A BENCH SCALE EXPERIMENTAL STUDY OF THE TREATMENT OF MILKING CENTRE EFFLUENT USING A SEQUENCING BATCH REACTOR

by

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ABSTRACT

Until recently, the management of milking parlour effluent has received very little attention. The wastewater produced by milking operations comprises mainly milk solids and manure and can impose environmental threats to nearby water bodies if not properly treated before disposal.

In this study, three bench-scale Sequencing Batch Biological Reactors were used to treat the UBC dairy barn milking centre wastewater. The experiment was designed to investigate the treatment efficiency of the reactors under different operating temperatures and different numbers of cycles employed per unit daily flow (for the same hydraulic retention time). Parameters studied included BOD_5, COD, Total Suspended Solids, NH_3-N, NO_2-NO_3-N and dissolved oxygen uptake.

It was concluded that very high and consistent treatment efficiency can be achieved by using a Sequencing Batch Biological Reactor to treat milking centre wastes. Over 90% BOD_5 removal was observed in the room temperature and $30°C$ reactors. Even in low operating temperatures of 3.7 and $10.5°C$, over 70% BOD_5 removal was attained. Removal of the other pollutional parameters studied was similarly excellent. Uncontrolled denitrification also occured to various degrees in all three reactors.

It was also concluded that within the range studied in this experiment (4 to 8 cycles per day), changing the number of cycles employed per unit daily flow did not have any significant effect on the treatment efficiency of the reactors.

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NOMENCLATURE AND ABBREVIATIONS

BOD or BOD_5 : 5-day Biochemical Oxygen Demand

COD : Chemical Oxygen Demand

COEFF. : Coefficient

CYC. : Cycle

D.O. : Dissolved Oxygen

EXPT. : Experiment

EFF. : Effluent

FILT. : Filtrate

INF. : Influent

ML : Mixed Liquor of Activated Sludge

MCRT : Mean Cell Residence Time

NH_3-N : Ammonia Nitrogen

NO_2-NO_3-N : Combined Nitrite & Nitrate Nitrogen

OP.PER. : Operating Period

SVI : Sludge Volume Index

S.V. : Settling Velocity

STD.DEV. : Statistical Standard Deviation

SS : Suspended Solids

TSS : Total Suspended Solids

TEMP. : Temperature

VSS : Volatile Suspended Solids

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I wish to dedicate this thesis to Maria.

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1. INTRODUCTION AND OBJECTIVES

There are increasing problems with the handling and disposal of the waste materials produced by modern dairy farms as a result of livestock concentration and increased proximity to expanding urban centres. Wastes from milking parlours comprise mainly milk residue and manure plus debris flushed from the parlour. The volume of cleaning water used is related to the size of the operation and the habit of individual operators. Great fluctuations in flow and concentration are normal in milking centre wastes.

When the manure waste of a dairy farm is handled as slurry, the wastewater from the milking parlour can be combined with the slurry for treatment (if any) and disposal (usually various land application methods). However, if the manure is to be handled in its semi-solid state, alternative treatment methods should be resorted to for the management of the milking parlour wastewater.

The high oxygen demand and the biochemical availability of milking parlour wastewater suggest biological treatment. Many biological systems have been investigated for this purpose. In general, aerobic processes have been found more satisfactory than anaerobic systems for this particular application (Lindley, 1979).

The sequencing batch reactor (SBR) is a modern version of the fill-and-draw systems which originated as early as 1914 (Irvine *et al.*, 1979). The concept was not viable in the wastewater treatment industry due to its requirement of a high degree of manual operator attention. However, with the availability of modern electronic control devices, interest in the application of SBR operations to wastewater treatment have revived, mainly as a result of the works of Irvine and Goronzy (Irvine 1979a, Goronzy 1979).

.With the operational difficulties removed, sequencing batch systems can become an attractive alternative to conventional continuous flow systems (CFS). Compared with continuous systems, SBR systems are more dynamic and flexible in terms of operation and are kinetically more advantageous (Irvine *et al.*, 1978,1979a,1980; Goronszy, 1979). The kinetics of a batch operated reactor resembles that of an ideal plug-flow reactor.

A SBR biological treatment unit operates periodically in a typical cycle of five phases : FILL (inflow of wastewater), REACT (aeration), SETTLE (quiescent sedimentation of biomass and solids), DRAW (outflow of treated effluent) and IDLE. The IDLE phase can be used to provide flexibility to the "active" phases of the cycle. Other treatment steps such as nitrification and denitrification can also be incorporated.

Although a unified approach to SBR design is yet to be developed, a number of studies have shown that the SBR is a successful system in small municipal applications (Irvine *et al.*,1979a,1983; Ketchum *et al.*, 1979). However, the applicability of the semi-batch process to specific agricultural operations has not yet been examined.

In this project, the treatment of milking-centre effluent by using the SBR system was studied in light of the following reasons :

1. Wastewater from milking centres has intermittent peak-flows (during flushing periods) and extremely small base flows. This kind of flow pattern can be easily synchronized with a semi-batch operation.

2. The highly fluctuating strength of the wastewater can be better handled by a batch system than a continuous one. As batch systems are not designed to operate under steady-state conditions, fluctuations in the forcing function can be better accomodated.

3. A SBR is kinetically similar to an ideal plug flow reactor and requires

only a fraction of the volume that would be required by a continuous flow reactor with similar substrate removal capacity. First cost analysis (Ketchum, 1979) has also shown that a SBR is economically superior to a CFS. A SBR system is therefore an economically and spatially attractive treatment option for small dairy farms located close to urban centres.

4. Semi-batch operations have always been familiar to designers of Chemical Engineering processes. With the operational difficulties now removed, it will very likely receive more attention in the wastewater treatment field. It is good timing now to examine its potential role in the agricultural industries which are facing increasing environmental problems with their operations.

OBJECTIVES

The primary objective of this study is to investigate the effect of temperature on the overall treatment efficiency of a Sequencing Batch Biological reactor receiving milking-centre effluent from the University of British Columbia Dairy Barn.

The secondary objective of this project is to investigate whether the number of treatment cycles employed to treat the same amount of wastewater flow has any significant effect on the treatment efficiency of a SBR.

2. LITERATURE REVIEW

2.1 MANAGEMENT AND CHARACTERISTICS OF MILKING-CENTRE WASTES

The liquid milking-centre wastes can be incorporated into the manure handling system of the dairy farm or treated separately. The combined management system is only recommendable if the manure is handled and treated as a slurry. Otherwise, oxidation ponds, spray irrigation and other direct land application methods have traditionally been used to handle the milking-centre wastewaters. The less legitamite method of direct discharge into nearby ditches and natural water bodies is also widely practiced in a lot of places. Experience with septic tanks and subsurface disposal fields have not been successful due to rapid plugging of soil leaching beds (Loehr 1977; Lindley 1979).

Direct land application of dairy wastes is an acceptable and simple method if sufficient land is economically available. Approximately 2.1 acres of grassed area is recommended (Muchmore *et al.* 1976) to be an acceptable area for a 60-head dairy operation except under heavy rainfall conditions.

The characteristics of milking-centre discharges can vary significantly in both concentration and quantity. They are affected by various factors such as the existence or absence of grates in the parlour, whether manure scraping is carried out before washing, and the type of cleaning and sanitizing equipment used.

Thirty-nine farmers in Connecticut surveyed by Lindley (Lindley 1979) estimated an average water usage of 9.7 l/day/cow in milking parlours. The average total solids and BOD_5 estimated from this flow-rate were 2400 mg/l and 1024 mg/l respectively. Lindley's own measurement from a

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dairy farm milking 140 cows averaged 15.5 l/day/cow in water usage, 1050 mg/l in BOD_5 concentration and 3875 mg/l in Total Solids (TS) concentration. His summary of other researchers' reports showed a wide range of average water usage from 6.8 l/day/cow to 189 l/day/cow.

2.2 TREATMENT OF DAIRY WASTES BY BATCH AERATION

Hoover and his co-workers (1951) observed the intermittent flow pattern of dairy wastes and proposed a fill-and-draw rapid aeration process with centrifugal separation as a treatment alternative. In their study (Hoover *et al.* 1951), a 12-litre tank was used to treat 9.6 litres of simulated dairy waste per day, at 30 °C. A reduction in *total* COD and BOD_5 of 50-60 % and 75 % respectively was observed. The corresponding COD and BOD_5 removal in the supernatant were 89 and 92 % respectively. A complete solids balance confirmed that about 50 % of the milk solids was assimilated, with the remaining 50 % oxidized to gain energy for this assimilation. The rate of assimilation (Hoover 1954) of the milk solids into the sludge mass was reported to be ten times the rate of oxidation when milk solids and sludge solids were aerated in the proportion of one to one.

Large quantities of COD can be stored in the sludge as glycogen-like substances (Porges 1955). These stored products can subsequently go through the process of assimilation and endogenous respiration. Porges used one gallon of 500 ppm sludge to act upon 4 gallons of dairy waste (1000 ppm milk) and found that conversion to cell material was completed in 6 hours and the sludge reduction rate was 1 % per hour from then on, with oxygen demand decreased to about 10 % of the initial rate.

A comparison of continuous-flow, daily fill-and-draw and batch aeration sludge digesters at low temperatures was carried out by Mavinic and Koers (1977). They found that the fill-and-draw operation was more efficient than the continuous-flow digester, in terms of VSS reduction, at 5 °C.

2.3 SEQUENCING BATCH REACTORS

Sequencing Batch Reactors (SBR) originally appeared as the fill-and-draw system in 1914 and 1915 when Arden and Lockett first discussed the concept of activated-sludge (Goronszy 1979). The semi-batch operation was soon replaced by continuous systems which minimized operational attention and diffuser clogging problems (Irvine et al. 1979b).

Hoover and Porges (Hoover 1951, 1953; Porges 1955, 1960) revived interest in using semi-batch operations to treat dairy wastes during the 1950's. However, this resurgence of interest was not sustained. The present-day interest in SBR started at the University of Notre Dame by Irvine and his co-workers. Their modern version of the semi-batch system emphasizes the employment of readily available control devices to overcome SBR's intrinsic demand of operational attention.

The contemporary concept of Sequencing Batch Biological Reactors developed by Irvine and his associates (Irvine 1979a, 1979b; Ketchum 1978) comprises five typical periods : FILL (receiving of raw waste), REACT, SETTLE, DRAW (outflow of effluent) and IDLE. A one-tank system can be used to handle intermittent flows and a multiple tank system can be used to handle continuous flows (Ketchum et al. 1979a). More sophisticated control devices such as level sensors, dissolved oxygen probes and turbidity meters can also be employed for a more flexible and dynamic

operation.

Numerous operational strategies can be adopted for a SBR; however, the effects of adjusting the operational variables can be quite different from those of continuous systems. Sludge age, an important operational parameter in continuous flow systems, does not play the same role in semi-batch operations. Kunz and Landis (Irvine *et al.* 1979) have shown that the actual performance of fill-and-draw reactors remained the same for sludge ages between 1-7 days.

Food to microorganism ratio (F/M), another major operational variable in continuous systems, can only reflect the loading rate of a SBR in a very crude manner because of the changing MLSS concentration and the anoxic conditions during FILL.

The filling rate of a SBR operation is a process variable that can affect the overall performance of the reactor. The kinetics of a relatively long and aerated fill period (Irvine *et al.* 1979) approximate that of a continuous flow system with variable volume. A SBR with a relatively short FILL phase resembles, kinetically, the steady-state conditions of a plug-flow reactor.

High substrate tension has been shown (Chudoba *et al.* 1973a,b) to be an effective control on the development of filamentous organisms. A high substrate tension can easily be created in a SBR by eliminating aeration during FILL to promote formation of storage products in the sludge (Ketchum *et al.* 1978, 1979a) and to improve settleability of the sludge mass. It should, however, be noted that these storage products should be utilized during REACT for proper system operation.

Irvine *et al.* (1980) utilized a cycle with 2h anoxic FILL, 3h aeration, followed by 3h anoxic stir and found that soluble organic carbon (measured

as Total Organic Carbon) removal was accompanied by a rise in organic carbon within the organism in the form of glycogen. They supposed that this stored glycogen acted as the electron donor during the anoxic stir period when both glycogen and oxidized nitrogen were utilized. They also noted that if the FILL period was kept anoxic, ammonia production at the beginning of aeration can exceed the rate of ammonia utilization.

Hissett et al. (1982) studied the effect of temperature on oxygen consumption during batch operation of piggery slurry at temperatures between 5 and 50 °C. They found that at temperatures between 5 and 40 °C, a shorter time period was required to reach peak microbial respiration rate at higher temperatures. Consequently, shorter treatment time was required. However, the variation in oxygen demand with time during aeration was higher at higher temperatures and would result in less efficient use of energy if the peak oxygen demand was to be met. They noted that more efficient use of energy can be accomplished at temperatures of 15 °C and below at the expense of longer total treatment time.

The application of SBRs in the treatment of specialized wastes has been studied. Alleman et al. (1979) used a simulated high-strength industrial wastewater consisted of Trypticase Soy Broth and found that at a Mean Cell Resident Time (MCRT) of four days, over 90% organic carbon removal was achieved but nitrification was nil. The reason given was that at short MCRT, the slow-growing nitrifiers were washed from the system. With the MCRT increased to 10 days, their reactor maintained 98+% oxidation of both the carbon and nitrogen species with a 2-hour anoxic FILL and a 4-hour aerobic react cycle. A transient load analysis (by increasing the influent waste strength by a factor of two for three cycles) conducted in the same study showed negligible change in carbon removal efficiency but

nitrification could not go to completion.

A maximum specific denitrification rate of 0.17 mg N/day/mg MLVSS was obtained (Alleman and Irvine 1978) by using Trypticase Soy Broth and a 6-litre reactor. Significant endogenous substrate utilization was observed during the denitrification period. Alleman and Irvine also noted that the majority of carbon uptake occured within the first 20-30 minutes of aeration.

Silverstein and Schroeder (1983) used a SBR with 4 to 6 hours of unaerated stir and accumulated endogenous substrate to achieve denitrification. Thay obtained a maximum denitrification rate of 0.014 g N/g MLSS.d. As much as 75% of organic compounds could be removed during the stir-only anoxic FILL by an adsorption process — the rate of the process appeared to be zero order with respect to substrate concentration and first order with respect to solids concentration. Storage of endogenous substrate in the sludge mass was evident but did not appear to be glycogen as Irvine et al. (1980) suggested.

Besides the activated-sludge process, other specific applications of SBRs have been studied. A laboratory study using Sequencing Batch Reactors for phosphorous reduction by chemical treatment (Ketchum and Liao 1979b) indicated the possibility of significant savings in chemical costs and tighter control over effluent quality.

In rural applications, a sequencing batch operated lagoon can eliminate algae growth because of its high mixed liquor solids content. Experimental results have shown that average BOD_5, SS and NH_3-N removal were all above 90 % (Irvine 1979b).

A two-tank SBR municipal treatment plant (Irvine et al. 1983) was converted from the existing continuous flow activated sludge plant in

Culver, Indiana (daily average design flow was 1400 cu.m/d). The converted SBR plant maintained the secondary effluent quality for an 18-month long evaluation period and was permanently adopted by the town of Culver. Operation experience at Culver showed that denitrification occured simultaneously with nitrification during FILL and REACT. A skimming device was also deemed necessary during cold weathers to avoid freezing of the scum build-up on the surface of the reactors.

First cost analysis (Ketchum et al. 1979a) carried out to compare the initial investment costs of sequencing batch operations and other conventional alternatives revealed that a SBR system is an economically attractive option. The cost summary calculated by Ketchum et al. is reproduced below :

ALTERNATIVE TREATMENT PLANT COST SUMMARY (Ketchum et al. 1979a)
Estimated Initial Investment Cost (US$)

	Sequencing Batch	Activated Sludge	Nonaerated Lagoon	Aerated Lagoon	Packaged Plant
Rural Community	101,000		232,000		153,000
Small Town	1,054,000	1,391,000	1,065,000	415,000	

The comparison was focused on a small rural community (design flow 0.004 cu.m/s) and a small town (design flow 0.04 cu.m/s). Because of their relatively small size, they are more likely to venture into new concepts of treatment facilities.

A 200-litre SBR system was installed at the University of British Columbia Dairy Barn to treat the milking centre wastewater. An average treatment efficiency of 86.5 % BOD_5 removal, 90.8 % SS removal and 61.8 % total nitrogen removal was achieved during a 8-month trial

operation period (Lo *et al.* 1985).

This literature review has shown that almost all the researches done to-date on Sequencing Batch operations have indicated that the SBR concept is a viable and economically attractive alternative to the conventional continuous flow activated-sludge process in BOD_5, SS and nitrogen removal, and chemical precipitation of phosphorus.

Although there are still uncertainties in the basis of SBR design and full-scale operation experience is still insufficient, these deficiencies can be overcome. The dynamic and flexible nature of SBR systems allows ample room for expansion and operational adjustments at minimal costs.

3. MATERIALS AND METHODS

3.1 INTRODUCTION

The major obstacles that prevented semi-batch operations such as Sequencing Batch Reactors (SBR) from dominating the wastewater treatment industry were difficulties in operational control and strategy planning. It is therefore imperative that due considerations are given to such factors when designing bench-scale or pilot-scale SBRs for investigative purposes. The specific cycle is one such operational strategy studied in this investigation.

The specific cycle is defined here as the total number of cycles employed to treat the same volume of wastewater per 24-hour period.

One often questioned practice in laboratory bench-scale studies of biological waste-treatment systems is the use of synthetic feed or substrate. Due to the sensitive and dynamic nature of biological treatment units, a system that operates successfully in treating synthetic wastes is not guaranteed a comparable performance in real-life situations where both concentration and composition of the wastewater can vary greatly. It was therefore decided to use "natural" milking-centre wastewater as the feeding material in this experimental study. The milking-centre effluent from the University of British Columbia (UBC) dairy barn was collected every four days. The wastewater collected was passed through a U.S. Series No.50 TYLER screen (0.295 mm openings) to remove the coarser solids (mainly undigested hay, bedding materials etc.) before storage in 4 °C. A maximum of four days of storage was allowed.

The UBC dairy barn milking room handles an average of 45 cows per milking cycle and the wastewater it produces includes the bulk milk tank flushwater and the clean-in-place (CIP) rinse water, which includes

detergent. These cleaning waters are discharged to a floor drain, carrying along with them some cow excrement, spilled milk, and feed. Two 246 l collection tanks were built at the outfall of the milking room floor drain (at the main waste-trench) to intercept this wastewater. The collected waste was then transferred, periodically, by a timer controlled 373 W submersible pump to two 246 l storage tanks. A 187 W centrifugal pump in turn transferred the wastewater to a pilot-scale SBR built for a seperate project. It is at this discharge point that the wastewater was collected for the bench-scale reactors used in this study.

A review of the literature showed that a lot of the research done on activated-sludge systems used filtrate parameters as the bases for observations. However, it is deemed more realistic to use supernatant parameters as the *treatment* indices. It is the supernatant that goes into the receiving body, not the filtrate. For example, an activated-sludge unit that produces an effluent of zero soluble BOD might cause more environmental damage than a system which removes only 50% of the soluble BOD if the supernatant of the former contains much more organic suspended solids than the latter one. For this reason, all parametric values reported in this study are supernatant values unless specifically stated otherwise.

3.2 EXPERIMENTAL SET-UP AND OPERATION

Three bench-scale reactors were set-up in the Bio-Resource Engineering waste-treatment laboratory of UBC. A schematic of the system can be found in fig. 3.01.

The three reactors, denoted as A, B and C, were frabricated from plexi-glass tubes 460 mm in height and 138 mm in diameter. The low temperature reactor "A" was seated in a 190 mm diameter plexi-glass

FIG. 3.01 : SCHEMATIC OF EXPERIMENTAL SET-UP
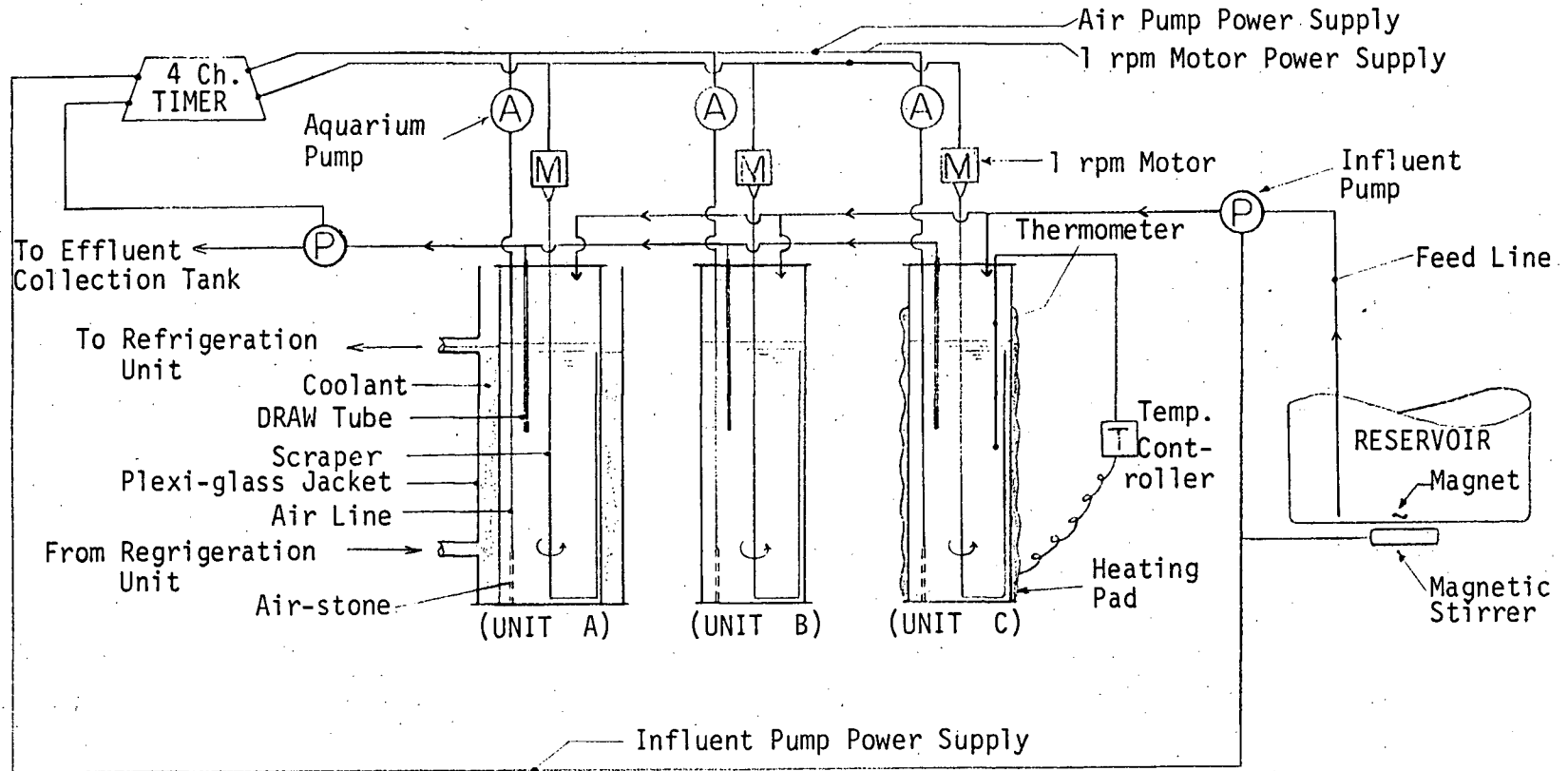
"jacket" which contained continuously circulated coolant from a Julabo F40 refrigeration unit. The high-temperature reactor "C" was wrapped by a heating pad connected to a temperature-feed-back controller. A thermometer dipped into reactor "C" was connected to the controller and triggered it to come on and off at approximately 29.9 $^{\circ}$C and 30.1 $^{\circ}$C respectively. The temperature of reactor "B" was not controlled by any means other than the air-conditioning of the laboratory which maintained an average ambient temperature of 21.8 $^{\circ}$C during the experimental period.

The entire experimental period lasted approximately 6 months. This period was further divided into 6 sub-periods; periods II, III and IV each had its own characteristic cycle length, yet all were treating the same daily amount of wastewater in order to study the effect of specific cycle. Period I was the start-up period. Periods V and VI were special periods included to test the performance of the reactor at very low · temperature (3.8 $^{\circ}$C) and high loading rate (influent flow-rate = 3.6 times the reactor volume per day). Table 4.01 summarizes most of the characteristics of these sub-periods. During all the experimental periods, the three reactors were subjected to the same amount of feed and the same cycle mode. Four operation cycles were employed throughout the experimental periods : 6 h cycle, 4 h cycle, 3 h cycle and a 2 h cycle. Each of the cycle modes (except for the 2 h cycle mode in which the IDLE phase was deleted) comprised all of the following five phases : FEED, REACT, SETTLE, DRAW and IDLE. A schematic representation of these cycle modes can be found in figure 3.02. Except wasting, all executions of the 5 phases were controlled automatically by a 4-channel table-top "ChronTrol" digital timer that included four independent AC receptacles.

TIME SCALE (h)

0   1   2   3   4   5   6

| F| REACT : 3 h 30 min | SETTLE 1 h 30 min | D | I | 6 h CYCLE MODE OP.PER. II & V

| F| REACT : 2 h 20 min | SETTLE : 1 h| D |I| 4 h CYCLE MODE OP.PER. III

F| REACT : 1 h 45 min | SETTLE | D| I| 3 h CYCLE MODE OP.PER. IV

|F | REACT 45 min| SETTLE   D| 2 h CYCLE MODE OP.PER. VI
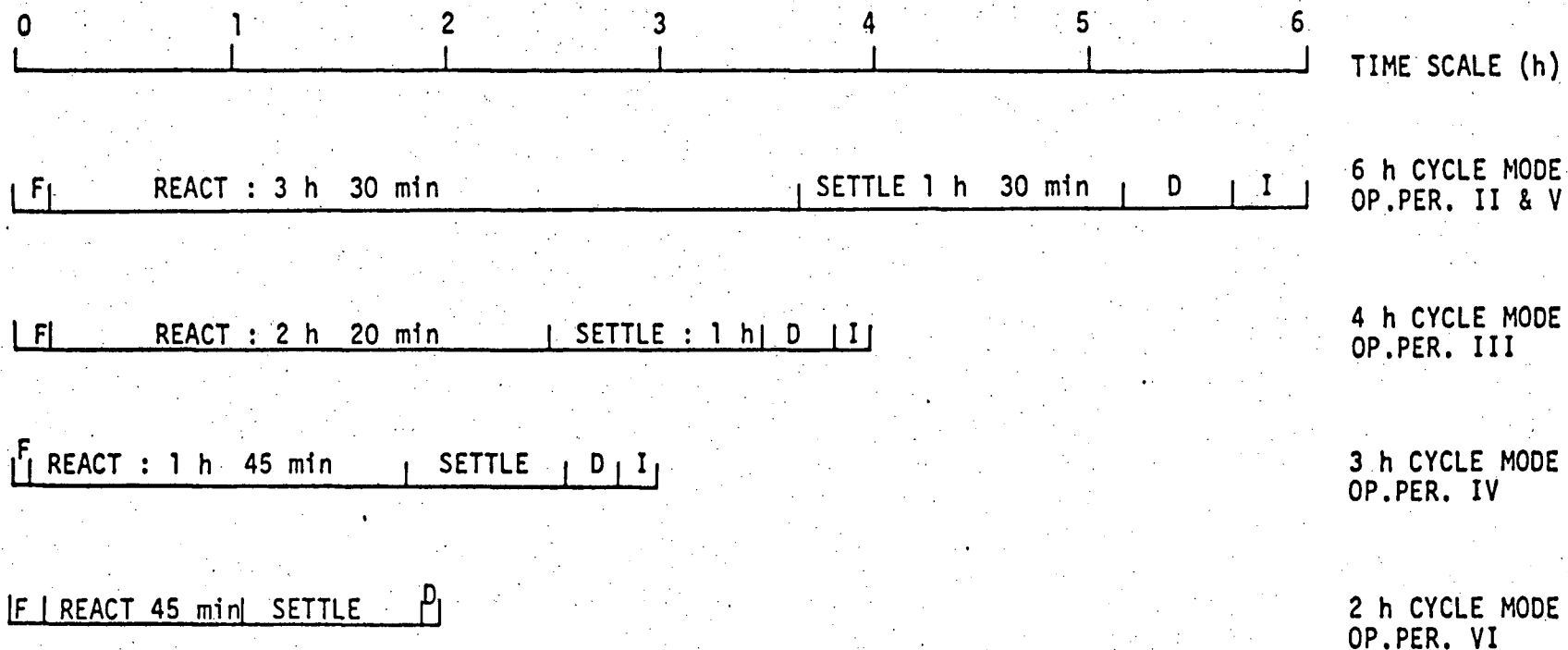
F : Non-aerated FILL period (1.5, 1.0, 0.75 & 1.5 litres/cycle for 6, 4, 3 & 2 h modes respectively)

D : DRAW phase; volume withdrawl per cycle same as FILL.

I : IDLE phase; non-aerated.

