and 30% plug flow.

Silveston et al. (1980) introduced the concept of flow contraction at the outlet. Their studies indicated that clarifier performance could be improved by the suppression of internal waves or vertical currents that may propagate back from the outlet weir.

Ives (1968), who mathematically characterized orthokinetic flocculation, introduced a new concept to minimize the effects of turbulence. The author proved that turbulence energy diminished rapidly as the flow passed upward through a sludge blanket. This was based on the theory that the sludge blanket could be regarded as an infinite series of flocculators, each δL thick (where L is the length of the sludge blanket). In the analysis, Ives emphasized that the inlet flow must be uniformly distributed across the inlet plan section and collection must be uniform at the outlet section.

(ii) Operational Conditions

Several operating conditions that influence effluent quality include: MLSS concentrations, solids loading, sludge blanket levels, and the method of sludge removal.

Norris et al. (1982) concluded that SESS levels in the effluent increase with a rise in MLSS concentration and/or solids loading. It was suggested that to minimize solids loading, one could convert an overloaded system to contact stabilization, step and tapered aeration processes (see Table 2.2 page 17).

In separating solids consisting of trickling filter effluent and return sludge, Norris et al. (1982) observed that SESS was at a minimum when MLSS was between 500 and 1500 mg/l. At higher MLSS concentrations, the solids removal efficiencies remained constant, resulting in higher SESS concentrations.

Bush (1983) suggested that it may be possible to lower MLSS concentrations and still maintain effluent quality (i.e. substrate concentration). Tuntoolavest and Grady, Jr. (1982) had shown that one way to improve effluent quality was to reduce the MLSS concentration by controlling the operating conditions in the aeration basin.

Pflanz (1968) and Munch et al. (1978) illustrated that solids loading is dependent on the overflow rate. It is rational to assume that the operation of a wastewater treatment plant using low MLSS concentrations and/or solids loading is desirable in both upflow and horizontal-type flow clarifiers.

The sludge blanket level and the method of sludge removal has a pronounced effect on the performance of a clarifier (Ghobrial, 1978). Parker (1983) suggested that the sludge blanket level be operated at a minimum level during average flow conditions. The author is in agreement with this since this not only decreases the possibility of high SESS but also minimizes the deterioration of sludge quality due to anoxic conditions. The minimum level in the new design model could be designed on the requirement to dissipate turbulent energy.

The depth of the clarifier should be based on the height of the sludge blanket resulting from an increase in solids during peak flow conditions. As suggested by Pflanz (1968), some consideration should be given to the SVI of the particular sludge.

The use of higher MLSS (example, pure oxygen systems) has resulted in the development of extensive research in solids flux theory. An even distribution of sludge removal is essential when considering the applicability of solids flux theory to design. Munch and Fitzpatrick (1978) postulated that some factor in the method of sludge removal

may be responsible for their observation that the actual solids flux capacity may fall below theoretical values predicted from batch settling volume tests. Mechanical scrapers and rotating hydraulic suction removal devices are commonly used in conventional clarifiers for sludge removal. The former option is considered to be less suitable since the sludge has to be moved from one blade to another in the raking mechanism. Katz et al. (1962) reported that critical velocities by these mechanisms may have shearing effects, resulting in resuspension of the sludge and causing it to flow as a thixotropic liquid to the removal point.

It appears that rotating hydraulic suction removal devices are more advantageous than scraping removal, as the sludge is removed more quickly. Data collected by Crosby and Bender (1980), however, indicated that commonly specified rotational speeds are excessive, resulting in solids overflow.

Boyle (1981) reported that an even distribution of sludge removal across the entire floor surface area is critical for improved clarifier performance. It is also an important consideration when using solids flux theory as a design parameter. Ironically, there is no sludge removal mechanism available which utilizes the entire floor area for the removal of solids.

(iii) Bioflocculation Considerations

Camp (1953) recognized the need for preflocculation and indicated that, to minimize floc breakup, the flocculation chambers should be constructed integrally with the clarifier. Several researchers have reported that gentle mixing and/or small velocity gradients improved effluent quality by enhancing flocculation through inter-particle collision (Katz et al., 1962; Dick, 1970; Parker et al., 1971; Norris

et al., 1982; and Parker, 1983).

Velocity gradients in the order of 2.5 sec^{-1} to 10.5 sec^{-1} were observed as suitable values to enhance flocculation (Ives, 1968; Parker et al., 1971). Further studies conducted by Japanese researchers indicated that flocculation in an upflow sludge blanket was enhanced at upflow velocities of 11 m/h to 15 m/h (Ives, 1968). There is substantial evidence that correlates detention time with flocculation enhancement (Ives, 1968). Norris et al. (1982) utilized a 20 minute residence time in the flocculation center well of their SCC unit.

Using the concept of stratified flow theory, the rate of collisions between particles, bioflocculation can be enhanced by the use of pulsating forces. Silveston et al. (1980) described this phenomenon by considering an element of fluid in a clarifier vessel flowing downward. If a random disturbance forces this fluid vertically upward, it enters into a region of less dense fluid. Subsequently, the fluid element, once overcoming the buoyancy forces, is accelerated downward by gravity, and will move past its original position into a region of higher density. Using the method of pulsating forces, one is actually forcing the particles into regions of higher density, hence, promoting inter-particle collison and enhancing the oportunity for bioflocculation.

Parker et al. (1971) proposed placing a mildly stirred flocculation step between the aeration basin and the secondary clarifier to promote bioflocculation of fine suspended solids. The mildly stirred flocculation basin could be modified as a downward flow channel (example, pipe) utilizing the concept of pulsating forces.

(iv) Summary of Hydraulic, Operational and Bioflocculation Conditions

A summary of the hydraulic, operational and bioflocculation conditions that should result in improved clarifier performance is shown in Table 2.6.

TABLE 2.6

A SUMMARY OF HYDRAULIC, OPERATIONAL AND BIOFLOCCULATION CONDITIONS APPLIED TO NEW DESIGN MODELS

CONDITIONS	DESIGN FEATURE
(A) HYDRAULIC CONDITIONS	
1. Less influence of overflow rate	1. Upflow mode
2. Prevent solids overflow	2. Deep clarifier (5.5 m)
3. Avoid density currents	3. Mixed liquor introduced at clarifier bottom
4. Minimize turbulence	4. Mixed liquor distributed uniformly at inlet and through sludge blanket
5. To reduce vertical currents	5. Flow contraction device at outlet
6. To prevent shortcircuiting	6. Uniform flow distribution at inlet
(B) OPERATIONAL CONDITIONS	
7. Reduce solids loading to clarifier	7. Operate at lower MLSS (1500-2000 mg/L)
8. Lower SESS concentrations	8. a) Low sludge blanket level during Q_{avg} (minimize turbulence) b) High sludge blanket level during Q_{peak} (storage)
(C) BIOFLOCCULATION CONDITIONS	
9. Uniform sludge removal	9. Sludge removal mechanism to remove sludge over entire floor area
10. Flocculation step between aeration basin and clarifier	10. Downward flow flocculation chamber constructed integrally with clarifier
11. Gentle mixing	11. Pulsating forces
12. Residence time in flocculation chamber	12. Varies (10-20 minutes)
13. Downward velocity	13. Varies (11 m/h-15 m/h)

A schematic of the proposed model incorporating the hydraulic, operating and bioflocculation design features summarized in Table 2.6 is shown on Figure 2.12.

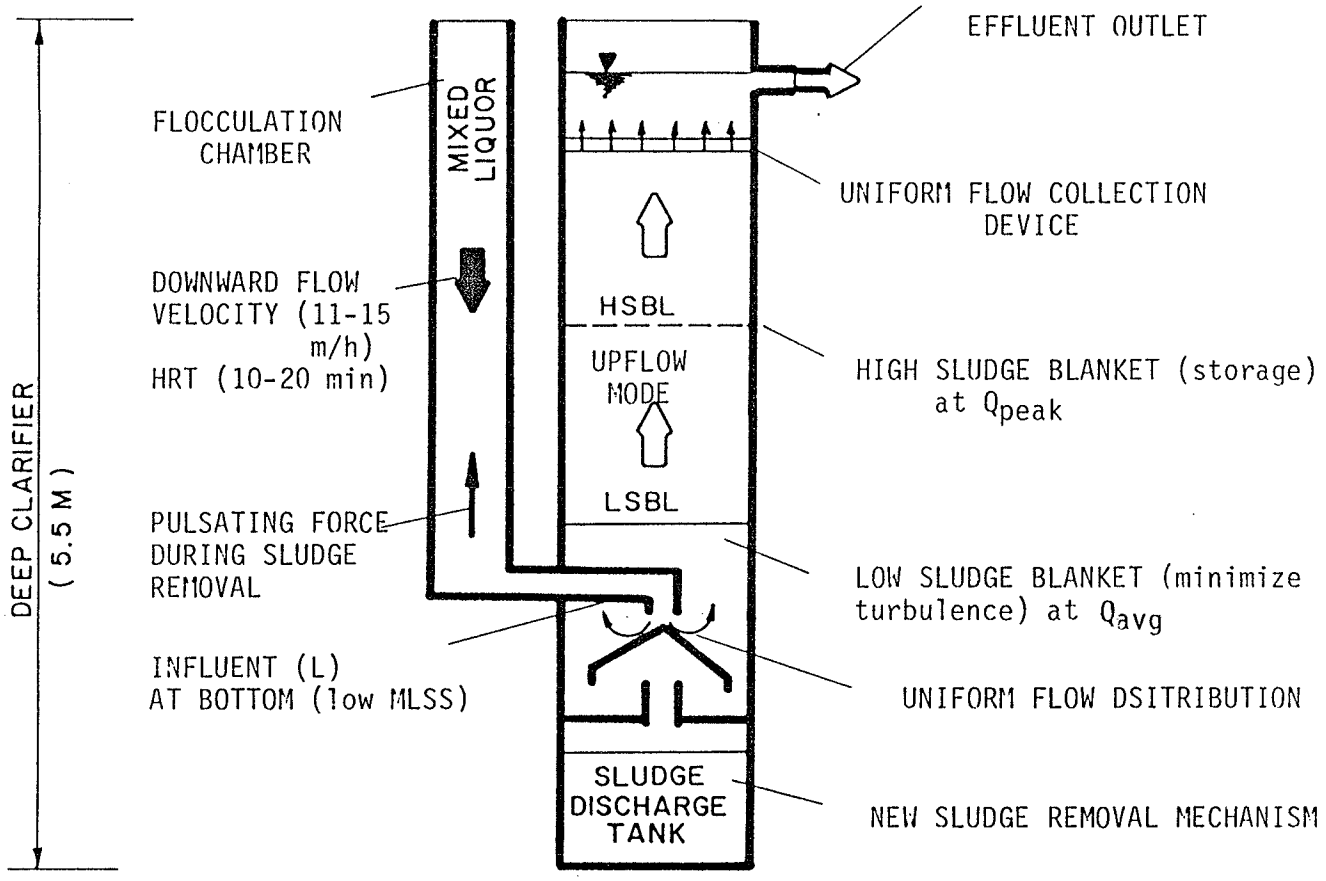


Figure 2.12 Schematic of proposed design model incorporating hydraulic, operating and bioflocculation concepts

2.4 CONTINUOUS FLUX UPFLOW CLARIFIER

2.4.1 Description and Operating Characteristics

The continuous flux upflow clarifier was developed by C.D. Hughes of Cowatt Systems Limited, Manitoba. A laboratory prototype of the continuous upflow clarifier (herein referred to as CFLUC) was

constructed by Dr. A.B. Sparling at the University of Manitoba. A description and operating characteristics of CFLUC is addressed in this section. A schematic of the upflow clarifier model is shown on Figure 2.13.

CFLUC consists of two basic components: the upflow clarifier, and the sludge removal mechanism. The clarifier component consists of the flocculation pipe, pressure lock cone, clarifier basin, filter deck, overflow weir, air supply, and air equalization tube. The sludge removal mechanism consists of the sludge discharge tube, sludge discharge tank, and the sludge effluent pipe.

The operation of the clarifier component is described as follows. Mixed liquor flows down the flocculation pipe and enters the clarifier at the bottom. The outlet of the flocculation pipe is located directly above the pressure lock cone. The pressure lock cone serves two functions. Firstly, the cone distributes the mixed liquor uniformly as it flows upward through the clarifier. Secondly, the mixed liquor is prevented from flowing into the sludge discharge tank by an air pressure lock located under the cone and in the sludge discharge tank. As the mixed liquor flows upward, flocculated sludge particles with a density greater than water will settle downward. A sludge blanket interface will form, with the clarified effluent leaving the unit via the overflow weir.

The formation of a sludge blanket is illustrated in Figure 2.14. The clarifier is normally referred to as operating in the clarification mode. During the clarification mode, the air supply valve is open and the air release valve is closed. The air pressure head in the sludge discharge tank is in equilibrium with the head of

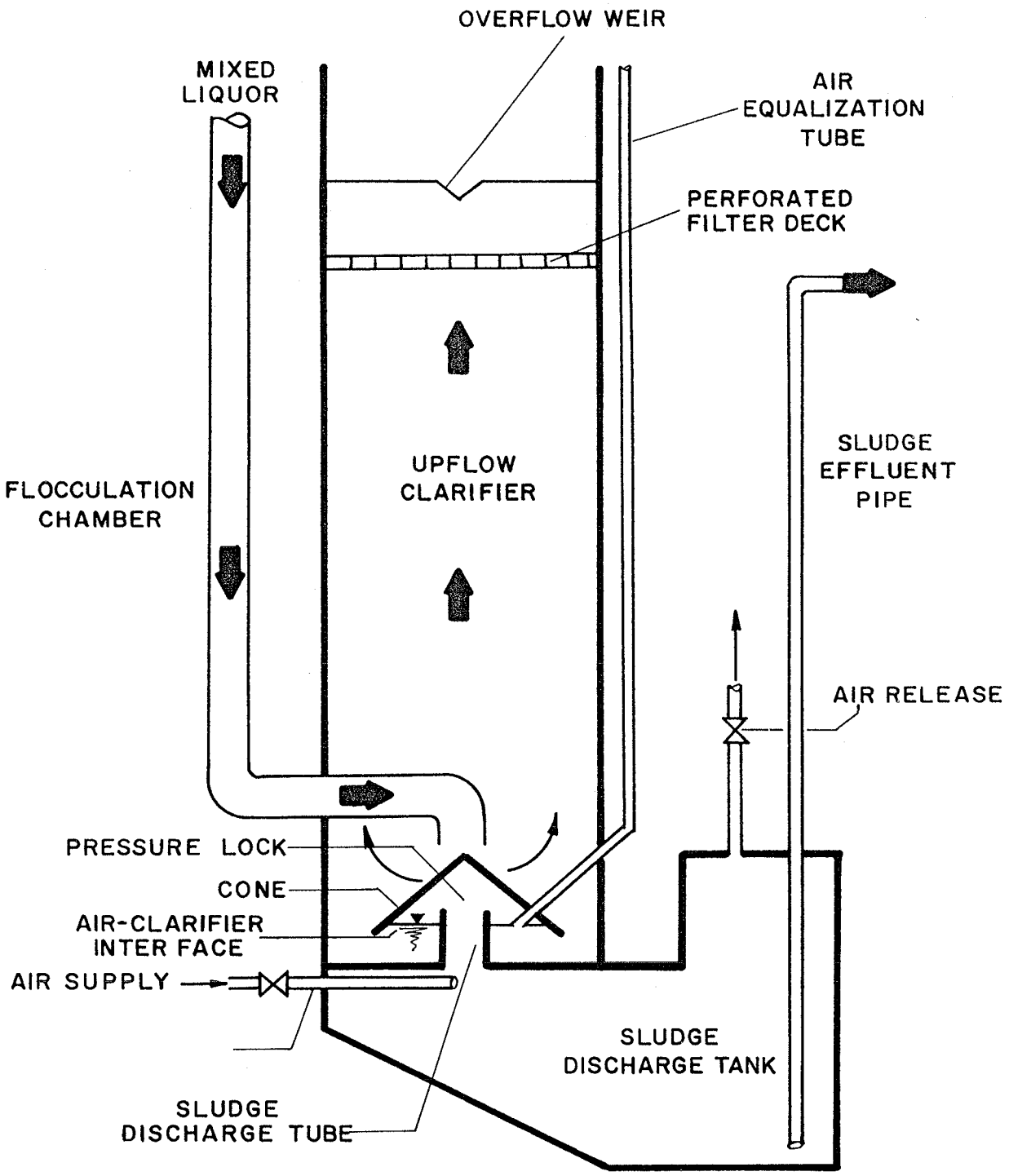


Figure 2.13 Schematic of continuous upflow clarifier system (after Sparling, 1983)

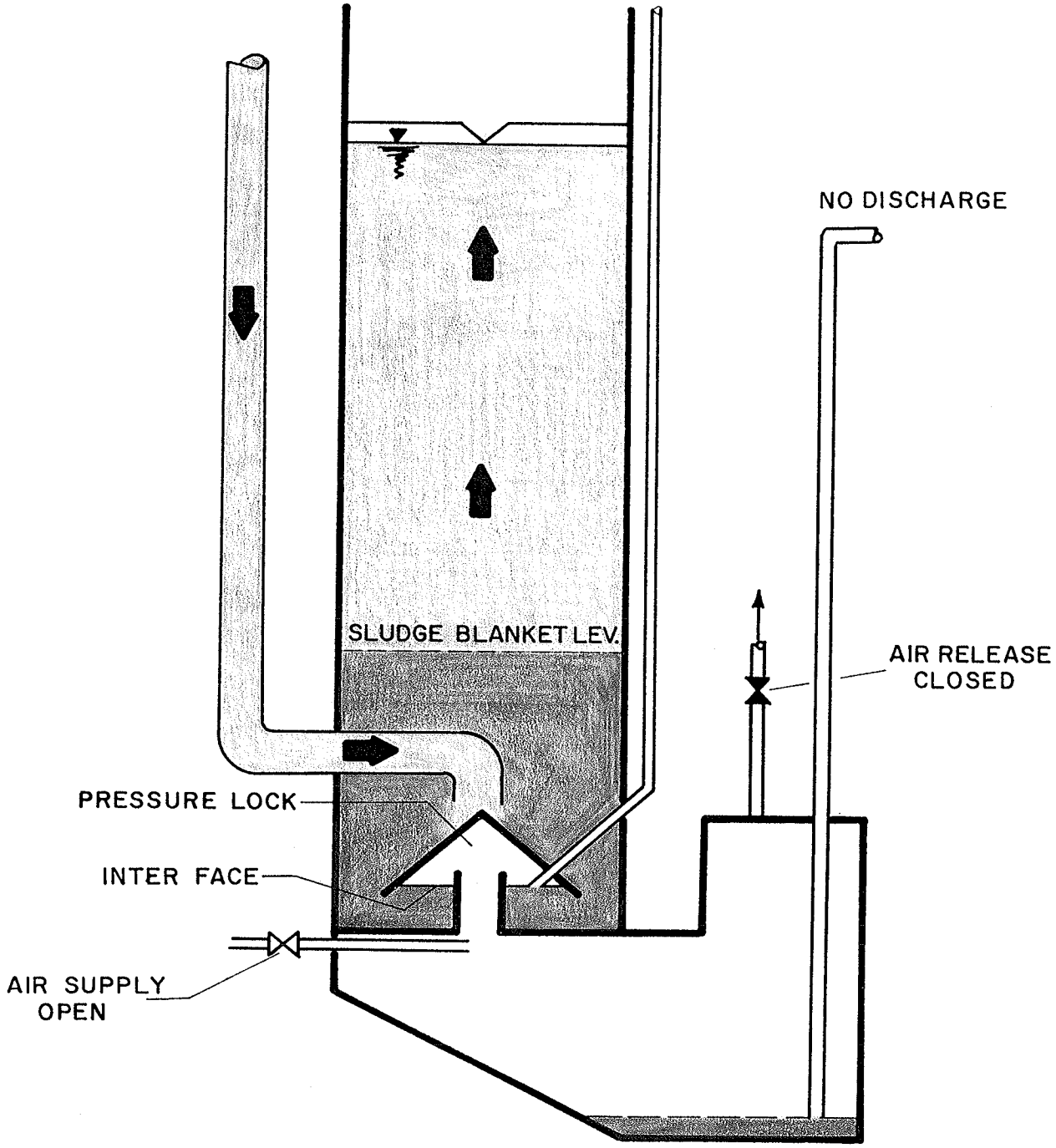


Figure 2.14 Schematic of CFLUC system operating in the clarification mode

water (and friction losses) in the clarifier vessel. This is accomplished through the use of air equalization tubes, whereby excess air escapes from beneath the cone through the equalization tube, thus maintaining a constant air-clarifier interface.

Periodically, activated sludge has to be removed from the clarifier to be either returned to the aeration basin or wasted from the system. To remove solids from the clarifier, the air supply valve is closed and the air release valve is opened. As shown in Figure 2.15 there is a drop in the sludge blanket level as the accummulated sludge flows under the cone, in a pulsated vacuum-type fashion, and into the sludge discharge tank. The rate of discharge is controlled by the air release rate, and the sludge volume to be removed by the duration of the air release. Sludge removal is usually accomplished with the use of timers and/or float switches.

Once the required volume of sludge to be removed is obtained, the air supply valve is reopened and the air release valve is closed. During the sludge removal process, the sludge blanket level and the liquid level will drop. Mixed liquor, however, flows continuously during sludge removal and, as a result, the liquid level rises in the clarifier to the overflow weir position after the sludge removal phase. As shown in Figure 2.16, the rise in static head and the new air supplied to the sludge discharge tank forces the slurry of solids up and out the sludge effluent pipe.

Once the sludge level drops in the sludge discharge tank to a controlled level, the sludge discharge through the sludge effleunt pipe ceases, and sludge removal is complete. As noted, we are back to the original clarification mode of operation.