FIG. 3.02 : CYCLE MODES OF THE EXPERIMENT

The operational details of the five phases were as follows :

FEED : The timer activated a triple-headed "Cole Parmer" peristaltic pump to deliver the appropriate amount (depending on the cycle) of wastewater from the reservoir to the reactors. A magnetic stirrer was activated simultaneously to mix the content in the feed reservoir. The feed inlets were rotated among the reactors on a daily basis to ensure that the reactors received the same average amount of feed over a long period of time, since one of the pump heads was constantly delivering about 3 % less than the other two heads despite repeated adjustments.

REACT : The feed pump was switched off by the timer which subsequently activated the aquarium pumps in approximately one minute.

Completely mixed condition was confirmed by visual inspection of the mixing pattern of 2x2x5 mm plastic chips and water during setting-up of the apparatus, and by using a submersible dissolved-oxygen probe to ensure uniform readings throughout the reactors during the experimental periods.

SETTLE : The 1-rpm motors were activated during sedimentation. Stainless steel wires connected to the motors continuously scraped the inner circumference of the reactors to assist flocculation and reduce arching effects due the the slenderness of the reactors.

DRAW : A triple-headed paristaltic pump drew the final supernatant effluents from the reactors through stainless steel tubes fixed at an appropriate level in the reactors. A few minutes of extra

draw time was allowed to ensure that the final draw-down level as controlled by the level of the tubes was reached before the effluent pump was turned off by the timer. The final effluents were ordinarily directed to a waste-tank for disposal except when sample collection was required for analysis.

IDLE :   The reactors simply sat IDLE after the draw pump was turned off, awaiting the beginning of the next cycle.

Filling of the reservoir (which contained one day's supply for operation periods I to V) and emptying of the effluent waste-tank were carried out manually on a daily basis. Wasting was also carried out manually by discharging the appropriate amount of mixed-liquor (according to the operation sludge-age) on a regular basis.

3.3 PARAMETRIC ANALYSIS

3.3.1 SAMPLING METHOD

Two sampling approaches were adopted for the analysis of the parameters : the influent-effluent analysis and the track analysis.

Influent-effluent analysis involved sampling of the feed from the reservoir at the time of FILL and the effluents produced by the reactors at time of DRAW. The influent samples were usually collected from the reservoir bucket, at a point right next to the inlets of the influent-pump tubing, in the middle of FILL. Enough wastewater was siphoned into a clean beaker first, then mixed and redistributed for preservation, or analysed immediately.

The effluent samples were collected by placing the discharge tubes of the reactors into three seperate 2 l beakers before DRAW. The batches of effluent produced by the reactors were usually collected in entirity. They were then stirred thoroughly in the beakers, then preserved or analysed right away.

For the track analysis samples, the following procedure was followed :

- Effluents from the *previous* cycle were collected.

- The influent feed was collected during FILL as previously described.

- Aeration of the three reactors were off-set by one minute, by delaying activation of the aquarium pumps manually.

- 120 ml of mixed-liquor (ML) from each reactor were siphoned into three seperate 120 ml beakers at various time intervals. Shorter sampling time-intervals were used in the beginning than at the end of the react phase. The first samples were generally taken within the first two minutes of aeration but at least one minute of aeration would be given to allow for sufficient mixing before sampling.

- Except for ML suspended-solids analyses, the samples collected were allowed to settle in the beaker for the designated SETTLE time. Samples of the supernatant were then pipetted out carefully for preservation or immediate analysis.

All BOD_5 and suspended-solids analyses were carried out immediately. If COD, NH_3-N and NO_2-NO_3-N analyses could not be carried out on the same day, they would be preserved in accordance to the procedures recommended in the Standard Methods For The Examination of Water and Wastewater (1975).

3.3.2 SAMPLE ANALYSIS

All 5-day Biochemical Oxygen Demand (BOD_5), Chemical Oxygen Demand (COD), Total Suspended-Solids (TSS), Volatile Suspended-Solids (VSS) and settling velocity analyses were carried out in accordance to the procedures recommended in the Standard Methods (1975) except for the following modifications :

1. TSS and VSS : Suspended-solids analyses were carried out regularly (usually every other day). Due to this high frequency of sampling, the tests were not done in duplicate. During the TSS and VSS track analyses and specific dissolved-oxygen uptake rate experiments, however, at least two duplicate. tests were carried out for each sample collected.

2. Settling velocity : A 385 ml plexi-glass cylinder (44.3 mm diameter) was used instead of the recommended 1-litre size because of the limited size of the bench-scale reactors. It was deemed not desirable to remove too much ML from the reactor for the period of time required to conduct the settling velocity test (usually over one hour during REACT for a duplicated settling velocity test).

Ammonia nitrogen (NH_3-N) and combined nitrite-nitrate nitrogen (NO_2-NO_3-N) of all samples were analysed with a Technicon Auto Analyzer II in accordance with the procedures recommended by the manufacturer.

3.3.3 DISSOLVED OXYGEN UPTAKE RATE

The dissolved oxygen (D.O.) uptake rate of the activated-sludge of the reactors was measured by using a submersible D.O. probe and a

YSI model 54 D.O. meter connected to a chart recorder. The chart speed was set to 1.0 cm/min.

The D.O. probe was carefully lowered to mid-depth of the reactor during the IDLE phase. Once aeration began, the chart recorder was turned on at the same time. When the D.O. level of the reactor reached a point above 3.0 mg/l, the air-pump was switched off manually and the mixed-liquor agitated manually or with a magnetic stirrer whenever possible (a magnetic stirrer could not be used in reactor A because of the presence of the cooling jacket). The D.O. concentration would then drop (usually quite linearly) and the oxygen uptake rate could be measured graphically from the recording chart. When the reactor D.O. level dropped to approximately 2.0 mg/l (which is the generally accepted limit for aerobic conditions), the air pump was turned on again until the D.O. concentration reached a suitable level, usually at least above 3.0 mg/l, and the above procedure would be repeated for another data point. A track analysis of other parameters, particularly BOD_5 and COD, could be carried out alongside the D.O. uptake experiment. It should be emphasized that throughout the test, the activated-sludge was kept in suspension either by aeration or by agitation or stirring.

The maximum initial D.O. uptake rate was in general very rapid and was thus very difficult to measure accurately insitu the reactors; the following procedure was therefore adopted to obtain the first data point on the D.O. uptake rate experiment :

-   500 ml of mixed-liquor was collected at the end of the react phase.

-   The content was allowed to sit in the beaker for the designated

SETTLE time.

- A proportionate amount of supernatant was then removed, for example, (1.5/5.0)x500 ml removed for a 6 h cycle.

- With the submersible D.O. probe seated properly in the beaker, the same amount of feed was added.

- The content in the beaker was then mixed immediately with a magnetic stirrer and the mixed-liquor D.O. concentration recorded on the chart recorder. The maximum D.O. uptake rate was subsequently measured graphically from the recording chart.

TSS and VSS levels of the reactor at the beginning and the end of aeration were taken and averaged for the computation of specific D.O. uptake rate of the activated-sludge in terms of mg/l D.O. uptake per min per g VSS.

4. RESULTS AND DISCUSSION

The experimental results of this study are presented and discussed in this section. A summary of the sub-divisions of the experimental period and their major features can be found in table 4.01. Frequent reference will be made to these "sub-periods" during the remaining discussions in this chapter. The schedules of the react-phase track analyses and the settling velocity tests are presented in tables 4.02 and 4.03 respectively.

4.1 BOD AND COD ANALYSES

4.1.1 BOD REMOVAL

Data obtained from 5-day Biochemical Oxygen Demand (BOD_5) removal analysis and statistical distribution of the results can be found in tables 4.04 and 4.05 respectively.

The mean BOD_5 removal efficiency of reactor A remained very consistent from operating period II to V, with mean % removal ranging from 90 to 92%. However, during operating period VI (3.7°C, 1.5 l treatment/2 h cycle), the average BOD_5 removal dropped sharply to 78%.

It is interesting to note that even during operating period V, when the operating temperature of reactor A was dropped to 3.7°C in a 1.5 l /6 h cycle, no appreciable change in the mean BOD_5 removal was observed. Only when this low temperature was coupled with an increased loading rate in period VI did reactor A show a diminution in its BOD_5 treatment efficiency of approximately 13%.

Reactor B can be regarded as the control unit as both its temperature and filled volume were maintained constant throughout the entire experimental period of this study. It consistently showed very

TABLE 4.01 : SUMMARY OF EXPERIMENTAL OPERATION

DATE (M.d)	PERIOD	REMARKS
1.10 TO 1.30	I	START-UP : ALL THREE REACTORS OPERATED UNDER IDENTICAL CONDITIONS AT 21.8 C
1.30 TO 4.28	II	6h CYCLE. AVERAGE TEMPERATURES AS FOLLOWS : A=10.5 C B=21.8 C AND REACTOR C=29.8 C. SETTLING PROBLEM DUE TO FEED SOURCE OCCURED FROM 3.06 TO 3.13. MEAN CELL RESIDENT TIME (MCRT) VARIED AS FOLLOWS : 20 d from 2.09 to 3.06 5.3 d from 3.06 to 3.21 8.3 d from 3.21 to 4.28 FILLED REACTOR VOLUME = 5.0 l EACH. TREATMENT VOLUME/CYCLE = 1.5 l PER REACTOR.
4.29 TO 5.29	III	4h CYCLE. AVERAGE TEMPERATURES AS IN (II). MCRT = 8.3 d from 4.29 to 5.02 = 16.7d from 5.02 to 5.29 FILLED REACTOR VOLUME = 5.0 l EACH. TREATMENT VOLUME/CYCLE = 1.0 l PER REACTOR.
5.30 TO 6.24	IV	3h CYCLE; TEMPERATURES AS IN (II) & (III). MCRT = 16.7 d. DENITRIFICATION EXPERIMENTS WERE CARRIED OUT DURING THIS PERIOD. FILLED REACTOR VOLUME = 5.0 l EACH. TREATMENT VOLUME/CYCLE = 0.75 l PER REACTOR.
6.25 TO 7.03	V	6h CYCLE. AVERAGE TEMPERATURES AS FOLLOWS : A=3.7 C B=21.6 C & C=30.0 C. MCRT = 16.7 d FILLED REACTOR VOLUME = 5.0 l FOR A & B = 4.5 l FOR C. TREATMENT VOLUME/CYCLE = 1.5 l PER REACTOR.
7.04 TO 7.05	VI	2h CYCLE. AVERAGE TEMPERATURES AS IN (V). FILLED REACTOR VOLUME AND TREATMENT VOLUME AS IN (V). A TOTAL OF 16 CYCLES OPERATED AND STUDIED.

TABLE 4.02 : SCHEDULE OF REACT-PHASE TRACK ANALYSES

TEST	DATE (M.d)	PERIOD	PARAMETERS MONITORED
1	2.14	II	TSS, VSS
2	2.28	II	pH OF MIXED LIQUOR, SUPERNATANT COD,BOD, NH_3-N, NO_2-NO_3-N.
3	4.10	II	TSS, VSS, SUPERNATANT BOD,COD, NH_3-N, NO_2-NO_3-N & TKN
4	4.24	II	DO CONCENTRATION NH_3-N, NO_2-NO_3-N.
5	5.19	III	REACTOR B : BOD,COD,NH_3-N, NO_2-NO_3-N AND NH_3-N, NO_2-NO_3-N & TKN THE CORRESPONDING OXYGEN UPTAKE RATES
6	5.22	III	REACTOR A : BOD,COD,NH_3-N, NO_2-NO_3-N AND THE CORRESPONDING OXYGEN UPTAKE RATES THE CORRESPONDING OXYGEN UPTAKE RATES
7	5.29	III	REACTOR C : BOD,COD,NH_3-N,NO_2-NO_3-N, AND THE CORRESPONDING OXYGEN UPTAKE RATES THE CORRESPONDING OXYGEN UPTAKE RATES
8	6.14	IV	COD,BOD,NH_3,NO_3,TSS,VSS
9	6.21	IV	OXYGEN UPTAKE RATES FOR ALL THREE REACTORS
10	6.28	V	OXYGEN UPTAKE RATES FOR ALL THREE REACTORS

TABLE 4.03 : SCHEDULE OF SETTLING VELOCITY TESTS

TEST #	DATE (M.d)	PERIOD	MCRT (d)	TSS(A) (mg/l)	TSS(B) (mg/l)	TSS(C) (mg/l)
1	2.09	II	20	4716	4588	3872
2	2.13	II	20	4516	5204	4440
3	3.18	II	5.3	3144	3668	3372
4	3.25	II	8.3	3416	4348	4532
5	5.01	III	8.3	3115	2648	2745
6	5.13	III	16.7	4022	2776	3150
7	6.03	IV	16.7	4826	4490	5964
8	6.12	IV	16.7	4798	5022	5156
9	6.26	V	16.7	5266	5220	5612
10	7.03	V	16.7	5324	3682	4074
11	7.05	VI	**	4784	4760	3942

** WASTE 300 ml MIXED LIQUOR PER FOUR CYCLES.

TABLE 4.04 : DATA OF BOD (5-DAYS) REMOVAL ANALYSIS

DATE (M.d)	PERIOD	INFLUENT BOD (mg/l)	EFFLUENT BOD (mg/l)			BOD REMOVAL (%)		
			A	B	C	A	B	C
1.20	I	197	34	36	43	83	82	78
2.14	II	140	14	12	13	90	91	91
2.28	II	324	26	11	13	92	97	96
4.03	II	375	23	6	29	94	98	92
4.10	II	276	25	18	19	91	93	93
4.25	II	238	15	7	8	94	97	97
5.08	III	250	28	15	11	89	94	96
5.13	III	325	12	6	5	96	98	98
5.19	III	260	12	8	5	95	97	98
5.22	III	212	16	3	4	92	99	98
5.28	III	223	30	22	14	87	90	94
5.29	III	246	31	18	19	87	93	92
6.07	IV	222	9	1	4	96	100	98
6.09	IV	309	27	11	6	91	96	98
6.14	IV	206	31	9	14	85	96	93
6.21	IV	308	21	13	12	93	96	96
6.25	IV	173	10	7	7	94	96	96
6.25	V	173	19	11	21	89	94	88
6.26	V	398	55	29	38	86	93	90
6.27	V	275	25	11	13	91	96	95
6.28	V	216	16	8	10	93	96	95
7.03	V	289	30	7	9	90	98	97
CYC# 1	VI	173	37	27	20	79	84	88
CYC# 2	VI	173	29	19	16	83	89	91
CYC# 3	VI	173	54	24	27	69	86	84
CYC#14	VI	255	46	19	28	82	93	89
CYC#16	VI	255	63	27	26	75	89	90

TABLE 4.05 : MEAN, STANDARD DEV. & RANGE OF BOD DATA

	PERIOD	INFLUENT BOD(mg/l)	EFFLUENT BOD (mg/l)			BOD REMOVAL (%)		
			A	B	C	A	B	C
MEAN	II	271	21	11	16	92	95	94
	III	253	22	12	10	91	95	96
	IV	244	20	8	9	92	97	96
	V	270	29	13	18	90	95	93
	VI	205	46	23	23	78	88	88
STD. DEV.	II	89	6	5	8	2	3	3
	III	40	9	8	6	4	3	3
	IV	62	10	5	4	4	2	2
	V	85	16	9	12	3	2	4
	VI	45	13	4	5	6	3	3
RANGE	II	235	12	12	21	4	7	6
	III	113	19	19	15	9	9	6
	IV	136	22	12	10	11	4	5
	V	225	39	22	29	7	5	9
	VI	82	34	8	12	14	9	7

high BOD$_5$ removal capacity from period II to V (95-97%). However, when the loading rate for reactor B was increased by three times during period VI, its BOD$_5$ treatment efficiency dropped considerably to 88%.

Except for the mean cell residence time (MCRT), operation periods II and V were identical operations for reactor B in terms of mean temperature, treatment volume and cycle length. The results indicated that the reactor was able to return to its original performance level after operating continuously for 5 months in three different modes. However, it should be noted that the high performance level was maintained throughout operations II and V.

Reactor C, the high temperature unit, also demonstrated a high degree of consistency and capability in BOD$_5$ removal throughout operation periods II to V. A 10% decrease in the total filled reactor volume imposed on C alone when operation changed from IV to V resulted in only 2.4% reduction in the reactor's treatment efficiency. When the loading rate was further increased in period VI by another 300%, the BOD$_5$ removal efficiency of C was reduced by approximately 7%.

To summarize, the following general observations were made based on the BOD$_5$ removal data :

1. During operating periods II to IV (refer to table 4.01 for operational characteristics), the room temperature unit (B) and the 30°C unit (C) exhibited similar BOD$_5$ treatment capacities. The overall means (%BOD$_5$ removal) for B and C during these operating periods were 95.7% and 95.4% respectively. A comparison of the means showed that this 0.3% difference is statistically

insignificant. On the other hand, the overall % BOD_5 removal in Reactor A during experimental periods II to IV was only 91.6%. Comparing this mean value with that of B showed that there was only a 0.01% chance that this difference was a result of chance error. It can therefore be concluded with 99.99% confidence level that reactor A was consistently less efficient than B and C in BOD_5 removal during operating periods II to IV.

2. The Standard Deviation (S.D.) and range of the data were very small for all the three reators (table 4.05), indicating that very consistent and reliable treatment performance can be expected from these Sequencing Batch Reactors.

3. The BOD_5 removal efficiencies of reactors A and C during operating period V were respectively 1.8 and 2.4 % lower than the overall average of periods II to IV (confidence level 89.8 and 89.4 % respectively). The BOD_5 treatment efficiency of reactor B was maintained during this period. The lower operating temperature in A and the 10 % reduction of hydraulic retention time in C were evidently the causes for the reduced treatment efficiency of these reactors.

4. The "Specific Cycle" of the treatment operation studied in this experiment did not seem to have any effect on the BOD_5 treatment efficiencies of the reactors. The effect of specific cycle will be discussed in more details in section 4.7.

4.1.2 COD REMOVAL

The data obtained from COD analysis are presented in table 4.06 and their statistical analysis is summarized in table 4.07. The results

displayed a trend very similar to that of the BOD_5 data except for the following deviations :

1. The COD treatment efficiencies in terms of percentage removal were consistently lower than that of BOD_5 for all the units. This could be due to the fact that some non-biodegradable organic substances which were neutral to the BOD_5 test were inevitably present in the effluents which contributed positively in a COD test.

2. The standard deviation and range of the COD removal data were overall slightly higher than that of the BOD_5 data. The COD data were more dispersed.

3. The COD removal efficiency of reactor A dropped 5% when the operation period changed from IV to V and the corresponding average operation temperature dropped from $10.5°C$ to $3.7°C$; but no further decline was observed when the cycle length was reduced in period VI. This can be taken as an indication that COD removal in this case is more sensitive to low temperature than BOD_5 removal (see discussion on unseeded BOD_5 test at different temperatures, section 4.14)

4. In terms of COD, a much greater fluctuation in influent strength was experienced. This fluctuation was also reflected in the effluent COD. A slightly higher percentage removal was also apparent in all three reactors during experimental period II when the influent COD was substantially higher than the rest. This is probably due to higher utilization of the REACT phase potential because of higher initial substrate concentration.

TABLE 4.06 : DATA OF COD REMOVAL ANALYSIS

DATE (M.d)	PERIOD	INF.COD (mg/l)	EFFLUENT COD (mg/l)			COD REMOVAL(%)		
			A	B	C	A	B	C
1.20	I	711	178	185	195	75	74	73
2.14	II	720	194	121	106	73	83	85
2.28	II	1760	363	244	277	79	86	84
3.15	II	2108	416	297	301	80	86	86
4.03	II	1940	360	252	284	81	87	87
4.10	II	577	123	54	55	79	91	91
4.25	II	872	153	74	92	83	92	89
5.08	III	679	197	156	113	71	77	83
5.13	III	606	107	113	115	82	81	81
5.19	III	971	113	99	73	88	90	93
5.22	III	708	214	94	96	70	87	86
5.28	III	836	213	129	102	75	85	88
5.29	III	854	208	111	108	76	87	87
6.07	IV	777	217	130	147	72	83	81
6.09	IV	809	234	149	144	71	82	82
6.14	IV	849	155	111	115	82	87	86
6.21	IV	965	215	154	142	78	84	85
6.25	IV	969	224	108	158	77	89	84
6.25	V	969	291	133	263	70	86	73
6.26	V	1114	382	189	101	66	83	91
6.27	V	939	267	173	178	72	82	81
6.28	V	830	224	149	158	73	82	81
7.03	V	743	208	129	132	72	83	82
CYC#1	VI	688	165	131	143	76	81	79
CYC#2	VI	688	175	131	153	75	81	78
CYC#3	VI	688	236	182	189	66	74	73
CYC#14	VI	722	225	166	173	69	77	76
CYC#16	VI	722	236	171	187	67	76	74

TABLE 4.07 : MEAN, STD. DEV. & RANGE OF COD DATA

	PERIOD	INFLUENT COD(mg/l)	EFFLUENT COD (mg/l)			COD REMOVAL (%)		
			A	B	C	A	B	C
MEAN	II	1330	268	174	186	79	88	87
	III	776	175	117	101	77	85	86
	IV	874	209	130	141	76	85	84
	V	919	274	155	166	71	83	82
	VI	702	207	156	169	71	78	76
STD. DEV.	II	680	126	103	113	3	3	3
	III	135	51	23	16	7	5	4
	IV	89	31	21	16	5	3	2
	V	141	69	26	61	3	2	6
	VI	19	35	24	20	5	3	3
RANGE	II	1531	293	243	246	10	9	7
	III	365	107	62	42	13	9	12
	IV	192	79	46	43	11	7	5
	V	371	174	60	162	7	4	18
	VI	34	71	51	46	10	7	6

4.1.3 REACT PHASE TRACK ANALYSIS OF SUPERNATANT BOD AND COD

As Sequencing Batch Reactors (SBR) do not operate under "steady-state" assumptions, it is important to have a clear conceptual picture of what goes on during the operation phases of a SBR. A "track analysis" monitors the parametric changes as a function of time and provides valuable insights into a semi-batch process.

The track analyses in this experiment were carried out for only the REACT phase because of its relatively dynamic nature. If the FILL phase was not "instantaneous" but extended over a significant portion of the cycle (Irvine *et al.*, 1979,1979a), the track investigation should be extended accordingly.

As with almost all analyses carried out in this study, only supernatant parameters were monitored in the track analyses. The reason is that supernatant samples provide a more realistic reflection of the quality of the effluent which is the final product of the entire operation. Filtrate sampling neglects potential effects such as sedimentation efficiency of the unit; effluents of different suspended solids levels , for instance, would exert different oxygen demands on the recieving water even though the oxygen demands of their filtrates might be the same.

A total of six COD & BOD_5 track analyses were carried out during experimental periods II to IV; three of these analyses also included the corresponding oxygen uptake rate.

Figures 4.01 to 4.09 show the supernatant BOD_5 and COD of all the three reactors from the beginning to the end of the aeration phase during these track analyses. All the three reactors showed a similar trend in both BOD_5 and COD removal — despite the big difference in

FIG. 4.01 : PLOT OF SUPERNATANT BOD VS. AERATION TIME IN
TRACK ANALYSIS #2 (OP.PER.II, 6h CYC., 1.5l/CYC.)

FIG. 4.02 : PLOT OF SUPERNATANT COD VS. AERATION TIME IN
TRACK ANALYSIS #2 (OP.PER.II, 6h CYC., 1.5l/CYC.)

FIG. 4.03 : PLOT OF SUPERNATANT BOD VS. AERATION TIME IN
TRACK ANALYSIS #3 (OP.PER.II, 6h CYC., 1.5l/CYC.)
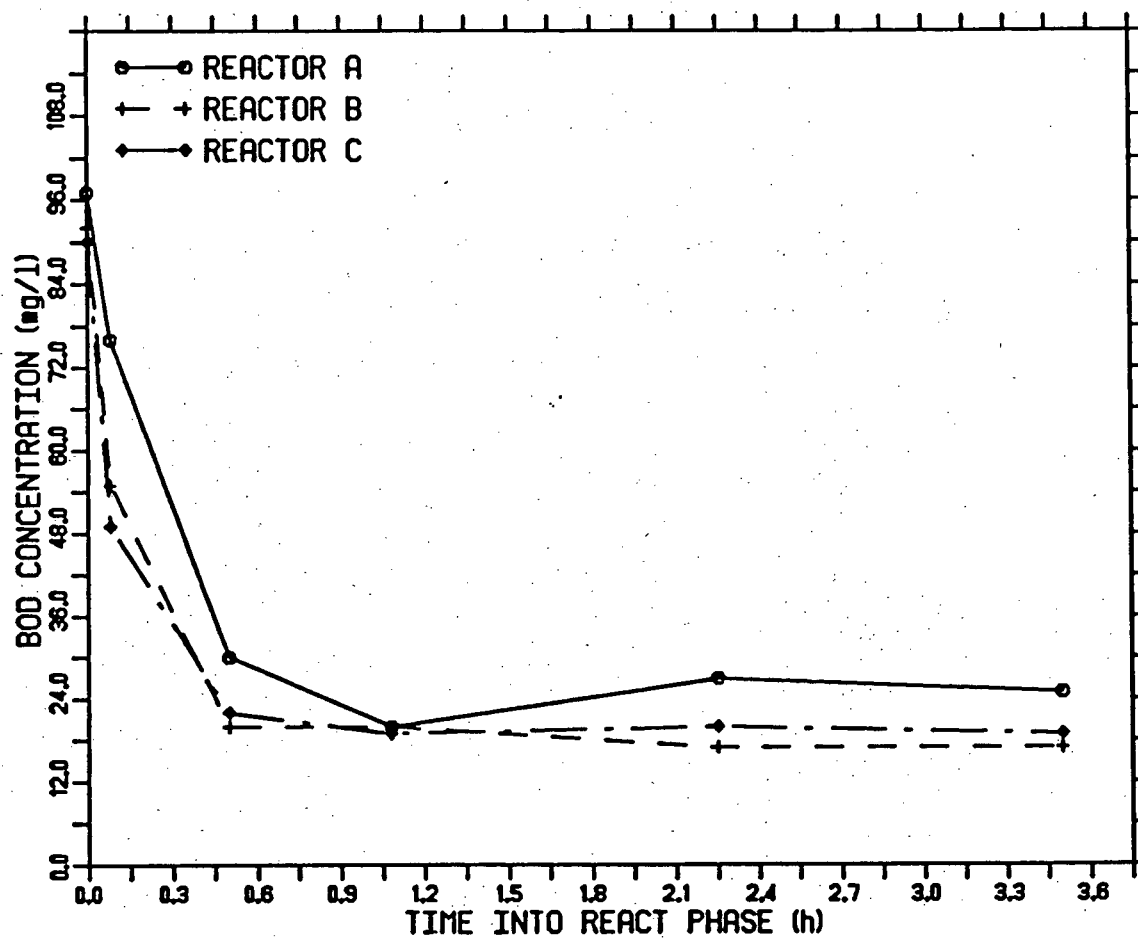
FIG. 4.04 : PLOT OF SUPERNATANT COD VS. AERATION TIME IN
            TRACK ANALYSIS #3 (OP.PER.II, 6h CYC., 1.5l/CYC.)

FIG. 4.05 : RESULTS OF TRACK ANALYSIS #5 (OP.PER.III, 4h CYC.,
1.01/CYC., REACTOR "B" ONLY)

FIG. 4.06 : RESULTS OF TRACK ANALYSIS #6 (OP.PER.III, 4h CYC.,
1.0 1/CYC., REACTOR "A" ONLY)
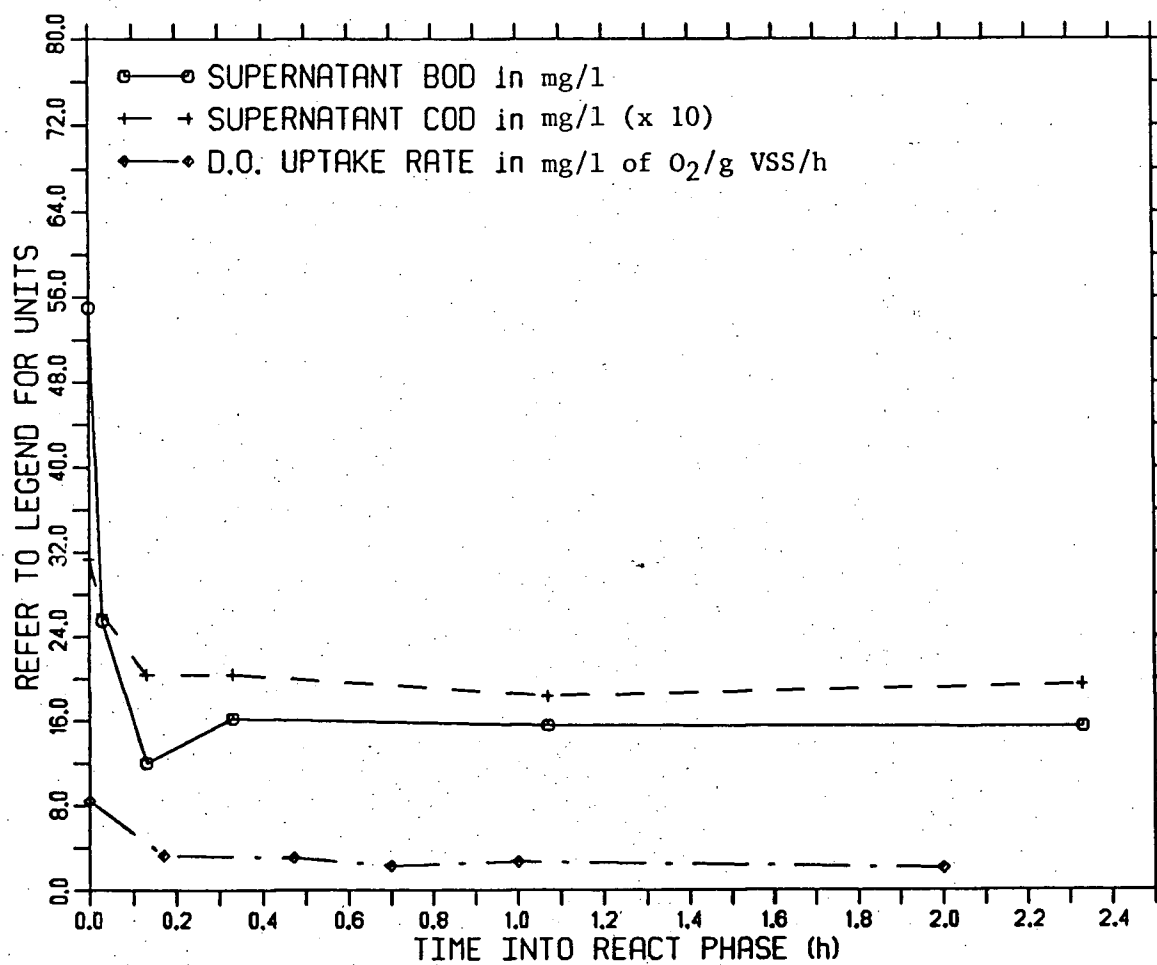
FIG. 4.07 : RESULTS OF TRACK ANALYSIS #7 (OP.PER.III, 4h CYC.,
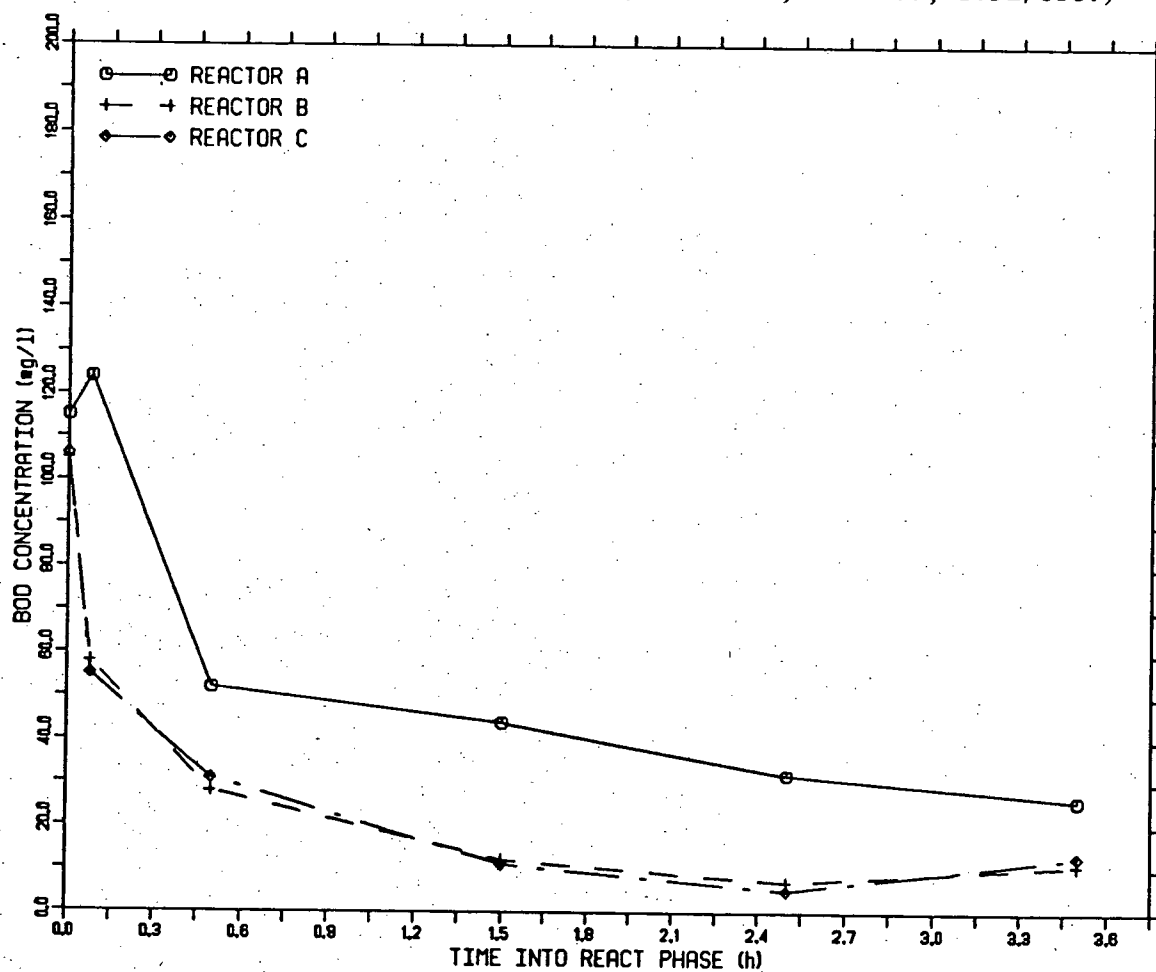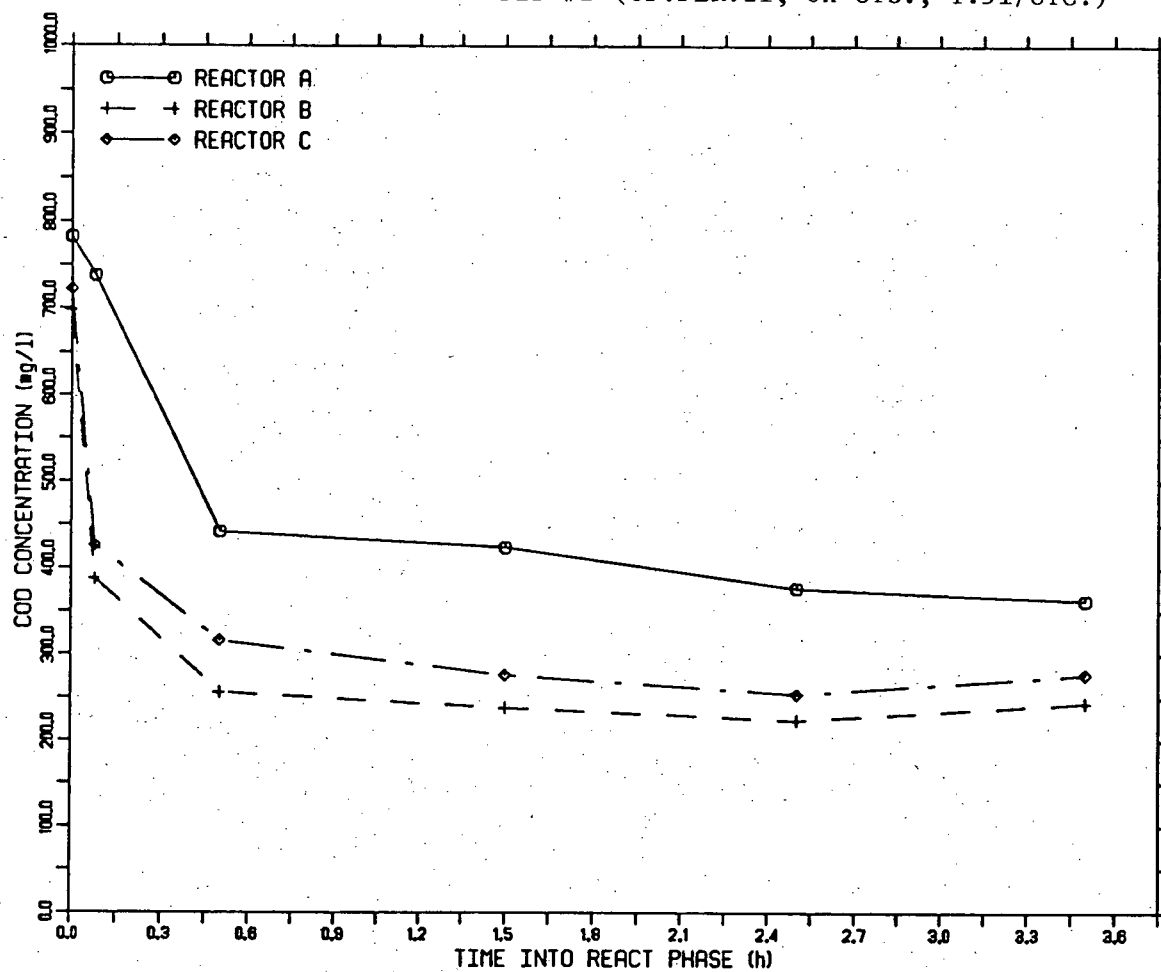1.0 1/CYC., REACTOR "C" ONLY)
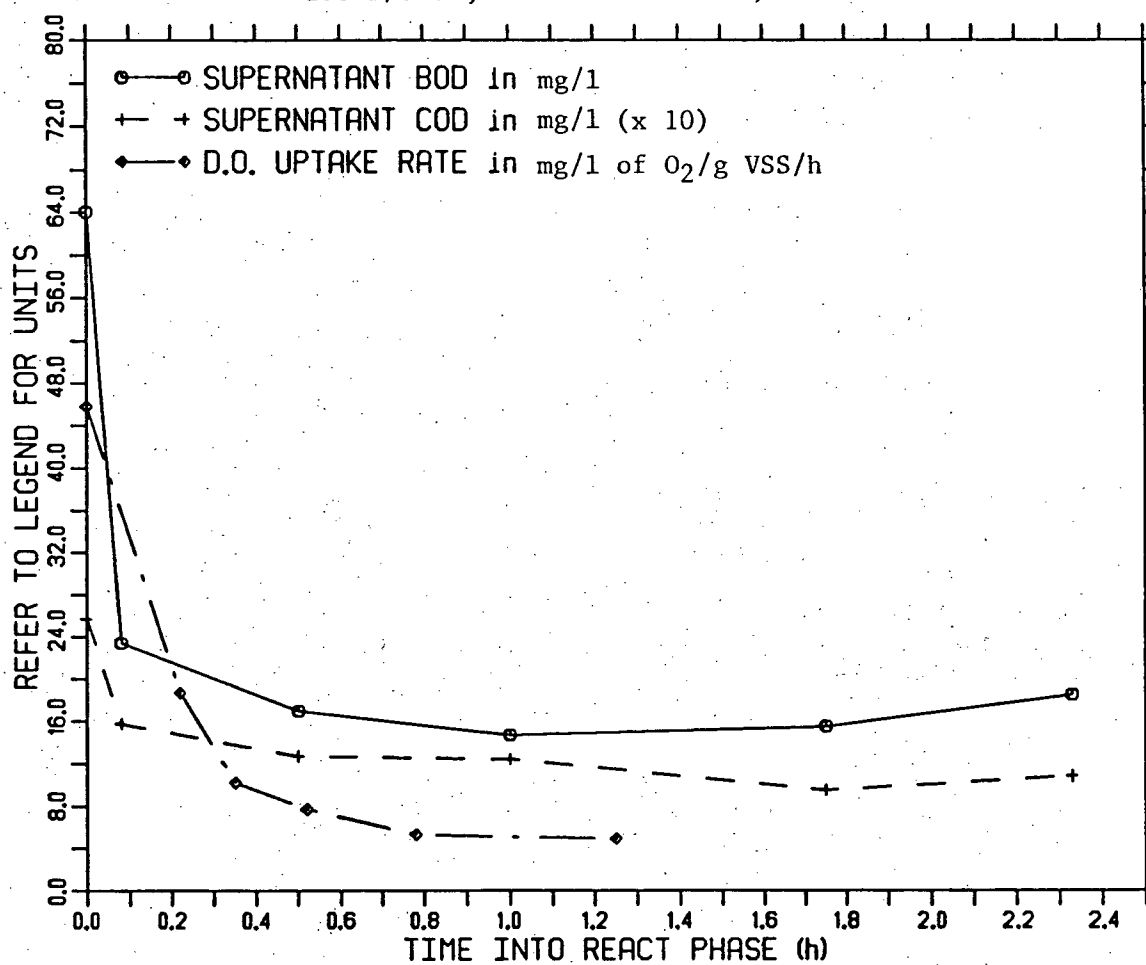
FIG. 4.08 : PLOT OF SUPERNATANT BOD VS. AERATION TIME IN TRACK ANALYSIS #8 (OP.PER.IV, 3h CYC., 0.75 1/CYC.)
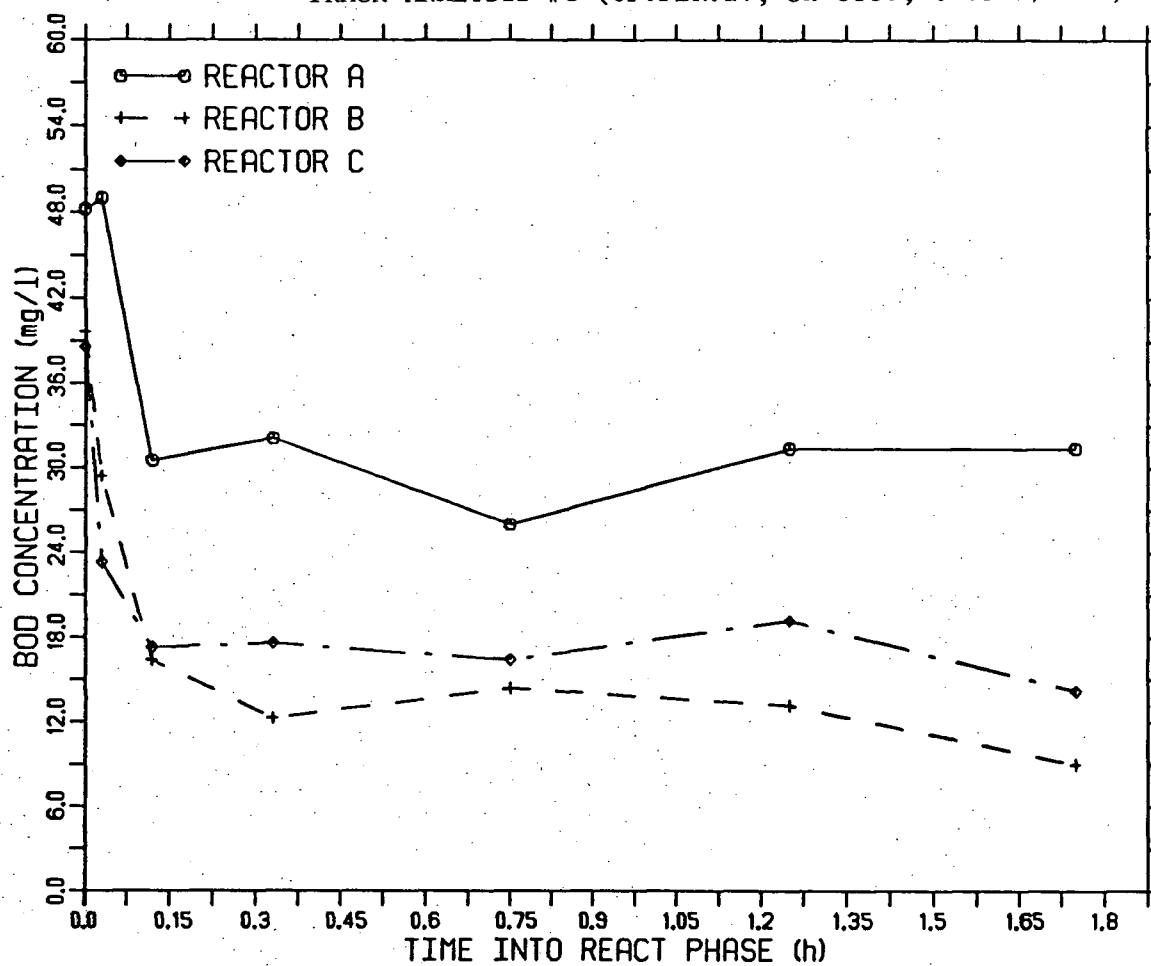
FIG. 4.09 : PLOT OF SUPERNATANT COD VS. AERATION TIME IN TRACK ANALYSIS #8 (OP.PER.IV, 3h CYC., 0.75 1/CYC.)
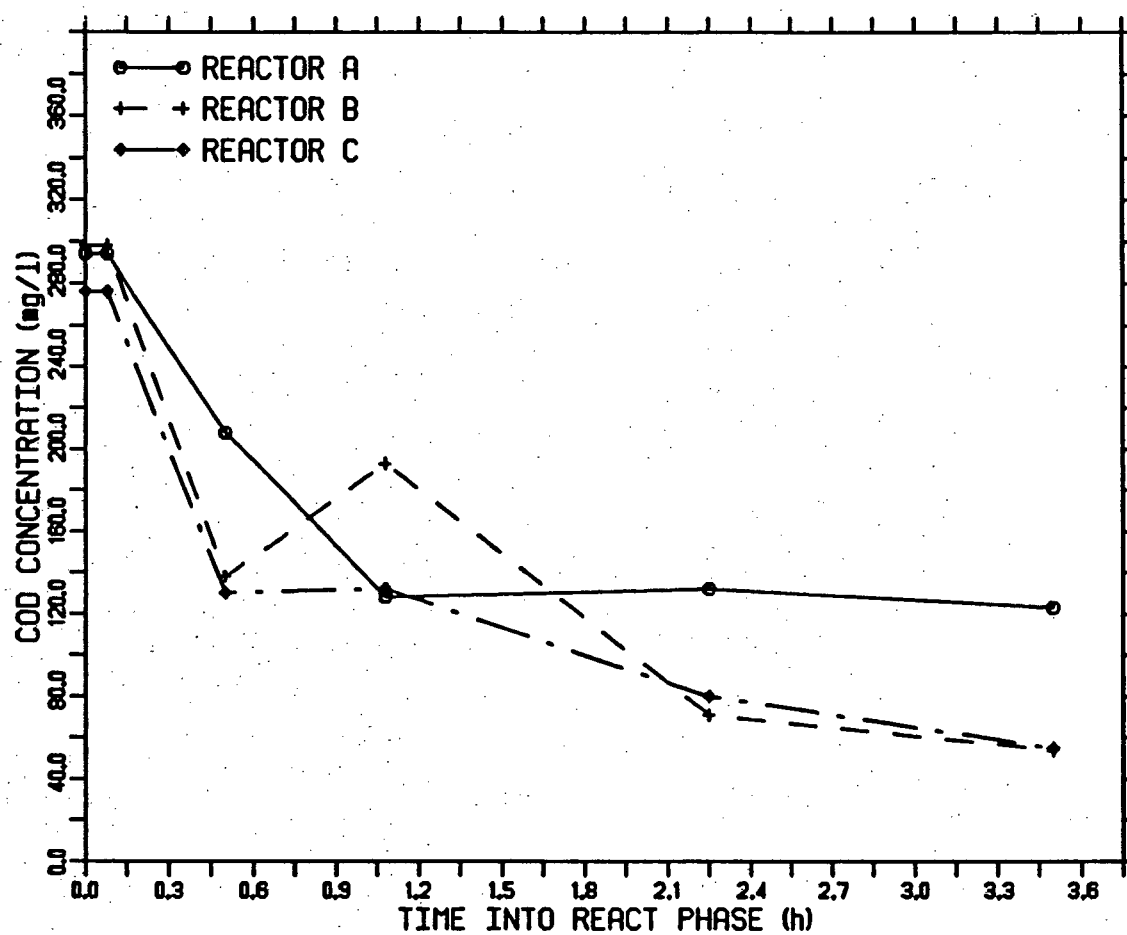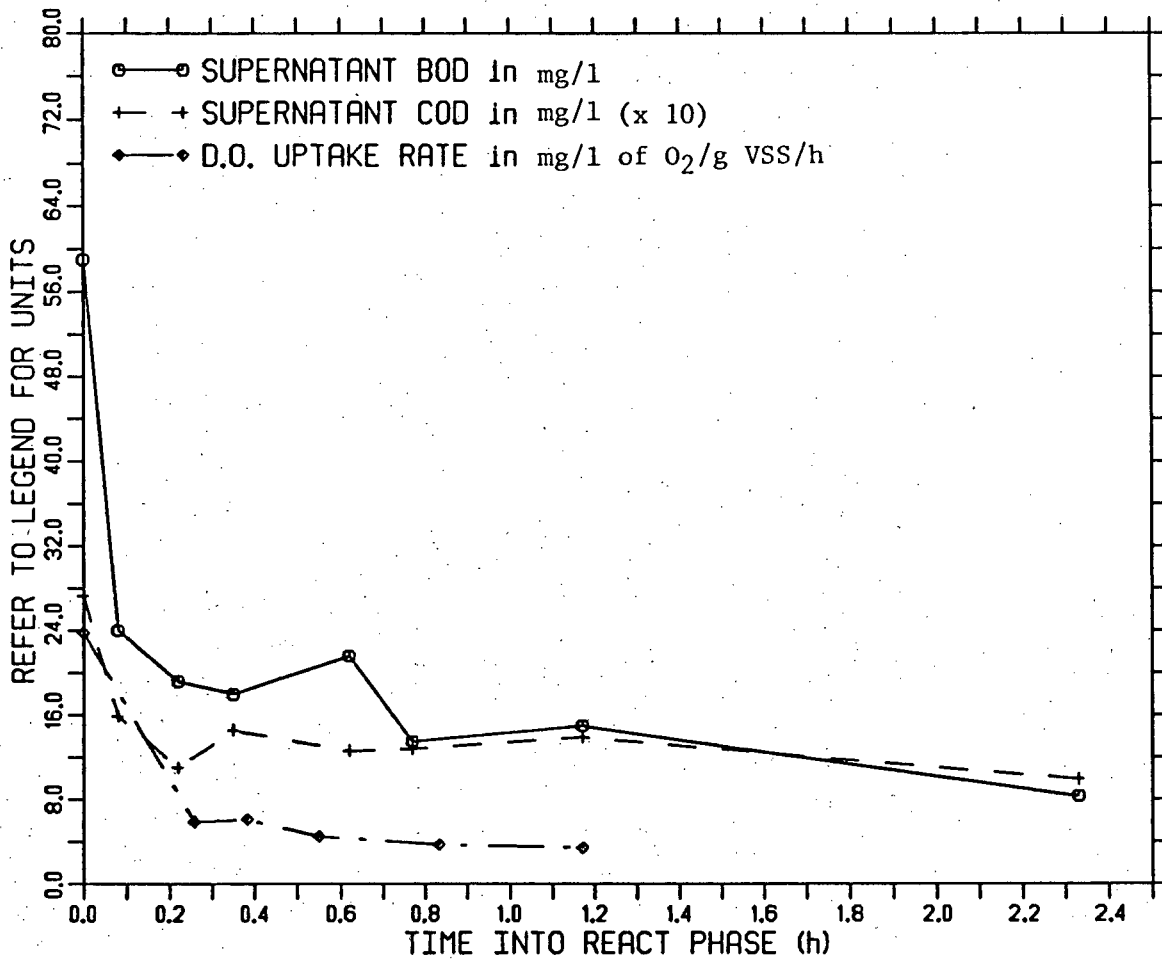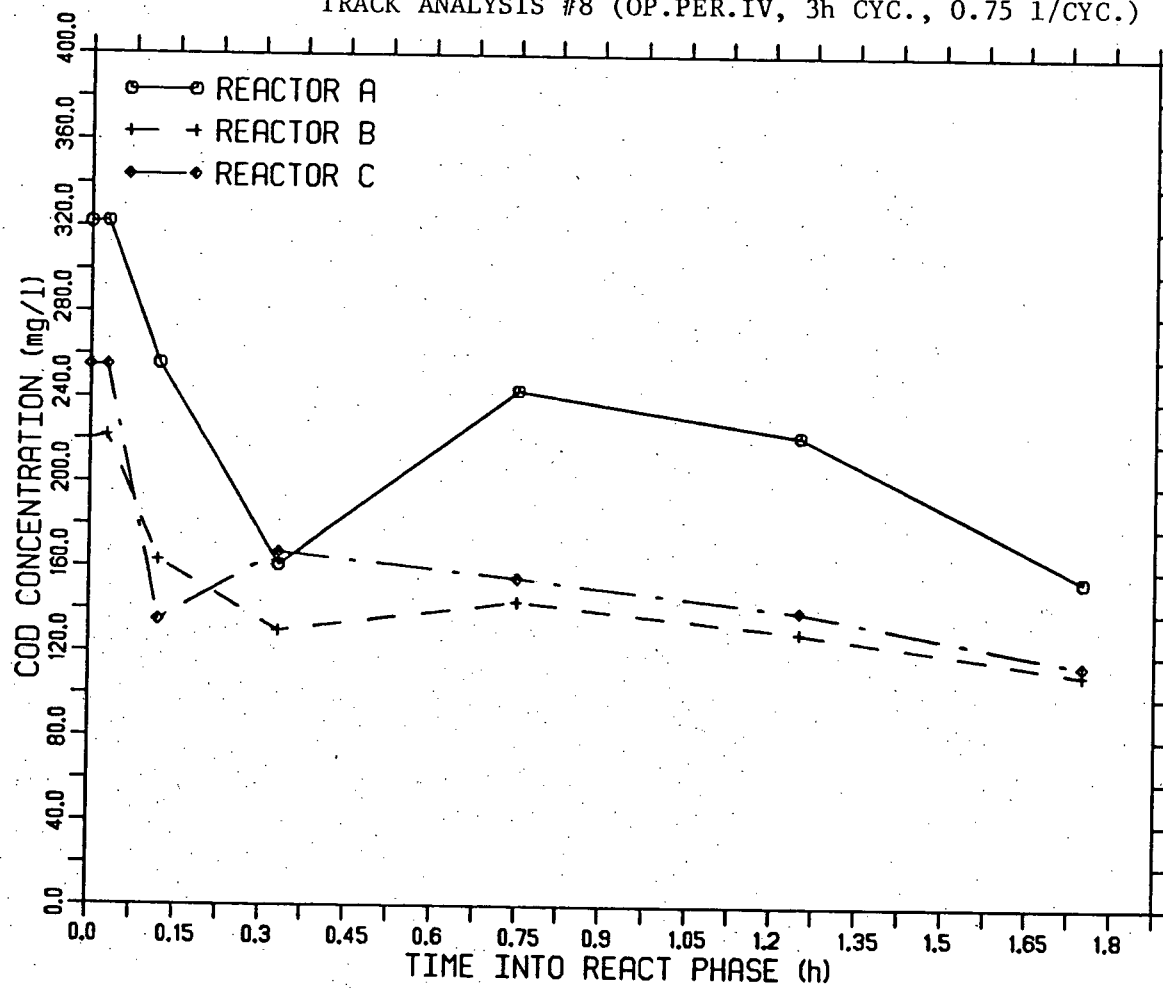
the initial substrate concentration for analyses #2 and #3, substrate removal was largely completed within the first 30 to 40 minutes of aeration. This observation is consistent with those reported by other researchers (Alleman et al., 1978,1979; Dennis and Irvine,1979; Hoepker and Schroeder,1979). After this initial period, substrate removal from the supernatant practically ceased. The residue COD that remained can be taken as the non-biodegradable portion of the wastewater.

4.1.3.1 REACTION KINETICS

The initial substrate removal rate constants can be estimated through results obtained from the track studies.

The initial reaction kinetics (before substrate removal is essentially completed) can be assumed to be an overall second order reaction – first order with respect to substrate concentration and first order with respect to biomass concentration (Dennis 1979). The kinetic equation can therefore be written as follows :

$$dC/dt = k.C.m \qquad <4.1>$$

Where

dC/dt   = rate of change of substrate conc.

k   = kinetic rate constant

C   = substrate conc.

m   = biomas conc.

t   = time

However, the change in biomass concentration is negligible as compared with the initial m and can be reasonably assumed constant

(Dennis and Irvine,1979; Irvine and Richter,1976). A pseudo first—order kinetics with respect to C can therefore be adopted :

$$dC/dt = k'.C \qquad \text{where } k' = k.m \qquad <4.2>$$

After rearrangement and integration,

$$\ln C - \ln C_0 = k'.t \qquad \text{where } C_0 = \text{initial substrate conc.}$$

Therefore a linear correlation of ln C and time would serve to verify the validity of the psuedo—first—order assumption. The kinetic constant (k) can be obtained by dividing the slope of the regression line (k') by the average biomass concentration during the test cycle.

This assumed kinetics is only valid during the initial stage of the REACT phase when active substrate consumption was taking place (generally within the first 30 min). The natural logarithm of the BOD_5 data obtained within this time frame was correlated, by using a calculator with built—in linear regression programme, with aeration time. The temperature coefficients for the kinetic constants were then computed by finding the slope of the regression line of k vs. temperature. The results of this computation are summarized in table 4.08.