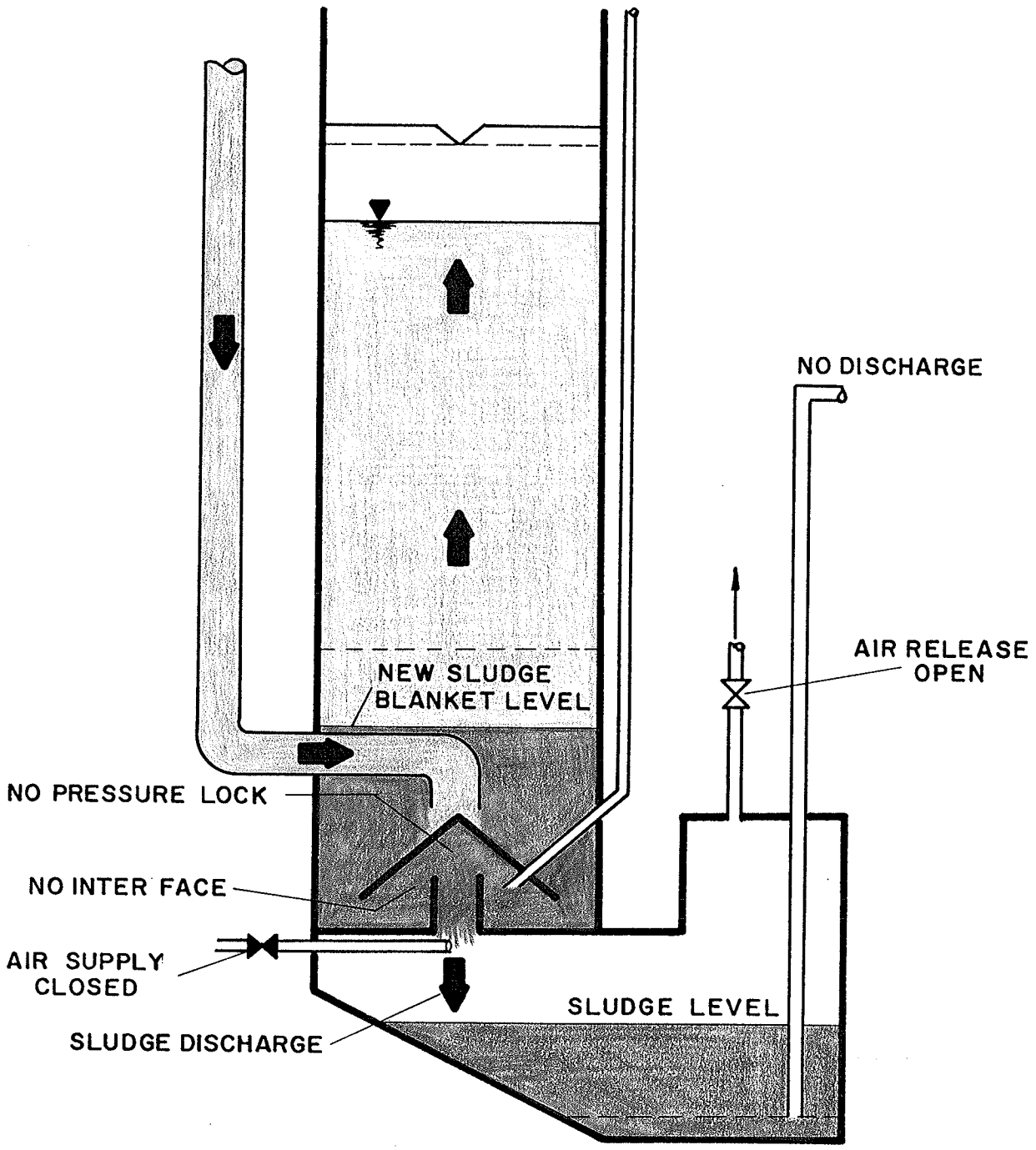


Figure 2.15 Sludge removal from clarifier component to sludge discharge tank

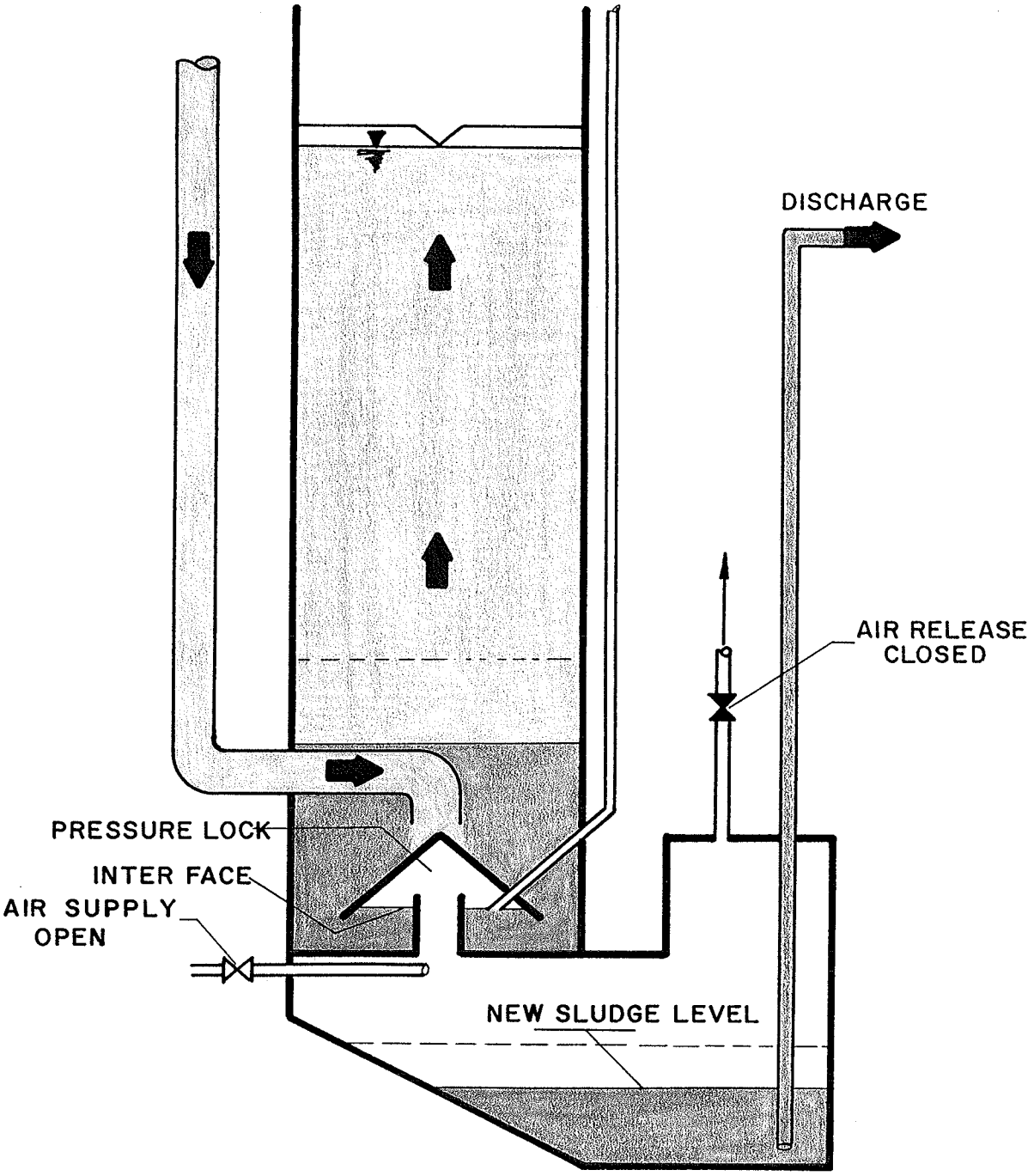


Figure 2.16 Sludge discharge mode in CFLUC System

The pulsed vacuum-type sludge withdrawal, which occurs in a matter of seconds, insures uniform sludge removal across the entire floor area of the clarifier bottom. The sludge blanket drops uniformly during solids removal without significant turbulence. This method of withdrawal has dual benefits of cleaning the filter deck in a backwash fashion, while enhancing bioflocculation in the sludge blanket and the flocculation chamber through the generation of pulsating forces.

2.4.2 CFLUC: Full Scale Application

Two CFLUC clarifiers, each 6.0 m x 8.0 m x 3.0 m (l x w x d), were constructed as part of a wastewater treatment facility for the Town of Souris, Manitoba. As shown in Figure 1.1, air is supplied to the clarifiers by 2 compressors with a rated output capacity of 1190 m^3/h each. The design flow and overflow rate through the clarifiers are 920 m^3/d and 19 $m^3/m^2.d$.

A cross-section of the secondary clarifiers is shown in Figure 2.17. The most noticeable feature of the clarifier is the multilplicity of cones. In each clarifier, there are twelve cones which are constructed of steel. Mixed liquor flows downward, through the flocculation chamber, and is transported by a series of plastic pipes to each cone. The cones are positoned to ensure uniform and complete solids removal.

The mixed liquor then flows through the rectangular filter deck consisting of perforated plastic sheets. The filter deck promotes uniform flow distribution at the outlet and reduces the effects of internal waves propagating back from the outlet weir to the sludge blanket.

The remaining design and operating features of the full-scale

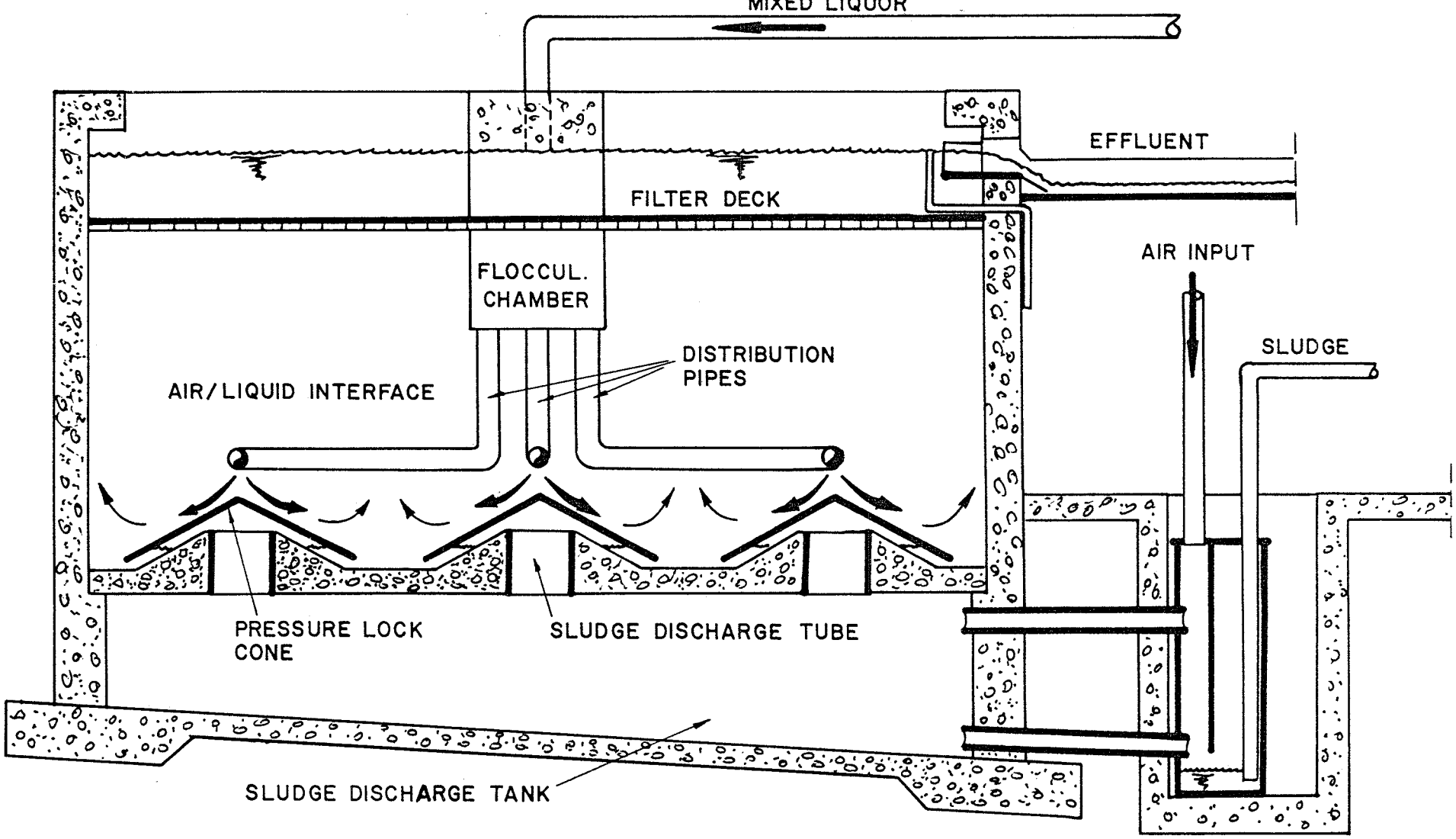


Figure 2.17 Cross-Section of upflow clarifier at Souris, Manitoba (after Poetker, 1984)

clarifier have been previously addressed in Section 2.4.1. The unique features of CFLUC are summarized in Table 2.7 (Sparling, 1984).

TABLE 2.7

FEATURES OF CONTINUOUS FLUX UPFLOW CLARIFIER

FEATURE	CONCEPT
1. Air pressure lock cone	1. Uniform inlet flow distribution of mixed liquor, avoiding the generation of mixed currents and short circuiting.
2. Sludge blanket	2. Minimize turbulence and enhance solids removal through the mechanism of flocculation and entrapment.
3. Pulsed sludge withdrawal	3. Uniform solids removal across clarifier bottom and cleans filter deck.
4. Flocculation chamber	4. Bioflocculation enhancement by pulsating forces (3).
5. Filter deck	5. Uniform outlet flow distribution, reduces propagation of internal waves.
6. Inlet flow	6. Inlet flow introduced at bottom to reduce vertical currents.
7. Upflow mode	7. Less influenced by overflow rate.

CHAPTER 3

EXPERIMENTAL APPROACH

3.1 SCOPE OF STUDY

In accomplishing the objectives outlined in Chapter 1, a laboratory study was conducted over a 135 day period. The selected operating control parameters measured during the experimental program are outlined in Table 3.1. The environmental conditions and operating control parameters were based on similar conditions for a full-scale plant at Souris, Manitoba.

TABLE 3.1

SCOPE OF STUDY - ENVIRONMENTAL CONDITIONS AND OPERATING CONTROL PARAMETERS

CFLUC SYSTEM COMPONENT	ENVIRONMENTAL CONDITIONS		PARAMETERS MONITORED	
1. AERATION BASIN	pH 7.0-7.5	F:M varies	MLSS	INFLUENT, COD
	DO 3.0-5.0 mg/L	SRT varies	MLVSS	INFLUENT, BOD
	MLSS 1500-2000	FLOW 576 L/d	SESS	EFFLUENT, COD
	Temp 20-25°C		SPSS	EFFLUENT, BOD
	HRT 1 day		TS (X_R)	SOLUBLE EFFLUENT
2. CFLUC COMPONENT	pH 7.0-7.5		TVS	SOLUBLE ML
	DO 1.0-2.0		SV1	TURBIDITY
	overflow rate	18.3 $m^3/m^2.d$	pH	DISSOLVED OXYGEN
	sludge wasting	varies	Q	AIRFLOW RATE
	Temp 20-25°C		Q_R	SOLIDS WASTED

The study was conducted to assess the performance of the CFLUC system in regards to activated-solids separation. The new concept

of upflow clarification in sludge separation offers the potential to improve effluent quality while minimizing capital and operating costs (Sparling, 1984).

3.2 MATERIALS AND METHODS

3.2.1 Apparatus

A CFLUC prototype used in previous studies by Sparling (1983) was employed for this study. The laboratory prototype consisted basically of two components; an aeration basin and an upflow clarifier. A sketch of the laboratory unit is shown in Figure 3.1.

The fiberglass aeration basin with an average inside diameter of 1.250 m, height of 0.750 m, held a total liquid volume of 576 L. The outlet of the aeration basin consisted of a 50 mm opening and was shielded with an aluminum baffle to reduce surging effects. Mixing in the aeration basin was provided with a laboratory mixer (Model # 103, T-Line Paddles and Accessories). Air was continuously supplied to the aeration basin from a laboratory air compressor. The air was distributed to four coarse bubble diffusers through a copper manifold. The four air diffusers were positioned at the bottom of the aeration basin (50 mm from basin bottom). The air flow rate was measured through a glass air/gas flow rator (series 10A 3500 - Fisher and Porter). A photograph of the aeration basin and air supply system is shown in Figure 3.2.

The synthetic raw waste entered the top of the aeration basin liquid level through two influent feed lines. Tap water was pumped from a continuously filling water reservoir through one of the feed lines. Concentrated influent feed was pumped through the second feed

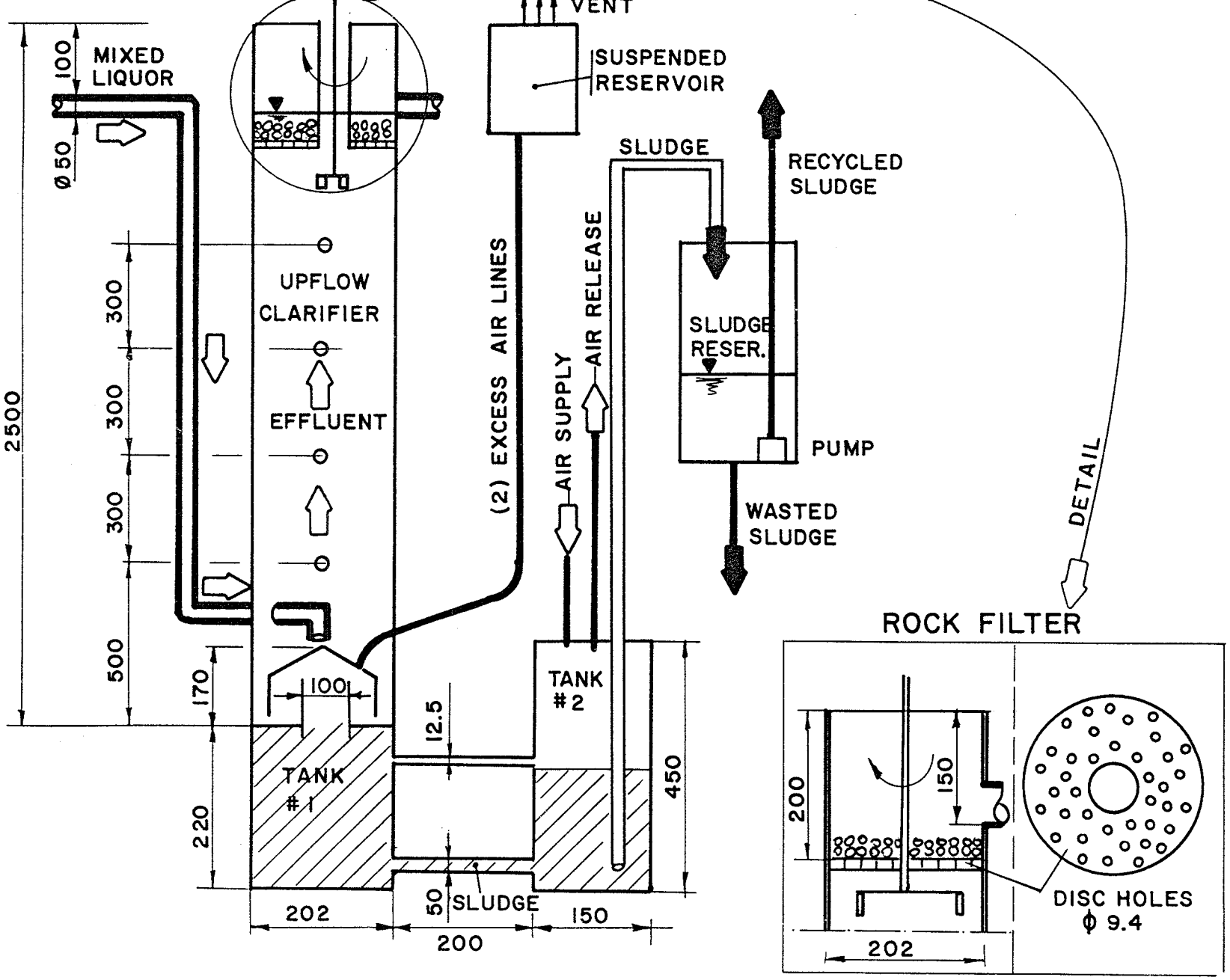


Figure 3.1 Sketch of a laboratory unit for a CFLUC System (N.T.S.)

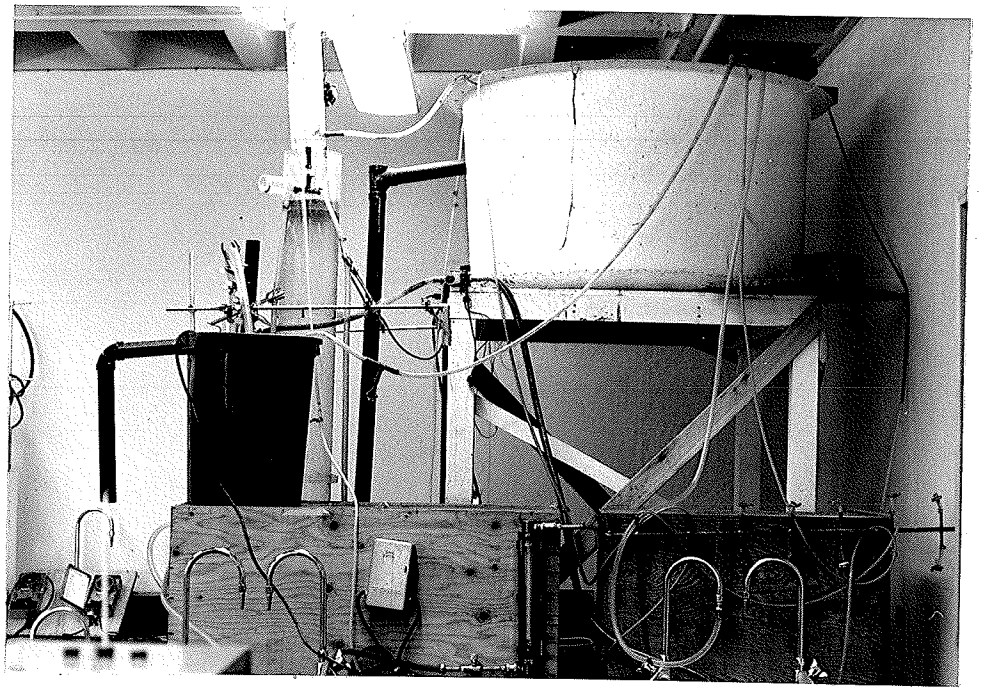


Figure 3.2 Photograph of air supply system and aeration basin

line and mixed with the tap water using a polypropylene tee connection. The concentrated influent storage reservoir was located in a refrigerator to prevent deterioration and souring of the synthetic raw waste.

Pumping was accomplished using a Masterflex® multi-channel variable speed pump (Model 7567, Cole-Parmer Instrument Company). Tygon tubing was used for all gas and liquid transport lines in the experimental unit. Pump membrane tubing was replaced weekly to reduce laboratory errors associated with varying flow rates resulting from worn tubing.

The mixed liquor flowed from the aeration basin through a 50 mm orifice located 80 mm from the top of the aeration basin. The mixed liquor flowed downward for a distance of 2.180 m through a 50 mm A.B.S. pipe. As shown on Figure 3.3, the mixed liquor entered the bottom of the upflow clarifier unit. The plexiglass upflow clarifier section has an outside diameter of 215 mm, inside diameter of 202 mm,

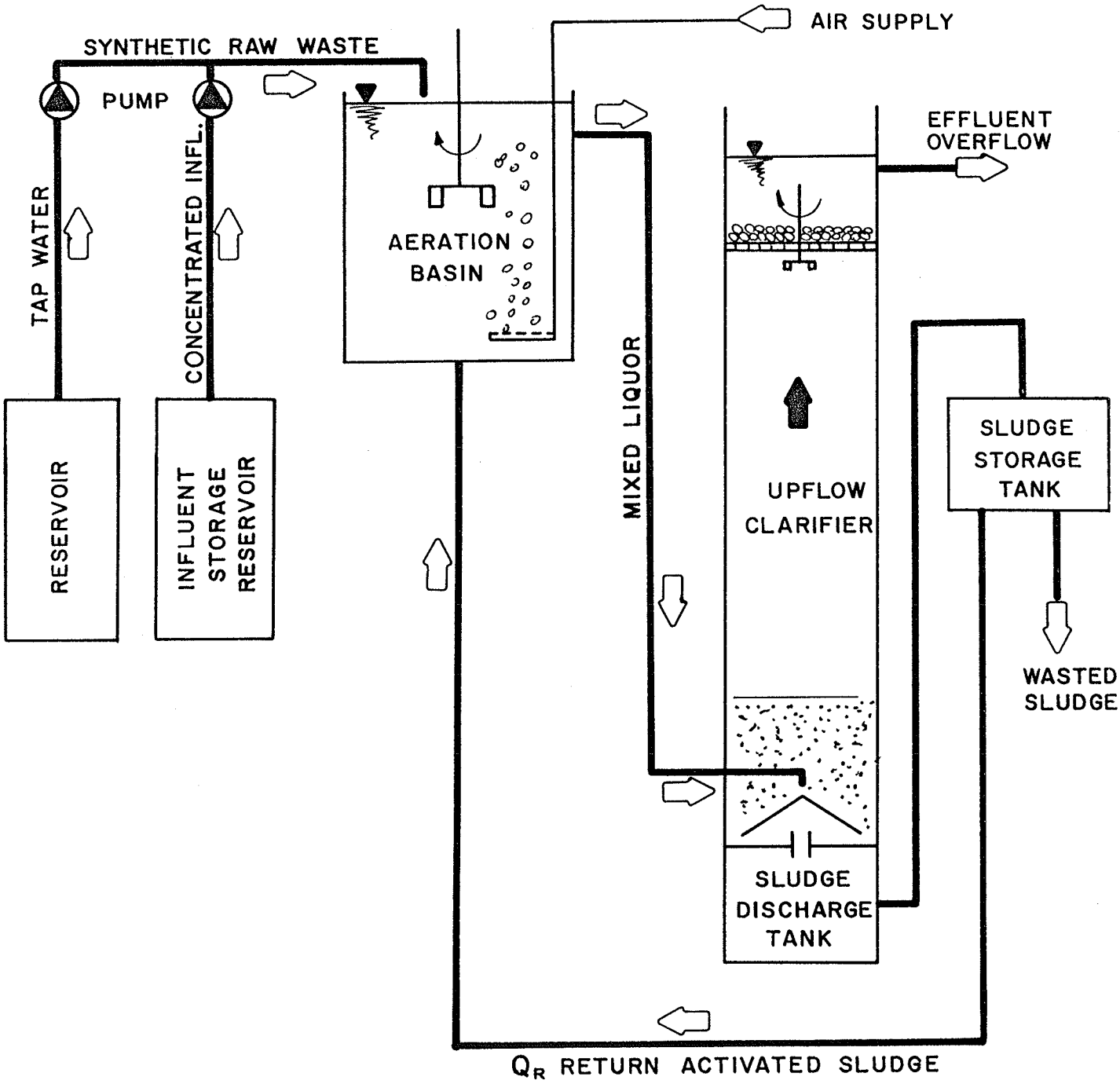


Figure 3.3 Detailed sketch of laboratory upflow clarifier section (N.T.S.)

and a total depth of 2.500 m. Four sampling points (12 mm ID nipples) were provided for testing purposes.

As the mixed liquor flowed upward, the solids were separated and the effluent passed through a rockfilter and out the overflow weir. A cross-section of the rock filter is shown in Figure 3.3. Graded granite rock was placed on a perforated acrylic disc (approximately 50 mm in thickness) to form the filter. The disc, which is 202 mm in diameter and 6.4 mm in depth, has 9 mm diameter holes drilled in at 25.4 mm centre. A tube, 38.1 mm in diameter, was attached to the disc to provide an access opening for a mixer (Model # 102 - T-Line Paddles and Accessories). The mixer was used to disperse floating solids that could accumulate under the filter.

As shown in Figure 3.3, air was supplied to sludge discharge tank # 2. The supply of air passed through an air pressure reducing valve (P_f = 82.8 kPa) and was transported to the sludge discharge tank via 12.7 mm I.D. tygon tubing. A cut-off valve was installed on the air supply line to control the volume of air transported to the tank. As a safety measure, the Masterflex® pump was connected to an air pressure activated switch. In case of a mechanical failure of the air compressor and/or corresponding reduction in air pressure the synthetic raw waste flow would be shut off automatically. Two excess air tubes (25 mm I.D.) were constructed integrally with the cone. To ensure an airtight seal, a silicon rubber compound was applied to all plexiglass-rubber interfaces. The excess air tubes extended upward from the clarifier wall into a suspended reservoir where the air was allowed to escape into the atmosphere.

Sludge was removed from the clarifier into sludge discharge

tank # 1 by the release of air from sludge discharge tank # 2. This was accomplished automatically using an air-gas solenoid valve (50 mm I.D. - 60 cycle - 120 volt, Ascoelectric Limited) and time switches (Model # 60M 800 1, Tork Time Control Inc.). Once the required volume of sludge was removed from the clarifier into tank # 1 and # 2, the air supply valve was opened and the air release valve was closed. The new air supplied to tank # 2 was transferred through a 12.5 mm connecting tube into tank # 1. The new air forced the sludge from tank # 1 into tank # 2, through a connecting tube (50 mm I.D.) located between the two tanks at the bottom of the clarifier. The sludge then flowed from tank # 2, through a sludge effluent pipe into a sludge reservoir. Sludge was manually wasted from the reservoir and/or pumped back to the aeration basin. This was accomplished using a submersible pump (Model # 1-42, Little Giant Pump Company) which was automatically controlled with a timer switch. The cyclic pump rate was selected based on a detention time of less than 30 minutes to avoid anoxic conditions and sludge deterioration in the sludge reservoir. A photograph of the continuously fed CFLUC system is shown in Figure 3.4.

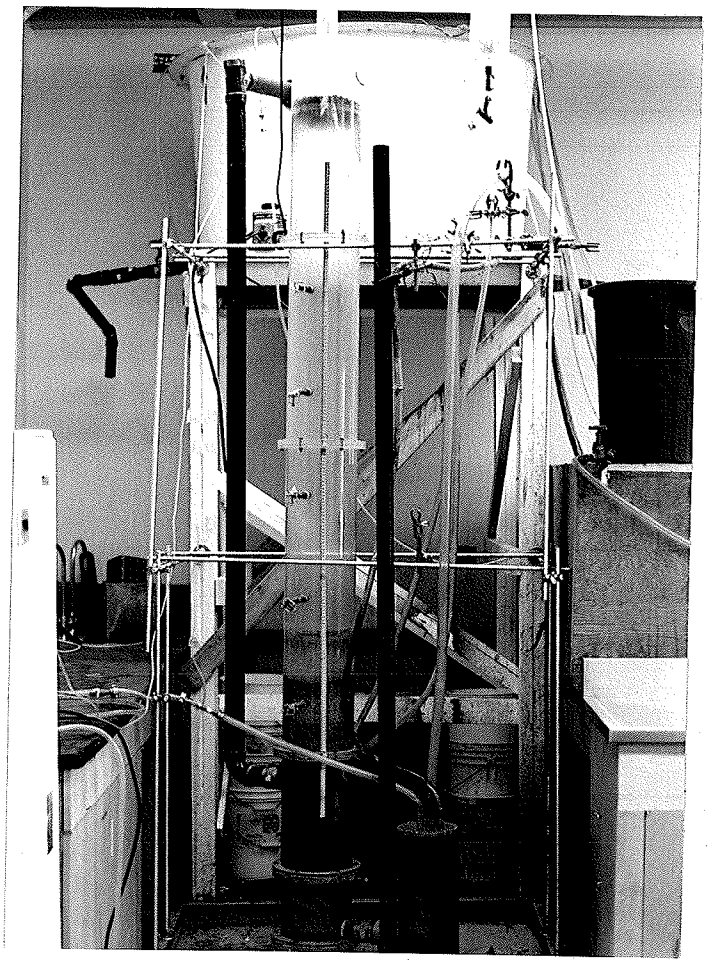


Figure 3.4 Photograph of upflow clarifier and sludge discharge tanks

3.2.2 Experimental Procedure

The experimental program was developed and conducted in two phases (herein referred to as phase A and phase B). Phase A of the experiment was termed the "start-up and acclimation phase". During this phase, the primary objective was to acclimate the seed sludge to the synthetic raw waste. Operating conditions for a specific F:M ratio were monitored during the acclimation phase. The F:M ratio selected during the acclimation period was an average of the range of F:M ratios that the experimental unit would be tested with during

the second phase of the program. A secondary objective was to characterize the synthetic raw waste and define the operating control parameters.

The primary objective during phase B of the experimental program was to evaluate the solids-separation performance and efficiencies of the laboratory unit at four different F:M ratios and/or SRT values. Since the biological operating parameters are directly related to the settling characteristics of the sludge, substrate removal efficiencies were also monitored to assess the biological activity and operating performance of the System. Figure 3.5 illustrates the overall time commitment in carrying out the experimental program. A further description of detailed experimental procedures during phase A and phase B are addressed in this section.

(i) Phase A - Start-up and Acclimation Phase Study

Return sludge from an extended aeration plant in Selkirk, Manitoba was utilized as seed for the laboratory unit during the acclimation phase. The average influent strength of the wastewater to the Selkirk plant was 360 mg/L COD, the HRT, MLSS and F:M ratio for the aforementioned plant was determined to be 19.5 h, 2800 mg/L, and 0.16 d^{-1}, respectively.

The first objective of the acclimation period study was to define the design and operating control parameters of the laboratory unit. These are summarized in Table 3.2

As a means of evaluating flow accuracy, the variance in flow rates (i.e. pumped by the Masterflex® pump) was checked at three different flow rates. The variations were calculated by conducting fifteen measurements with a graduated cylinder over a one hour time

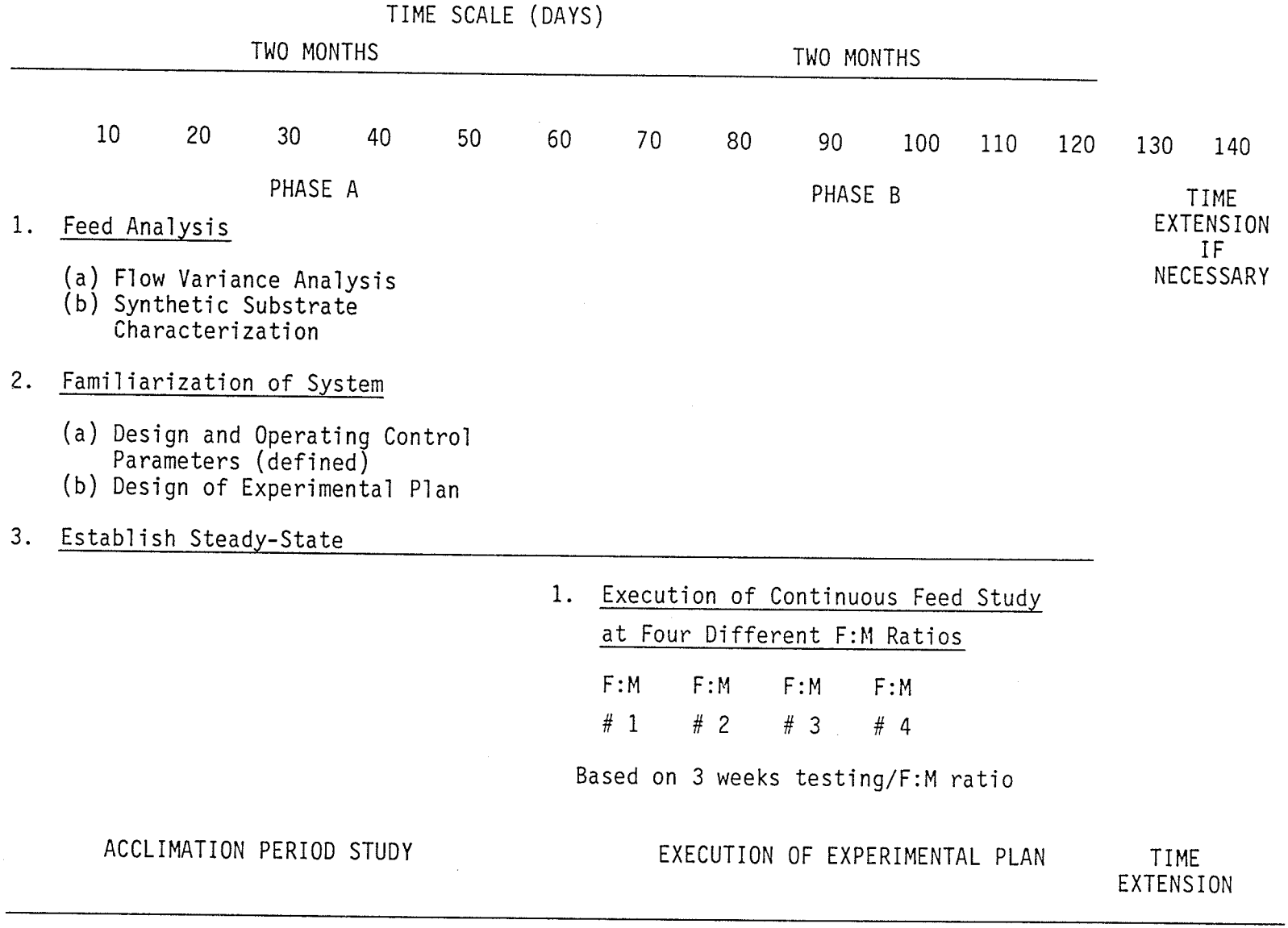


FIGURE 3.5 HISTOGRAM OF EXPERIMENTAL PROGRAM; PHASE A AND PHASE B

TABLE 3.2 DESIGN AND OPERATING CONTROL PARAMETERS FOR THE LABORATORY UNIT

AERATION BASIN	VOLUME Litres	INFLUENT FLOW L/d	INFLUENT FLOW mL/min	HRT h	MLVSS mg/L	So COD mg/L	F:M COD d^{-1}
	576	576	400	24	1500-2500	400	.21

UPFLOW CLARIFIER	AREA m^2	HEIGHT m	OVERFLOW $m^3/m^2.d$	RATE m/h	HRT h	SOLIDS LOADING kg/m^2h
	0.0324	2.33	18.3	0.8	3.1	1.1-1.9

FLOCCULATION CHAMBER	AREA cm^2	HEIGHT m	DOWNWARD cm/min	VELOCITY m/h	HRT min
	20.3	2.18	19.7	11.8	11.0

TABLE 3.3 FLOW VARIATIONS AT DIFFERENT OVERFLOW RATES

Q	OVERFLOW RATE	TEST RUN (One Hour Period)															AVERAGE Q
mL/min^{-1}	$m^3/m^2.day$	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	mL/min
400	18.3	390	390	390	395	400	400	400	410	400	400	400	395	405	410	410	399.7
800	36.6	805	805	805	805	810	810	815	820	820	820	820	820	820	820	820	813.6
1200	54.9	1200	1200	1210	1210	1210	1215	1210	1215	1220	1210	1210	1210	1210	1210	1210	1210

period. The results are tabulated in Table 3.3.

There appeared to be little variance in flow rate at the design flow of 400 mL/min. Since all test runs were conducted at this flow rate, the control accuracy was within 400 ± 2.5 mL/min. This is acceptable. However, to minimize the effects of this variation and possible flow variations due to worn tubing, the flow rates were monitored daily before sampling and adjusted when necessary.

The second objective of the acclimation phase study was to characterize the synthetic raw waste. The synthetic raw waste consisted primarily of milk solids supplemented with additional nutrients. A concentrated stock feed solution was prepared which was fed continuously and mixed with tapwater, to the desired influent raw waste concentration. The concentrated stock feed solution consisted of various constituents to provide the microorganisms with a substrate of organic carbon, nitrogen, phosphorous, miscellaneous nutrients and alkalinity. The organic carbon was provided in the form of skim milk solids (Modern Dairies Brand), beet extract, (Difco Brand), and yeast extract (BBL Brand). The composition of the dry milk solids is illustrated in Table 3.4.

TABLE 3.4

COMPOSITION OF DRY MILK SOLIDS

CONSTITUENT	PERCENT BY WEIGHT	CONCENTRATION OF STOCK FEED g/L
Butterfat	0.8	-
Protein	35.0	-
Carbohydrates	55.0	-
Organic Carbon	-	23.4 O_2
Nitrogen (N)	-	1.33
Phosphorous (PO_4)	-	0.27

Additional inorganic (nitrogen) was provided in the form of ammonium sulphate and inorganic phosphorous in the form of mono and dibasic potassium phosphate. The additional nutrients (excluding milk solids) and their respective concentrations are shown in Table 3.5.

Soda Ash (Na_2CO_3) was added to provide the necessary alkalinity required for pH control. Approximately 80 grams of alkalinity ($CaCO_3$) was added per gram of nitrogen. This quantity was based on alkalinity requirements for nitrification and pH control. The soda ash solution was added directly to the aeration tank so as to avoid chemical precipates in the stock feed solution.

Trace metals were added to the stock feed solution. A concentrated trace metal solution of four litres in volume was prepared. A volume of one mL of trace metal solution was added per litre of stock feed solution. As illustrated in Table 3.6, this resulted in a trace metal concentration of 6.6 mg/L for each trace metal added in the stock

TABLE 3.5

COMPONENTS AND CONCENTRATIONS OF ADDITIONAL NUTRIENTS IN STOCK FEED SOLUTION

COMPONENT	COMPONENT CONCENTRATION g/L	CONCENTRATION OF DESIRED CONSTITUENT g/L
Organic Carbon		
Beef Extract	0.60	3.37 O_2
Yeast Extract	0.15	trace
Inorganic Nutrients		
$(NH_4)_2\ SO_4$	7.86	1.67 NH_4-N
KH_2PO_4	1.45	0.33 PO_4
$K_2H\ (PO_4)$	1.90	0.33 PO_4
$CaCL_2.2H_2O$	0.98	0.27 Ca
$MgSO_4.7H_2O$	2.03	0.20 Mg
$MnSO_4.7H_2O$	0.21	0.20 Mn
Alkalinity		
Na_2CO_3	25.25	23.8 $CaCO_3$

TABLE 3.6

TRACE METAL CONCENTRATIONS IN STOCK FEED SOLUTION

CHEMICAL ADDED	CHEMICAL CONCENTRATION (g/L)	TRACE METAL CONSTITUENT	TRACE METAL CONCENTRATION (g/L)
$FeCL_3$	19.25	Fe	6.6
$MnSO_4.H_2O$	20.57	Mn	6.6
$CoCL_2$	14.62	Co	6.6
$CuSO_4.5H_2O$	25.23	Cu	6.6
$H_3\ BO_3$	36.62	B	6.6
$Zn\ SO_4$	16.50	Zn	6.6
$(NH_4)_6MO_7O_{24}4H_2O$	12.20	M	6.6
$ALCL_3$	33.00	Al	6.6

feed solution.

The stock feed solution was prepared in 20 L volume quantities and stored in a refrigerator at 4°C. During the acclimation period, dilute raw waste feed was prepared in concentrations ranging from 350 to 400 mg/L COD. A dilution factor (P) of 66.67 (6 ml of stock feed per 400 ml of tap water) was utilized. The average COD:N:P ratio of the diluted feed solution was 100:11:3. The concentrations of the constituents of the diluted raw waste feed are shown in Table 3.7.

An analysis of the diluted synthetic raw waste for COD, nitrogen and alkalinity concentrations was conducted to confirm the values shown in Table 3.7. The results of the analysis are tabulated in Table 3.8.

The acclimation phase commenced on August 24, and terminated October 27. A summary of data collected over the 65 day acclimation period study is shown in Table A.1, Appendix A. The operating conditions of the system were stabilized by September 14. The solids retention time during the period September 14 to October 14 was estimated to be 26 days and was based on periodic solids wasting calculations shown in Table A.2, Appendix A. A clarifier upset, attributed to power failure in the laboratory, resulted in the loss of 45% of the solids on October 15. The concentrated feed rate was increased to increase the concentration of bacterial cells in the system. Stability was achieved by October 27. The rapid assimilation of the substrate and cell generation confirmed that the microorganisms were well acclimated to the substrate.

(ii) Phase B - Execution of Experimental Plan

Subsequent to sludge growth and acclimation, the second phase

TABLE 3.7

CONCENTRATIONS OF DILUTED RAW WASTE CONSTITUENTS

CONSTITUENT	CONCENTRATION STOCK FEED (g/L)	DILUTION FACTOR	CONCENTRATION DILUTED RAW WASTE mg/L
COD	26.74	66.7	400
Total Nitrogen	3.0	66.7	45
Organic N	1.33	66.7	20
Inorganic N	1.67	66.7	25
Total Phosphorous	0.93	66.7	14
Organic P	0.27	66.7	4
Inorganic P	0.66	66.7	10
Calcium	0.27	66.7	4
Magnesium	0.20	66.7	3
Manganese	0.20	66.7	3
Yeast Extract	0.15	66.7	trace
Trace Metals	6.6	66.7	0.1
Alkalinity ($CaCO_3$)	23.81	66.7	357

TABLE 3.8 ANALYSIS OF DILUTED SYNTHETIC RAW WASTE

DATE	ACTUAL	NITROGEN			pH	ALKALINITY
	COD mg/L	NH_3-N	ORGANIC N	TKN-N		$CaCO_3$
Sept. 8/84	411.0	-	-	-	-	-
Oct. 4/84	403.0	-	-	-	9.4	-
Oct. 6/84	407.0	-	-	-	9.1	389
Oct. 7/84	-	24.0	22.0	46.0	-	-

of the experimental plan was to evaluate the solids-separation performance and the efficiency of the CFLUC system at various organic loadings. An outline of a tentative experimental design plan is shown in Table 3.9. All testing was conducted at the same HRT.

TABLE 3.9 EXPERIMENTAL DESIGN PLAN: PHASE B

WEEK	TIME PERIOD PERIOD	MLSS mg/L	So(COD) mg/L	F:M (COD) $days^{-1}$	HRT days	RUN #
1	Oct 28-Nov 3	1500-2000	140	0.09-0.12	1.0	B1 AC
2	Nov 4-Nov 10	1500-2000	140	0.09-0.12	1.0	B1 T
3	Nov 11-Nov 17	1500-2000	350	0.22-0.29	1.0	B2 AC
4	Nov 18-Nov 24	1500-2000	350	0.22-0.29	1.0	B2 T
5	Nov 25-Dec 1	1500-2000	875	0.55-0.73	1.0	B3 AC
6	Dec 2-Dec 8	1500-2000	875	0.55-0.73	1.0	B3 T
7	Dec 9-Dec 15	2000-2500	1850	0.93-1.16	1.0	B4 AC
8	Dec 16-Dec 22	2000-2500	1850	0.93-1.16	1.0	B4 T

Previous studies by Klapwijk et al. (1981) for upflow clarifiers indicated that the best results for test data is that the test run be conducted for at least three times the SRT.