The high correlation coefficients for the computed temperature coefficients indicate that the BOD_5 kinetic constants were significantly temperature dependent. The "specific cycle", on the other hand, did not seem to affect the k values in a linear manner. It may be noted that for both units A and B, k was lowest in operation period II, higher in IV, and highest in III. For unit C, k seemed to be on an ascending trend as the operation proceeded from period II to period IV. However, no conclusion can be drawn about the effect of "specific cycle" on k with the limited

TABLE 4.08 : KINETIC COEFFICIENTS OF BOD REMOVAL
COMPUTED FROM LINEAR REGRESSION OF
ln(BOD) vs REACTION TIME

TRACK RUN #	PERIOD	KINETIC COEFFICIENT,k (1/mg.d) (WITH CORRELATION COEFF. IN BRACKETS)			TEMPERATURE COEFF. OF k (1/mg.d.C)
		A (10.5 C)	B (21.8 C)	C (29.8 C)	
2	II	0.009(.97)	0.010(.95)	0.010(.92)	5.75×10^{-5}(.92)
3	II	0.014(1.0)	0.017(.98)	0.021(.97)	3.56×10^{-4}(.98)
5	III	---	0.055(.96)	---	
6	III	0.042(.91)	---	---	$\Big\} 8.50 \times 10^{-4}$(.97)
7	III	---	---	0.058(1.0)	
8	IV	0.023(.95)	0.041(1.0)	0.090(1.0)	3.45×10^{-3}(.94)

amount of data available.

The temperature coefficients, however, appeared to have a positive relationship with the "specific cycle" (table 4.08). There are no obvious theoretical reason for this trend. It is suspected that this relationship may be partly due to experimental errors introduced by the temperature dilution effect of feeding. The amount of room-temperature influent fed to the reactors were 1.5, 1.0 and 0.75 l per cycle for operation periods II, III and IV respectively. Readings taken during the react phase of operation period II showed that the initial temperature of reactors A and C could change by as much as $3^{\circ}C$ during the beginning of REACT. The equilibrium temperatures were usually attained after approximately 30 min of aeration. This dilution effect was proportionately smaller during operation periods III and IV.

It is therefore expected that the smaller temperature errors during the shorter cycles were partly reflected by the larger temperature coefficients (i.e. stronger temperature dependency) as operation proceeded from period II to IV.

4.1.3.2 OXYGEN UPTAKE RATE

The dissolved oxygen (D.O.) uptake rate per gram of volatile suspended solids measured during track runs 5, 6 and 7 (fig. 4.05, 4.06, 4.07) assumed the same general shape of the corresponding BOD_5 and COD curves. As the initial BOD_5 concentrations were quite close during these three separate runs (approximately 60 mg/l), correlation of the maximum D.O. uptake rate with operation temperature was made possible. The temperature coefficient thus calculated was 1.91 mg/l of D.O. per g VSS per h per $^{\circ}C$ (slope of the regression line of the maximum D.O. uptake rate vs. temperature). The correlation coefficient

was 0.98. The built-in function of a Hewlett-Packard hp11c calculator was utilized to carry out the regression and correlation.

Similar correlations for track analyses 9 and 10 (fig. 4.10, 4.11, 4.12, 4.13) yielded temperature coefficients of 1.564 and 1.393 mg/l.g.h.$^\circ$C respectively. The corresponding correlation coefficients were 0.68 and 0.99. The initial BOD_5 concentrations during track tests #9 and 10 were approximately 94 mg/l and 75 mg/l respectively.

4.1.4 THE <u>UNSEEDED BOD VS. COD ANALYSIS FOR SAMPLES FROM</u> DIFFERENT TEMPERATURES

By comparing the BOD_5 removal data (table 4.06) and the COD removal summary (table 4.07), the following observations were made :

1. From period II to V, the overall mean COD removal (the overall average of the mean % removal values from period II to V) for reactors A, B and C were 75.8, 85.3 and 84.8% respectively. The corresponding BOD_5 removal were 91.3, 95.5 and 94.8% respectively. The overall difference between A and B was 9.5% in terms of COD removal and only 4.2% in terms of BOD_5 removal. On the other hand, the difference between B and C were much smaller, being 0.5 % for COD and 0.7 % for BOD_5 treatment efficiency.

2. The mean % BOD_5 removal of reactor A during operating period II and V were 92% and 90% respectively; despite the considerable temperature difference (II was a 6h cycle at 10.5°C and period V was a 6h cycle at 3.7°C), the difference in BOD_5 treatment efficiency was only 2%. The corresponding mean %COD removal of A was 79% and 71% for operation periods II and V respectively, showing a much bigger difference of 8%.

FIG. 4.10 : PLOT OF GROSS D.O. UPTAKE RATE VS. AERATION TIME IN
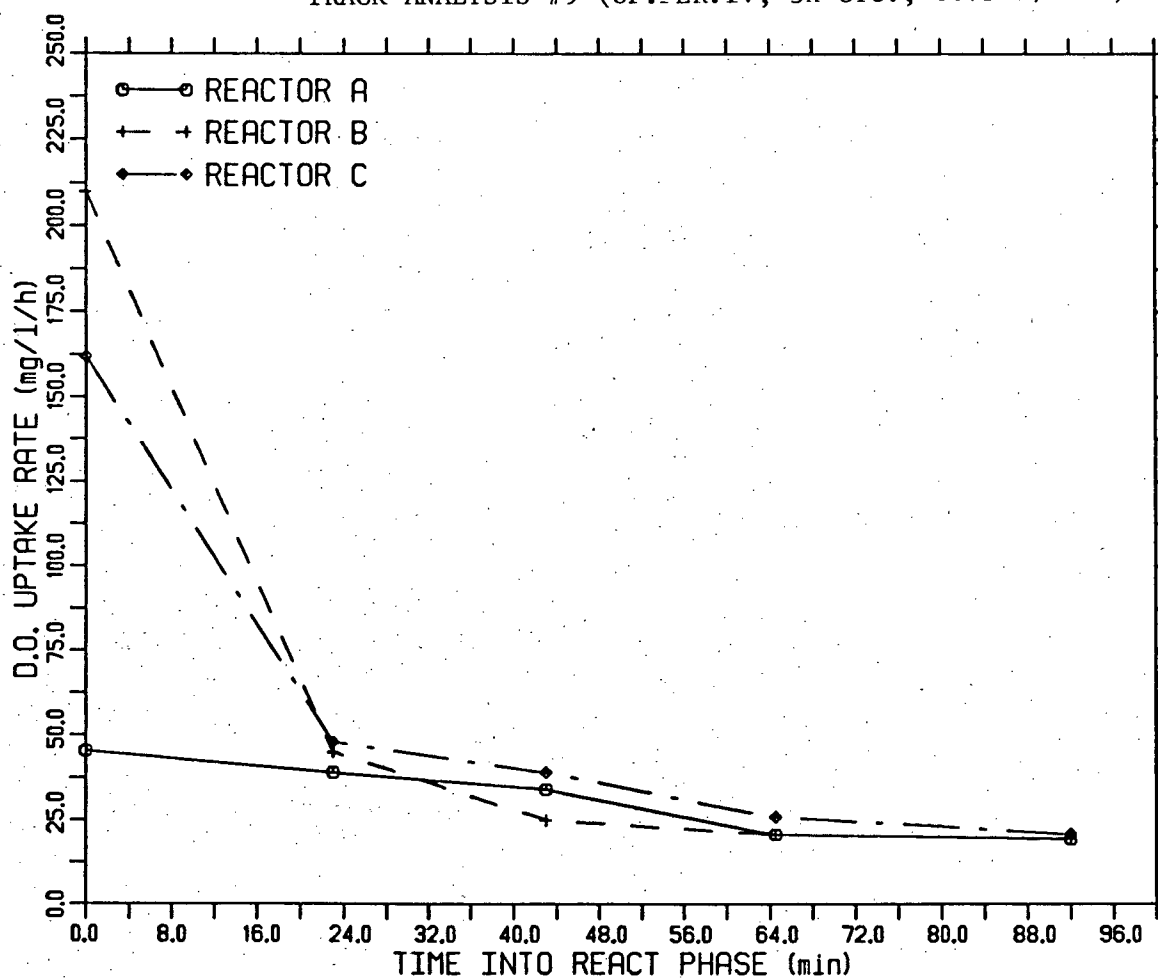TRACK ANALYSIS #9 (OP.PER.IV, 3h CYC., 0.75 1/CYC.)

FIG. 4.11 : PLOT OF SPECIFIC D.O. UPTAKE RATE VS. AERATION TIME
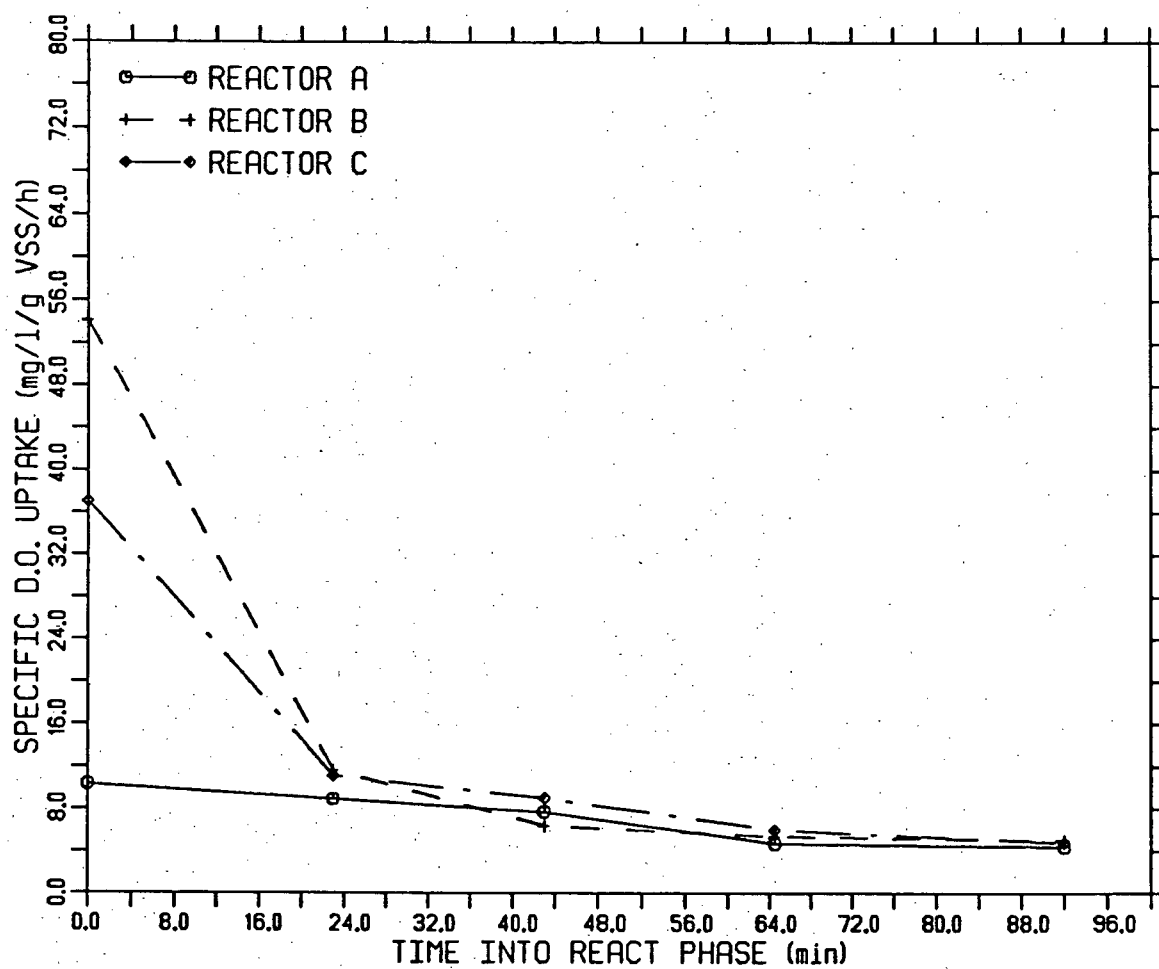IN TRACK ANALYSIS #9 (OP.PER.IV, 3h CYC., 0.75 1/CYC.)

FIG. 4.12 : PLOT OF GROSS D.O. UPTAKE RATE VS. AERATION TIME IN
TRACK ANALYSIS #10 (OP.PER.IV, 3h CYC., 0.75 1/CYC.)
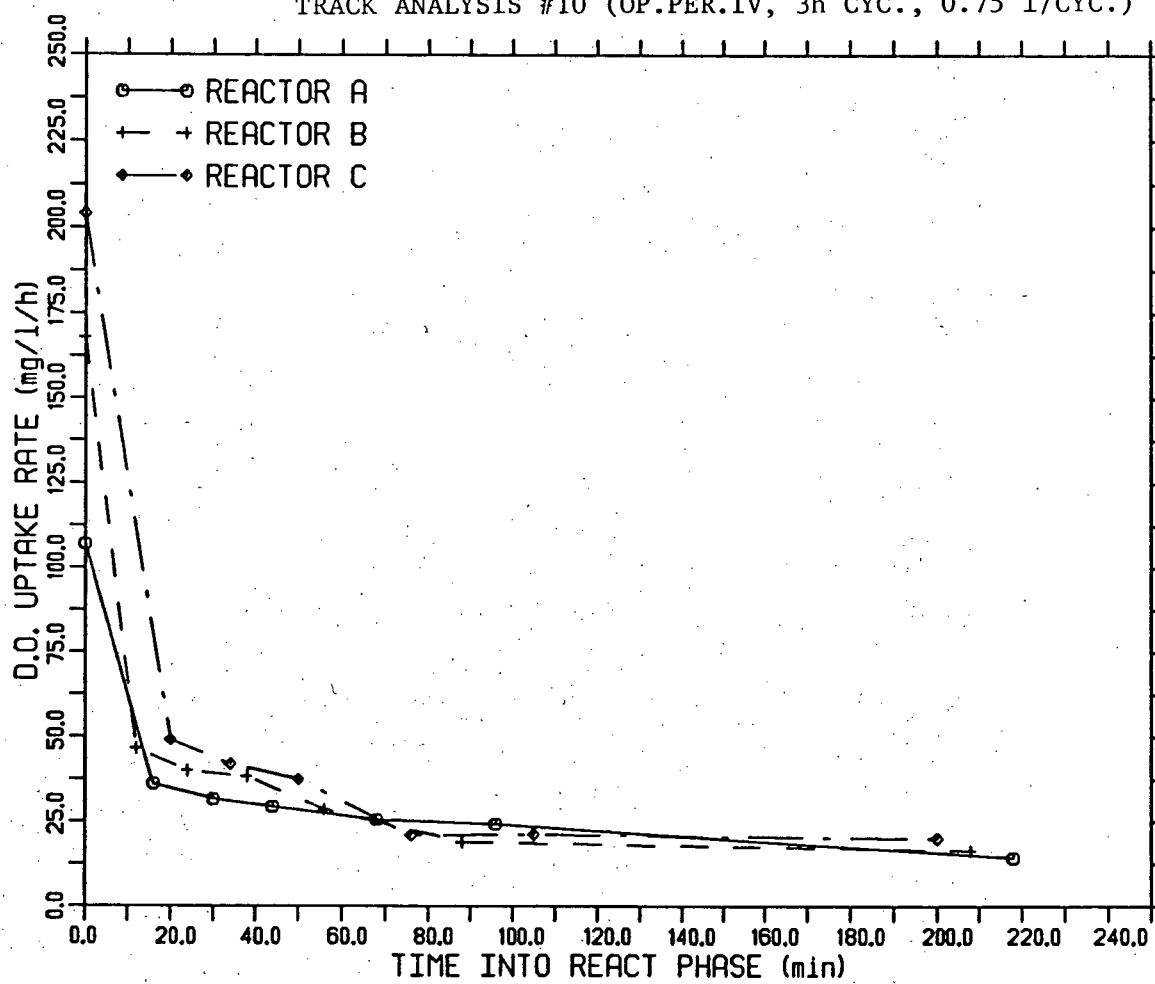
FIG. 4.13 : PLOT OF SPECIFIC D.O. UPTAKE RATE VS. AERATION TIME
IN TRACK ANALYSIS #10 (OP.PER.IV, 3h CYC., 0.75 1/CYC.)
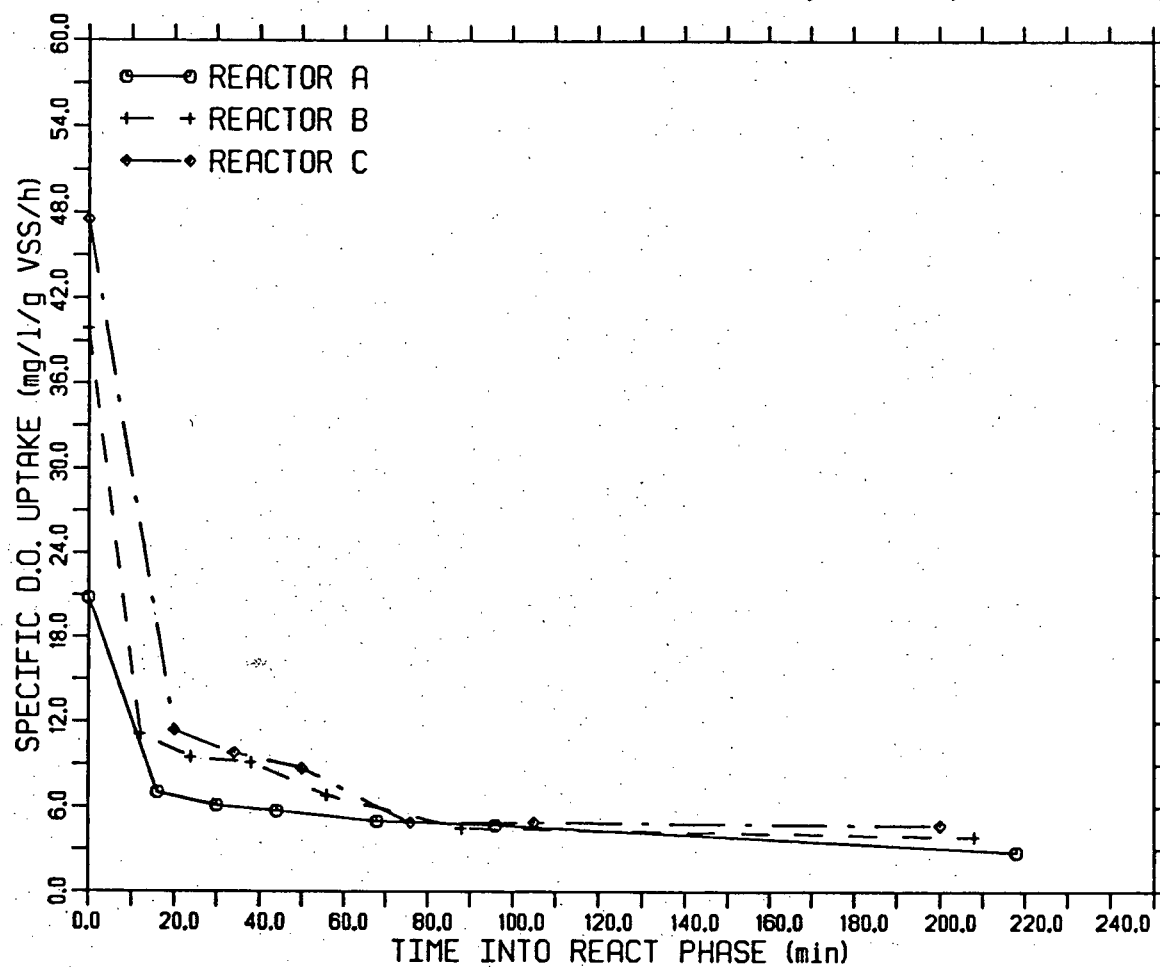
3. From (1) and (2) above, it appears that at low temperatures (10.5°C and 3.7°C) the organic carbon removal capacity of a SBR as indicated by the COD test displayed more temperature sensitivity than the unseeded BOD_5 test.

Because of the consistency of this observed trend, it was deemed a worthwhile exercise to develop a model or explanation that would be consistent with the observed phenomena.

First of all, an activated-sludge unit can be pictured as a complex ecosystem consisting of a myriad of microorganisms. It is reasonable to assume that the entire cross section of this population is responsible for the overall degree of "treatment" achieved by the activated-sludge unit.

The next assumption is that activated-sludge units in general display different treatment efficiencies under different operating temperatures because of the following reasons :

1. The metabolic rate of the microorganisms is temperature dependent; the substrate consumption *rate* of the microorganisms is lower at lower temperatures.

2. Different types of microorganisms dominate the microbial population under different ambient temperatures; and these different groups of microbes consume different kinds of substrate materials preferentially.

Reason (1) above is a popular view on the effect of temperature on the treatment performance of activated-sludge units, particularly for continuous-flow systems. However, reason (1) standing alone implies that given enough reaction time, the final BOD_5 removal of an

activated-sludge unit would be the same regardless of the operation temperature, as the metabolic rate only affects the rate at which substrates are removed, *not* the ultimate amount. Futhermore, this temperature effect should be equally reflected by the BOD_5 and COD data. However, as mentioned previously, results tabulated in table 4.07 show that the final COD removal capacity of A lagged behind those of B and C by a considerably larger percentage. Furthermore, reactor A suffered a more dramatic set-back in COD removal than in BOD_5 removal when operation changed from IV (10.5°C) to V (3.7°C). The lower %COD removal to %BOD_5 removal ratio during the low temperature operations indicates that more organic matters had escaped biomass consumption than the unseeded BOD_5 index suggested.

A possible reason for this discrepancy is proposed as follows :

The unseeded BOD bottle is, in terms of microbial population, a microcosm of the activated-sludge reactor. Any kind of microorganisms not present in the reactor will therefore not be available in an unseeded BOD bottle. In another words, any organic matter that escaped microbial consumption due to the absence of the major microbes that consume this kind of substrate (because of the low operating temperatures) will not be further oxidized in an unseeded BOD test. However, the presence of these organic matters will still be detected by a COD test. A schematic of this proposed model is presented in figure 4.14.

Further experimental work with both seeded and unseeded BOD tests would be required to support the hypothesis of this model. However, based on the development of the model, the adequacy of an unseeded BOD_5 test is questioned. Sewage effluents are usually

GROUPS OF MICROORGANISMS

GROUPS OF ORGANIC MATTERS

A   B
a   b
      X
C   D
c   d

ACTIVATED SLUDGE
REACTOR AT TEMPERATURE T1

AFTER EXCESSIVE
REACTION TIME

X

EFFLUENT

COD = X

BOD = 0

X
b   a
  c
    d

INFLUENT
COD = X+a+b+c+d
BOD = a+b+c+d
(X comprise non-
  biodegradable
  organic matters)

A   B
a   b
      d
C   X
c   D

REACTOR AT
TEMPERATURE T2
(GROUP 'D' ORGANISMS
NOT VIABLE AT T2)

AFTER EXCESSIVE
REACTION TIME

X
d

EFFLUENT

COD = X+d

UNSEEDED BOD
= 0 because
organism 'D'
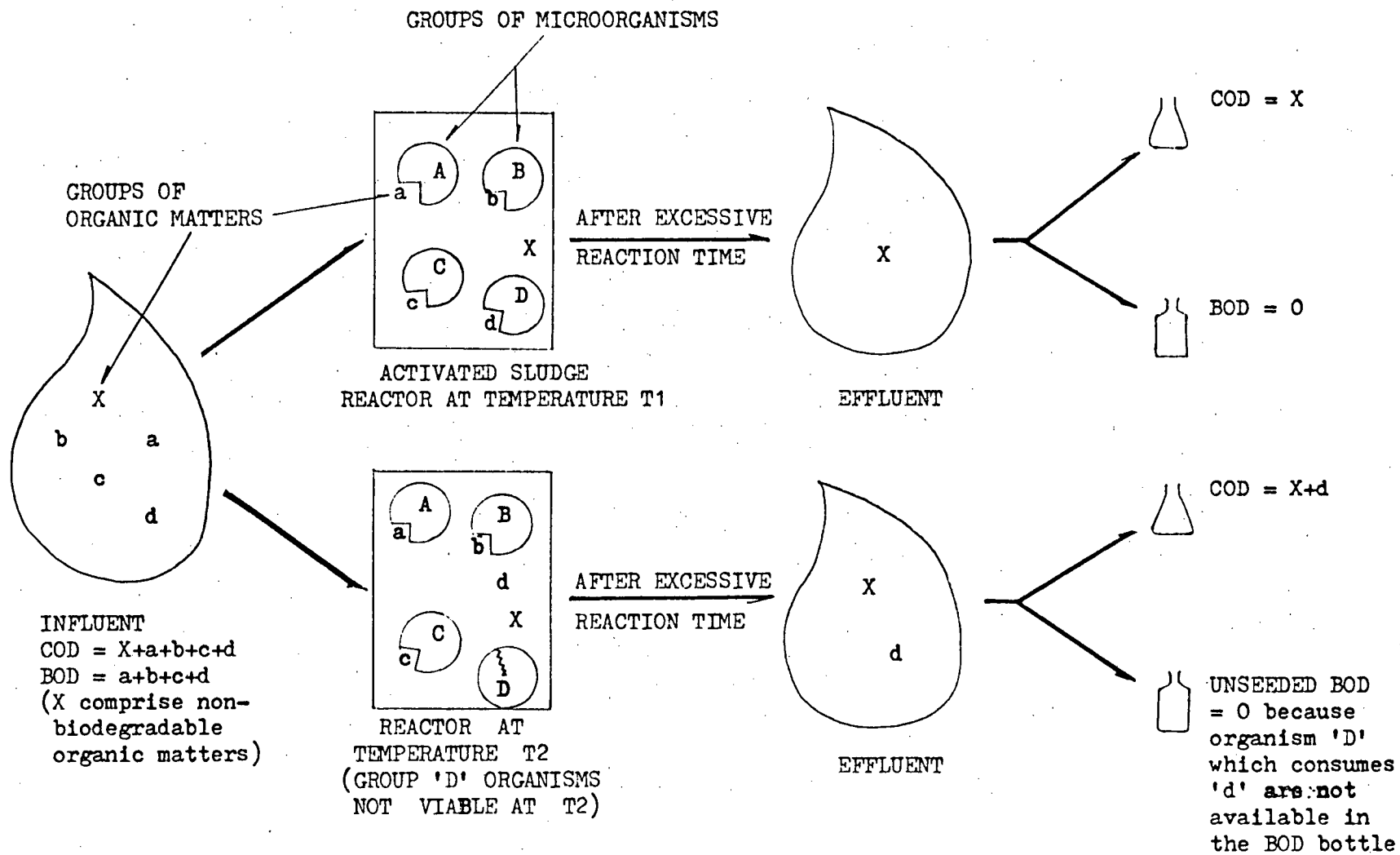which consumes
'd' are not
available in
the BOD bottle

FIG. 4.14 : PROPOSED MODEL OF UNSEEDED BOD ANALYSIS VS. COD ANALYSIS UNDER DIFFERENT
OPERATION TEMPERATURES

assumed to contain sufficient microorganisms for a BOD_5 test. In fact, they are often used as the "seed" for samples from other sources. But from the experience of this study, it appears that great care should be exercised when using the BOD_5 index alone as an oxygen demand indicator; particularly when cross-temperature data are being examined for design or comparison purposes.

4.2 SUPERNATANT AMMONIA AND COMBINED NITRITE-NITRATE NITROGEN ANALYSIS

4.2.1 AMMONIA REMOVAL

Data obtained from NH_3-N and combined NO_2-NO_3-N analyses are tabulated in tables 4.09 and 4.10 respectively. The statistical summary of the data can be found in tables 4.11 and 4.12. During operation periods II, III and IV, the ammonia removal capacity of A remained the lowest while reactor C showed consistently the highest efficiency.

NH_3-N removal in an activated-sludge system is largely attributed to three different processes : oxidation to nitrite (which concentration is expected to be negligible in most aerobic systems) and nitrate (the nitrification process); biomass assimilation; and stripping or desorption by aeration.

All the above three processes are temperature dependent to various degrees. Numerous researches done with pure cultures (Painter, 1970) have concluded that the optimum growth rate of *nitrosomonas* is at $30°C$, although the reported growth-rate constant varies over a wide range.