Each experiment, however, would take too long so Klapwijk conducted his experiments for at least three times the HRT before collecting data. In this experimental plan, a period of six days or six times the HRT was selected as a guideline for the acclimation periods (Run # B1 AC, B2 AC, B3 AC, and B4 AC) between test runs (Run # B1 T, B2 T, B3 T and B4 T). By monitoring parameters such as MLVSS, SVI and waste removal efficiencies, one could accurately define the time periods

corresponding to the acclimation period and steady state conditions. Adjustments were made to the experimental design plan in test runs where longer or shorter acclimation time periods were required to achieve steady state operating conditions.

Testing for phase B of the experimental study commenced on October 20, 1983 and terminated January 5, 1984. Two additional weeks of acclimation were required in test runs # B2 T and B4 T to achieve steady state conditions.

Over the 70 day experimental study, test data were collected for the four F:M ratios and/or SRT values being investigated. The sampling procedures and analytical measurements for the study are addressed in the following section.

3.3 SAMPLING AND ANALYSIS PROGRAM

A sampling and analysis program was developed for the experimental study and is shown in Table 3.10.

Most analytical measurements were conducted twice daily (09 00 and 16 00 h) to accurately monitor and define environmental conditions in the system. All tests, excluding the COD test, were analyzed immediately, after withdrawal from the aeration basin and the upflow clarifier. Samples for the COD test were acidified with sulphuric acid to pH 2.0. The COD samples were analysed within 48 hours of preservation.

Samples for the SESS test were withdrawn from the overflow weir. On alternate days, samples were also withdrawn at various sample port locations in the clarifier (above the sludge blanket). These samples were analyzed to determine the influence of the rock filter

and the sludge blanket depth on SESS concentrations.

TABLE 3.10 TIME SCHEDULE FOR ANALYTICAL MEASUREMENTS

COMPONENT	ANALYSIS PARAMETER	SAMPLING TIMES DAILY	WEEKLY	ANALYSIS PARAMETER	SAMPLING TIMES DAILY	WEEKLY
AERATION BASIN	Flow Rate	XX*		BOD_5 (Feed)		XX
	pH	XX		COD (Feed)	X	
	DO	XX		COD (Feed) Soluble	X	
	MLSS	XXX		Air Flow	XX	
	MLVSS	XXX		Temperature	XX	
	SVI	XX		Microscopic Examination	X	
UPFLOW CLARIFIER	pH	XX				
	DO	XX		TS Sludge	XX	
	Temperature	XX		TVS Sludge	XX	
	SESS	XX				
	SPSS		XXXX	Sludge Discharge	XX	
	Turbidity (EFF)	XX		Sludge Height	XX	
	BOD_5 (EFF)		XX	Sludge Wasting	varies	
	COD (EFF)	X				
	COD (EFF) Soluble	X				

* NOTE: X denotes the number of times samples taken in a day and/or week.

Sludge samples were obtained during discharge from the sludge effluent pipe and were analyzed for total and total volatile solids. The data was analyzed to assist in the determination of sludge wasting and recycle rates. Periodically, solids were withdrawn from the sample port locations to determine the average solids concentration in the fluidized sludge blanket.

All glassware was washed with hot soapy water and rinsed with tap water. The glassware was then washed with concentrated

chromic-sulphuric acid cleaning solution, rinsed with tap water, distilled water and dried at 103°C in a drying oven.

3.4 ANALYTICAL TECHNIQUE

3.4.1 Flow Rates

The flow rates of the concentrated feed lines and dilution water (tap water) were checked twice daily using a graduated cylinder and a stop watch. Three measurements were taken and the appropriate flow adjustments were made on the Masterflex® multi-channel variable speed pump. A substantial varying flow rate was usually an indication of worn pump membrane tubing.

3.4.2 pH

The pH of the mixed liquor and effluent was determined by the glass electrode method given in Standard Methods (anon., 1980). A Radiometer Type PHM 29 b pH meter was used in conjunction with two electrodes. The pH electode and reference electrode was manufactured by Fisher Scientific. The pH electrode was a universal glass pH electrode (Cat. No. 13-639-3) while the reference electrode was a calomel reverse-sleeve reference electrode (Cat. No. 13-639-61). Prior to pH determinations, the pH meter was calibrated with a set of buffer solutions at the temperature of the samples to be analyzed. The accuracy of the pH meter was established by the manufacturer to be ± 0.03 pH units.

3.4.3 Dissolved Oxygen

The membrane electrode method given in Standard Methods (Anon.,

1980) was used. The dissolved oxygen was measured using a YSI Model 51B oxygen meter and a YSI Model 5739 membrane electrode. The oxygen meter was calibrated against a water sample of known dissolved oxygen concentration, as determined by the modified azide idometric method. The membrane electrode system can obtain an accuracy of ± 0.1 mg/L DO.

3.4.4 Non-Filtrable and Volatile Residue

The mixed liquor suspended solids (MLSS) and non-filtrable residues (SESS) were determined using Method 209 0 "Total Non-filtrable Residue dried at 103-105°C". The mixed liquor volatile suspended solids (MLVSS) were determined using Method 209 E "Total Volatile and Fixed Residue at 550°C". Both methods are described in Standard Methods (Anon., 1980). Duplicates of each sample were analyzed immediately after collection.

3.4.5 Total and Volatile Solids

Sludge samples were analyzed for total and volatile solids using Method 209 G, "Volatile and Fixed Matter in Non-filtrable and in Solid and Semisolid samples" as described in Standard Methods (Anon., 1980). Duplicates were analyzed for each sample.

3.4.6 Sludge Volume Index

The sludge volume index test was utilized as a routine process control parameter to monitor the settling characteristics of the activated sludge. The sludge volume index was measured in accordance with Method 213 C "Sludge Volume Index" as described in Standard Methods (Anon., 1980).

3.4.7 Supernatant Suspended Solids (SPSS)

The mixed liquor suspension for the SVI test was allowed to settle for an additional 30 minutes. A supernatant sample was drawn off using a syphon. The SPSS concentration was determined in accordance with Section 3.4.4.

3.4.8 Turbidity

The turbidity of the effluent was measured using a nephelometric turbidimeter (Model DRT-150) as manufactured by H.F. Instruments Limited, Ontario, Canada. The "Nephelometric Method", section 214 A as described in Standard Methods (Anon., 1980) was used for determination of turbidity.

3.4.9 Chemical Oxygen Demand

All chemical oxygen demand test determinations employed a colorimetric method described by Knechtel (1978). The samples were placed in Kimax 25 x 150 mm culture tubes. Digestion reagents were added and the Kimax tubes were then tightly capped with teflon-lined bakelite caps. The culture tubes were inverted three times to ensure uniform mixing of the contents and placed in a 150°C oven for three hours to digest the oxygen demanding material. After cooling in a water bath (this practise has recently been discontinued due to safety reasons), the tubes were placed in a Baush and Lomb, Spectronic 20 spectrophotometer and absorbance and transmittance readings were taken at a 600 mm wavelength. A COD analysis was conducted on two replicates of each sample and standards. Each set of potassium acid phtalate standards were measured to obtain a calibration curve. The COD of

the unknown samples were determined from the calibration curve. Total and soluble COD determinations were determined from the mixed liquor sample. In order to determine removal efficiencies, COD determinations were also conducted on influent and effluent samples.

3.5.0 Biochemical Oxygen Demand

The BOD_5 of the influent and effluent samples were determined using Method 507 "Oxygen Demand Biochemical" as described in Standard Methods (Anon., 1980). The samples were analyzed immediately, therefore provisions for preserving the samples were not required. The dissolved oxygen measurements conducted on the prepared samples and dilution water blanks were determined using the method outlined in Section 3.4.3.

3.5.1 Microscopic Examination

Microscopic examination of the mixed liquor was conducted on a daily basis to identify major microorganism types. The examination of microorganisms gave a quick indication of treatment efficiencies and operation in accordance with a biological growth curve. A Bausch and Lomb, Balplan microscope was utilized for the test.

CHAPTER 4

RESULTS

4.1 SUMMARY OF SOLIDS SEPARATION DATA

The results of the solids separating performance of the CFLUC system at various F:M ratios are presented in Tables 4.1 to 4.4, inclusive. The tabulated data represents the average daily values of the analytical results during steady state operating conditions for each test run. As discussed in Section 3.2.2, one could accurately define the time periods corresponding to acclimation and steady state operating conditions by plotting the operating parameters: SRT, MLSS and waste removal efficiencies versus time. As shown on Figure 4.1, the aforementioned operating parameters remained relatively constant with time during steady state operating conditions. Data tabulated for runs B1 AC, B2 AC, B3 AC and B4 AC have been excluded from further analysis but are shown in Tables B.1 to B.4, Appendix B, inclusive. The daily results presented in Tables 4.1 to 4.4 have been averaged and are sumamrized in Table 4.5.

As shown in Table 4.5, the environmental conditions (ph, Do and Temp.) remain relatively constant during the test periods. The only operating parameter that was significantly varied was the organic loading and MLSS concentrations. A summary of the F:M ratios and SRT values corresponding to each test run are shown in Table 4.6.

In determining the F:M ratios corresponding to each test run, an analysis of the influent characteristics and removal efficiencies was conducted. The results were tabulated and are shown in Tables C.1 to C.4, Appendix C, inclusive. The influent has been characterized

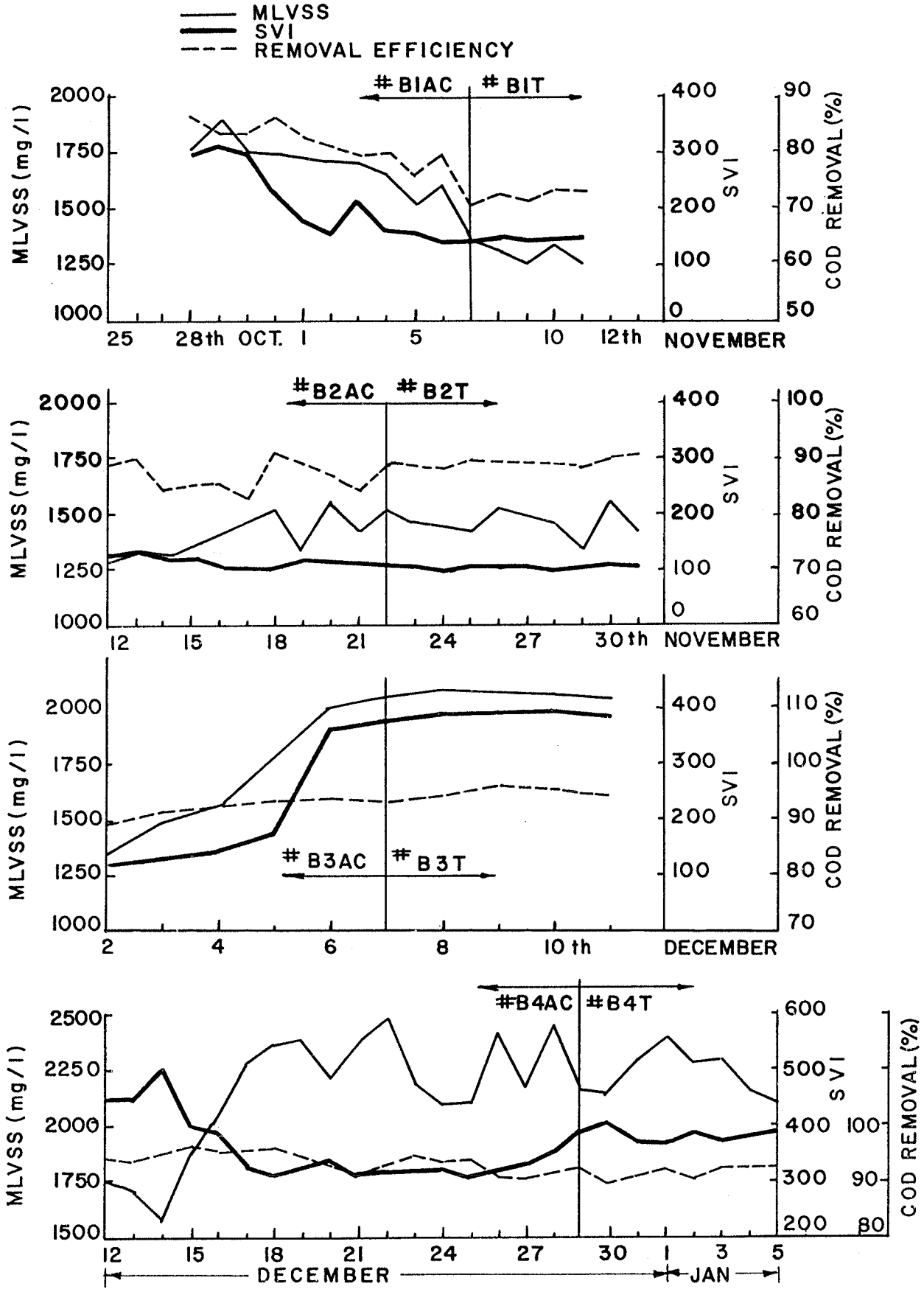


Figure 4.1 Daily MLVSS, SVI and removal efficiency measurements

TABLE 4.1 STEADY STATE OPERATING CONDITIONS RUN # B1T; F:M = 0.10 g COD/g MLVSS.d

DATE		AERATION BASIN						UPFLOW CLARIFIER						
MONTH	DAY	pH	DO mg/L	TEMP °C	MLSS mg/L	MLVSS mg/L	SVI mg/L	pH	DO mg/L	TEMP °C	SESS mg/L	TURBIDITY N.T.U.	SPSS mg/L	REMOVAL %
Nov	7	7.55	5.3	18.0	1705	1368	141	7.75	3.3	19.5	8.9	2.1	-	99.5
Nov	8	7.50	5.0	18.0	1608	1335	149	7.70	2.9	20.0	8.6	2.2	34.0	99.5
Nov	9	7.48	4.2	18.0	1490	1255	142	7.70	2.9	20.0	8.8	2.9	37.5	99.4
Nov	10	7.45	5.3	18.0	1517	1337	144	7.68	3.6	20.0	8.8	2.6	40.0	99.4
Nov	11	7.50	6.0	17.5	1525	1255	144	7.65	4.0	19.5	11.5	3.4	35.0	99.2

TABLE 4.2 STEADY STATE OPERATING CONDITIONS RUN # B2T; F:M = 0.24 g COD/g MLVSS.d

DATE		AERATION BASIN						UPFLOW CLARIFIER					
MONTH	DAY	pH	DO mg/L	TEMP °C	MLSS mg/L	MLVSS mg/L	SVI mg/L	pH	DO mg/L	TEMP °C	SESS mg/L	SPSS mg/L	S.S REMOVAL
Nov	22	7.63	6.0	16.5	1978	1523	103	7.75	2.4	18.5	14.2	44.5	99.3
Nov	23	7.63	5.9	16.0	1818	1460	102	7.75	2.4	18.7	22.4	520	98.8
Nov	24	7.63	6.0	16.0	1762	1437	96	7.75	2.5	18.3	21.9	74.0	98.8
Nov	25	7.53	4.9	16.0	1685	1415	105	7.70	1.7	18.5	16.9	44.0	99.0
Nov	26	7.68	6.6	16.0	1767	1507	101	7.77	2.9	18.5	22.4	54.0	98.7
Nov	27	7.65	5.9	16.0	1773	1485	102	7.75	2.7	18.0	21.1	53.0	98.8
Nov	28	7.82	6.2	16.0	1655	1452	198	7.90	3.3	18.0	24.3	72.5	98.5
Nov	29	7.60	5.9	16.0	1635	1327	101	7.63	2.6	18.2	19.2	52.0	98.8
Nov	30	7.50	5.6	16.0	1738	1553	105	7.68	2.6	18.3	15.4	48.0	99.1
Dec	1	7.50	6.5	16.0	1767	1405	104	7.55	2.3	18.0	19.3	54.0	98.9

TABLE 4.3 STEADY STATE OPERATING CONDITIONS RUN # B3T; F:M = 0.41 g COD/g MLVSS.d

DATE		AERATION BASIN						UPFLOW CLARIFIER					
MONTH	DAY	pH	DO mg/L	TEMP °C	MLSS mg/L	MLVSS mg/L	SVI mg/L	pH	DO mg/L	TEMP °C	SESS mg/L	SPSS mg/L	S.S REMOVAL
Dec	7	7.66	4.1	15.5	2457	2075	381	7.70	1.1	17.5	7.4	7.1	99.7
Dec	8	7.67	5.1	16.5	2455	2110	390	7.73	1.3	17.5	6.7	8.8	99.7
Dec	9	7.63	4.9	15.5	2478	2070	391	7.60	1.3	17.7	6.5	6.7	99.7
Dec	10	7.60	4.7	15.2	2437	2060	394	7.70	1.5	17.0	10.3	11.7	99.5
Dec	11	7.60	4.2	15.0	2500	2039	383	7.50	0.9	17.0	6.5	9.0	99.7

TABLE 4.4 STEADY STATE OPERATING CONDITIONS RUN # B4T; F:M = 0.83 g COD/g MLVSS.d

DATE		AERATION BASIN						UPFLOW CLARIFIER					
MONTH	DAY	pH	DO mg/L	TEMP °C	MLSS mg/L	MLVSS mg/L	SVI mg/L	pH	DO mg/L	TEMP °C	SESS mg/L	SPSS mg/L	S.S REMOVAL
Dec	29	7.75	4.9	15.0	2442	2143	389	7.78	1.0	17.0	37.0	31.0	98.5
Dec	30	7.68	5.1	15.0	2313	2163	409	7.75	0.7	17.0	34.0	34.2	98.5
Dec	31	7.60	4.4	15.0	2567	2287	372	7.68	0.8	17.0	41.0	-	98.4
Jan	1	7.60	4.4	15.0	2637	2377	365	7.65	0.8	17.0	39.5	-	98.5
Jan	2	7.63	6.0	15.0	2510	2140	382	7.63	0.8	18.0	43.0	-	98.3
Jan	3	7.60	4.3	15.0	2667	2290	364	7.70	0.8	18.0	44.0	-	98.4
Jan	4	7.70	4.2	15.5	2537	2140	376	7.75	0.7	18.0	44.3	-	98.3
Jan	5	7.70	3.0	16.0	2493	-	381	7.80	0.6	18.5	33.3	36.5	98.7

TABLE 4.5 SUMMARY OF AVERAGE RESULTS DURING STEADY-STATE CONDITIONS

TEST	AERATION BASIN						UPFLOW CLARIFIER					
RUN	pH	DO mg/L	TEMP °C	MLSS mg/L	MLVSS mg/L	SVI mg/L	pH	DO mg/L	TEMP °C	SESS mg/L	SPSS mg/L	SS REMOVAL %
B1T	7.50	5.2	17.9	1569	1310	144	7.70	3.3	19.8	9.3	36.6	99.4
B2T	7.62	6.0	16.1	1758	1456	102	7.72	2.5	18.3	19.7	54.8	98.9
B3T	7.63	4.6	15.5	2465	2071	388	7.65	1.2	17.3	7.5	8.7	99.7
B4T	7.66	4.5	15.2	2521	2220	380	7.72	0.8	17.6	39.5	33.9	98.5

TABLE 4.6 SUMMARY OF F:M RATIOS AND SRT VALUES

TEST	INFLUENT COD mg/L	INFLUENT BOD mg/L	MLVSS mg/L	HRT h	F:M RATIO gCOD/gMLVSS.d	F:M RATIO gBOD/gMLVSS.d	SRT d
B1T	132	87	1310	24	0.10	0.07	net growth = 0 α
B2T	346	210	1456	24	0.24	0.14	27.1
B3T	857	552	2071	24	0.41	0.27	9.1
B4T	1837	906	2220	24	0.83	0.41	4.2

in terms of COD and BOD. During the course of the testing, BOD/COD relationships were developed to reduce extensive BOD testing. The average BOD_5/COD value was 0.60 during the test program.

In determining the SRT values corresponding to each test run, a solids analysis was conducted to determine the sludge wasting rates and the characteristics of the return sludge. The results were tabulated and are shown in Tables C.5 to C.8 inclusive, Appendix C. Typical calculations for the F:M ratios, and SRT values are shown in Appendix C.

4.2 Evaluation of Biokinetic Constants

An evaluation of the biokinetic constants was conducted to confirm the validity of F:M ratios and SRT values shown in Table 4.6. As discussed in Section 2.2.3, the SRT is related to the reciprocal of the net microbial growth rate by the following equation:

$$\frac{1}{SRT} = Y \left[\frac{F:M}{100} \right] E - K_d \qquad (7)$$

A plot of the specific oxygen utilization rate, [F:M/100]E, versus the specific growth rate, [1/SRT], should result in a straight-line function. The straight line function would indicate that the yield constant, Y, was identical for each selected F:M ratio and would also verify that steady-state conditions had been achieved in all test runs. The derived biokinetic constants (Y and Kd) could also be compared with other literature sources to further verify the test data.

The first step in evaluating the biokinetic constants was to determine the process efficiency (E in percent). The process efficiency can be defined by the following equation:

$$\frac{SO-S}{So} \times 100 = E \qquad (8)$$

The process efficiency is determined by measuring the soluble BOD_5 and/or soluble COD in the influent and effleunt. In this test program, the COD test was the most convenient and accurate method of measuring process efficiency, however, not without its limitations. The COD test measures both biodegradable and non-biodegradable organic matter. It is, however, only the biodegradable portion that influences the biochemical process. The derivation of biokinetic constants in previous kinetic models have been developed solely on the soluble biodegradable organic fraction. Grady and Lim (1980) concluded that the efficiency of the process can be corrected for the non-biodegradable COD by subtracting this value from the influent and soluble effluent concentrations. This can be mathematically expressed as follows:

$$So = Co - Ci$$

$$S = C - Ci \qquad (9)$$

where: Ci = non biodegrabale COD

Co = influent concentration

C = soluble COD effluent concentration

A summary of the corrected process efficiencies at the various F:M ratios and/or SRT values is shown in Table 4.7.

There is a decrease in the corrected process efficiencies with a substantial increase in the F:M ratios. The corrected values for the process efficiency were used in the development of the arithmetic plot shown on Figure 4.2.

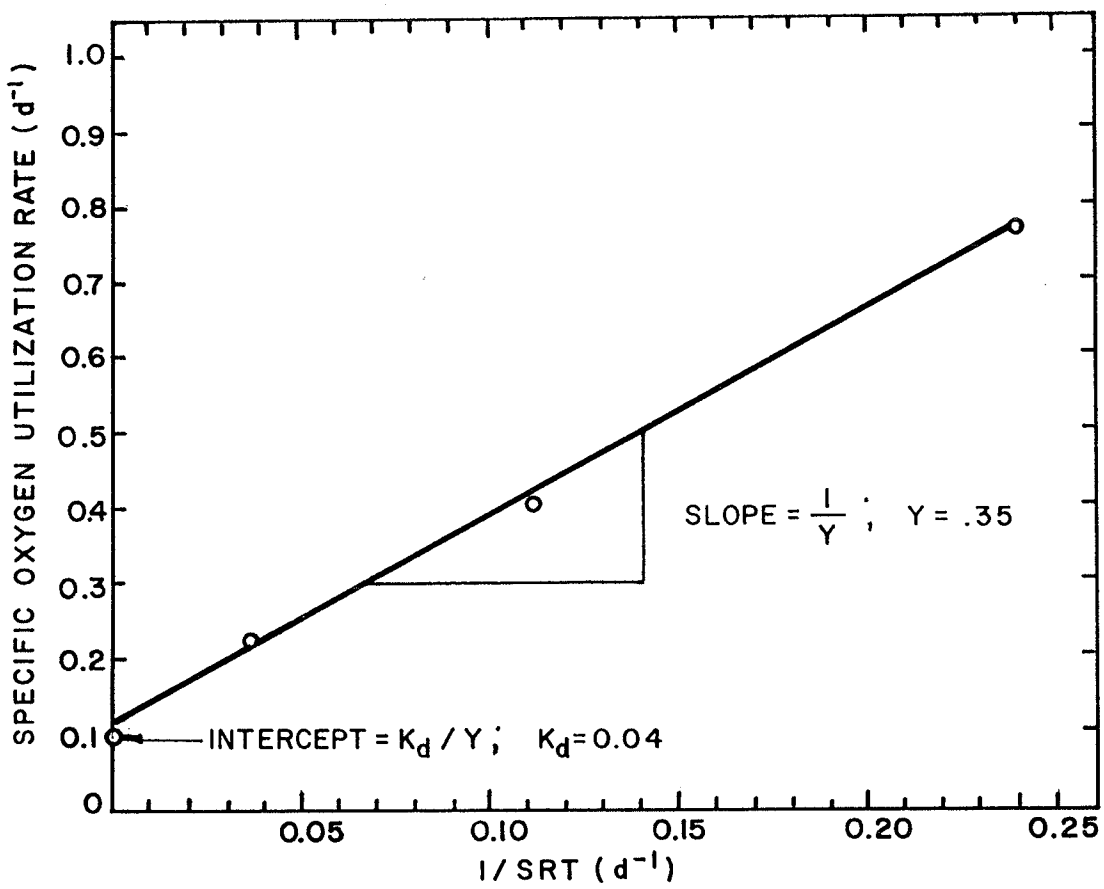


Figure 4.2 Determination of biokinetic constants, Y and Kd

The biokinetic constants, Y and Kd, derived from the plot were determined to be 0.35 mg/mg and 0.04 $days^{-1}$, respectively. It is interesting to note that the specific growth rate was zero at a

TABLE 4.7

CORRECTED PROCESS EFFICIENCIES AT VARIOUS F:M RATIOS AND SRT VALUES

TEST RUN	SRT d	F:M gCOD/gMLVSS.d	Co mg/L	C mg/L	E EFFICIENCY $\frac{(CO-C)}{Co} \times 100$	Ci mg/L	CORRECTED EFFICIENCY So-S/SO x 100 (%)
B1T	-	0.10	132	37	72.0	37	≃ 100.0
B2T	27.0	0.24	346	37	89.3	37	≃ 100.0
B3T	9.0	0.41	857	44	94.9	37	99.2
B4T	4.2	0.83	1837	152	91.7	37	93.7

specific oxygen utilization rate of 0.12 d^{-1}. Theoretically, this would be equivalent to an SRT value of infinity. Typical biokinetic constants reported by Grady and Lim (1980) for skim milk substrates were:

(1) Yield Coefficient (Y) - 0.38 gram cells/gram COD removed

(2) Decay Ceofficient (Kd) - 0.05 d^{-1}

The comparison of the biokinetic constants confirmed the F:M ratios and SRT values determined in phase B of the experimental study. The constant yield and decay coefficients also confirmed that steady-state conditions had been achieved in all test runs.

4.3 ZONE SETTLING VELOCITIES

The average zone settling velocities were determined for each F:M ratio (during steady-state conditions) and are shown in Table 4.8.

TABLE 4.8 AVERAGE ZONE SETTLING VELOCITIES AT VARIOUS F:M RATIOS

RUN #	F:M gCOD/gMLVSS.d	ZSV in/h	ZSV cm/h
B1T	0.10	114	290
B2T	0.24	170	431
B3T	0.41	1.5	3.9
B4T	0.83	0.9	2.3

A computer program was utilized to plot the daily solids interface versus time curves. Typical plots representing the average zone settling velocities at each F:M ratio are shown in Figures 4.3 to 4.5, inclusive. Daily results are shown in Tables D.1, Appendix D.

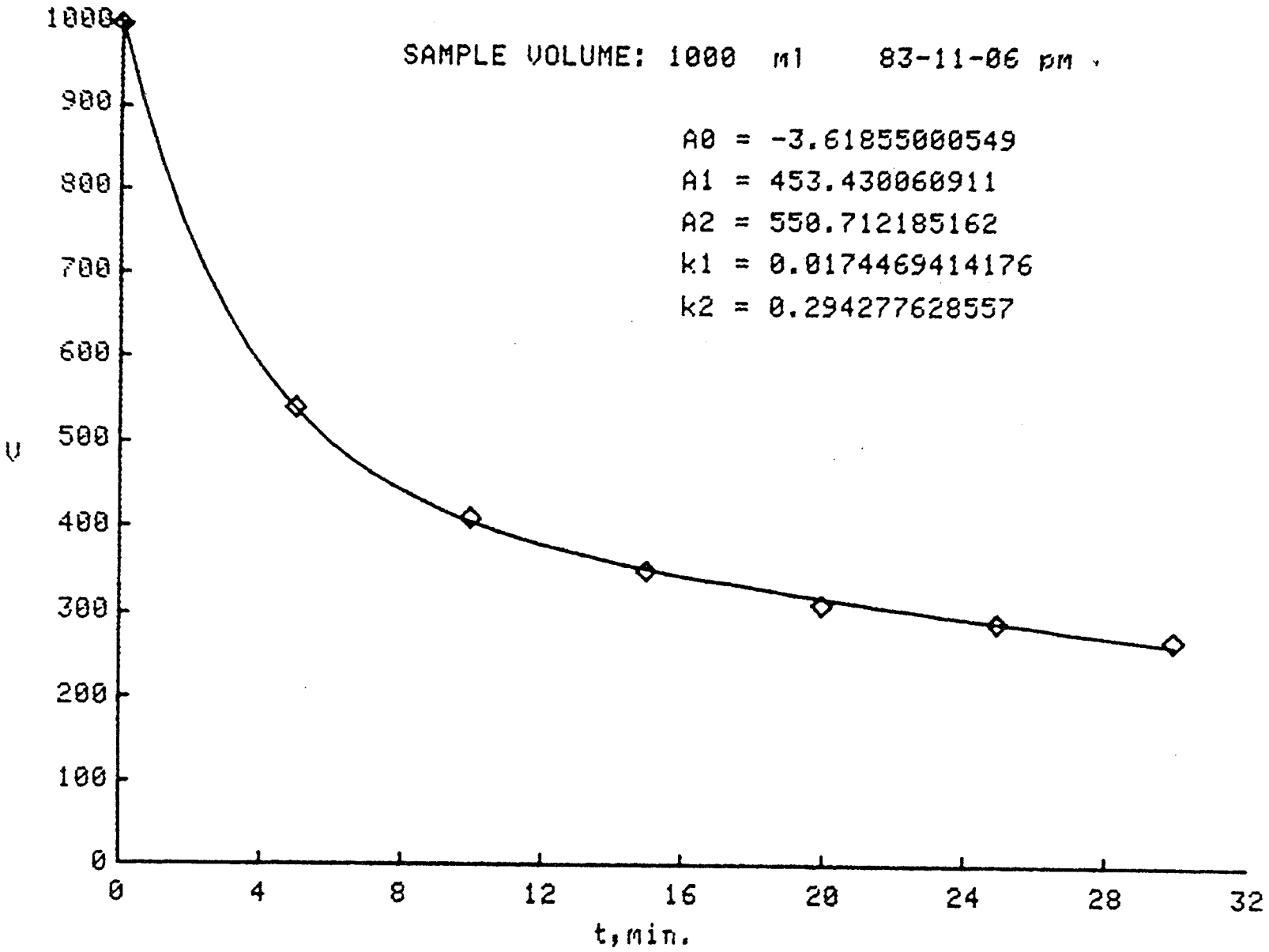


Figure 4.3 Typical average zone settling velocity plot; run # B1T
F:M = 0.10 g COD/g MLVSS.d

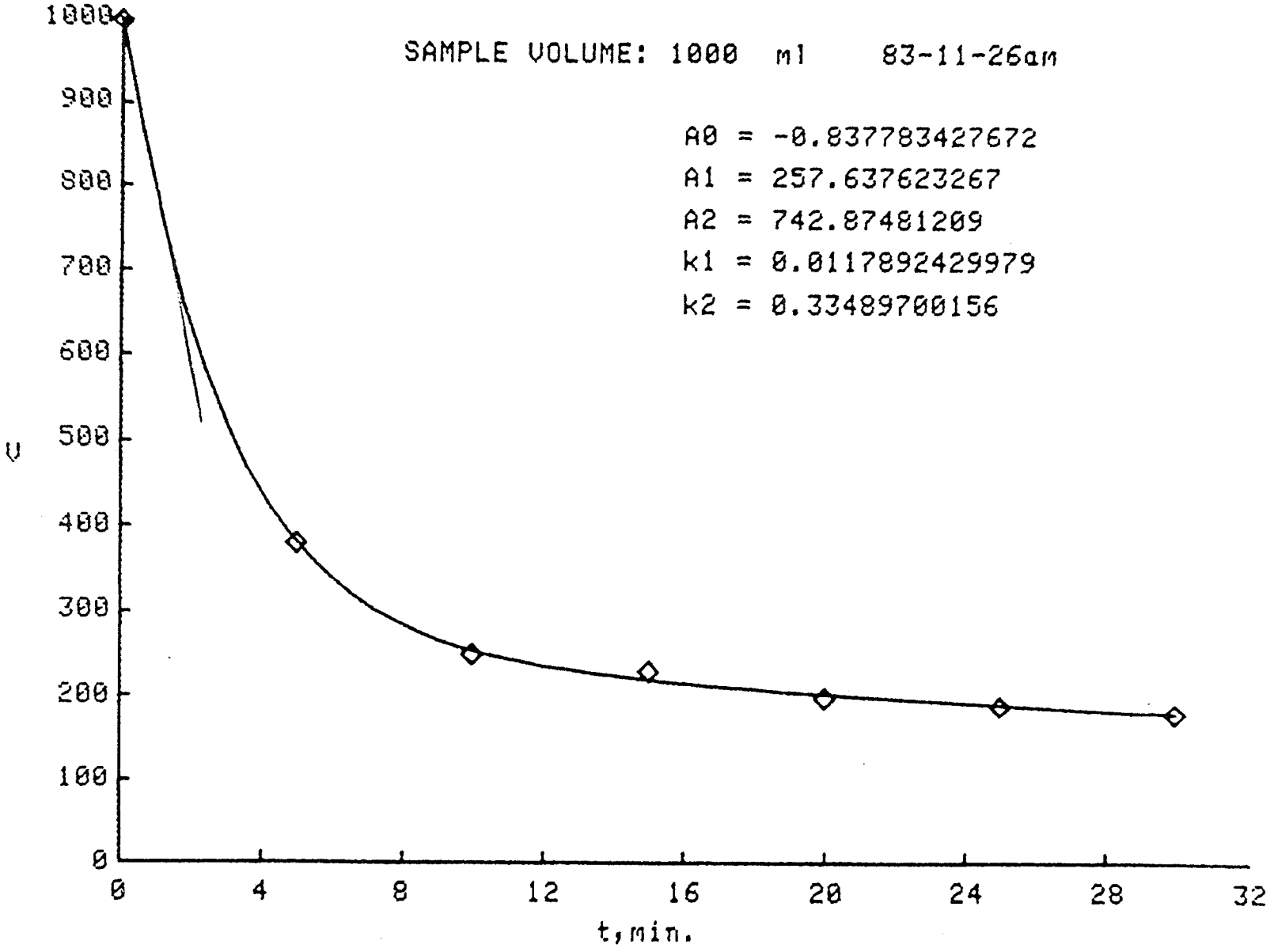


Figure 4.4 Typical average zone settling velocity plot; run # B2T
F:M = 0.24 g COD/g MLVSS.d

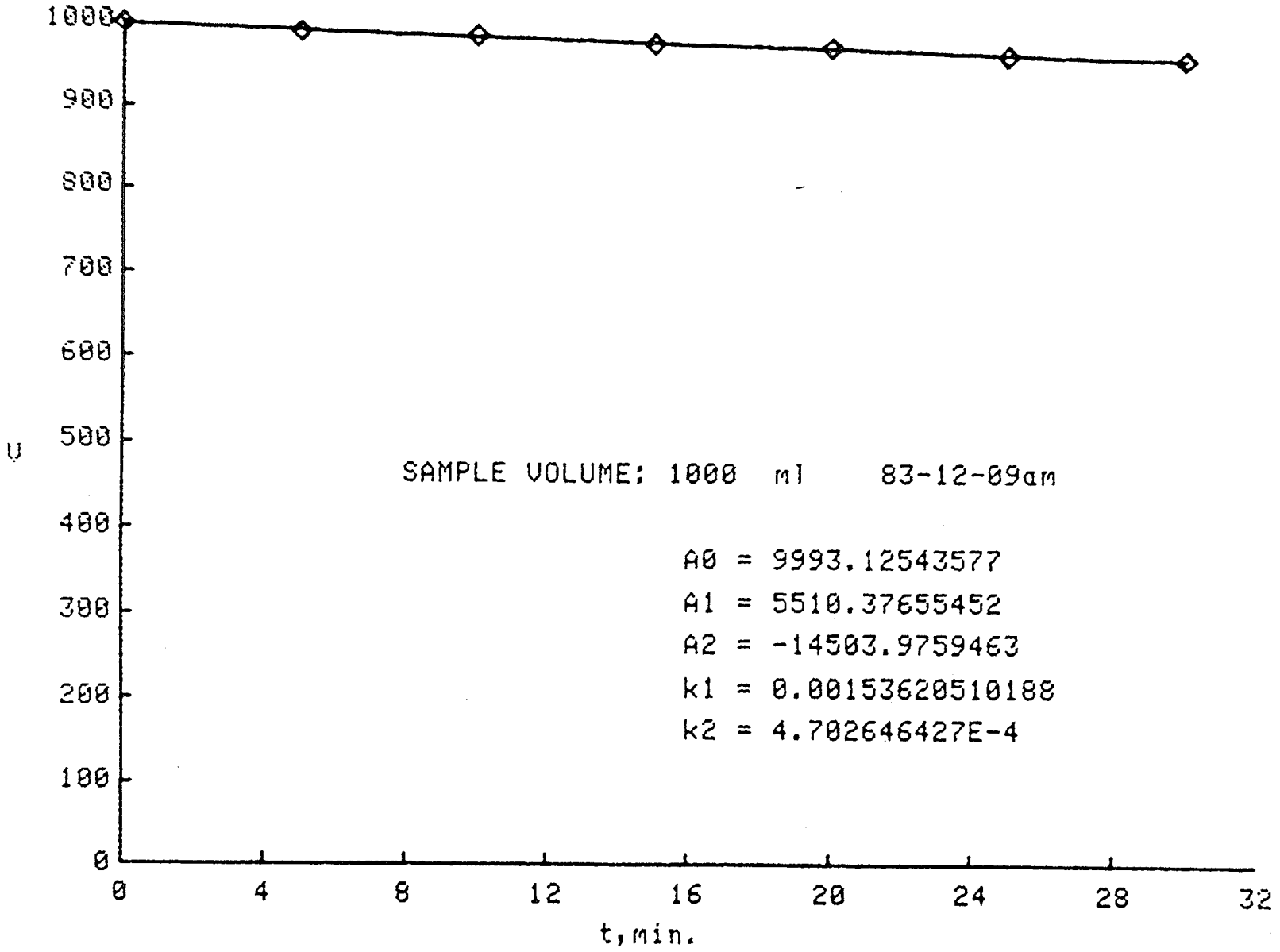


Figure 4.5 Typical average zone settling velocity plot; run # B3T
F:M = 0.41 g COD/g MLVSS.d

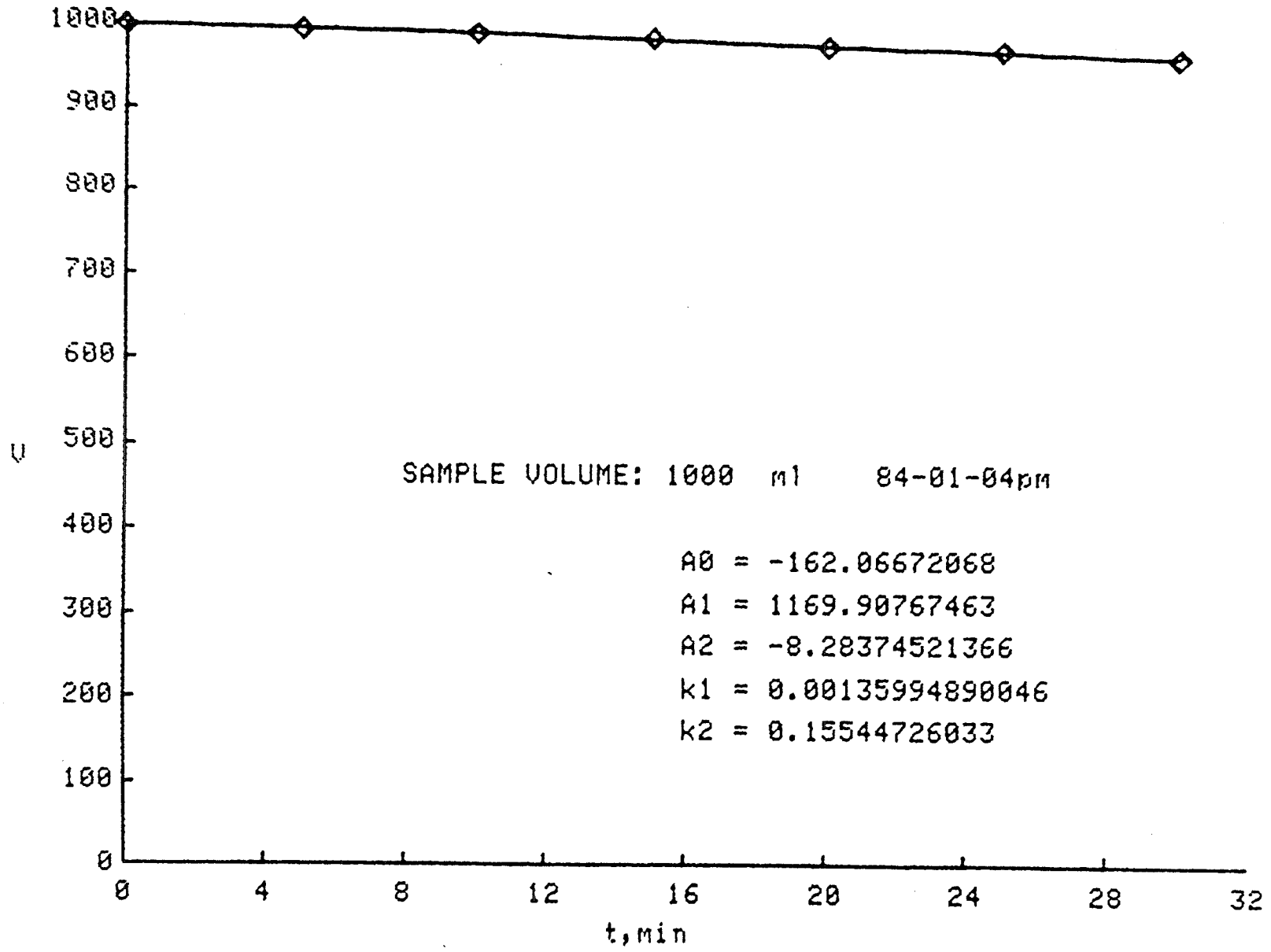


Figure 4.6 Typical average zone settling velocity plot; run # B4T
F:M = 0.83 g COD/g MLVSS.d

4.4 SLUDGE BLANKET AND SLUDGE RECYCLE CHARACTERISTICS

To determine the effect of sludge blanket height in the upflow clarifer on effluent quality, miscellaneous suspended solids measurements were taken from sample ports in the clarifier (i.e. at various heights from the sludge blanket). These values were then compared to SESS measurements at the overflow weir. The results are shown in Table 4.9.

A comparison of the suspended solids at the various sample point locations to SESS measurements indicates little improvement in effluent quality above the sludge blanket level. Also, an increase in sludge blanket height does not appear to improve effluent quality.

Total solids were also conducted on sludge samples at various sample points in the sludge blanket to define the characteristics of the fluidized and thickened portions of the sludge blanket. The results are tabulated in Table 4.10. The results indicate that the fluidized sludge blanket is characterized by a constant distribution of suspended solids, regardless of sludge blanket height.

The purpose of the return of sludge is to maintain a sufficient concentration of activated sludge in the aeration tank so that the required degree of treatment can be obtained. In general, the return-sludge pumps in conventional clarifiers are set so that the return flow is approximately equal to the percentage ratio of the volume occupied by the settleable solids to the volume of supernatant liquid after settling for 30 minutes in a 1000 ml graduated cylinder (Metcalf and Eddy, Inc., 1979). The sludge recycle rates calculated as above were compared with actual sludge recycle rates determined for the laboratory unit. The average results obtained for each test run under

TABLE 4.9

THE EFFECT OF SLUDGE BLANKET & CLARIFIER HEIGHT ON EFFLUENT QUALITY

RUN #	DATE	SLUDGE BLANKET HEIGHT cm	SAMPLE LOCATION HEIGHT cm	SAMPLE SS mg/L	SESS* mg/L
B1T	Nov 8	30	34	15.0	8.2
B1T	Nov 10	29	66	13.0	8.8
B2T	Nov 23	34	156	25.0	21.7
B2T	Nov 27	27	156	22.5	21.5
B3T	Dec 7	50	156	5.6	6.0
B3T	Dec 9	66	156	5.7	4.3
B4AC	Dec 20	48	66	15.5	13.0
B4AC	Dec 22	54	66	18.3	18.5
B4AC	Dec 25	68	96	16.9	14.4
B4AC	Dec 26	48	66	40.0	36.5
B4AC	Dec 27	21	34	40.0	36.5
B4AC	Dec 28	21	34	29.5	25.0
B4T	Dec 29	37	66	36.0	34.0
B4T	Dec 30	47	66	36.5	35.0
B4T	Dec 31	67	96	42.0	36.5
B4T	Jan 1	72	96	37.5	32.5
B4T	Jan 2	59	96	42.5	43.0
B4T	Jan 3	67	96	49.5	42.0
B4T	Jan 4	55	66	44.0	40.5
B4T	Jan 5	43	66	35.5	29.0

* NOTE: The overflow weir (SESS) is at 218 cm above datum

TABLE 4.10

CHARACTERISTICS OF THE FLUIDIZED AND THICKENED PORTIONS OF THE SLUDGE BLANKET

RUN #	DATE	SLUDGE BLANKET HEIGHT (cm)	FLUIDIZED T.S. %	THICKENED T.S. %
A1AC	Oct 28	45	0.47	0.81
	Oct 29	35	0.46	1.02
	Oct 31	45	0.45	0.85
	Nov 1	70	0.45	0.94
	" 2	50	0.47	0.95
	" 3	48	0.47	0.96
	" 4	45	0.45	0.97
	" 5	44	0.49	1.07

TABLE 4.11

CALCULATED VERSUS ACTUAL SLUDGE RECYCLE RATES

RUN #	F:M gCOD/gMLVSS.d	CALCULATED RECYCLE RATE Q_R/Q	ACTUAL RECYCLE RATE Q_R/Q
B1T	0.10	0.29	0.20
B2T	0.24	0.22	0.20
B3T	0.41	21.21	0.58
B4T	0.83	28.80	0.24

steady state operating conditions are shown in Table 4.11.