TABLE 4.09 : DATA OF AMMONIA NITROGEN ANALYSIS

DATE (M.d)	PERIOD	INF.NH$_3$ (mg/l)	EFF. NH$_3$-N (mg/l)			NH$_3$-N REMOVAL (%)		
			A	B	C	A	B	C
2.14	II	13.7	12.5	1.3	1.0	8.8	90.5	92.7
2.28	II	52.0	31.0	2.2	2.2	40.4	95.8	95.8
3.14	II	70.5	27.4	2.5	2.2	61.1	96.5	96.9
4.10	II	11.8	1.5	1.4	0.9	87.3	88.1	92.4
5.08	III	19.0	3.0	0.0	0.0	84.2	100.0	100.0
5.13	III	10.7	1.8	0.7	0.7	83.2	93.5	93.5
5.19	III	14.9	7.8	3.4	1.5	47.7	77.2	89.9
5.22	III	13.6	5.8	0.4	0.0	57.4	97.1	100.0
5.28	III	13.5	2.3	0.8	0.7	83.0	94.1	94.8
5.29	III	16.7	1.6	0.4	0.4	90.4	97.6	97.6
6.07	IV	10.8	1.6	0.8	0.6	85.2	92.6	94.4
6.09	IV	20.0	4.3	1.0	0.5	78.5	80.0	90.0
6.14	IV	17.6	2.5	0.5	0.4	85.8	97.2	97.7
6.21	IV	15.9	1.3	0.9	0.7	91.8	94.3	95.6
6.25	IV	13.5	1.9	1.4	1.4	85.9	89.6	89.6
6.25	V	13.5	1.3	0.3	0.6	90.4	97.8	95.6
6.26	V	14.1	2.3	1.5	1.5	83.7	89.4	89.4
6.27	V	15.4	2.9	0.7	1.2	81.2	95.5	92.2
6.28	V	14.2	1.6	1.1	1.4	88.7	92.3	90.1
7.03	V	17.7	6.4	1.4	1.4	63.8	92.1	92.1
CYC #1	VI	19.1	6.8	1.3	4.1	64.4	93.2	78.5
CYC #2	VI	19.1	8.9	1.7	3.9	53.4	91.1	79.6
CYC #4	VI	19.1	10.4	1.5	1.9	45.5	92.1	90.1
CYC#14	VI	18.2	10.8	1.5	1.5	40.7	91.8	91.8
CYC#16	VI	18.2	10.8	1.8	0.7	40.7	90.1	96.2

TABLE 4.10 : DATA OF NITRIFICATION ANALYSIS

DATE	PERIOD	INF. NO_x-N (mg/l)	EFFLUENT NO_x-N (mg/l)			NITRIFICATION * NH_3 REMOVAL		
(M.d)			A	B	C	A	B	C
2.14	II	0.2	0.9	7.3	7.0	0.57	0.58	0.54
2.28	II	0.4	4.3	15.0	13.0	0.19	0.29	0.25
3.14	II	1.0	5.5	19.0	14.0	0.10	0.26	0.19
4.10	II	0.0	0.2	0.7	1.5	0.02	0.07	0.14
5.08	III	0.0	5.2	7.8	6.9	0.33	0.41	0.36
5.13	III	0.5	4.3	6.9	6.9	0.43	0.64	0.64
5.22	III	0.3	2.1	2.3	1.6	0.23	0.15	0.10
5.28	III	0.1	3.9	3.3	2.9	0.34	0.25	0.22
5.29	III	0.0	2.5	2.3	2.4	0.17	0.14	0.15
6.09	IV	0.8	8.8	8.6	4.5	0.51	0.41	0.19
6.14	IV	1.3	4.3	6.9	6.6	0.20	0.33	0.31
6.21	IV	0.0	1.4	1.4	1.7	0.10	0.09	0.11
6.25	IV	0.5	5.2	6.0	5.3	0.41	0.45	0.40
6.25	V	0.5	5.3	6.0	5.6	0.39	0.42	0.40
6.26	V	0.1	4.4	5.2	5.0	0.36	0.40	0.39
6.27	V	0.5	4.1	5.2	4.9	0.29	0.32	0.31
6.28	V	0.4	0.7	3.1	1.8	0.02	0.21	0.11
7.03	V	0.3	3.4	8.0	5.4	0.28	0.48	0.32
CYC #1	VI	0.4	2.8	8.2	4.2	0.20	0.44	0.26
CYC #2	VI	0.4	1.0	5.3	3.6	0.06	0.28	0.21
CYC #4	VI	0.4	1.0	5.1	3.7	0.07	0.27	0.19
CYC#14	VI	0.4	0.5	2.7	2.1	0.02	0.14	0.10
CYC#16	VI	0.4	0.7	2.5	2.4	0.05	0.13	0.12

* COMPUTED AS (EFF.NO_x-N - INF.NO_x-N)/(INF.NH_3-N - EFF.NH_3-N)
** NO_x-N IS THE COMBINED NITRITE-NITRATE NITROGEN

TABLE 4.11: MEAN, STD. DEV. & RANGE OF NH_3-N DATA

	PERIOD	INF.NH_3 (mg/l)	EFF. NH_3-N (mg/l)			NH_3-N REMOVAL (%)		
			A	B	C	A	B	C
MEAN	II	37.0	18.1	1.9	1.6	49.4	92.7	94.5
	III	14.7	3.7	1.0	0.6	74.3	93.3	96.0
	IV	15.6	2.3	0.9	0.7	85.4	90.7	93.5
	V	15.0	2.9	1.0	2.4	81.6	93.4	91.9
	VI	18.7	9.5	1.6	2.4	48.9	91.7	87.2
STD. DEV.	II	29.0	13.7	0.6	0.7	33.2	4.1	2.2
	III	2.9	2.5	1.2	0.6	17.4	8.2	4.0
	IV	3.6	1.2	0.3	0.4	4.7	6.6	3.5
	V	1.7	2.1	0.5	1.5	10.6	3.3	2.4
	VI	1.7	1.7	1.2	1.5	10.1	1.2	7.8
RANGE	II	58.7	29.5	1.2	1.3	78.5	8.4	4.5
	III	8.3	6.2	3.4	1.5	42.7	22.8	10.1
	IV	6.5	1.2	0.9	1.0	13.3	17.2	8.1
	V	4.2	5.1	1.2	3.4	26.6	8.4	2.7
	VI	0.9	4.0	0.5	0.9	23.7	3.1	17.7

TABLE 4.12 : MEAN, STANDARD DEVIATION & RANGE OF
NITRIFICATION DATA

	PERIOD	INF. NO_x-N (mg/l)	EFF. NO_x-N (mg/l)			NITRIFICATION * NH_3 REMOVAL		
			A	B	C	A	B	C
MEAN	II	0.4	2.7	10.5	8.9	0.22	0.30	0.28
	III	0.2	3.6	4.5	4.1	0.30	0.32	0.29
	IV	0.7	4.9	5.7	4.5	0.31	0.32	0.25
	V	0.4	3.6	5.5	4.5	0.27	0.37	0.31
	VI	0.4	1.2	5.0	3.2	0.08	0.25	0.18
STD. DEV.	II	0.4	2.6	8.1	5.8	0.24	0.21	0.18
	III	0.2	1.3	2.6	2.6	0.10	0.21	0.22
	IV	0.5	3.1	3.1	2.1	0.19	0.16	0.13
	V	0.2	1.7	1.8	1.6	0.15	0.10	0.12
	VI	0.0	0.9	2.3	0.9	0.07	0.13	0.07
RANGE	II	1.0	5.3	18.3	12.5	0.55	0.51	0.40
	III	0.5	3.1	5.5	5.3	0.26	0.40	0.54
	IV	1.3	7.4	7.2	4.9	0.41	0.36	0.29
	V	0.4	4.6	4.9	3.8	0.37	0.27	0.29
	VI	0.0	2.3	5.7	2.1	0.18	0.31	0.16

* COMPUTED AS (EFF. NO_x - INF. NO_x)/(INF. NH_3 - EFF. NH_3)
** NO_x-N IS THE COMBINED NITRITE-NITRATE NITROGEN

TABLE 4.13 : INITIAL AMMONIUM REMOVAL RATE DURING
REACT PHASE

TRACK TEST#	PERIOD	INITIAL AMMONIA REMOVAL RATE (mg/l.g VSS.h) (WITH CORRELATION COEFFICIENTS IN BRACKETS)		
		A	B	C
2	II	0.55(0.95)	1.61(0.96)	1.83(0.95)
3	II	0.59(0.98)	1.28(0.99)	1.88(0.99)
5	III	----	1.50(0.91)	----
6	III	0.14(0.92)	----	----
7	III	----	----	1.07(1.00)
8	IV	0.24(1.00)	1.51(0.99)	1.46(1.00)

The accelerated growth rate and metabolism of the biomass at elevated temperatures are well known phenomena and are expected to cause higher NH_3-N assimilation rate. Stripping of ammonia by aeration is a mass transfer process which is highly affected by the ambient temperature and pH. In fact one of the practical limitations of desorption as a potential ammonia removal process is its inability to operate at ambient temperatures near freezing.

In light of the above factors, the temperature dependency exhibited by the data in periods II, III and IV is deemed logical.

During operation periods V and VI, the NH_3-N removal efficiency of reactor C fell short of B, possibly because of the reduced retention time of unit C during these operations. Reactor A showed a surprisingly high NH_3-N removal power during operation period V when the temperature was lowered to 3.7 °C. However, when this low operation temperature was coupled with decreased retention time in operating period VI (2h cycle, 1.5 l treated/cycle), the NH_3-N treatment efficiency of reactor A dropped sharply to 48.9 %.

4.2.2 NITRIFICATION

Nitrification is mainly the result of a pollutant removal reaction — ammonia oxidation. It is therefore meaningless to express nitrification with respect to the initial influent NO_2-NO_3-N level. Instead, the extend of nitrification was expressed as a fraction of the amount of NH_3-N removed (tables 4.10, 4.12). The remaining fraction of NH_3-N removed from the wastewater is assumed to be largely consumed for biomass assimilation, lost through denitrification (see 4.2.4) and, volatilized through aeration.

From the data summary in table 4.12, it is apparant that reactor B experienced the highest degree of nitrification per unit of NH_3-N removed (from now on denoted as "nitrification ratio") throughout all the operation periods. However, this interpretation of the nitrification data cannot be conclusive unless the ammonia removal data is also studied simultaneously. An examination of table 4.11 shows that the effluent NH_3-N level of reactor C was consistently lower than that of B during operation periods II to IV, indicating that more NH_3-N was removed or converted in reactor C than in reactor B during these periods.

The fact that the mean nitrification ratio in B is higher than that in C may be due to the following reasons:

1. Higher assimilation in C because of a metabolically more active biomass.

2. C had higher denitrification power. This assumption was verified by some of the track analyses (fig. 4.16, 4.20).

3. More desorption of ammonia in reactor C due to its higher temperature.

During operation periods V and VI, the effluent NH_3-N level in C exceeded that of B (i.e. lower NH_3-N removal in C); however, the effluent NO_2-NO_3-N level of C was sufficiently low to keep its nitrification ratio below that of B. Reactor A, on the other hand, maintained a fairly high nitrification ratio from period II to V. Since the effluent NH_3-N level of A was consistently higher than that of the other two reactors and the effluent NO_2-NO_3-N concentration was in general lower, the high nitrification ratio suggests that the previously discussed nitrogen removal processes, namely nitrification,

denitrification, assimilation and stripping, were occuring relatively slowly in reactor A.

When the low temperature $(3.7^{\circ}C)$ was coupled with short retention time in operation VI, the nitrification ratio of reactor A dropped sharply to 0.08. The mean effluent NH_3-N and NO_2-NO_3-N concentrations during this period were 9.5 and 1.2 mg/l respectively, indicating very low NH_3-N oxidation and removal. It should be noted that during operation period VI, the raw data (tables 4.09 & 4.10) for NH_3-N removal and nitrification ratio in reactor A showed a declining trend from cycle #1 to cycle #16. The same applied for reactor B and reactor C. Unfortunately, period VI could not be operated over a longer period of time due to operational restrictions and limitations in this bench scale study.

4.2.3 TRACK ANALYSIS

Concentrations of NH_3-N and NO_2-NO_3-N as a function of aeration time were measured during operation periods II, III and IV. The results are plotted in figures 4.15 to 4.20.

The following general observations were made from the experimental results :

1. The ammonium oxidation capacities of units B and C were very similar while that of A was consistently lower in all the tests conducted.

2. For all the three reactors, ammonium consumption by the activated-sludge approximated a linear function of aeration time at supernatant NH_3-N concentrations greater than approximately 1 to 2 mg/l. An inflection occurs at this concentration level and

FIG. 4.15 : PLOT OF AMMONIA AND NITRITE-NITRATE NITROGEN VS. AERATION TIME IN TRACK ANALYSIS #2 (OP.PER.II, 6h CYC., 1.51/CYC.)
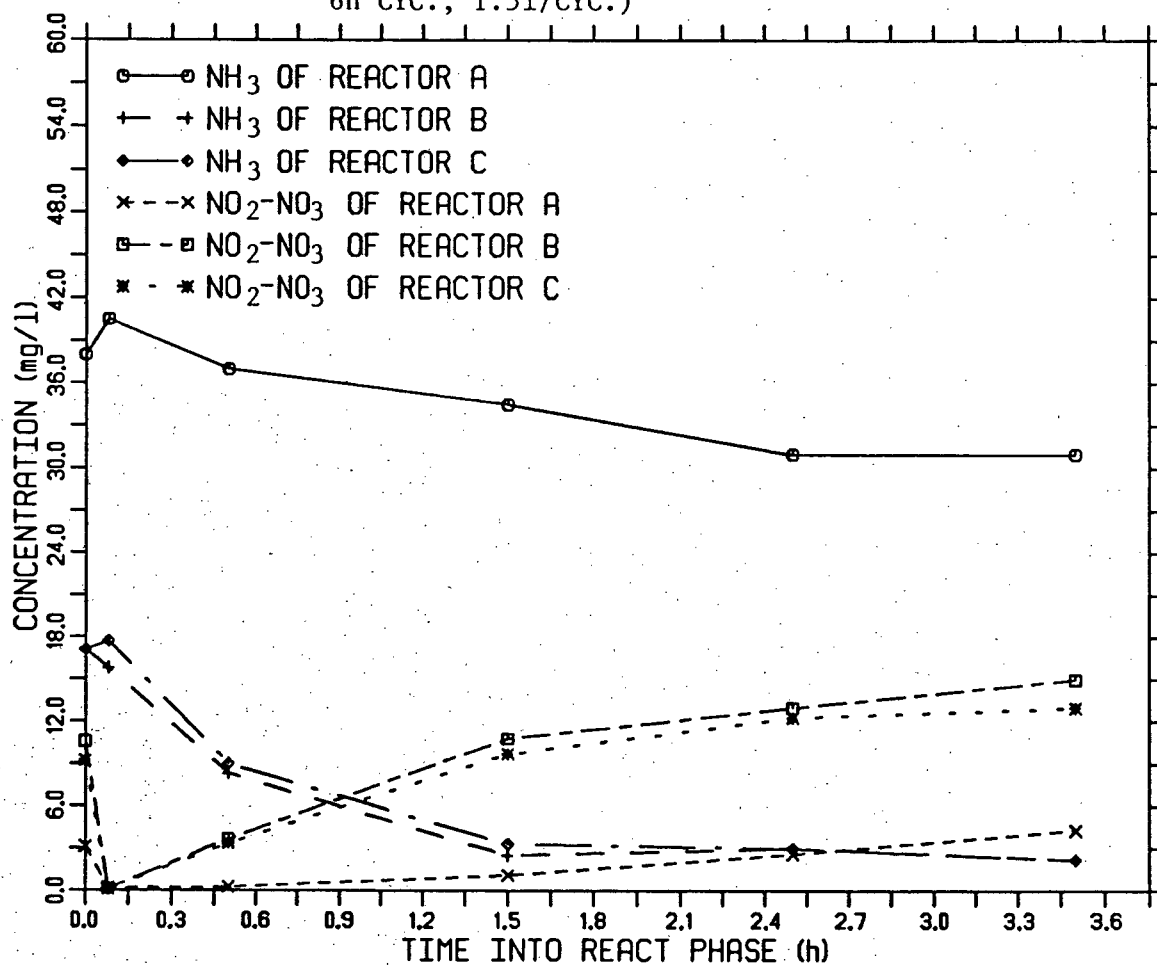
FIG. 4.16 : PLOT OF AMMONIA AND NITRITE-NITRATE NITROGEN VS.
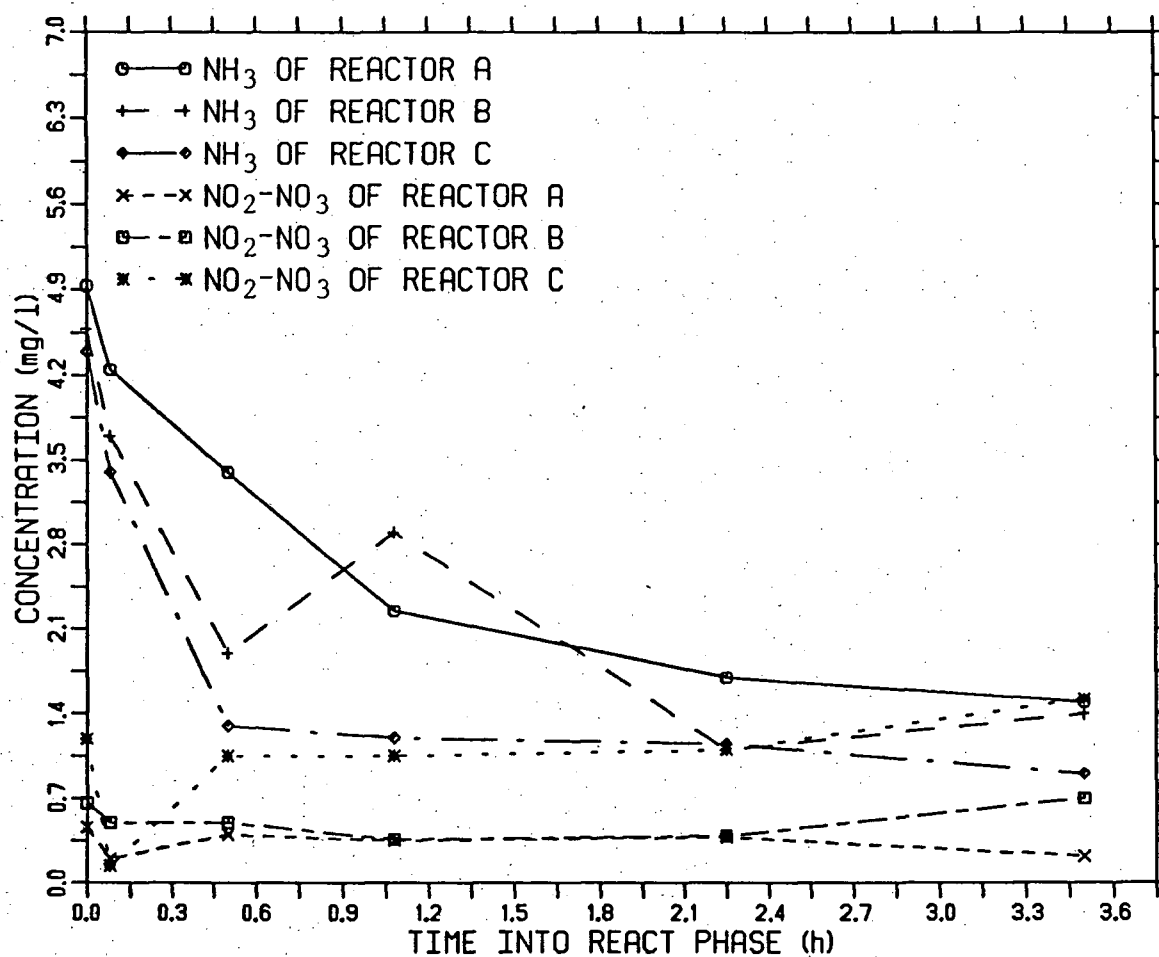AERATION TIME IN TRACK ANALYSIS #3 (OP.PER.II,
6h CYC., 1.5 1/CYC.)

FIG. 4.17 : PLOT OF AMMONIA AND NITRITE-NITRATE NITROGEN VS.
AERATION TIME IN TRACK ANALYSIS #5 (OP.PER.III,
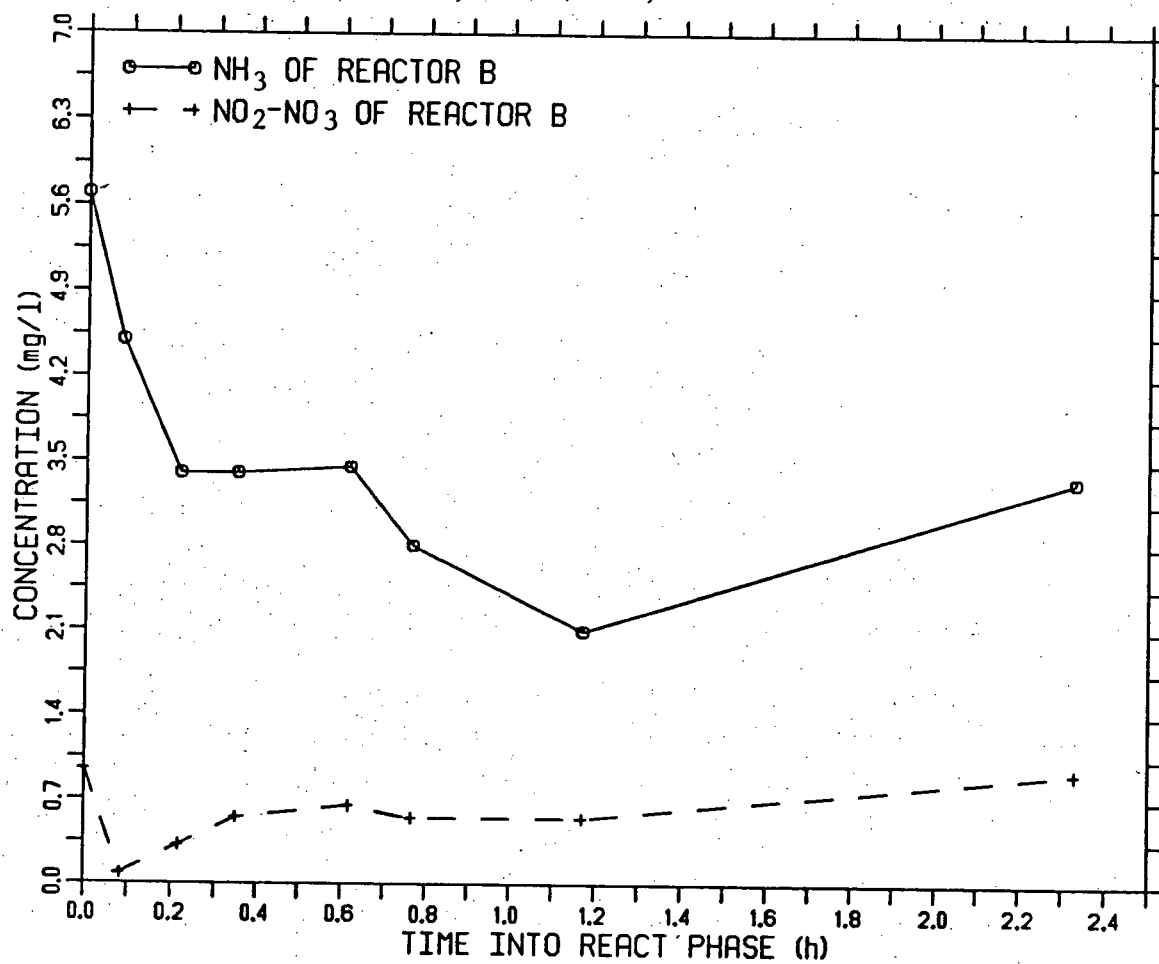4h CYC., 1.0 1/CYC.)

FIG. 4.18 : PLOT OF AMMONIA AND NITRITE-NITRATE NITROGEN VS.
AERATION TIME IN TRACK ANALYSIS #6 (OP.PER.III,
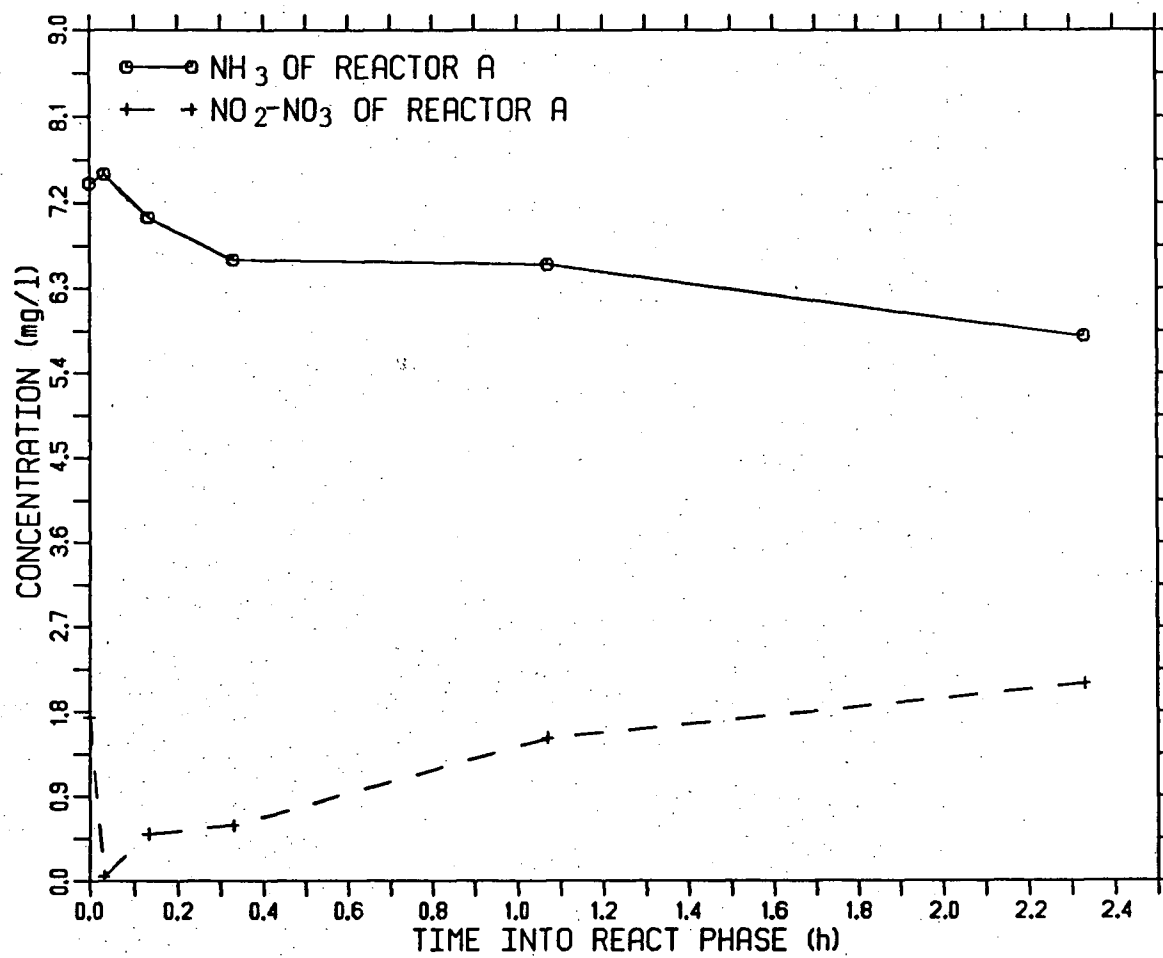4h CYC., 1.0 1/CYC.)

FIG. 4.19 : PLOT OF AMMONIA AND NITRITE-NITRATE NITROGEN VS. AERATION TIME IN TRACK ANALYSIS #7 (OP.PER.III, 4h CYC., 1.0 1/CYC.)
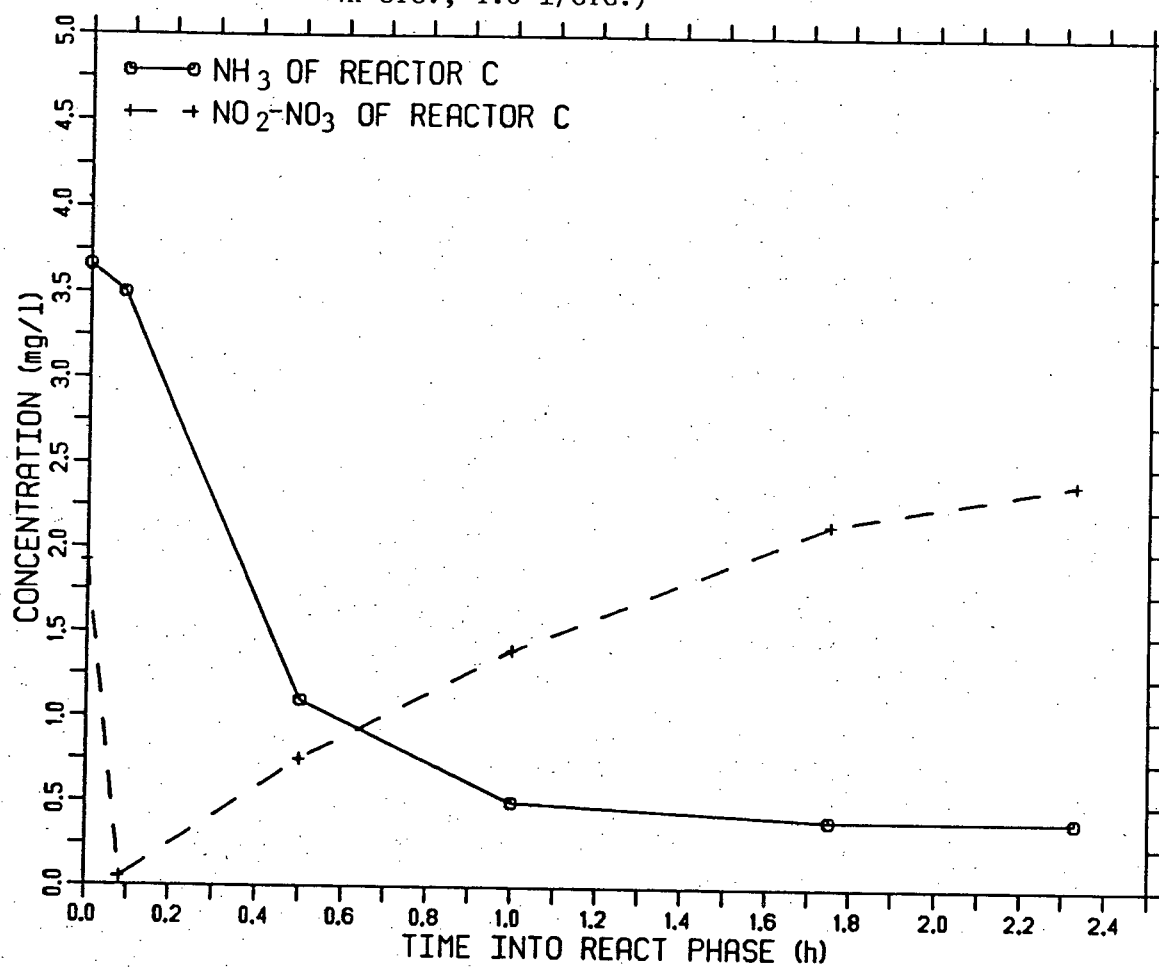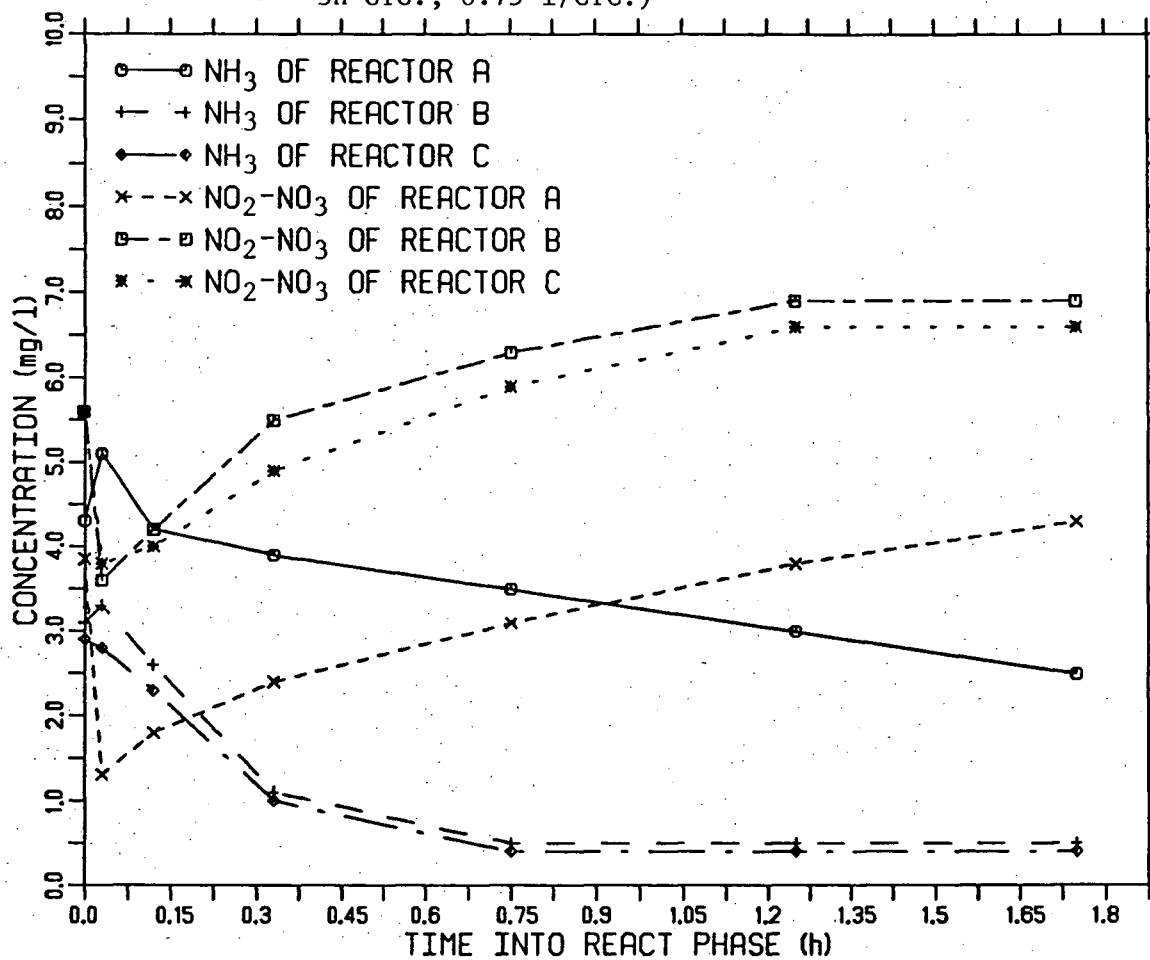
FIG. 4.20 : PLOT OF AMMONIA AND NITRITE-NITRATE NITROGEN VS.
AERATION TIME IN TRACK ANALYSIS #8 (OP.PER.IV,
3h CYC., 0.75 1/CYC.)

ammonium oxidation proceeded at a slower rate for the rest of the aeration phase.

3. Instantaneous denitrification occured during the beginning of the react phase — when feed was added and mixing befell through aeration.

4. Except for the initial denitrification 'dent' on the NO_2-NO_3-N curve, the shape of the nitrification curve in general changed in response to that of ammonium oxidation.

The specific rates of ammonia removal (reported as mg/l NH_3-N removed per g VSS per h) during the track experiments were computed and summarized in table 4.13. The corresponding average nitrification rate varied between 8 to 37% of that of ammonium removal (see table 4.12). This oxidation rate was computed by linear regression of the data points obtained prior to the levelling off of the NH_3-N curve : only the initial removal rate was computed.

From the results of table 4.13, it is quite obvious that B and C were more efficient, in terms of NH_3-N removal, than A in all the operation periods; but the difference between B and C was more erratic and unpredictable.

4.2.4 DENITRIFICATION EXPERIMENTS

Results from the track analyses of NH_3-N and NO_2-NO_3-N indicated the occurance of uncontrolled denitrification in the reactors during the beginning of the aeration phase. Consequently, a few independent investigations were carried out to further reveal the nature and extend of this denitrification process.

Due to the independent nature of these experiments, it is deemed more suitable to present the experimental procedures and results together in this section.

4.2.4.1 DENITRIFICATION EXPT.#1 (7/6/84, OPER. PERIOD IV)

OBJECTIVE : To deterimine whether feeding was necessary for denitrification to occur.

PROCEDURE : (Same for all three reactors)

- Removed 600 ml of mixed-liquor (ML) approximately 2 min before end of aeration.

- Let the ML settle in a beaker for 45 min (the assigned SETTLE phase length for operating period IV).

- Drew 10 ml of supernatant from the beaker after the assigned sedimentation period. Preserved with $HgCl_2$ for analysis by autoanalyser.

- Mixed the remaining ML with magnetic stirrer for three minutes.

- Drew 20 ml of ML, filtered into test tubes containing $HgCl_2$ to preserve for analysis.

- Collected approximately 200 ml of feed from the reservoir.

- Added 100 ml of the feed into the beaker containing the remaining ML while mixing with magnetic stirrer continued.

- Collected 20 ml of ML after 1 and 3 minutes of mixing with feed, filtered into test tubes with $HgCl_2$ to preserve for analysis.

- Preserved the remaining of the feed collected from the reservoir for BOD_5 NH_3-N and NO_2-NO_3-N analyses.

RESULTS : tabulated in table 4.14

CONCLUSIONS : The results showed that very little denitrification took place when settled ML was stirred without the introduction of new

feed. This small amount of denitrification may be the result of residual substrate utilization. When a proportionate amount of feed was added and the mixture stirred, further denitrification occured to a much higher degree.

4.2.4.2 DENITRIFICATION EXPT.#2 (8/6/84, OPER. PERIOD IV)

OBJECTIVE : To determine the extent of denitrification by using influent with artificially added nitrate.

PROCEDURE : (only reactor B was studied)

- Sampled and preserved influent feed and final effluent of previous cycle for analysis.

- Added 100.0 ml of standard nitrate solution (845 ppm) to the reactor at the end of FILL.

- Sampled ML at close intervals during the first 30 min for analysis.

RESULTS : See fig. 4.21

DISCUSSION : The result of this experiment shows that denitrification occured only to a limited extend and that it only occured during the beginning of aeration. Figure 4.21 shows that the NO_2-NO_3-N concentration took an initial dive, then increased again to 16.9 mg/l and stayed there until the end of the cycle. The NH_3-N concentration dropped steadily from 4.15 mg/l at the beginning of aeration to 1.00 at the end of SETTLE. However, this drop in NH_3-N concentration was not accompanied by a corresponding increase in NO_2-NO_3-N level as was the case in all the other track analyses. It is therefore suspected that the high initial NO_2-NO_3-N level might have some inhibitory effects on the nitrifiers (Loehr,1977); the ammonium removal observed during the remaining aeration period was therefore suspected to be a result of the other previously discussed processes (see 4.2.1).

TABLE 4.14 : RESULTS OF DENITRIFICATION EXPT. #1

SAMPLE		NH_3-N (mg/l)	NO_2-NO_3-N (mg/l)	D.O.(mg/l)
FEED		11.50	0.70	
SUPERNATANT	A	1.60	0.72	
AFTER 45 min	B	0.80	1.20	
SETTLE	C	0.60	1.05	
NO FEED,	A	1.90	0.40	2.80
MIXED 3 min	B	0.70	1.08	3.20
AFTER SETTLE	C	1.40	0.98	3.00
THEORETICAL	A	3.08	0.72	
VALUE AFTER	B	2.40	1.13	
FEED	C	2.23	1.00	
MIX 1 min	A	3.30	0.25	0.40
AFTER FEED	B	2.50	0.55	0.30
	C	2.40	0.55	0.30
MIX 3 min	A	2.80	0.00	0.10
AFTER FEED	B	1.90	0.30	0.05
	C	1.70	0.40	0.05

FIG. 4.21 : RESULTS OF DENITRIFICATION EXPERIMENT #2 (8/6/84)
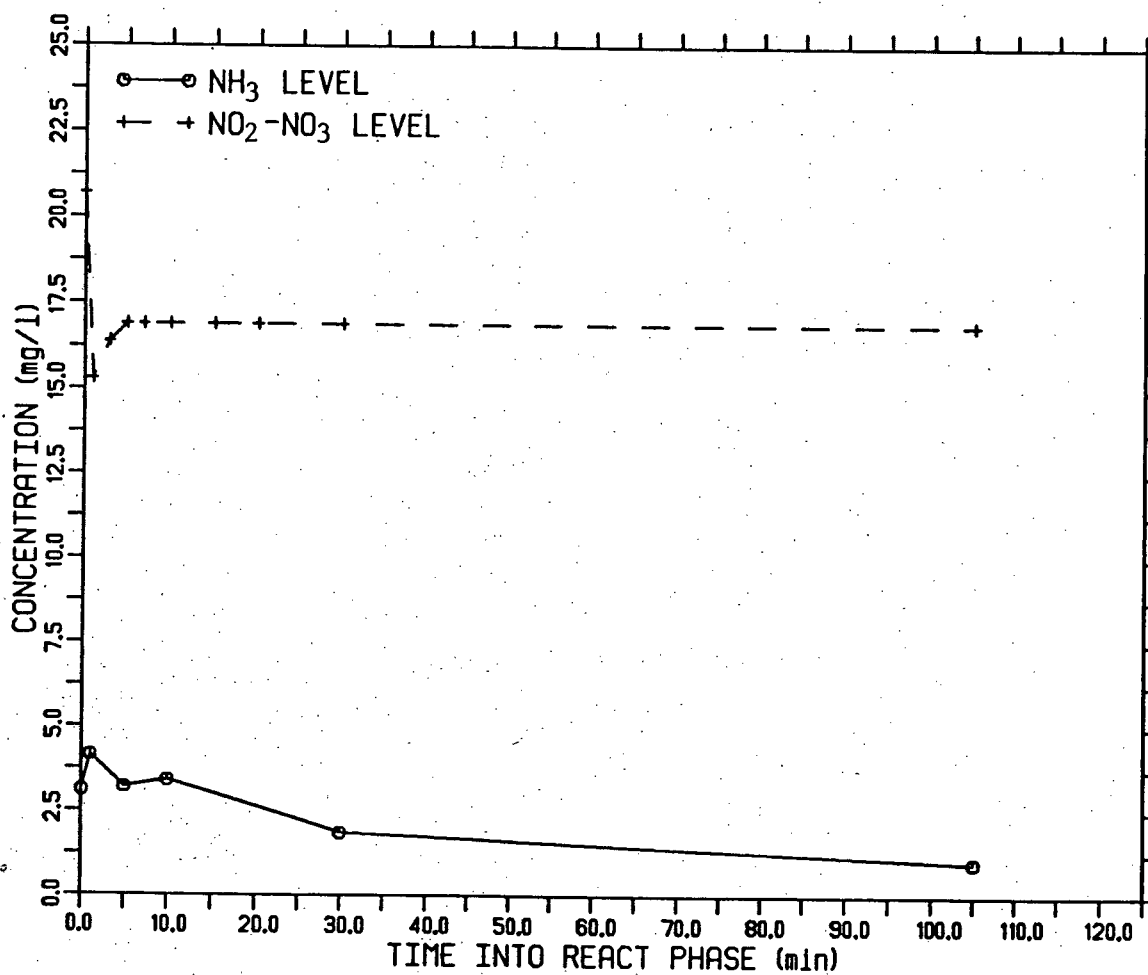
Only reactor B was used in this experiment; however, the same amount of artificial nitrate was also added to reactors A and C to maintain their comparable "life-histories" with B.

4.3 SETTLING VELOCITY

4.3.1 GENERAL DISCUSSION

A total of 11 settling velocity tests were carried out, covering operation periods II to VI. A schedule of the settling velocity tests can be found in table 4.03 and the results of the tests are plotted in figures 4.22 to 4.32. The flocculant settling velocity of the interface was determined graphically and tabulated in table 4.15 (see FIG.4.22 for example). Table 4.15 also includes the "lag time, t_l " for the three reactors. The lag time is the initial time lapse required by the mixed-liquor to form a distinct settling interface and to begin sedimentation at the flocculant rate.

As expected of typical systems that contain high concentration of suspended solids, both hindered settling (type 3) and compression settling (type 4) were observed in addition to discrete and flocculant settling in almost all the tests carried out. The only exceptions were observed in reactor A during operation periods V and VI ($3.7^\circ C$). During these operations, very little discrete and flocculant settling were observed in reactor A. The mixed-liquors of A sampled during these periods took a longer time to flocculate , and when they did, a very sharp and distinct interface would form and the sludge settled slowly through the test cylinder. The supernatant that formed as sedimentation proceeded contained almost no visible discrete flocs but was much

FIG. 4.22 : PLOT OF SLUDGE INTERFACE IN S.V.TEST#1
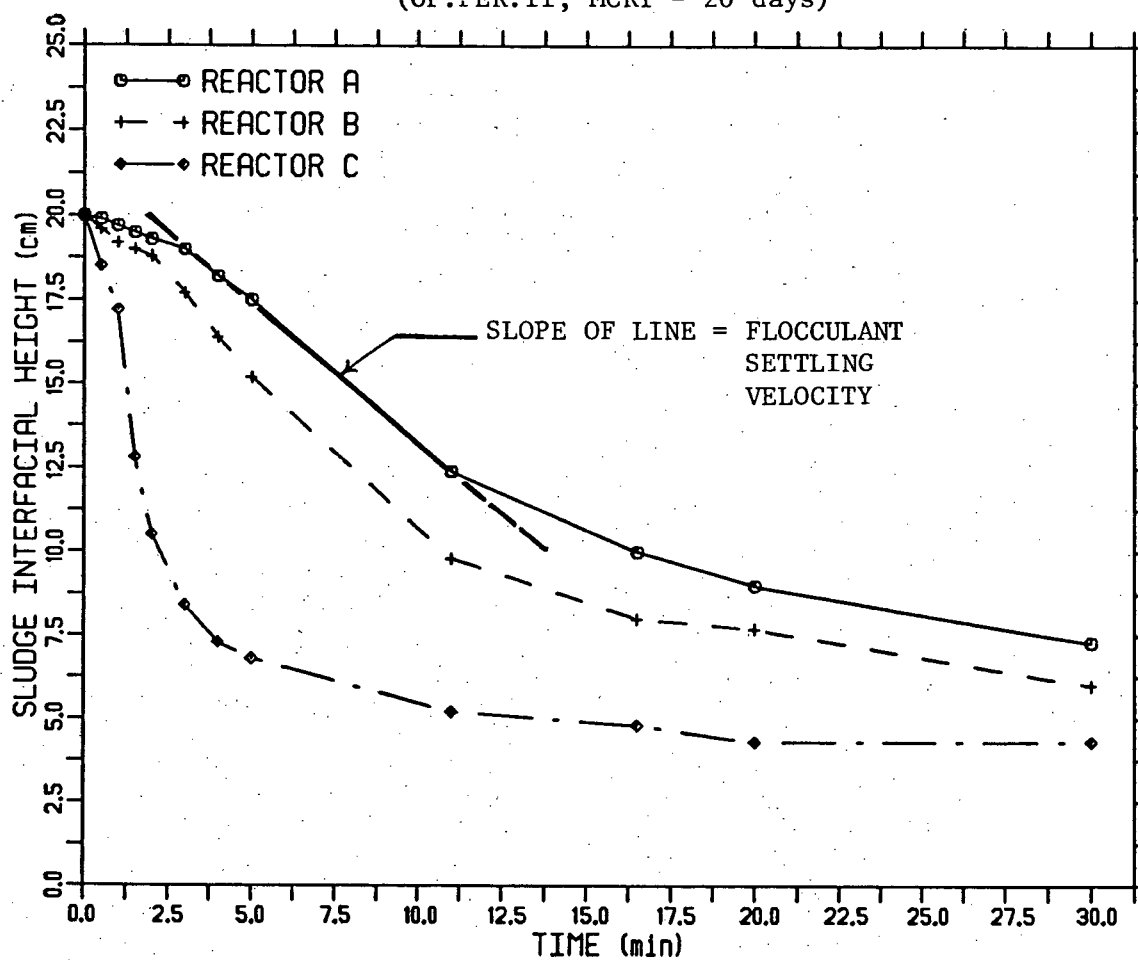(OP.PER.II, MCRT - 20 days)

FIG. 4.23 : PLOT OF SLUDGE INTERFACE IN S.V. TEST #2
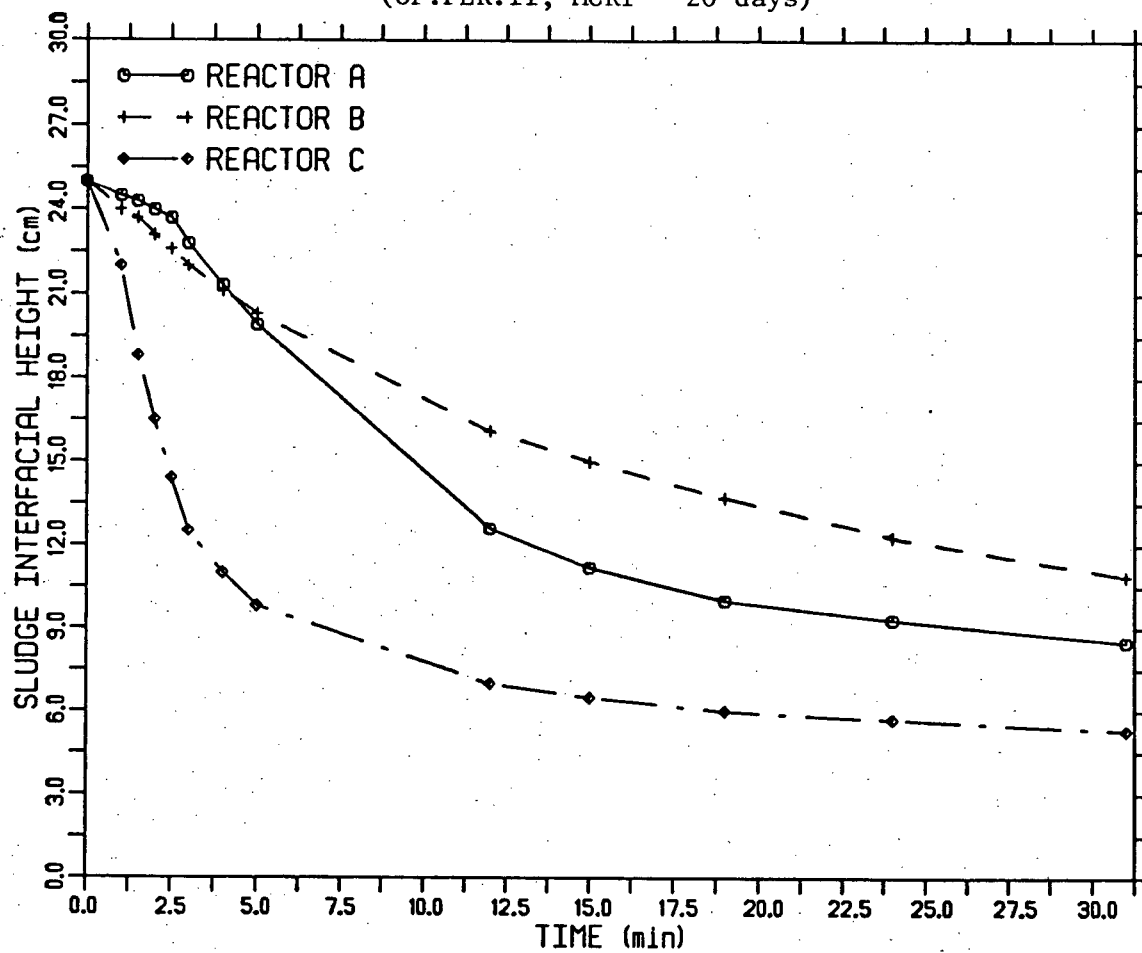(OP.PER.II, MCRT - 20 days)

FIG. 4.24 : PLOT OF SLUDGE INTERFACE IN S.V. TEST #3
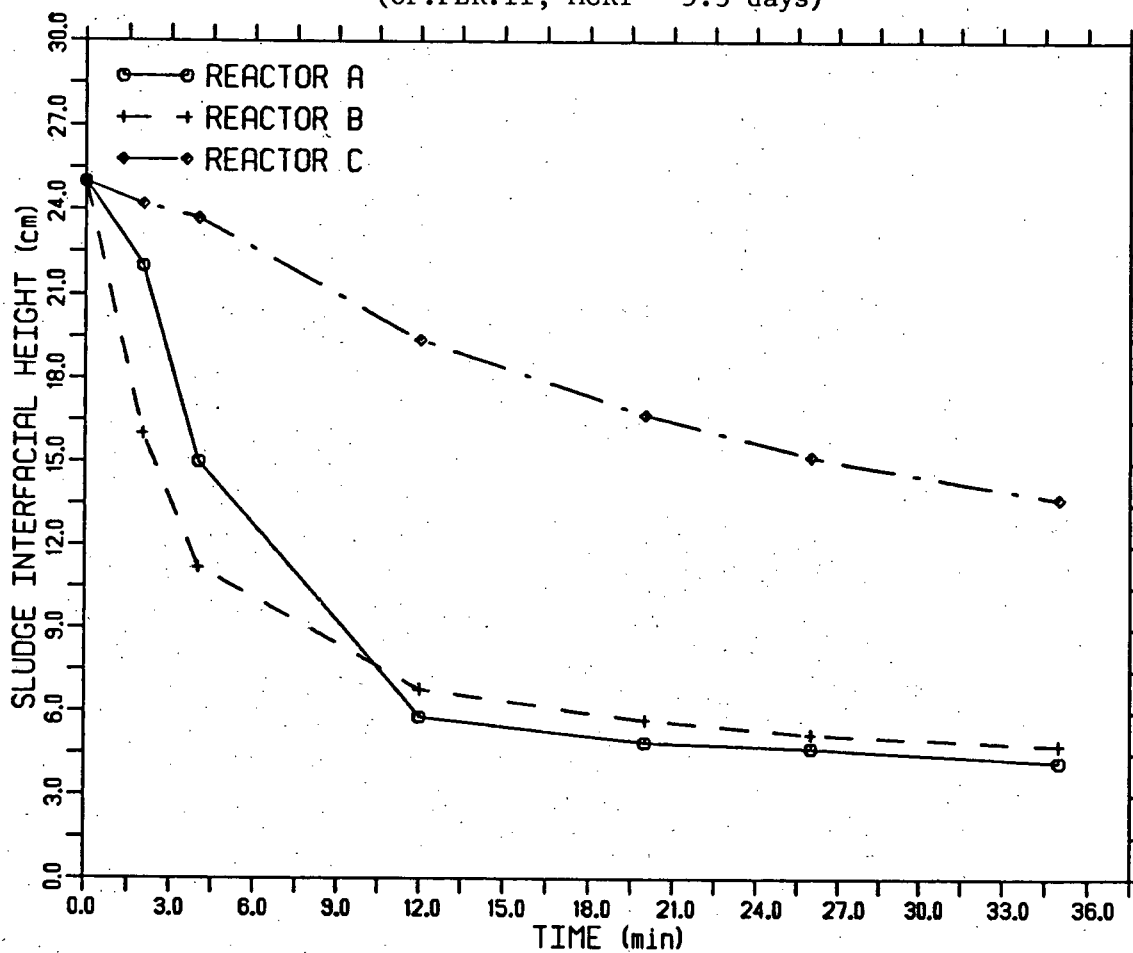(OP.PER.II, MCRT - 5.3 days)

FIG. 4.25 : PLOT OF SLUDGE INTERFACE IN S.V. TEST #4
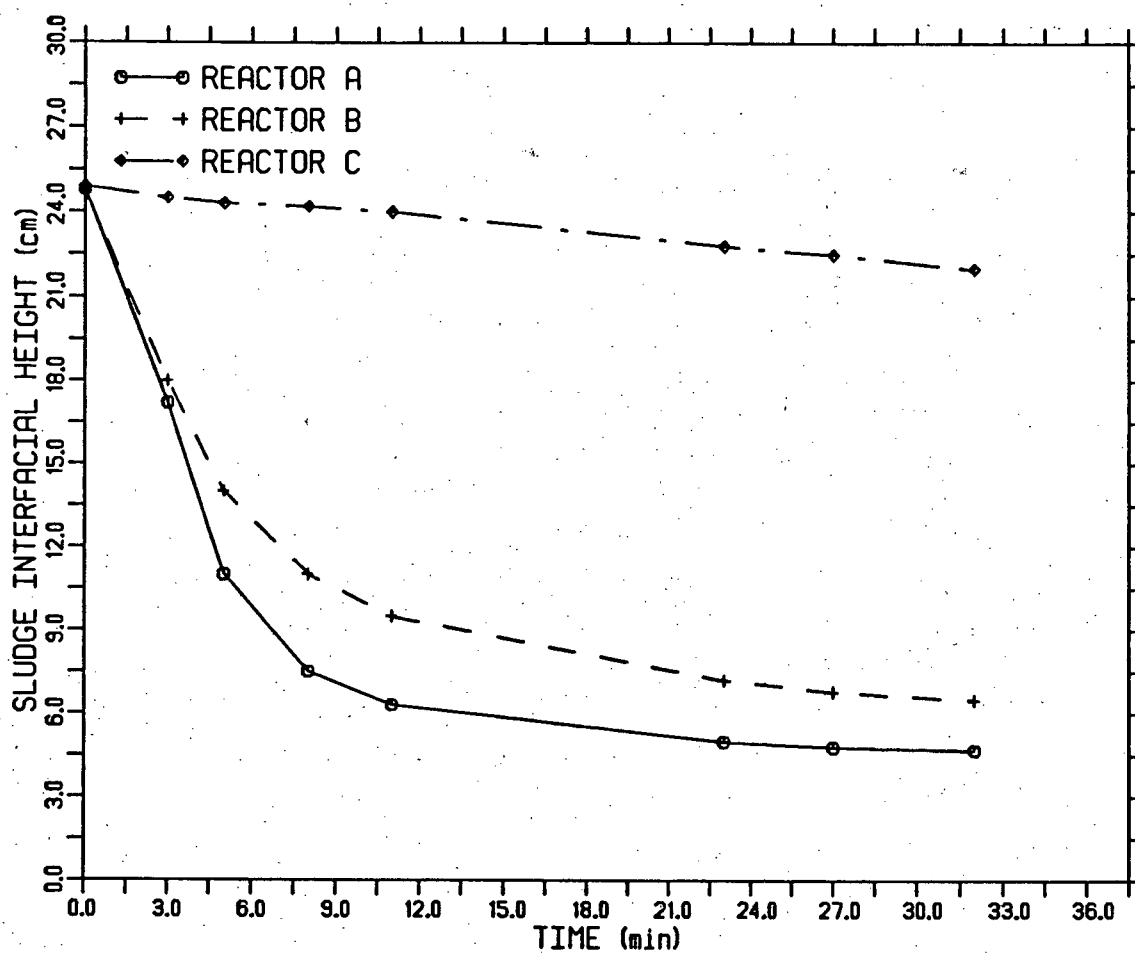(OP.PER.II, MCRT - 8.3 days)