The results indicate the calculated sludge recycle rates in test runs B3 T and B4 T would be 40 to 120 times greater than that actually obtained in the laboratory unit. As shown on Figures 4.5 and 4.6, this is attributed to the poor settling bulking characteristics of the sludge. The enhancement of bioflocculation in the upflow clarifier, however, results in significantly reduced sludge recycle rates which permits the solids-separation of bulking sludges.

CHAPTER 5

DISCUSSION

5.1 PROCESS EFFICIENCY

The efficiency of a secondary treatment process is normally described in terms of the concentrations of BOD_5 and suspended solids in the secondary effluent. The present secondary effluent requirement specifies a 30-day average BOD_5 of 30 mg/L and a suspended solids concentration of 30 mg/L. The effluent from biological treatment processes consists of soluble and insoluble effluent BOD_5. The soluble BOD_5 is mainly a result of soluble organics that escape treatment in the aeration tank whereas the insoluble BOD_5 is due to poor solids-separation in the secondary clarifiers. The removal of soluble BOD_5 can be readily achieved in most cases by controlling operating parameters such as the F:M ratio and SRT values. Over the past 10 years, there has been a general recognition among investigators (Pipes, 1979; Eckenfelder, 1967; Loehr and de Navarra, 1969) that most of the effluent BOD is a result of insoluble BOD_5 caused by excessive loss of solids from the secondary clarifier.

A plot of BOD_5 and soluble COD removal efficiencies versus various organic loadings are shown in Figure 5.1. The BOD_5 and soluble COD removal efficiencies ranged from 95 to 99%. The 30 mg/L standard for effluent BOD_5 was only exceeded at an F:M ratio of 0.83 d^{-1}. The influent organic concentration to the CFLUC system was in the order of 900 mg/L BOD_5.

The solids separation efficiency of the upflow clarifier at various F:M ratios is shown in Figure 5.2. The plot of SESS

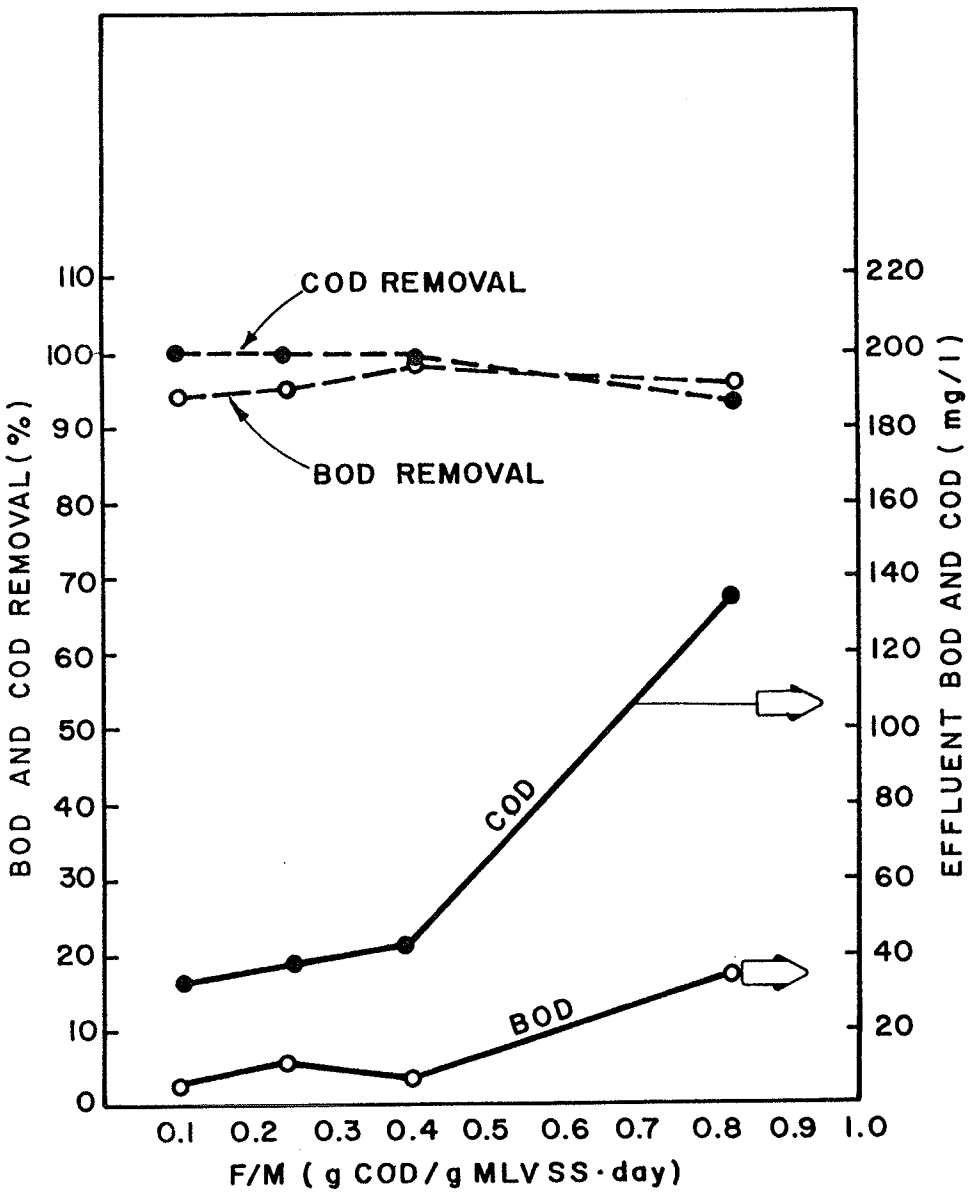


Figure 5.1 BOD_5 and COD process efficiencies

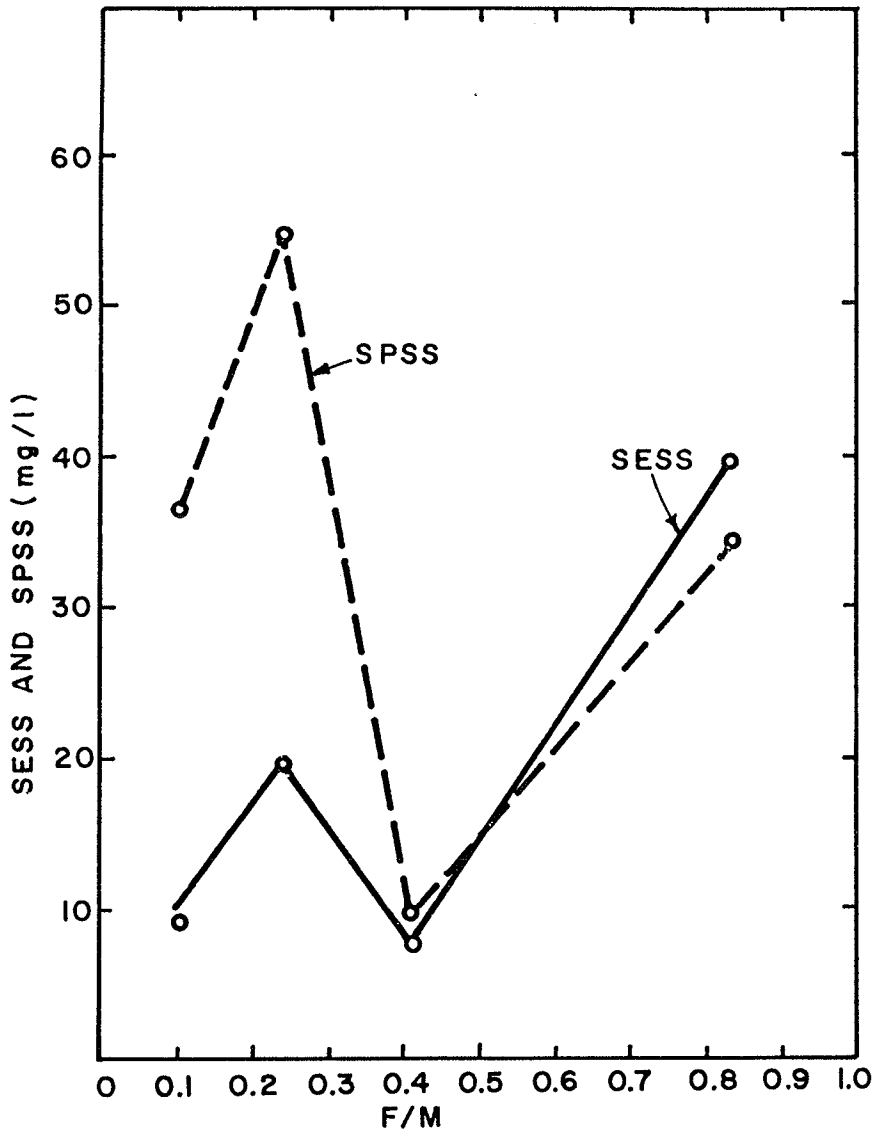


Figure 5.2 SESS and SPSS versus F:M ratios

illustrates that the suspended solids in the effluent were well below the 30 mg/L standard (less than 20 mg/L and as low as 8 mg/L) for three of the organic loadings ranging from 0.10 to 0.41 d^{-1}. Similar to Figure 5.1, the standard was exceeded at an F:M ratio of 0.83 d^{-1}. However, the average SESS concentration at this high organic loading was only 40 mg/L. Very few plants have F:M ratios exceeding 0.4 d^{-1} (Pipes, 1979). The flexibility of the CFLUC system to meet present secondary effluent requirements for such a wide range of organic loadings has great engineering significance.

A plot of supernatant suspended solids (SPSS) on Figure 5.2 illustrates that SPSS values exceed SESS values at F:M ratios of 0.10 and 0.41 d^{-1}. The upflow clarifier is capable of removing 70% of the suspended solids that are left after the sludge is settled under quiescent conditions in a 1000 ml cylinder. This phenomenon seems to indicate that enhanced bioflocculation and/or the filtration effect (enmeshment) of the sludge blanket may be responsible for the reduction in pinpoint floc. It is interesting to note that at higher organic loadings (0.41 and 0.83 d^{-1}) the SESS values were approximately equal to the SESS values. The solids-separation performance of the upflow clarifier for a wide range of organic loadings exceed or equal the performance efficiencies predicted by quiescent batch settling tests.

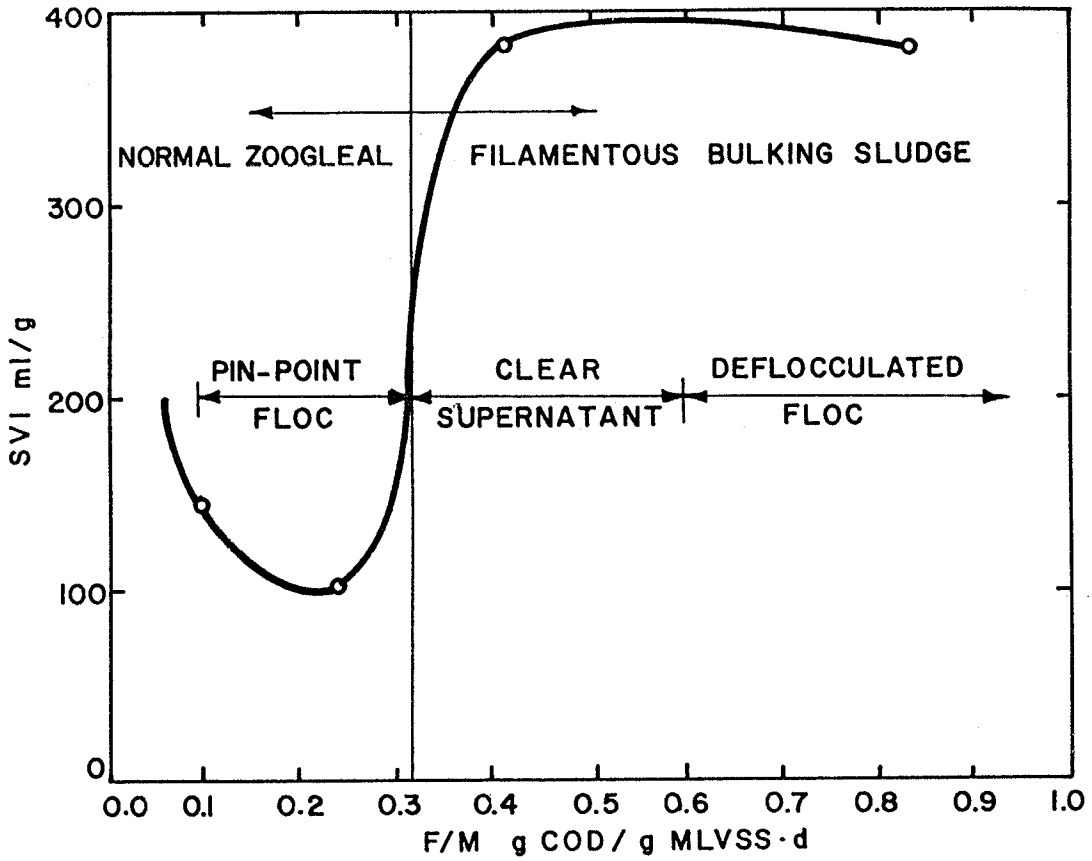
The solids-separation performance of a treatment process is dependent not only on the design of the secondary clarifier but also on the settling characteristics of the sludge. As indicated in Section 2.23, a relationship exists between the settling characteristics of the sludge (expressed as SVI index) and operating control parameters (F:M ratios and/or SRT values). The relationships developed in this

study are shown in Figure 5.3.

The curves are in agreement with a study completed by Pipes (1979) who developed a relationship shown in Figure 5.4 from data collected over a 12-year period from 32 municipal wastewater treatment plants located in the United States.

Similar to the aforementioned study, pinpoint floc was observed at low organic loadings (0.10 to 0.24 d^{-1}). As shown in Figure 5.5, the zone settling velocities ranged from 291 to 431 cm/h, indicating a good settling sludge. Bosogni and Lawrence (1970) reported average zone settling velocities of 457 cm/h and also observed pinpoint floc for similar organic loadings. A microscopic examination of the activated sludge at low organic loading (0.10 to 0.24 d^{-1}) revealed a healthy normal zoogleal population with eucaryotic populations of stalked ciliates, rotifers and little filamentous bacteria.

As noted in Chapter 4, the SRT corresponding to an F:M ratio of 0.10 days was calculated to be infinity since no solids were wasted from the system at the specific organic loading. The accuracy of these results are questionable due to the large scale of the pilot plant. For example, sludge samples and mixed liquor samples for testing purposes were indirectly wasted from the system. It is estimated (see dashed line on Figure 5.3), that the actual SRT would be in the range of 35 to 40 days. Barnard (1978) and Pitman (1980) had shown that when reaching SRT's of 50 to 60 days filamentous forms that can use the remaining polysaccaride material as food may proliferate. This phenomenon was actually observed by the writer on earlier nitrification studies conducted for the CFLUC system. Wasting of activated sludge to within an SRT of 30 days rectified the problem of filamentous



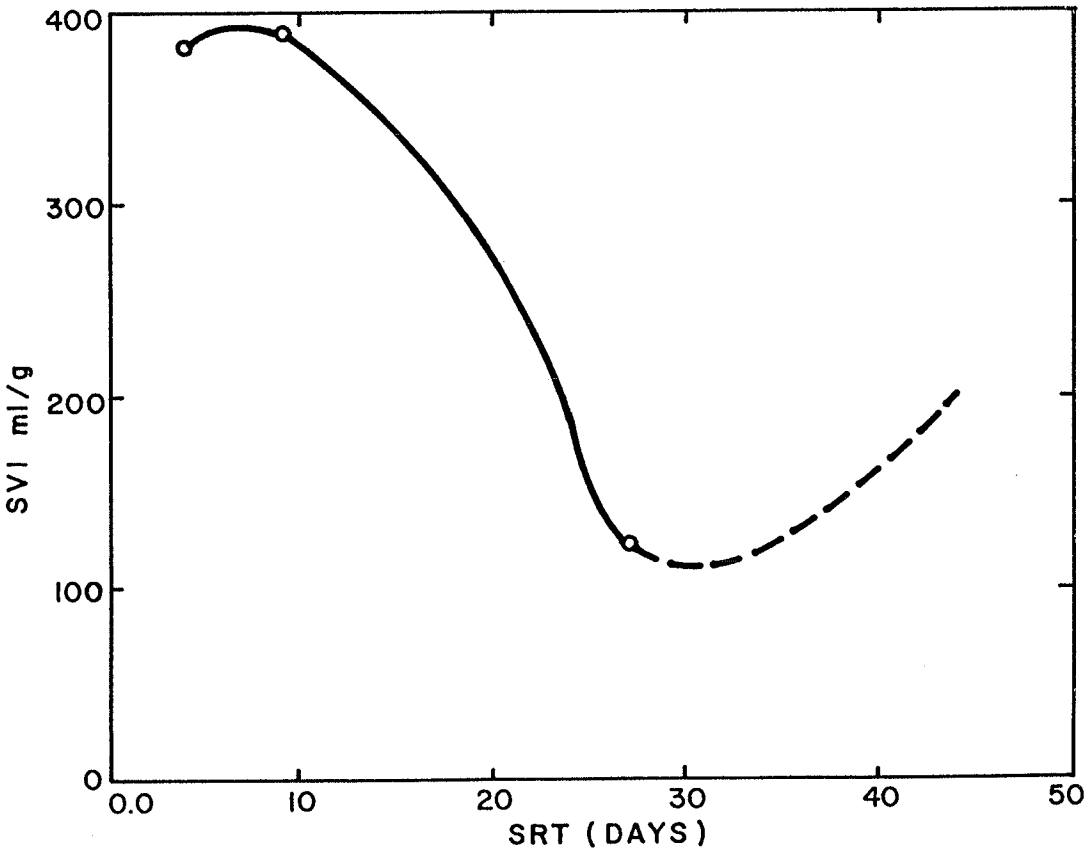


Figure 5.3 Sludge volume index versus F:M ratios and SRT values

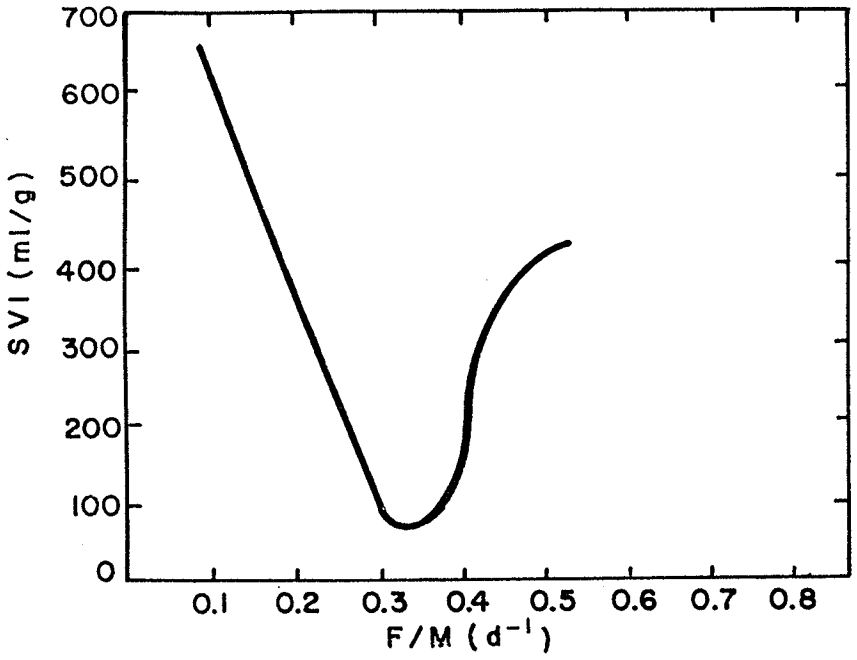


Figure 5.4 Sludge volume index versus F:M ratio (after Pipes, 1979)

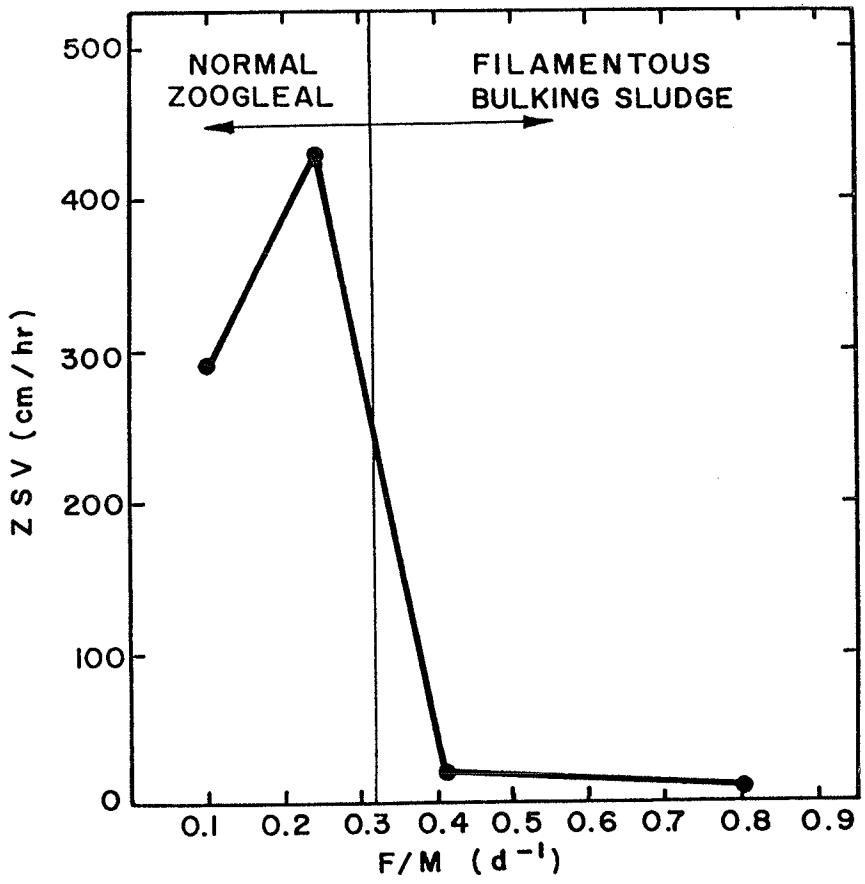


Figure 5.5 Zone settling velocities versus F:M ratios

organisms in the aeration basin.

At high organic loadings (0.41 to 0.83 d^{-1}), the sludges could be characterized as severe bulking sludges with very poor settling characteristics. At an F:M ratio of 0.41 d^{-1} or an SRT value of 9 days, the sludge volume index was 388 mL/g. A clear supernatant with low SPSS concentrations (8.7 mg/L) was observed. This characteristic is typical of filamentous bulking sludges (Pipes, 1979). Filaments were observed from microscopic examinations as sticking out from the floc particles. A photograph of the filaments are shown in Figure 5.6.