FIG. 4.26 : PLOT OF SLUDGE INTERFACE IN S.V. TEST #5
(OP.PER.III, MCRT - 8.3 days)

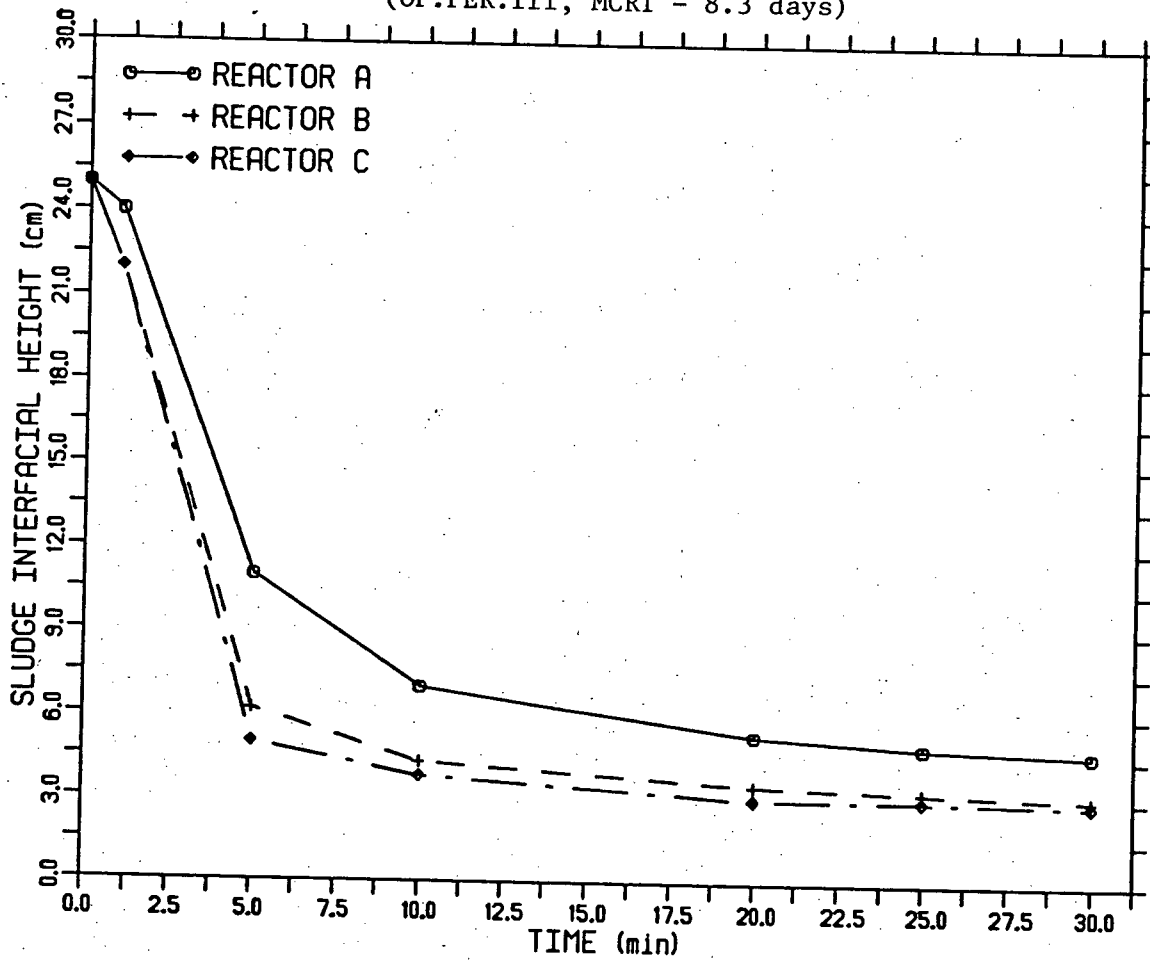FIG. 4.27 : PLOT OF SLUDGE INTERFACE IN S.V. TEST #6
(OP.PER. III, MCRT - 16.7 days)

FIG. 4.28 : PLOT OF SLUDGE INTERFACE IN S.V. TEST #7
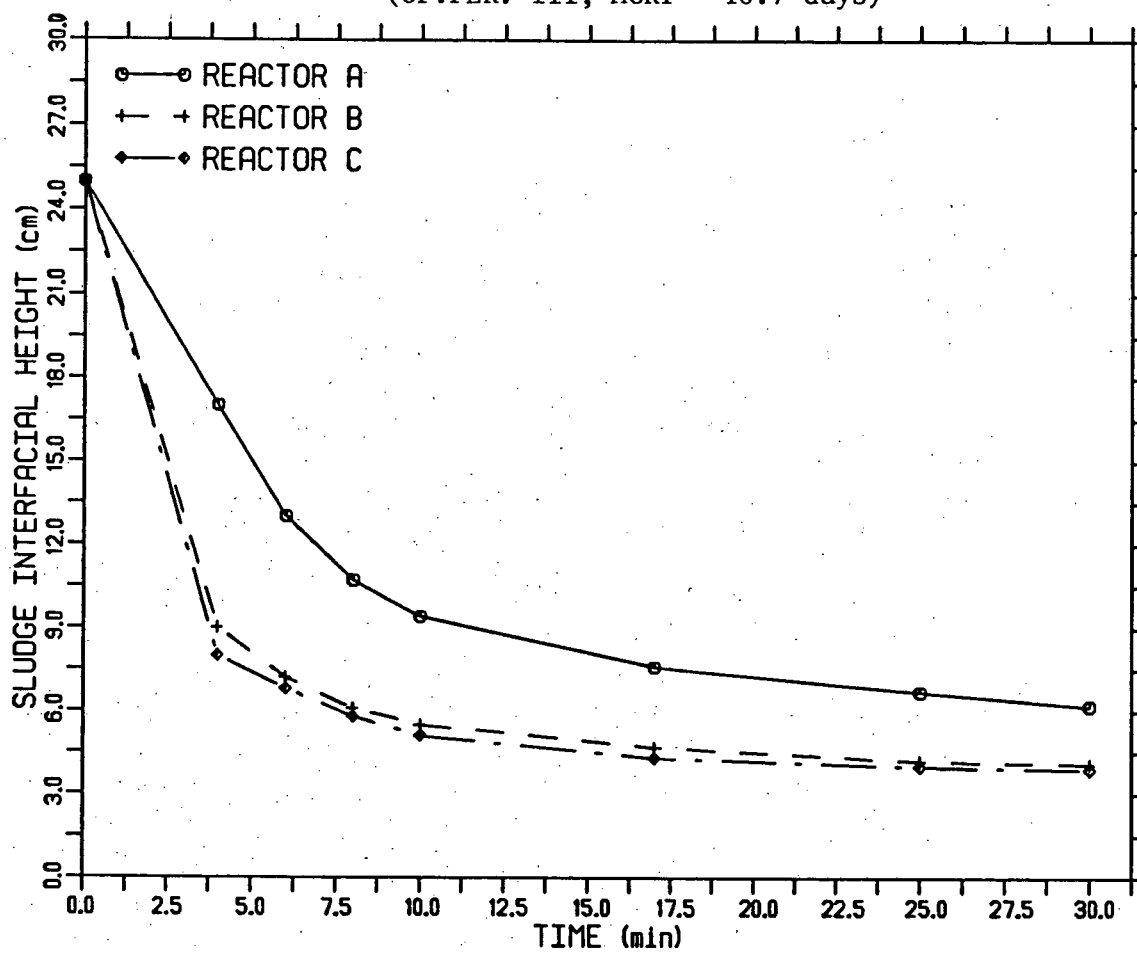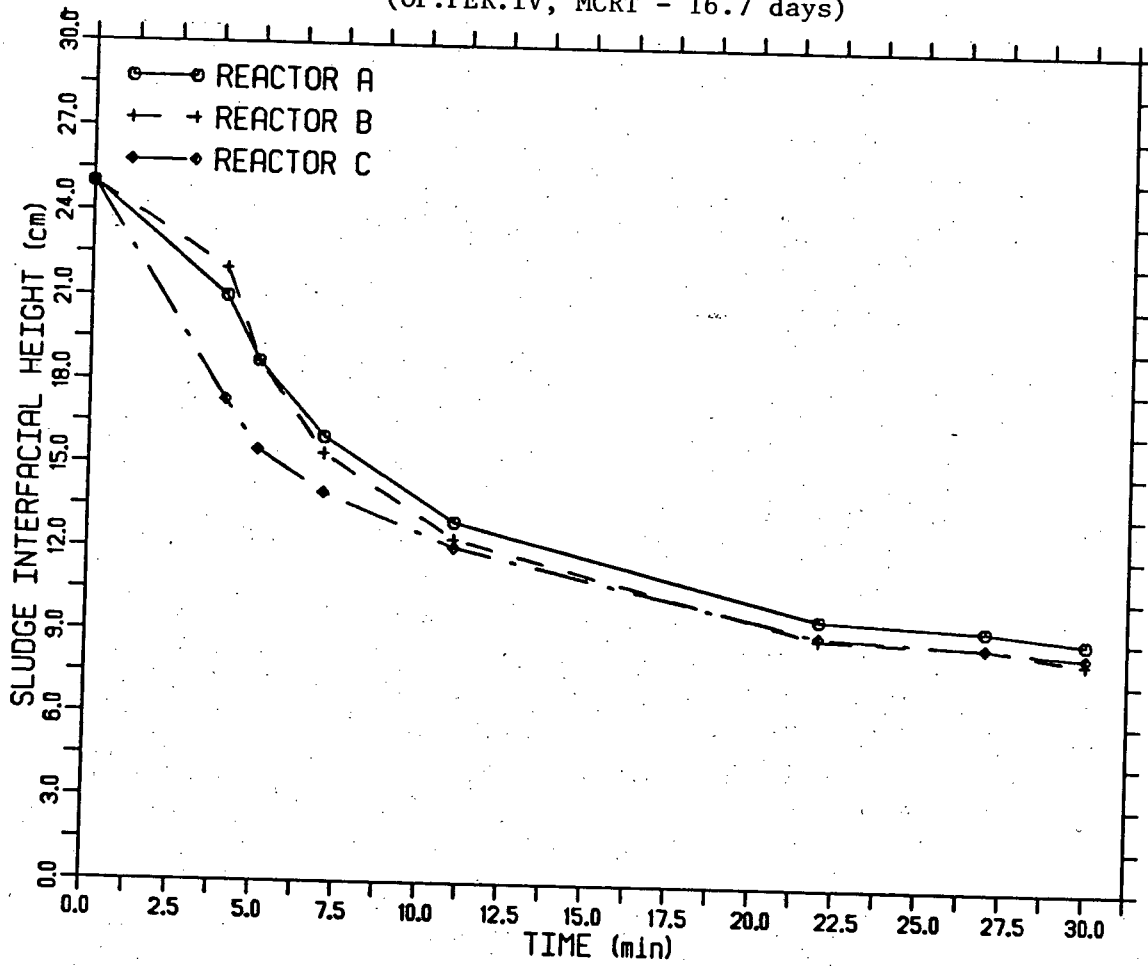(OP.PER.IV, MCRT - 16.7 days)

FIG. 4.29 : PLOT OF SLUDGE INTERFACE IN S.V. TEST #8
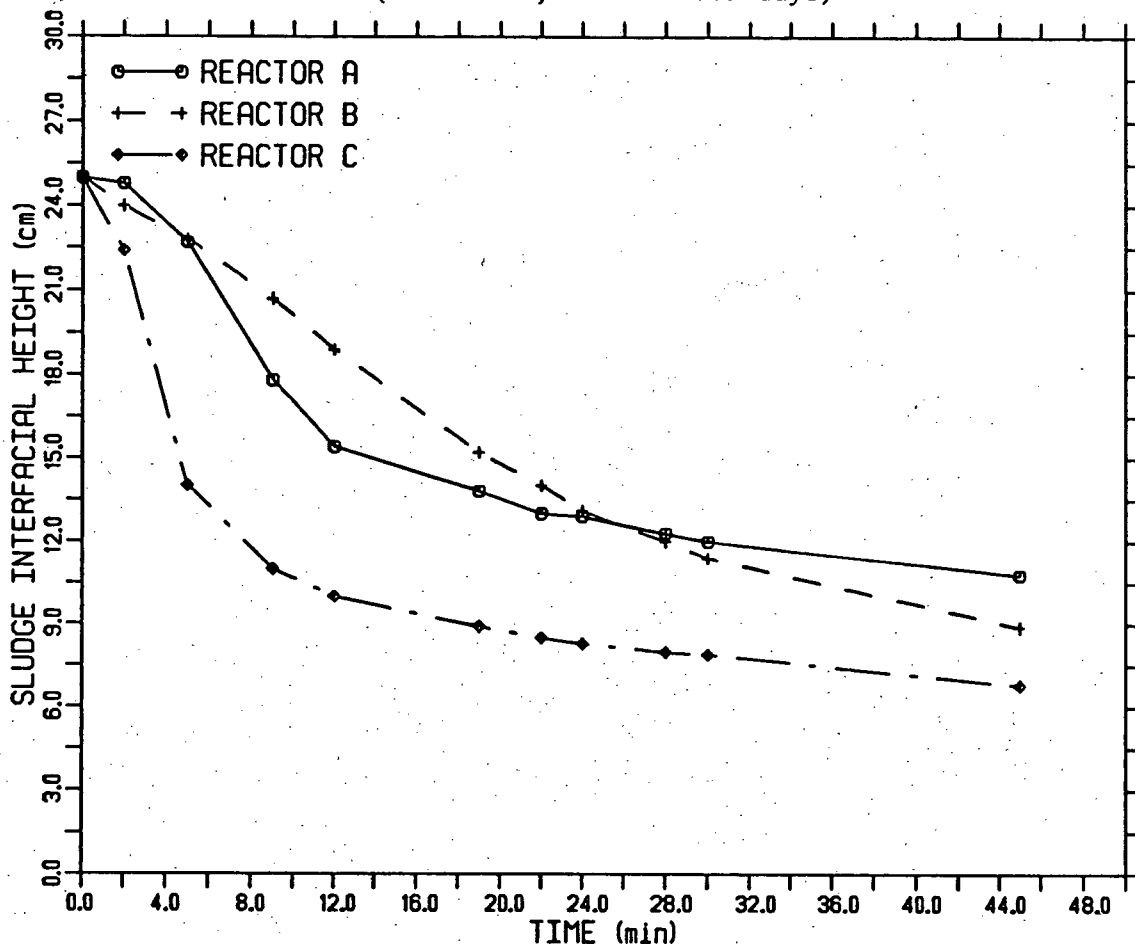(OP.PER.IV, MCRT - 16.7 days)

FIG. 4.30 : PLOT OF SLUDGE INTERFACE IN S.V. TEST #9
(OP.PER.V, MCRT - 16.7 days)
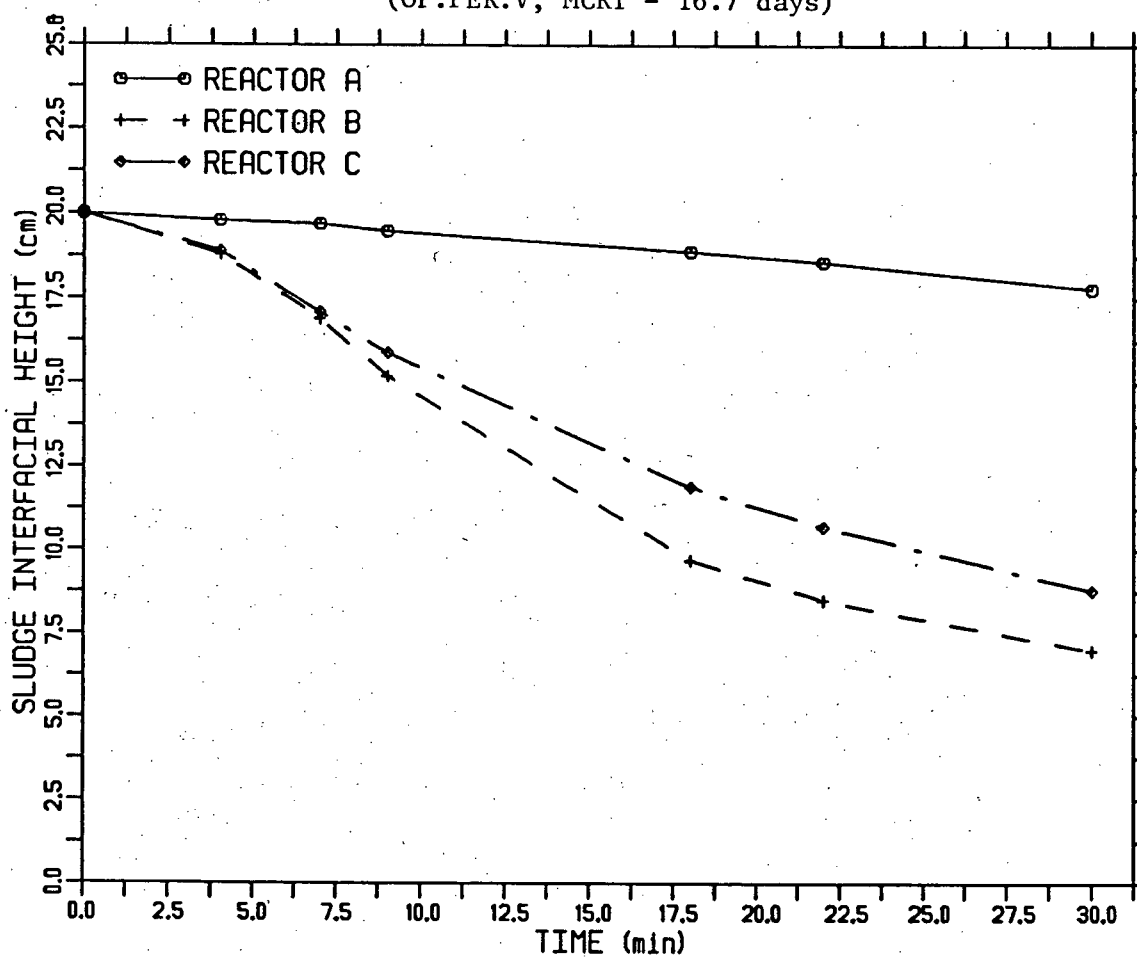
FIG. 4.31 : PLOT OF SLUDGE INTERFACE IN S.V. TEST #10
(OP.PER.V, MCRT - 16.7 days)
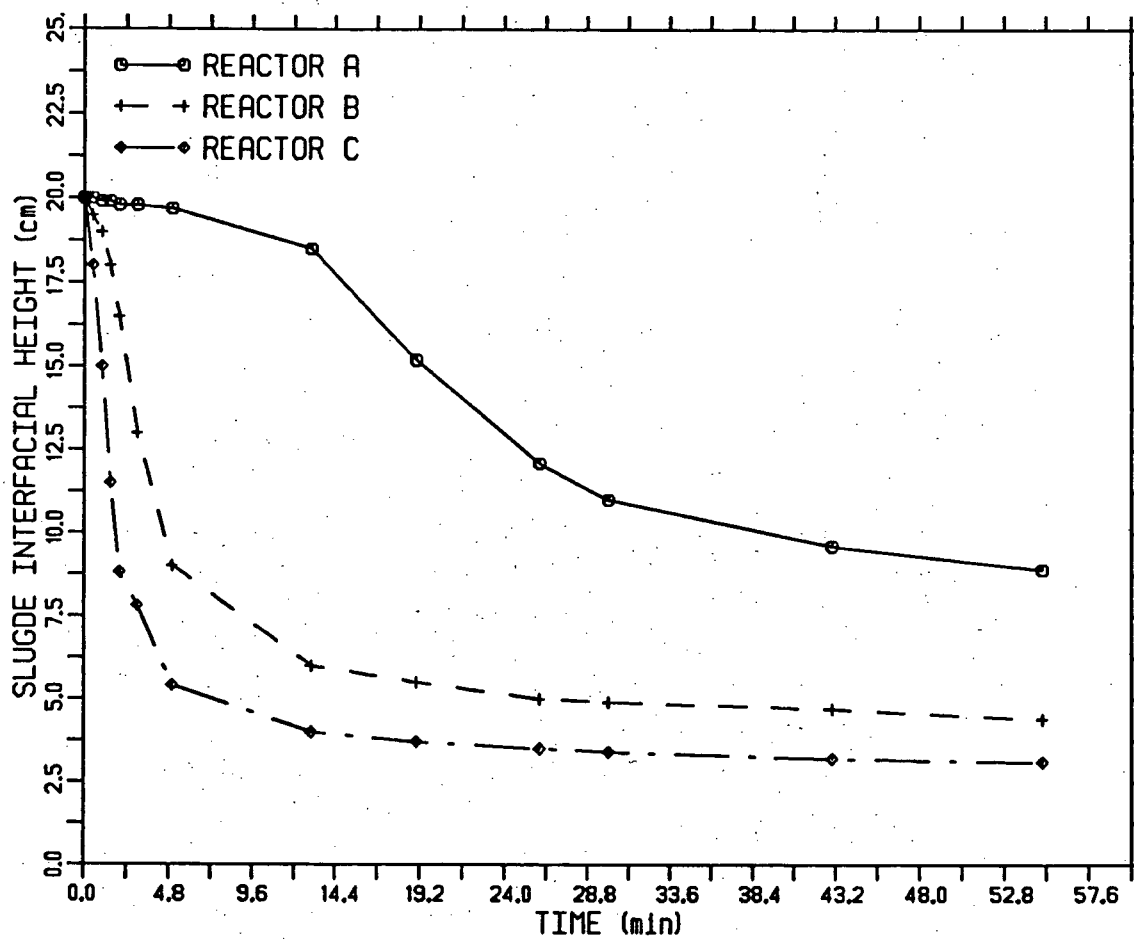
FIG.4.32 : PLOT OF SLUDGE INTERFACE (S.V. TEST#11 - TEST
CARRIED OUT INSITU THE REACTORS, OP.PER.VI)
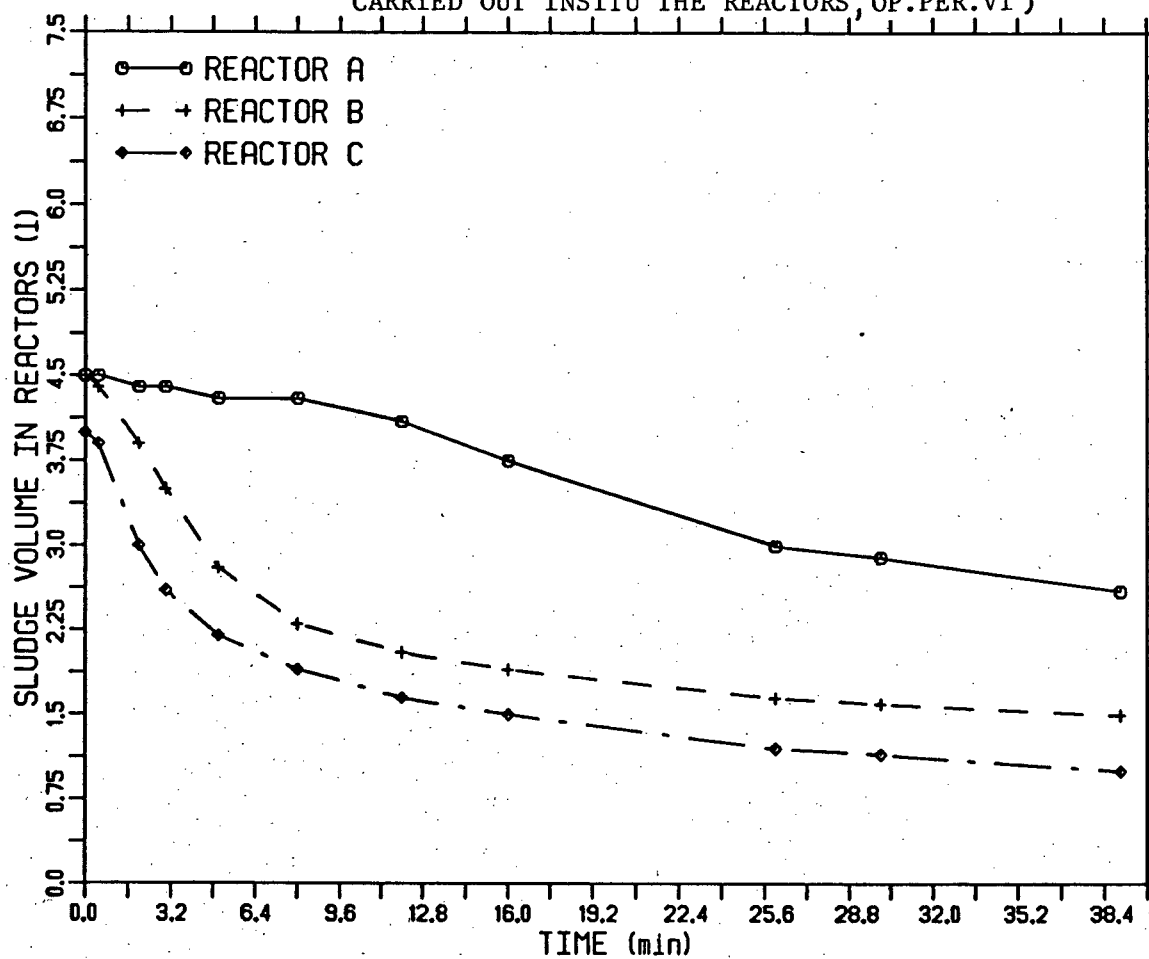
TABLE 4.15 : RESULTS OF SETTLING VELOCITY TESTS

TEST#	LAG TIME (min)			FLOCCULANT SETTLING VELOCITY (cm/min)		
	A	B	C	A	B	C
1	2.0	2.0	1.0	0.8	1.1	9.1
2	3.0	2.0	1.0	1.1	0.7	4.8
3	2.0	0.0	4.0	3.5	3.5	0.7
4	0.0	0.0	0.0	2.8	2.1	0.1
5	1.0	0.0	0.0	3.3	3.8	4.0
6	0.0	0.0	0.0	2.0	4.0	4.3
7	3.0	3.0	0.0	1.6	1.7	1.9
8	2.0	0.0	1.0	1.2	0.5	2.8
9	4.0	4.0	4.0	0.1	0.7	0.6
10	5.0	0.5	0.0	0.2	2.6	6.1
11	8.0	0.0	0.0	0.4	2.6	3.5
MEAN	2.7	1.1	1.0	1.6	2.1	3.5
STD.DEV.	2.3	1.5	1.6	1.2	1.3	2.7
RANGE	8.0	4.0	4.0	3.4	3.5	9.0

darker in colour than those of B and C.

The sharp temperature difference between the tested ML of A in periods V and VI, and the ambient room temperature in which the tests were carried out was expected to have only mild effects on the test results. Initial and final temperature of the tested ML were rarely differed by more than $4^{\circ}C$ during the 30 min test time. This assumption was verified by observations of the reactors, which revealed similar settling characteristics.

The computed settling velocities of all the three reactors seem eratic in both magnitude and nature. However, not once in the entire experimental period did any of the reactors suffered operational failure because the sludge could not settle past the outlet level at time of DRAW; even during periods when extremely high Sludge Volume Indices (SVI) were experienced by some of the reactors (see fig.4.40, 4.41).

4.3.2 DESIGN CONSIDERATIONS

Unlike the design of continuous-flow sedimentation units, batch sedimentation systems are rarely discussed with a unified approach in conventional water and wastewater treatment texts. This is due partly to the fact that batch systems are not yet considered to be a conventional process in the wastewater treatment industry, and maybe due partly to its simple nature.

However, the simplicity of the design of a batch sedimentation system does not render it less important. On the contrary, inadequate design of the sedimentation process of a semi-batch operation can mean failure of the entire system if the effluent is drawn from a fixed level. In which case insufficient settling would result in pumping

out part of the concentrated sludge.

Some foundamental differences exist between a continuous-flow settling unit and a SBR. It is therefore important to bear the following conceptual differences in mind when designing a sequencing batch reactor :

1. Continuous-flow sedimentation is carried out in a unit seperate from the aeration basin. In a SBR, the reactor serves both as the aeration chamber and the sedimentation tank. A SBR therefore has to accommodate the often conflicting requirements of an aeration chamber (which favours a relatively high height to horizontal area ratio for effective mass transfer) and that of a sedimentation tank (which usually favours maximization of horizontal cross-section area).