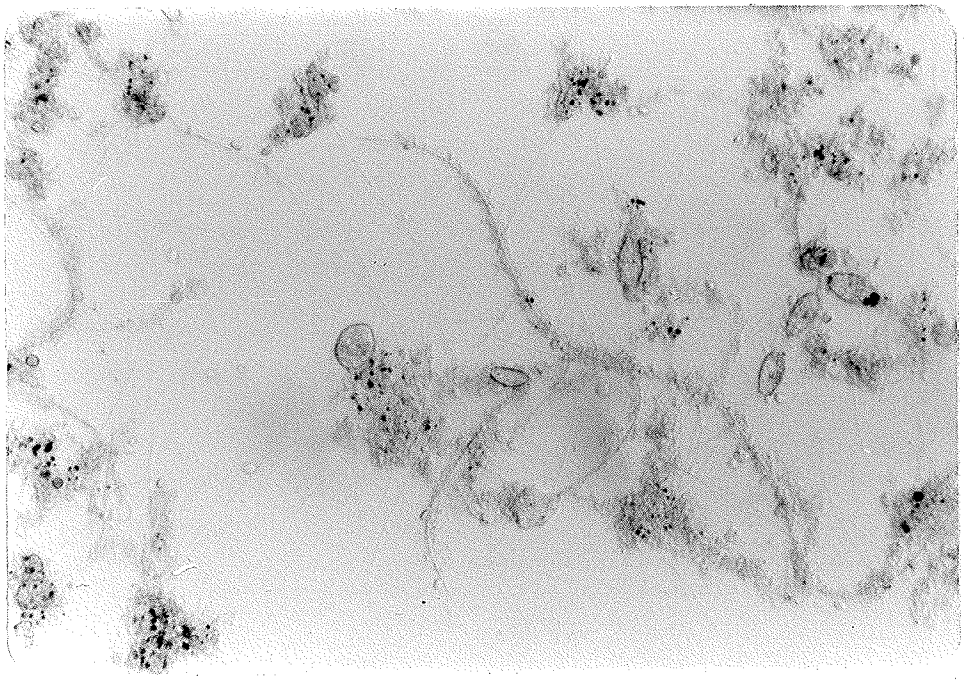


Figure 5.6 Photograph of filaments protruding from floc particles at an F:M ratio of 0.41 d^{-3}

The zone settling velocities decreased significantly (by a factor of 100) as compared to settling characteristics at lower organic loadings (see Figure 5.5).

At an F:M ratio of 0.83 d^{-1}, the sludge was still bulking (SVI = 380 mL/g) but deflocculated particles, much smaller than pinpoint floc, were observed in the supernatant. There was excessive filamentous growth and few eucaryotes. Pipes (1979) and Bosogni and Lawrence (1970), also observed predominatly dispersed growth at high F:M ratios (i.e. greater than 0.40 d^{-1}). Starkey and Karr (1984) concluded that at an F:M ratio of 0.8 g BOD_5/g MLVSS.d, low exocellular polymer production and few eucaryotes were responsible for deflocculated particles and dispersed growth.

The efficiency of the upflow clarifier can be described in terms of the settling characteristics which have been previously evaluated at various organic loadings in the CFLUC system. It appears that at low organic loadings (0.10 d^{-1} to 0.24 d^{-1}) the upflow clarifier is capable of removing approximately 70% of the pinpoint floc particles that remain in the supernatant during quiescent batch settling tests. This is due to enhanced bioflocculation in the upflow clarifier.

At higher organic loadings, the upflow clarifier is capable of separating filamentous-type bulking sludges. The efficiency of the upflow clarifier, however, is similar to that achieved from quiescent batch settling tests. The flexibility of the CFLUC system to separate activated sludges for a wide range of organic loadings and meet environmental regulations is significant.

5.2 OPERATING CHARACTERISTIC AND DESIGN CRITERIA

As shown in Section 4.3, the sludge blanket height does not appear to improve effluent quality (i.e. in terms of SESS concentrations). It is reasonable to assume that the bioflocculation process is taking place through physical, bio-conditioning mechanisms in the downward flow channel. More research is required to define and determine the mechanisms responsible for bioflocculation enhancement. As shown in Section 4.4, the enhancement of bioflocculation in the upflow clarifier results in a significant reduction in sludge recycle rates which permits the solids-separation of bulking sludges at high organic loadings.

The characteristics of the sludge blankets suggest that a low sludge blanket level be maintained in the upflow clarifier. This operating characteristic would be advantageous for two reasons. Firstly, anoxic conditions which could result in denitrification and the formation of nitrogen gas bubbles in the sludge blanket could be prevented. This would reduce the occurrence of a poor quality effluent caused by the buoying of sludge particles to the surface. Secondly, the dissolved oxygen levels in the clarifier would increase, resulting in operating with lower DO levels in the aeration basin. Presently, a dissolved oxygen concentration of 5.0 - 6.0 mg/L is required in the aeration tank to maintain a dissolved oxygen level of 0.6 mg/L (re-operating a sludge blanket at 1/3 the height of the clarifier). Lowering the sludge blanket level in the clarifier could permit operating dissolved oxygen levels of 2-3 mg/L in the aeration tank with adequate dissolved oxygen levels in the clarifier to prevent denitrification. Ultimately, a lower sludge blanket level would reflect a reduction in operating

costs due to lower energy consumption.

The design criteria discussed in Chapter 2, based on guidelines established in the Federal Republic of Germany, has direct application to the CFLUC system. A summary of the actual overflow rates versus recommended overflow rates for each F:M ratio is shown in Table 5.1. At lower organic loadings (F:M = 0.10 to 0.24 d^{-1}) greater overflow rates may be utilized. The German guidelines recommend overflow rates, three times greater than those used in this study. The overflow rates used in this study were compared with overflow rates used for horizontal-flow type clarifiers for the separation of extended aeration sludges (see Table 2.3). As concluded in Section 2.3.2, the use of lower MLSS concentrations has definite operating advantages in that greater overflow rates may be utilized.

TABLE 5.1

ACTUAL OVERFLOW RATE VERSES ATV GUIDELINE RECOMMENDED OVERFLOW RATES

OPERATING CHARACTERISTIC		ACTUAL OVERFLOW RATE	ACTUAL MLSS	VS_V	ATV GUIDELINE RECOMMENDED OVERFLOW RATE
F:M (d^{-1})	SVI mL/g	m/h	mg/L	ml/L	m/h
0.10	144	0.8	1569	226	2.28
0.24	102	0.8	1758	179	2.60
0.41	388	0.8	2465	956	0.40
0.83	380	0.8	2521	958	0.40

At higher organic loadings, greater overflow rates were utilized in the laboratory unit (0.8 m/h versus 0.4 m/h) as compared

to surface loading recommended in the German guidelines. Similar results by Resch (1980) concluded that actual plants incorporating upflow clarifiers in Germany yield much higher permissible surface loadings than given by the ATV-guidelines.

More research is required to verify the design guidelines on overflow rates established by the Federal Republic of Germany for upflow clarifiers. The possibility of increasing the overflow rate by a factor of three for good settling sludges has great engineering significance. Smaller clarifier units would be required in the future resulting in a significant reduction in capital costs.

Extensive investigations by Pflanz (1969) led to the German guidelines on dimensioning upflow secondary clarifiers, as shown in Table 2.5. Pflanz recommended solids loading rates to achieve an SESS of 30 mg/L for various SVI values.

For example, Pflanz recommended a solids loading of 0.8 - 1.1 $kg/m^2.h$ for a sludge with an SVI of 300 mL/g. The actual solids loading to the CFLUC clarifier was 1.83 $kg/m^2.h$ for an SVI of 380 mL/g (i.e. F:M = 0.83 d^{-1}).

An interesting observation was made on December 12-14 inclusive (Run # B4 AC). During this time period (see Table B.4, Appendix B), a problem with frozen concentrated feed resulted in SVI's exceeding 500 ml/g. As expected, the sludge blanket rose to the overflow weir, confirming that clarifier area and depth is directly related to the SVI. As observed by the writer and confirmed by Resch (1980), increasing the sludge recycle rate did not lower the sludge blanket but only aggravated the problem by increasing the overflow rate. The only solution was to reduce the overflow rate by one-half, resulting in

a solids loading rate of 0.9 $kg/m^2.hr$. Once the frozen feed problem was corrected and the SVI improved to 383 ml/g (Dec. 17) the overflow rate and solids loading rate were increased to their original values. This observation confirmed that the separation of bulking sludges in an upflow clarifier is controlled by SVI, overflow rate, MLSS and the dimensions of the clarifier.

The characteristic of a bulking sludge is that a clear supernatant low in suspended solids is produced. The capability to separate a bulking sludge has great engineering significance since the results will yield a high quality effluent.

CHAPTER 6

ENGINEERING SIGNIFICANCE

Based on the results of the study, the CFLUC system appears to have the potential to separate all types of sludges for a wide range of organic loadings (F:M = 0.10 - 0.83 g COD/g MLVSS.d). At low organic loadings the upflow clarifier is capable of removing 70% of the pinpoint floc particles that remain in the supernatant during quiescent batch settling tests. At high organic loadings, the upflow clarifier has the capability of separating filamentous-type bulking sludges. The significance of the results is that no system has been capable of separating bulking sludges due to the large sludge recycle rates required. It is hypothesized that due to the mechanism of bioflocculation and/or physical bio-conditioning in the downward flow channel of the clarifier, a significant reduction in sludge recycle rates can be achieved, thus permitting the separation of severe bulking sludges (380-388 mL/g).

The potential of having the capability to separate all types of sludges over a wide range of organic loadings gives the designer and operator great flexibility in the design and operation of a wastewater treatment facility.

The study results seem to indicate that the solids-separation efficiencies of the CFLUC system far exceed similar capabilities of conventional clarifiers. For example, the upflow clarifier appears to have the potential to separate severe bulking sludges (SVI: 380-388 mL/g) at similar overflow rates used in conventional clarifiers which are used to separate normal extended aeration sludges (see Table

2.3). More research is required to verify the pilot plant studies (eg. full-scale plant studies) due to the engineering significance of the above findings.

The efficiencies of the CFLUC system can be compared to advanced solids-separation techniques such as filtration, flotation, tube and lamella separators. The study results seem to indicate that the CFLUC system does approach the ideal design model proposed by Camp (1953) for an "ideal sedimentation basin".

CHAPTER 7

CONCLUSIONS

The results of this study lead to the following conclusions:

1. Based on the operating control parameters selected to simulate conditions at Souris, Manitoba, the CFLUC system can effectively treat a synthetic dairy waste at F:M ratios of 0.10, 0.24, 0.41 and 0.83 g COD/g MLVSS.d.

2. The SRT values corresponding to the aforementioned F:M ratios were determined to be approximately 40, 27, 9 and 4 days, respectively.

3. An evaluation of the biokinetic constants confirmed that steady-state operating conditions had been achieved for each operating condition.

4. Normal zoogleal sludges with pinpoint floc were observed at low F:M ratios of 0.10 and 0.24 d^{-1}. The upflow clarifier could remove approximately 70% of the pinpoint floc that remained after quiescent batch settling in a 1000 ml cylinder. The sludges at these loadings could be characterized as typical extended aeration sludges with good settling properties.

5. Filamentous bulking sludges were observed at F:M ratios of 0.41 and 0.83 d^{-1}. The upflow clarifier was still capable of separating the bulking sludge, however, the effluent quality (in terms of SESS) was similar to that occurring under quiescent batch settling tests.

6. The BOD_5 and soluble COD removal efficiencies ranged from 90 to 99% for the different operating conditions, indicating good biological activity in the aeration basin at all F:M ratios investigated.

7. Dissolved oxygen concentrations ranging from 5.0 - 6.0 mg/l are required in the aeration basin to prevent anoxic conditions in the clarifier (i.e. operating a sludge blanket at 1/3 the clarifier height).

8. A higher sludge blanket does not appear to improve effluent quality (in terms of SESS).

9. A low sludge blanket level and MLSS concentration is desirable in the operation of the CFLUC system.

10. The separation of normal zoogleal sludges in upflow clarifiers is less influenced by overflow rate as compared to recommended design values for conventional clarifiers.

11. The separation of bulking sludge is controlled by the following operating control parameters SVI, MLSS and the overflow rate.

12. The actual sludge recycle rates in the pilot plant during the separation of bulking sludges was significantly lower in magnitude (40-120 times) as compared to the calculated sludge recycle rates from batch settling tests. It is hypothesized that this phenomenon is due to bioflocculation enhancement and/or physical bio-conditioning in the downward flow channel.

13. The study results seem to indicate that the solids separation efficiencies of the CFLUC system at Souris, Manitoba would exceed previous treatment efficiencies that have been achieved by conventional clarifiers.

14. At the full-scale plant in Souris, Manitoba, there is the potential to separate all types of sludges (including filamentous bulking sludges) for a wide range of organic loading conditions. This advantage gives the operator greater flexibility in separating sludges with different settling characteristics (eg. shock loads).

15. The study results seem to indicate that the upflow clarifier incorporated in the CFLUC system simulates the ideal design model proposed by Camp (1953) for an "ideal sedimentation basin".

CHAPTER 8

SUGGESTIONS FOR FURTHER STUDY

Based on the findings of this study, the following topics are suggested as possible subjects for further study:

1. Investigate the mechanism of bio-flocculation and/or physical conditioning in the downward flow channel of the upflow clarifier.

2. Due to the engineering significance associated with separating a filamentous bulking sludge, more studies in a full scale plant are required to verify the results of the pilot plant studies.

3. Develop design criteria and guidelines for overflow and solids loading rates in the CFLUC system. These guidelines can be compared to recommended rates used in the Federal Republic of Germany.

4. Conduct studies to determine the application of upflow clarifiers in water treatment and/or the treatment of industrial wastes.

5. Apply solids flux theory to the CFLUC system in order to evaluate the potential and efficiencies of the new sludge removal mechanism.

6. Investigate methods of increasing denitrification rates in the upflow clarifier to convert the clarifier to a biological denitrification reactor.

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APPENDIX A

TABLE A.1 ACCLIMATION PERIOD STUDY RESULTS - PHASE A

TIME		AERATION BASIN pH	MLSS mg/L	MLVSS mg/L	DO mg/L	TEMP °C	NFR mg/L	SVI mL/g	UPFLOW CLARIFIER T.S. Sludge %	DO mg/L	TEMP °C	pH
Aug	24	7.5	4630	3180	2.8	21.0	46.0	-	1.0	0.3	22.5	
	25	7.5	3630	2510	5.4	22.5	25.0	220	1.5	0.7	23.5	
	26	7.4	4020	2810	-	-	15.0	-	0.9	-	-	
	27	7.5	-	-	1.5	23.5	-	-	-	0.9	24.5	
	28	7.5	-	-	-	-	-	-	-	-	-	
	29	7.5	-	-	-	-	-	-	-	-	-	
	30	7.4	3270	2390	3.8	23.5	8.2	229	0.91	0.6	24.5	
	31	6.3	3090	2225	-	-	9.0	178	-	-	-	
Sept	1	7.4	2460	1825	3.2	23.0	21.6	159	1.33	0.7	24.0	
	2	7.3	2635	1950	3.3	24.0	21.5	-	1.32	0.7	25.0	
	3	7.3	-	-	3.1	24.5	-	-	-	0.7	25.5	
	4	8.3	-	-	2.0	23.5	-	-	-	0.7	24.5	
	5	7.6	-	-	3.4	21.5	-	-	-	0.7	22.5	
	6	7.8	1950	1455	3.4	22.0	7.4	328	1.30	0.8	23.0	
	7	7.4	1985	1540	3.8	21.5	15.7	252	1.11	0.8	22.5	
	8	7.6	2000	1580	3.9	21.5	35.2	375	1.24	0.8	22.5	
	9	7.4	2375	1860	-	-	-	244	-	-	-	
	10	7.6	-	-	6.0	20.0	-	-	-	1.0	21.0	
	11	6.8	-	-	6.0	20.0	-	-	-	0.9	21.0	
	12	7.5	2580	1995	4.8	20.0	49.0	451	1.17	0.9	21.0	
	13	7.6	-	-	2.4	20.0	-	-	-	0.6	21.0	
	14	7.5	2329	1847	4.2	20.0	2.8	-	0.85	0.9	20.5	
	15	7.6	2515	1980	3.8	20.0	13.0	318	0.85	0.8	21.0	
	16	7.9	-	-	6.2	20.0	-	-	-	1.2	20.5	
	17	7.4	-	-	1.4	20.0	-	-	-	0.8	20.5	
	18	7.6	-	-	3.8	18.5	-	-	-	0.9	19.0	
	19	7.5	2310	1865	2.8	18.5	6.6	390	0.85	0.8	19.0	
	20	7.5	-	-	2.5	19.0	-	-	-	-	-	
	21	-	-	-	-	-	-	-	-	-	-	
	22	7.7	2380	1890	4.5	19.0	4.6	-	0.68	1.2	20.0	
	23	7.4	2170	1725	1.1	20.0	3.0	300	0.77	0.6	20.5	
	24	7.2	-	-	4.5	20.0	-	-	-	1.3	21.0	

TABLE A.1 (Continued)

TIME	AERATION BASIN pH	MLSS mg/L	MLVSS mg/L	DO mg/L	TEMP °C	NFR mg/L	SVI mL/g	UPFLOW CLARIFIER T.S. Sludge %	DO mg/L	TEMP °C	pH
Sept 25	7.7	2160	1785	-	-	1.4	-	0.82	-	-	-
26	7.7	2290	1660	5.2	20.0	1.0	349	0.72	1.9	21.0	-
27	7.5	-	-	4.5	20.0	-	-	-	1.8	20.5	-
28	7.8	-	-	4.5	21.0	-	-		1.0	23.0	-
29	7.6	2350	1920	2.0	20.0	6.7	383	0.70	1.9	21.0	7.6
30	7.4	-	-	-	-	-	-	-	-	-	-
Oct 1	7.6	-	-	-	-	-	-	-	-	-	-
2	7.6	-	-	-	-	-	-	-	-	-	-
3	7.6	1695	1410	-	-	4.0	324	1.0	-	-	7.5
4	7.6	2145	-	4.9	19.0	13.0	186	0.90	1.6	20.0	7.5
5	7.6	-	-	-	-	-	-	-	-	-	7.6
6	7.3	2275	1860	4.4	20.0	5.5	299	0.81	0.8	21.0	7.5
7	7.7	2135	1755	-	-	3.4	337	-	-	-	7.6
8	7.6	-	-	-	-	-	-	-	-	-	-
9	7.6	-	-	4.0	20.0	-	-	-	0.9	21.0	-
10	7.7	-	-	-	-	-	-	-	-	-	-
11	7.5	2270	1890	-	-	-	345	-	-	-	7.7
12	7.5	2370	1930	-	-	-	295	-	-	-	7.7
13	7.6	2405	1995	-	-	1.7	380	0.79	-	-	7.5
14	7.6	-	-	-	-	-	-	-	-	-	7.5
15	LOST SOLIDS (CLARIFIER UPSET)										
16	7.5	-	-	-	-	-	-	-	-	-	-
17	7.7	1800	1500	-	-	-	-	-	-	-	-
18	7.6	1945	1625	3.4	19.0	-	-	1.00	1.6	20.0	7.7
19	7.7	2005	1655	-	-	6.0	174	-	-	-	7.7
20	7.8	2325	1960	-	-	4.7	215	-	-	-	8.0
21	7.6	2410	2130	-	-	5.7	257	-	-	-	7.7
22	7.7	-	-	-	-	-	-	-	-	-	-
23	7.7	-	-	-	-	-	-	-	-	-	-
24	-	-	-	-	-	-	-	-	-	-	-
25	7.6	2690	2325	-	-	3.7	249	0.92	-	-	7.6
26	7.7	2580	2150	6.6	18.5	4.7	322	0.90	3.0	20.0	7.6
27	7.6	2355	1965	6.8	18.0	-	314	0.85	3.8	20.0	7.7

TABLE A.2 SOLIDS RETENTION TIME DURING ACCLIMATION PERIOD STUDY

DATE	Q_{WR} litres	X_R mg/L	MLVSS mg/L	CALCULATIONS
Sept 14	20	8500	1847	$\theta_C = \frac{\forall \text{ MLVSS}}{Q_{WR} X_R}$
22	20	6800	1890	
24	40	8000 E	1755 E	$Q_{WR} = 165/30 = 5.5$ L/day
28	30	7000	1920	$\forall = 547$ litres
29	30	7000	1920	
Oct 14	25	7900	1995	$\therefore Q_C = \frac{(547)(1888)}{(5.5)(7533)}$
TOTAL:	165			$\theta_C = 26$ days
AVERAGE:		7533	1889	

NOTE: The SRT of 26 days during the acclimation period study was approximately equal to the SRT of 27 days for Test Run # B2T. This was expected since the organic loadings were identical (0.24 g COD/g MLVSS.d). The duplication of results provided a check to the accuracy of the experimental study results.

APPENDIX B

TABLE B.1 ACCLIMATION RUN # B1 AC

DATE	AERATION BASIN						UPFLOW CLARIFIER				
Mth/Day	pH	DO mg/L	TEMP °C	MLSS mg/L	MLVSS mg/L	SVI mL/g	pH	DO mg/L	TEMP °C	SESS mg/L	TURBIDITY N.T.U.
OCT 28	7.5	5.3	18.3	2113	1766	296	7.5	2.7	20.3	1.9	3.1
OCT 29	7.6	5.7	18.3	2343	1898	313	7.8	2.3	20.3	5.8	5.2
OCT 30	7.6	4.9	18.0	2113	1753	308	7.6	2.2	19.0	1.3	2.2
OCT 31	7.5	4.2	18.0	2130	1753	228	7.5	2.4	19.5	2.3	2.0
NOV 1	7.5	4.7	18.0	2060	1718	181	7.6	1.6	19.8	2.2	1.5
NOV 2	7.5	4.6	18.5	2087	1690	154	7.7	2.2	20.5	3.0	1.1
NOV 3	7.4	5.5	18.0	2003	1682	218	7.5	2.4	19.5	4.5	1.7
NOV 4	7.4	5.2	18.0	1983	1655	161	7.5	3.3	19.5	3.5	3.0
NOV 5	7.4	4.4	18.0	1862	1517	154	7.5	1.9	19.8	3.0	1.2
NOV 6	7.4	4.8	18.0	1798	1603	140	7.6	1.7	20.3	5.0	2.0

TABLE B.2 ACCLIMATION RUN # B2 AC

DATE	AERATION BASIN						UPFLOW CLARIFIER					
Mth/Day	pH	DO mg/L	TEMP °C	MLSS mg/L	MLVSS mg/L	SVI mL/g	pH	DO mg/L	TEMP °C	SESS mg/L	SPSS mg/L	TURBIDITY N.T.U.
NOV 12	7.53	3.5	17.3	1535	1233	127	7.75	1.7	19.3	11.6	46.0	4.9
NOV 13	7.65	5.4	17.0	1655	1340	133	7.75	1.9	18.5	17.5	47.0	3.5
NOV 14	7.67	5.4	17.0	1570	1295	119	7.87	2.9	19.0	12.6	47.0	4.8
NOV 15	7.55	4.6	17.0	1558	1337	120	7.70	2.2	19.0	10.7	39.5	6.3
NOV 16	7.55	4.6	17.0	1693	1397	105	7.63	2.1	19.0	10.5	40.5	5.8
NOV 17	7.55	6.2	16.5	1810	1448	100	7.78	2.2	18.7	16.2	54.5	8.3
NOV 18	7.65	6.1	17.0	1843	1537	103	7.75	2.0	19.0	19.4	62.5	6.5
NOV 19	7.60	5.9	17.0	1670	1340	119	7.65	2.9	19.0	16.5	48.0	-
NOV 20	7.70	6.3	16.0	1790	1540	117	7.80	2.9	18.0	18.5	46.0	9.2
NOV 21	7.67	6.1	16.0	1735	1418	114	7.83	3.4	18.3	17.4	44.0	8.4

TABLE B.3 ACCLIMATION PERIOD RUN # B3 AC

DATE	AERATION BASIN						UPFLOW CLARIFIER				
Mth/Day	pH	DO mg/L	TEMP °C	MLSS mg/L	MLVSS mg/L	SVI mL/g	pH	DO mg/L	TEMP °C	SESS mg/L	SPSS mg/L
DEC 2	7.53	5.6	15.8	1658	1345	115	7.60	2.0	18	14.8	53.5
DEC 3	7.50	4.6	15.0	1785	1500	126	7.68	1.4	17	14.5	42.0
DEC 4	7.63	4.8	15.0	1873	1558	144	7.70	1.2	17	16.0	40.8
DEC 5	7.73	5.6	15.0	2151	1773	172	7.73	1.5	17	15.6	49.3
DEC 6	7.73	5.5	15.0	2467	2000	368	7.73	1.4	17	14.3	14.2

TABLE B.4 ACCLIMATION PERIOD RUN # B4 AC

DATE	AERATION BASIN						UPFLOW CLARIFIER				
Mth/Day	pH	DO mg/L	TEMP °C	MLSS mg/L	MLVSS mg/L	SVI mL/g	pH	DO mg/L	TEMP °C	SESS mg/L	SPSS mg/L
* DEC 12	7.55	4.7	15.0	2117	1755	449	7.67	1.9	17.3	2.3	-
* DEC 13	7.60	4.5	15.3	2165	1708	448	7.65	1.5	17.5	6.0	-
* DEC 14	7.65	4.1	16.3	1895	1562	504	7.70	1.5	17.5	6.8	-
DEC 15	7.35	2.3	17.0	2448	1993	398	7.32	-	-	4.9	-
DEC 16	8.00	3.5	16.7	2500	2030	383	7.85	1.0	19.3	5.2	-
DEC 17	8.00	4.8	16.0	2975	2278	322	8.03	1.1	18.8	10.6	-
DEC 18	8.00	4.2	15.7	3085	2345	308	7.95	1.0	18.0	13.7	-
DEC 19	7.78	5.0	15.3	2976	2375	321	7.88	1.4	17.7	14.0	-
DEC 20	7.63	5.5	15.0	2640	2213	335	7.60	1.1	17.0	13.0	-
DEC 21	7.75	4.8	15.0	2768	2278	310	7.75	1.2	16.7	15.8	-
DEC 22	7.70	4.4	15.0	2865	2475	315	7.77	1.4	16.7	15.4	-
DEC 23	7.73	4.6	15.0	2578	2170	307	7.78	1.0	17.0	10.4	10.7
DEC 24	7.75	4.4	15.0	2610	2083	321	7.80	1.1	16.8	11.0	9.7
DEC 25	7.65	4.6	15.0	2760	2103	300	7.70	1.3	16.5	14.4	19.7
DEC 26	7.65	4.3	14.8	2873	2395	316	7.65	1.1	17.0	31.3	-
DEC 27	7.63	3.5	14.5	2666	2168	327	7.78	1.1	16.5	31.5	30.5
DEC 28	7.68	3.8	15.0	2632	2468	351	7.75	0.9	17.0	27.0	36.0

* Note Frozen Feed

APPENDIX C

TABLE C.1 BOD AND COD ANALYSIS - RUN # B1 AC AND B1 T

	DATE	INFLUENT (COD)				EFFLUENT (COD)			INFLUENT (BOD)			EFFLUENT BOD
		(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)	(9)	(10)	
TEST PERIOD	MTH/DAY	FEED mg/L	MIXED LIQUOR mg/L	FILTERED M.L. mg/L	REMOVAL %(1)-(3) x100(1)	EFFLUENT mg/L	FILTERED EFFLUENT mg/L	REMOVAL % (1)-(5) x100(1)	FEED mg/L	BOD/COD (1)/(8)	EFFLUENT mg/L	% REMOVAL (8)-(10) x100(8)
# B1AC	OCT 28	180	2360	25	86.1	25	25	86.1				
	OCT 29	180	2600	30	83.3	42	42	76.7	124	.69		
	OCT 30	180	2500	29	83.9	20	27	88.9				
	OCT 31	180	2300	24	86.7	24	28	86.7				
	NOV 1	175	2500	30	82.9	30	30	82.9	110	.63		
	NOV 2	155	2490	30	80.6	37	38	76.1				
	NOV 3	170	2300	35	79.4	33	-	80.6				
	NOV 4	140	2264	28	80.0	43	35	75.0				
	NOV 5	143	2148	35	75.5	40	19	79.7				
	NOV 6	146	1900	30	79.5	28	25	82.9	100	.68		
# B1T	NOV 7	135	1900	40	70.4	36	33	75.6	89EST	.66EST	5	94EST
	NOV 8	135	1828	37	72.6	46	29	78.5	86EST	.64		
	NOV 9	121	1628	35	71.1	49	33	72.7	80EST	.66EST	3	96EST
	NOV 10	133	1940	36	72.9	39	33	75.2	88EST	.66EST		
	NOV 11	137	1938	37	73.0	66	30	78.1	90EST	.66EST	6	93EST
AVERAGE # B1T Steady-State		132	1847	37	72.0	47	32	76.0	87EST	.66*	5	94EST

*NOTE: (EST) - Estimate
Average of Oct 28, Nov 1, Nov 6 and Nov 8. Used in calculating Estimated Influent BOD volume and % removal values.

TABLE C.2 BOD AND COD ANALYSIS - RUN # B2 AC AND B2 T

	DATE	INFLUENT (COD)				EFFLUENT (COD)			INFLUENT (BOD)			EFFLUENT BOI
TEST PERIOD	MTH/DAY	(1) FEED mg/L	(2) MIXED LIQUOR mg/L	(3) FILTERED M.L. mg/L	(4) REMOVAL %(1)-(3) x100(1)	(5) EFFLUENT mg/L	(6) FILTERED EFFLUENT mg/L	(7) REMOVAL % (1)-(5) x100(1)	(8) FEED mg/L	(9) BOD/COD (1)/(8)	(10) EFFLUENT mg/L	% REMOVAL (8)-(10) x100(8)
# B2AC	NOV 12	450	1938	49	89.1	55	42	87.7	300	.66		
	NOV 13	458	1840	45	90.2	70	45	84.7				
	NOV 14	313	-	48	84.7	58	48	81.5			8.0	
	NOV 15	400	-	43	85.7	65	40	78.3				
	NOV 16	351	-	50	85.8	70	50	80.1			6.0	
	NOV 17	360	-	62	82.8	80	65	77.8	213	.59		
	NOV 18	390	-	35	91.0	50	35	87.2				
	NOV 19	315	-	35	88.9	50	35	84.1			6.0	
	NOV 20	310	-	40	87.1	50	40	83.9	189	.61		
	NOV 21	315	-	50	84.1	62	40	80.3				
# B2T	NOV 22	320	-	35	89.1	50	36	84.4	195EST	.61EST		
	NOV 23	313	-	37	88.2	60	36	80.8	191EST	.61EST		
	NOV 24	313	-	37	88.2	65	40	79.2	191EST	.61EST	11.0	94EST
	NOV 25	353	-	37	89.5	58	35	83.6	208	.59		
	NOV 26	343	-	37	89.2	62	37	81.9	209EST	.61EST	10.0	95EST
	NOV 27	340	-	37	89.1	62	40	81.8	208	.61		
	NOV 28	340	-	37	89.1	67	35	80.3	208EST	.61	9.0	96EST
	NOV 29	353	-	42	88.1	62	40	82.4	215EST	.61EST		
	NOV 30	348	-	35	90.0	53	36	84.8	212EST	.61EST		
	DEC 1	435	-	40	90.8	65	45	85.1	265EST	.61EST	12.0	95EST
AVERAGE # B2T Steady-State		346	-	37	89.1	60	38	82.4	210	.61*	11	95

*NOTE: (EST) - Estimate
Average of Nov 12, Nov 17, Nov 19, Nov 25 and Nov 28.

TABLE C.3 BOD AND COD ANALYSIS - RUN # B3 AC AND B3 T

TEST PERIOD	DATE MTH/DAY	INFLUENT (COD) (1) FEED mg/L	(2) MIXED LIQUOR mg/L	(3) FILTERED M.L. mg/L	(4) REMOVAL %(1)-(3) x100(1)	EFFLUENT (COD) (5) EFFLUENT mg/L	(6) FILTERED EFFLUENT mg/L	(7) REMOVAL % (1)-(5) x100(1)	INFLUENT (BOD) (8) FEED mg/L	(9) BOD/COD (1)/(8)	(10) EFFLUENT mg/L	EFFLUENT BOD % REMOVAL (8)-(10) x100(8)
# B3AC	DEC 2	450	-	47	89.6	60	45	86.7			12	
	DEC 3	540	-	45	91.7	65	46	88.0				
	DEC 4	630	-	50	92.1	74	54	88.3				
	DEC 5	700	-	47	93.3	65	47	90.7			14	
	DEC 6	800	-	50	93.8	65	45	91.9	535	.67		
# B3T	DEC 7	800	-	50	93.8	60	50	92.5	520EST	.65EST		
	DEC 8	885	-	50	94.4	60	55	93.2	575EST	.65EST	7	98.8EST
	DEC 9	894	-	35	96.1	40	37	95.5	578	.65		
	DEC 10	845	-	37	95.6	39	35	95.4	549EST	.65EST	7	98.7EST
	DEC 11	860	-	50	94.2	65	37	92.4	540	.63		
AVERAGE # B3T Steady-State		857	-	44	94.8	53	43	93.8	552	.65*	7	98.8

*NOTE: (EST) - Estimate
Average of Dec 6, Dec 9, and Dec 11.

TABLE C.4 BOD AND COD ANALYSIS - RUN # B4 AC AND B4 T

	DATE	INFLUENT (COD)				EFFLUENT (COD)			INFLUENT (BOD)			EFFLUENT BOI
TEST PERIOD	MTH/DAY	(1) FEED mg/L	(2) MIXED LIQUOR mg/L	(3) FILTERED M.L. mg/L	(4) REMOVAL %(1)-(3) x100(1)	(5) EFFLUENT mg/L	(6) FILTERED EFFLUENT mg/L	(7) REMOVAL % (1)-(5) x100(1)	(8) FEED mg/L	(9) BOD/COD (1)/(8)	(10) EFFLUENT mg/L	% REMOVAL (8)-(10) x100(8)
# B4AC	DEC 12	845	-	45	94.7	40	37	95.3				
	DEC 13	875	-	55	93.7	50	45	94.3				
	DEC 14	2223	-	-	-	-	-	-				
	DEC 15	3600EST	-	115	96.8	50	48	98.6				
	DEC 16	3600EST	-	180	95.0	200	195	94.4				
	DEC 17	3600EST	-	150	95.8	130	115	96.4				
	DEC 18	3600	-	130	96.4	130	115	96.8				
	DEC 19	2640EST	-	150	94.3	140	130	95.1			24.5	
	DEC 20	1800EST	-	120	93.3	140	138	92.2	970	.54		
	DEC 21	1800	-	-	-	125	95	93.1				
	DEC 22	1800EST	-	125	93.1	120	110	93.9				
	DEC 23	1830	-	95	94.8	110	95	94.8				
	DEC 24	1840	-	115	93.8	95	88	94.8			28.0	
	DEC 25	1740	-	115	93.4	125	100	92.8	816	.47		
	DEC 26	1820	-	175	90.4	205	155	88.7				
	DEC 27	1840	-	180	90.2	200	168	89.1				
	DEC 28	1800	-	160	91.1	190	150	89.4				
# B4T	DEC 29	1800	-	140	92.2	165	130	90.8	882EST	.49EST		
	DEC 30	1740	-	187	89.3	197	145	88.7	853EST	.49EST	40.0	95.3
	DEC 31	1880	-	170	91.0	185	140	90.2	972EST	.52		
	JAN 1	1840	-	145	92.1	200	145	89.1	902EST	.49EST		
	JAN 2	1900	-	180	90.5	215	155	88.7	882EST	.46EST	42.0	95.2
	JAN 3	1900	-	140	92.6	175	130	90.8	931	.49		
	JAN 4	-	-	125	-	175	-	-	-	-	24.0	-
	JAN 5	1800	-	130	92.8	140	105	92.2	815	.45		
AVERAGE # B4T		1837	-	152	91.5	182	136	90.1	906	.49	35.3	95.3

TABLE C.5 SOLIDS ANALYSIS - RUN # B1 AC and # B1 T

TEST PERIOD	DATE Mth/Day	FLUIDIZED BED HEIGHT cm	(1) TS (X_R) %	TVS %	RECYCLE (Q_R) L/h	$\frac{Q_R}{Q}$ %	(2) WASTING (Q_W) L/d	(3) MLVSS (X) mg/L	(4) SRT Days	MICROSCOPIC EXAMINATION
# B1AC	OCT 28	45	0.69	0.55	8.8	.37	-	1766	-	During test period # B1 AC
	OCT 29	35	0.88	0.69	7.8	.33	-	1898	-	populations included
	OCT 30	44	0.95	0.72	8.8	.37	-	1753	-	stalked cilicites,
	OCT 31	45	0.82	0.66	7.8	.33	-	1753	-	rotifers, nematodes.
	NOV 1	70	0.94	0.78	9.7	.37	-	1718	-	Filamentous microorganisms
	NOV 2	50	0.88	0.70	9.7	.37	-	1690	-	began to decrease on
	NOV 3	48	0.89	0.72	9.7	.37	-	1682	-	November 4th.
	NOV 4	45	0.93	0.75	8.8	.37	3	1655	-	
	NOV 5	44	1.01	0.82	7.8	.33	3	1517	-	
	NOV 6	44	.84	0.68	8.8	.37	-	1603	-	
# B1T	NOV 7	43	.93	0.73	6.8	.28	-	1368	-	Many rotifers, less stalked
	NOV 8	33	1.00	0.81	3.9	.16	-	1335	-	ciliates and little
	NOV 9	38	1.08	0.86	4.9	.20	-	1255	-	filamentous during
	NOV 10	33	1.09	0.87	4.9	.20	-	1337	-	test period B1 T.
	NOV 11	20	.80	0.62	3.9	.16	-	1255	-	
AVERAGE # B1 T Steady-State		33	.98	.78	4.3	.20	0	1310		Death rate = growth rate no wasting

TABLE C.6 SOLIDS ANALYSIS - RUN # B2 AC and # B2 T

TEST PERIOD	DATE Mth/Day	FLUIDIZED BED HEIGHT cm	TS (X_R) %	TVS %	RECYCLE (Q_R) L/h	$\frac{Q_R}{Q}$ %	WASTING (Q_W) L	MLVSS (X) mg/L	SRT Days	MICROSCOPIC EXAMINATION
# B2AC	NOV 12	32	1.04	0.81	3.9	0.16	-	1233	-	More rotifers than stalks
	NOV 13	44	0.99	0.77	6.8	0.28	-	1340	-	little filamentous
	NOV 14	34	1.20	0.92	5.8	0.24	-	1295	-	no nematodes
	NOV 15	23	1.27	1.01	2.9	0.12	-	1337	-	More stalked ciliates
	NOV 16	21	1.33	1.05	3.9	0.16	-	1397	-	than rotifers. Pinpoint
	NOV 17	34	1.22	0.95	5.8	0.24	-	1448	-	floc particles visible
	NOV 18	30	1.02	0.79	5.8	0.24	-	1537	-	in effluent. Crisp
	NOV 19	24	1.32	1.04	2.9	0.12	-	1340	-	white foam in aeration
	NOV 20	33	0.79	0.61	5.8	0.24	-	1540	-	basin.
	NOV 21	25	1.14	0.88	4.9	0.20	-	1418	-	
# B2T	NOV 22	33	1.10	0.85	4.9	0.20	4	1523	26	Many free swimmers,
	NOV 23	28	1.22	0.94	4.9	0.20	4	1460	22	stalked ciliates and
	NOV 24	33	1.22	0.94	3.9	0.16	4	1437	22	rotifers evident.
	NOV 25	35	1.32	1.04	4.9	0.20	-	1415	-	Mixed culture of
	NOV 26	31	1.11	0.87	5.8	0.24	3	1507	33	microorganisms
	NOV 27	31	1.37	1.09	5.8	0.24	3	1485	26	little to no filamentous
**	NOV 28	24	1.23	0.98	4.9	0.20	**	1452		pinpoint floc
**	NOV 29	39	1.10	0.94	3.9	0.16	**	1327		evident in effluent
	NOV 30	36	1.13	0.91	3.9	0.16	3	1553	33	
	DEC 1	39	1.26	1.00	3.9	0.16	3	1405	27	
AVERAGE # B2T Steady-State		33	1.21	0.96	4.7	0.20	TOTAL 24	1456	27	$SRT = \frac{\bar{V}x \text{ MLVSS}}{Q_W \times TVS}$

** Reactor upset - excluded wasting data

TABLE C.7 SOLIDS ANALYSIS - RUN # B3 AC and # B3 T

TEST PERIOD	DATE Mth/Day	FLUIDIZED BED HEIGHT cm	TS (X_R) %	TVS %	RECYCLE (Q_R) L/h	$\frac{Q_R}{Q}$ %	WASTING (Q_W) L	MLVSS (X) mg/L	SRT Days	MICROSCOPIC EXAMINATION
# B3AC	DEC 2	36	1.00	0.78	4.9	0.20	3	1345	-	Transition from little
	DEC 3	29	0.65	0.49	4.9	0.20	4	1500	-	filamentous to considerable
	DEC 4	44	1.00	0.78	5.9	0.24	5	1558	-	filamentous on December 6th.
	DEC 5	42	0.94	0.73	7.8	0.31	11	1773	-	Few rotifers, more stalked
	DEC 6	63	0.89	0.67	9.7	0.39	15	2000	-	ciliates
# B3T	DEC 7	70	0.67	0.50	13.1	0.52	25	2075	10	Sludge bulking supernatant
	DEC 8	85	0.69	0.52	24.6	0.58	30	2110	8	clear, considerable
	DEC 9	85	0.67	0.50	14.6	0.58	30	2070	8	filamentous, many free
	DEC 10	87	0.56	0.41	14.6	0.58	30	2060	10	swimmers, few rotifers,
	DEC 11	94	0.50	0.36	16.0	0.64	50	2039	7	some stalked ciliates
AVERAGE # B3T Steady-State		70	0.62	0.46	14.6	0.58	33	2071	9	$SRT = \frac{\forall \times MLVSS}{Q_W \times TVS}$

TABLE C.8 SOLIDS ANALYSIS - RUN # B4 AC and # B4 T

TEST PERIOD	DATE Mth/Day	FLUIDIZED BED HEIGHT cm	TS (X_R) %	TVS %	RECYCLE (Q_R) L/h	$\frac{Q_R}{Q}$ %	WASTING (Q_W) L	MLVSS (X) mg/L	SRT Days	MICROSCOPIC EXAMINATION
# B4AC	DEC 12	141	0.53	0.38	16.0	0.64	70	1755	3.8	(1) many free swimmers
	DEC 13	180	0.47	0.32	11.6	0.46	70	1708	4.4	and stalked ciliates
	DEC 14	131	0.46	0.32	7.8	0.60 *	20	1562	-	(2) considerable filamentous
	DEC 15	119	0.54	0.35	7.8	0.60	60	1993	3.9	(3) sludge a light
	DEC 16	113	0.69	0.45	5.9	0.45	40	2030	6.5	orange brown color
	DEC 17	49	0.85	0.56	5.9	0.45	40	2278	5.9	(4) sludge is bulking
	DEC 18	47	0.93	0.62	3.8	0.29 *	55	2345	4.0	(5) protozoa consuming
	DEC 19	37	1.13	0.82	3.8	0.15	50	2375	3.3	bacteria attached
	DEC 20	77	1.03	0.81	6.8	0.27	40	2213	4.0	to filamentous
	DEC 21	95	0.78	0.60	6.8	0.27	40	2278	5.5	(6) aeration basin foaming
	DEC 22	39	0.88	0.70	5.9	0.24	40	2475	5.1	
	DEC 23	126	0.95	0.76	8.7	0.35	50	2170	3.3	
	DEC 24	56	0.85	0.69	10.7	0.43	40	2083	4.3	
	DEC 25	40	0.88	0.70	8.7	0.35	40	2103	4.3	
	DEC 26	45	0.84	0.67	8.7	0.35	50	2395	4.1	
	DEC 27	33	0.88	0.73	9.7	0.39	50	2168	3.4	
	DEC 28	35	0.86	0.70	7.8	0.31	40	2468	5.1	< 4.0 avg. >
# B4T	DEC 29	44	1.17	0.99	5.9	0.24	40	2143	3.1	
	DEC 30	44	1.04	0.87	5.9	0.24	35	2163	4.1	
	DEC 31	56	0.94	0.79	5.9	0.24	35	2287	4.8	
	JAN 1	53	0.83	0.68	5.9	0.24	45	2377	4.5	
	JAN 2	51	0.88	0.73	5.9	0.24	45	2140	3.8	
	JAN 3	43	0.90	0.72	5.9	0.24	40	2290	4.6	
	JAN 4	45	0.81	0.64	5.9	0.24	40	2140	4.8	
	JAN 5	37	0.92	0.76	5.9	0.24	40	-	-	NOTE * - Reduced Q by ½
AVERAGE # B4T Steady-State		47	0.94	0.77	5.9	0.24	40	2220	4.2	

APPENDIX D

TABLE D.1 DAILY ZONE SETTLING VELOCITIES AT VARIOUS F:M RATIOS

RUN #	DATE	ZSV cm/h
B1 T	11-06	270
	11-07	270
F:M = 0.10	11-08	210
	11-09	235
	11-10	235
	11-11	411
	11-12	396
AVERAGE		290

RUN #	DATE	ZSV cm/h
B2 T	11-22	500
	11-23	389
F:M = 0.24	11-24	274
	11-25	468
	11-26	429
	11-27	444
	11-28	546
	11-29	471
	11-30	389
	12-01	396
AVERAGE		431

RUN #	DATE	ZSV cm/h
B3 T	12-07	5.4
	12-08	5.5
F:M = 0.41	12-09	3.0
	12-10	3.0
	12-11	2.7
AVERAGE		3.9

RUN #	DATE	ZSV cm/h
B4 T	12-29	4.7
	12-30	3.0
F:M = 0.83	12-31	2.1
	1-01	1.5
	1-02	1.0
	1-03	1.0
	1-04	2.5
	1-05	2.5
AVERAGE		2.3