2. There is no inflow into a SBR during sedimentation.

3. Outflow of supernatant in a continuous system is usually by overflow; the final level of the sludge blanket is therefore not as critical except for sludge concentration purposes. Effluent withdrawal in a SBR maybe carried out from a fixed level in the reactor; the final level of the sludge blanket is therefore more crucial to the success of the entire operation.

4. Underflow sludge withdrawal is a continuous process in conventional activated sludge systems and has definite effects on the sedimentation process. Sludge withdrawal from a SBR is logically carried out at the end of SETTLE and does not directly affect the sedimentation process.

5. For the above reasons, standard design procedures for clarifiers and/or sludge concentration units such as those described in

Metcalf and Eddy (2nd Ed.,1979) or Keinath *et al.*,1976 cannot be applied directly to a SBR design without appropriate modifications.

Before a unified approach to the design of Sequencing Batch Reactors is fully developed, it is advisable to gain additional operation experience and information through pilot scale studies before finalizing a SBR design. Other considerations, such as the problem with freezing in cold climates for a long settle phase and possibly denitrification potentials, should always be taken into account when designing sequencing batch reactors.

4.4 SUSPENDED SOLIDS ANALYSIS

4.4.0.1 MIXED-LIQUOR SUSPENDED SOLIDS

The Mixed-Liquor Suspended Solids (MLSS) level of the reactors was monitored regularly throughout the experimental period and the results are plotted in figures 4.33, 4.34,and 4.35. Figures 4.33 and 4.34 are the continuous records of the Mixed-Liquor Total Suspended Solids (MLTSS) and the Mixed-Liquor Volatile Suspended Solids (MLVSS) respectively. Fig.4.35 is a superimposition of 4.33 and 4.34 to show how the MLVSS responded to the fluctuation in total suspended solids. The operation Mean Cell Resident Time (MCRT) and the corresponding operation periods are also shown at the bottom of fig.4.35. Sampling for the purpose of this record were generally carried out near the end of the react phase.

From figure 4.35, it can be seen that the quasi-steady state MLSS levels of the reactors responded gradually to the MCRT. It would therefore be possible to control the solid level of a SBR to some degree by adjusting the MCRT. However, it should be noted that while

FIG. 4.33 : TSS RECORD FOR THE FULL PERIOD OF THE EXPT. STUDY

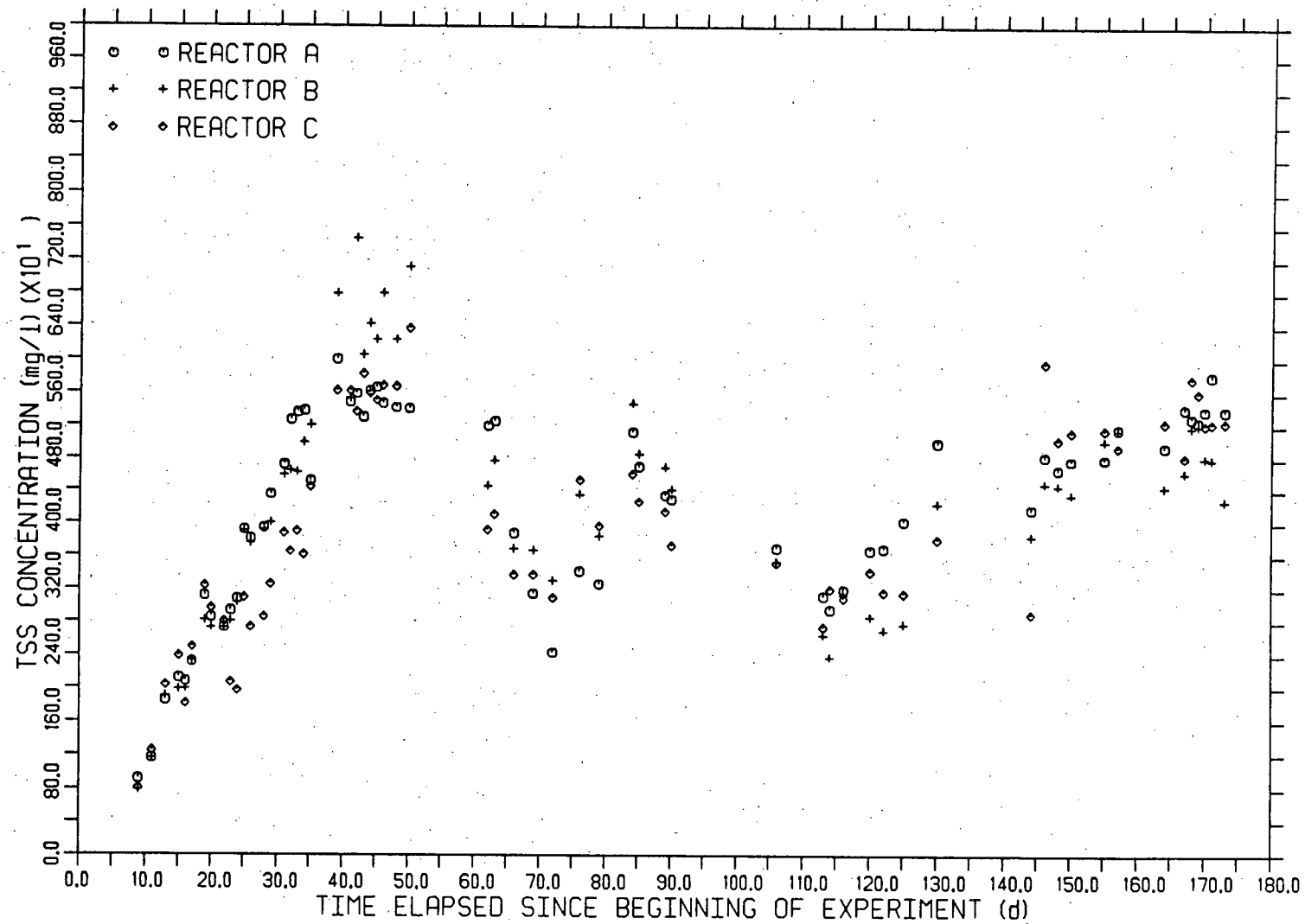FIG. 4.34 : VSS RECORD FOR THE FULL PERIOD OF THE EXPT. STUDY

⊙ REACTOR A
+ REACTOR B
◊ REACTOR C

VSS CONCENTRATION (mg/l) (X10¹)

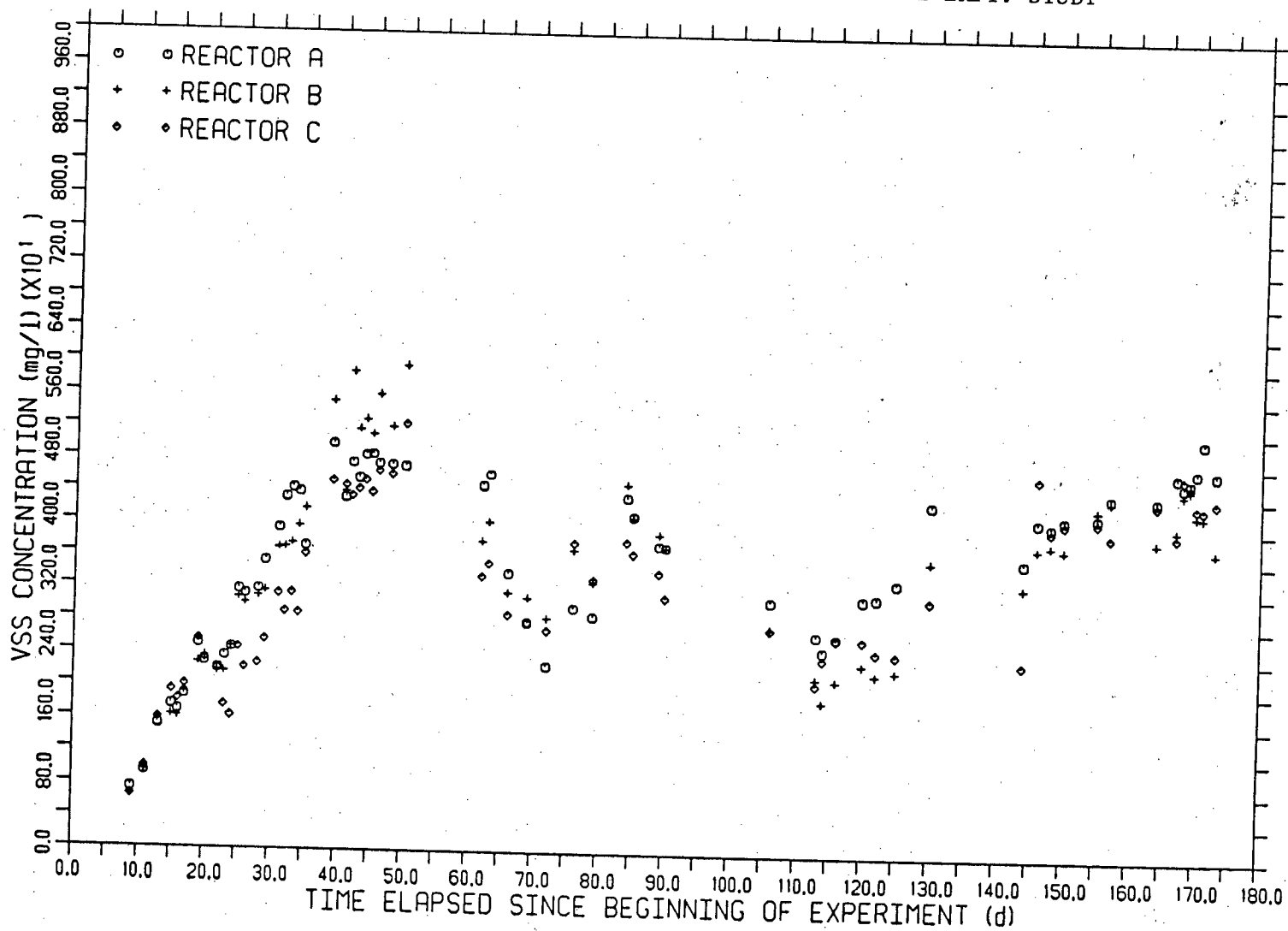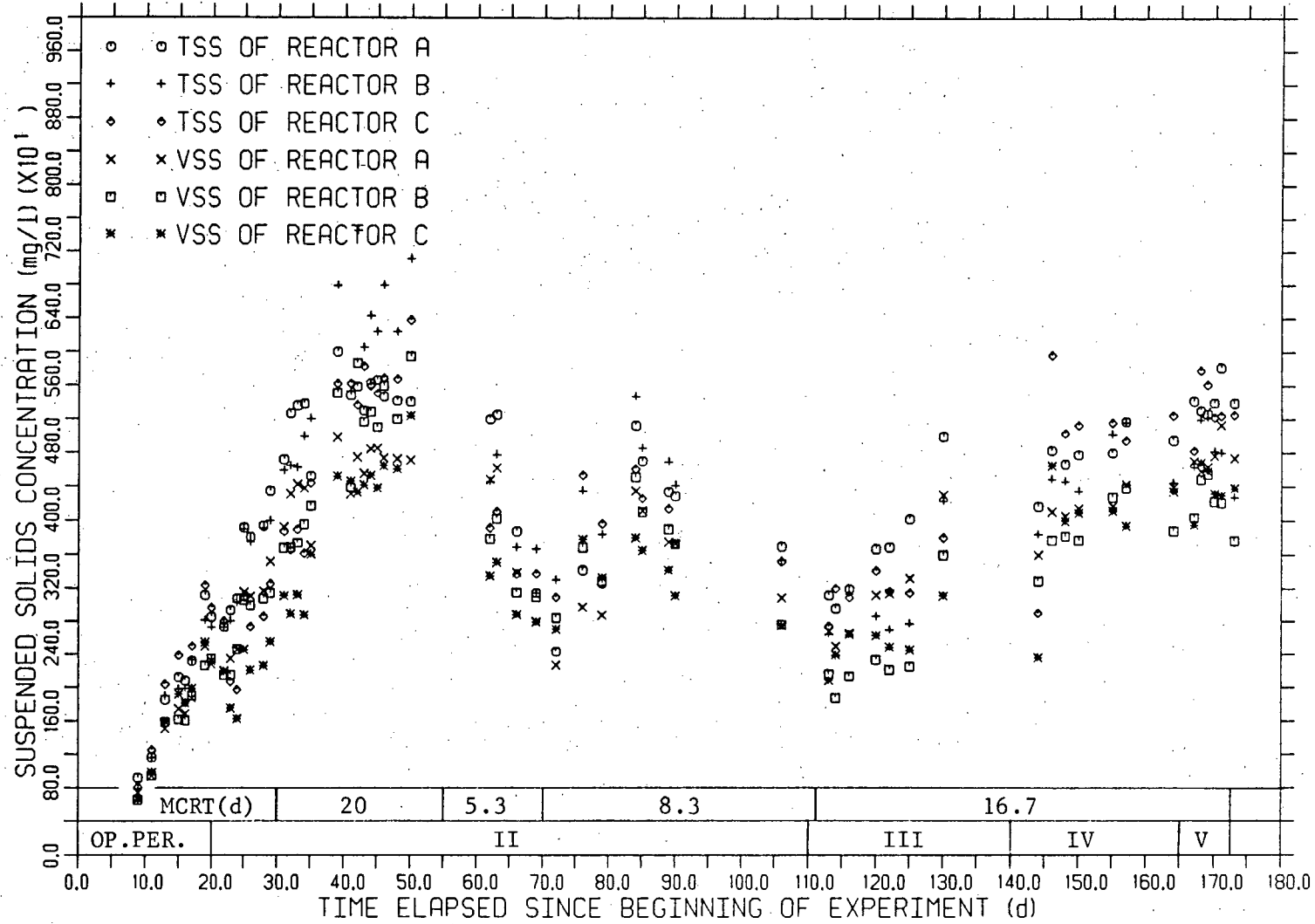TIME ELAPSED SINCE BEGINNING OF EXPERIMENT (d)

FIG. 4.35 : COMBINED TSS AND VSS RECORD FOR THE FULL PERIOD
OF THE EXPERIMENTAL STUDY

MCRT (which is simply the reactor filled volume divided by the ML wastage rate) controls the solid level to a certain extend, the maximum allowable solid level puts an upper limit on the MCRT; the maximum SS level is in turn dictated by the designed draw-down level of the reactor. The minimum MCRT, on the other hand, is dictated by such factors as settleability and generation time (Loehr,1977; Painter,1970). Should conflicting requirements occur, the reactor would have to be sized-up to decrease the maximum draw-down level.

Three MLSS track analyses were carried out during experimental periods II and IV. The results are plotted in figures 4.36, 4.37 and 4.38. Figure 4.36 shows that the MLTSS and MLVSS did not vary significantly during aeration. This is consistent with observations made by other researchers (Dennis and Irvine 1979; Irvine and Richter 1976). Results from figures 4.37 and 4.38 show greater fluctuation, particularly during the beginning of aeration. This fluctuation, however, is suspected to be due to sampling errors resulting from incomplete mixing of the chunkier sludge mass during the beginning of aeration.

4.4.0.2 INFLUENT-EFFLUENT TSS ANALYSIS

Results of the influent-effluent TSS analysis can be found in table 4.16. The statistical summary of the data can be found in table 4.17.

From table 4.17, it is quite clear that given the same sedimentation time, the ability of reactor A to remove SS was inferior to that of B and C. From observations made during the settling velocity tests, good settling characteristics were in general noted in all the reactors. The higher effluent SS in reactor A is therefore assumed to be due to fine suspensions that were too light to settle through the

FIG. 4.36 : PLOT OF TSS AND VSS VS. AERATION TIME IN TRACK ANALYSIS #1 (OP.PER.II, 6h CYC., 1.5 1/CYC.)
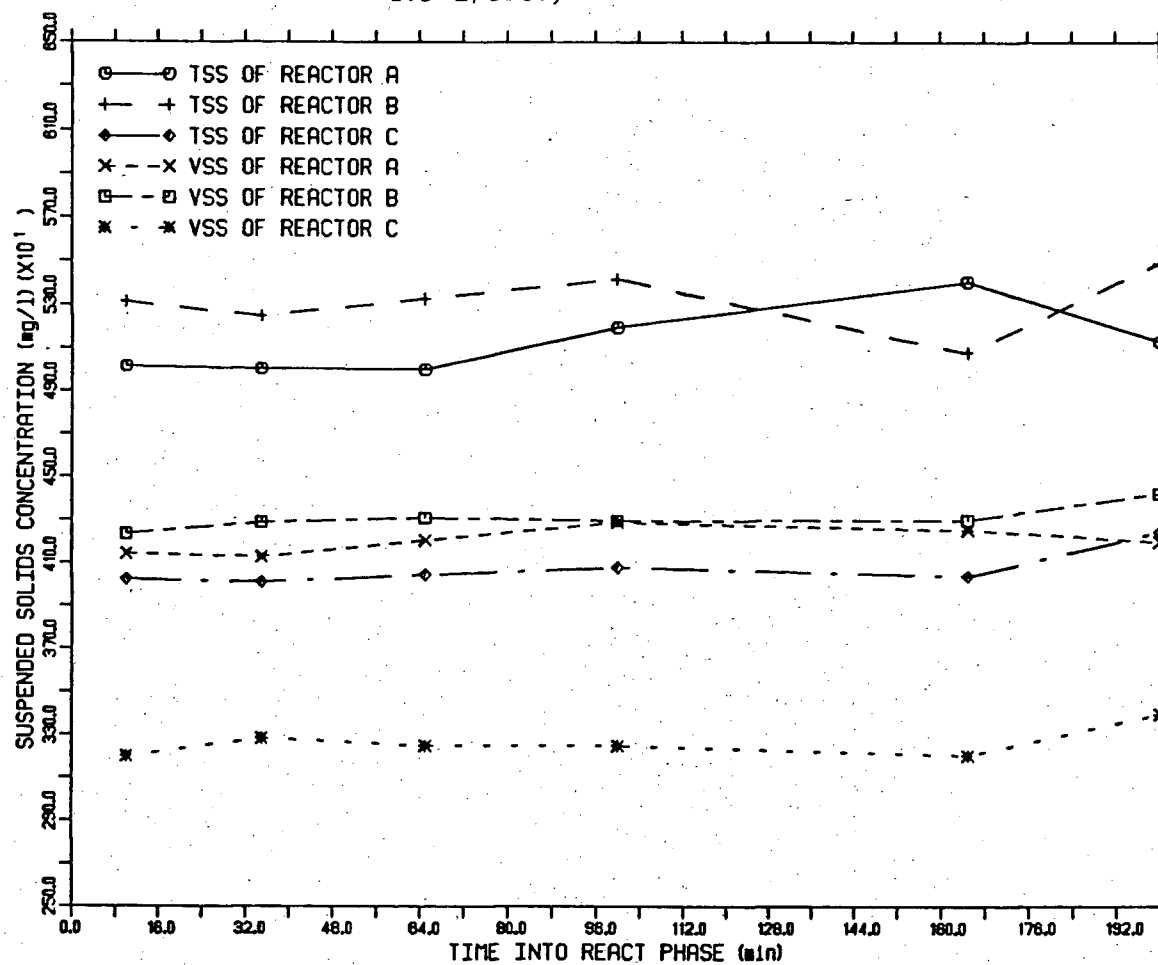
FIG. 4.37 : PLOT OF TSS AND VSS VS. AERATION TIME IN
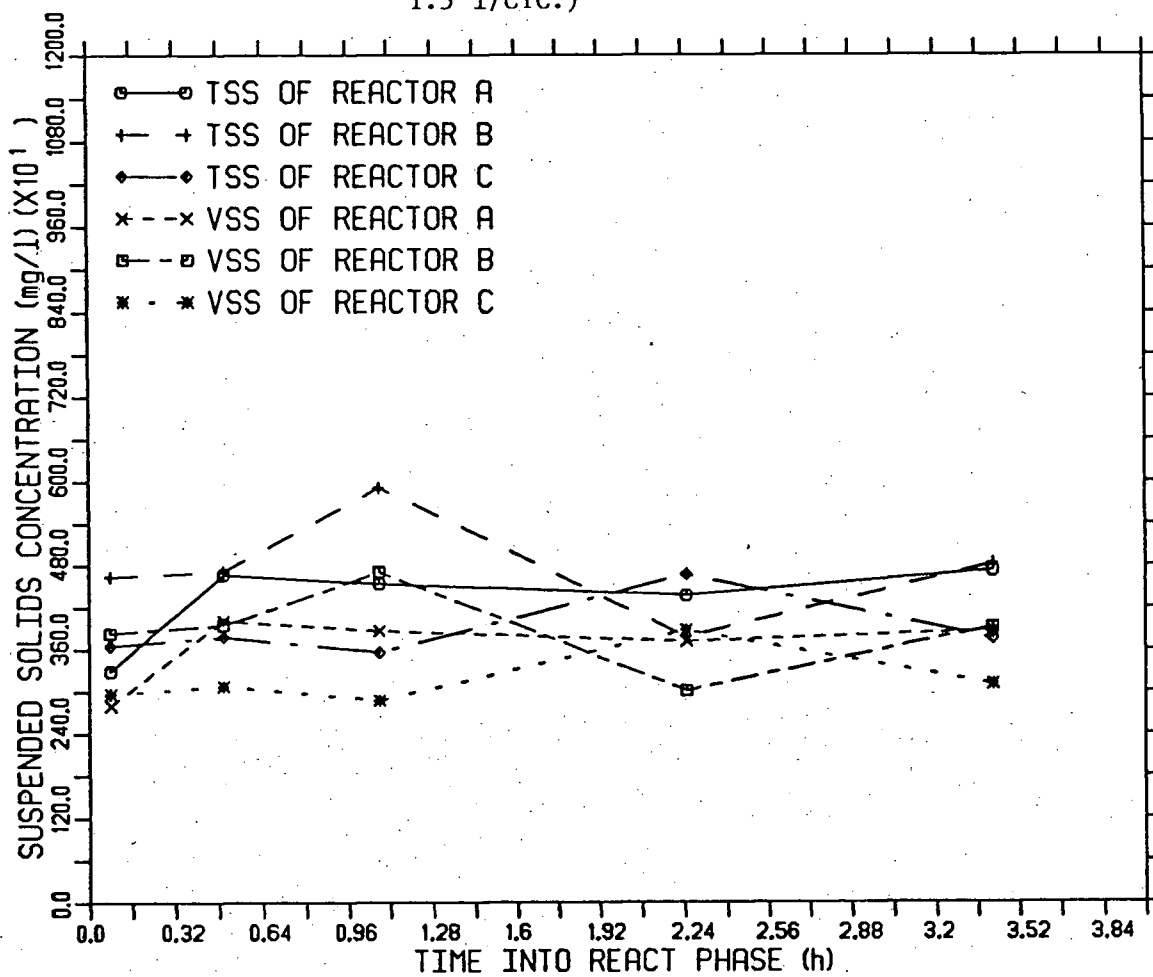TRACK ANALYSIS #3 (OP.PER.II, 6h CYC.,
1.5 1/CYC.)

FIG. 4.38 : PLOT OF TSS AND VSS VS. AERATION TIME IN TRACK ANALYSIS #8 (OP.PER.IV, 3h CYC., 0.75 1/CYC.)
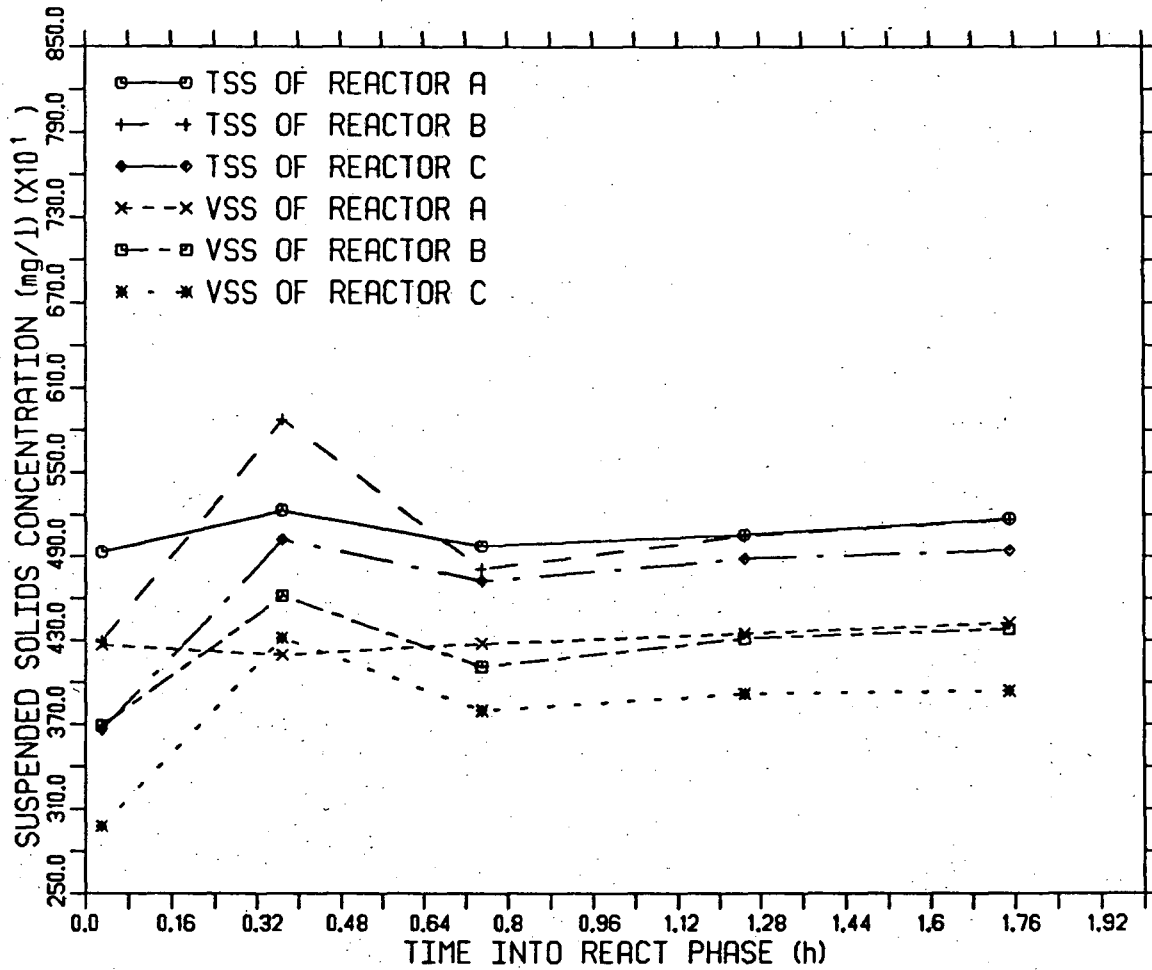
TABLE 4.16 : DATA OF TOTAL SUSPENDED SOLIDS (TSS) ANALYSIS

DATE (M.d)	PERIOD	INF.TSS (mg/l)	EFF. TSS (mg/l)			TSS REMOVAL (%)		
			A	B	C	A	B	C
2.14	II	236	39	6	6	83	97	97
2.28	II	---	60	38	50	--	--	--
4.03	II	1596	244	112	240	85	93	85
5.08	III	173	52	36	21	70	79	88
5.13	III	193	26	8	14	87	96	93
5.28	III	271	19	4	4	93	99	99
6.09	IV	68	11	2	1	84	97	99
6.21	IV	303	67	17	89	78	94	71
6.27	V	207	32	17	21	85	92	90
7.03	V	173	24	9	9	86	95	95
CYC# 4	VI	173	33	26	19	81	85	89
CYC#14	VI	180	36	12	17	80	93	91

TABLE 4.17 : SUMMARY OF TSS REMOVAL DATA

PERIOD	AVERAGE INFLUENT TSS(mg/l)	AVERAGE EFF. TSS (mg/l)			AVERAGE TSS REMOVAL (%)		
		A	B	C	A	B	C
II	916	114	52	99	84	95	91
III	212	32	16	13	83	91	93
IV	186	39	10	45	81	96	85
V	190	28	13	15	86	94	93
VI	177	35	19	18	81	89	90

more viscous and dense fluid of the low temperature supernatant within the given sedimentation time.

No notable difference in TSS removal capacity was noted between reactors B and C.

4.5 SLUDGE VOLUME INDEX

The Sludge Volume Index (SVI) is defined as "the volume in millilitres occupied by one gram of activated-sludge mixed-liquor solids, dry weight, after settling for 30 min in a 1000-ml graduated cylinder" in Metcalf & Eddy's text (1979).

The SVI reported in this study, as is commonly done in practice, is taken to be the volume ratio of the sludge after 30 min of settling, divided by the TSS concentration of the mixed-liquor expressed in g/ml. The SVI here is therefore determined as follows :

$$SVI = (V_2/V_1)/(TSS \text{ in g per ml})$$

V_1 and V_2 are the initial volume of the mixed-liquor and the final volume of the sludge after 30 min of settling, respectively.

The SVI of the reactors were monitored frequently and recorded in fig. 4.39. The actual % reduction in sludge volume at the end of the 30-min tests was also recorded in fig. 4.40.

The SVI is frequently used in continuous-system designs to determine the sludge recycling rate. In semi-batch designs, it can be used to estimate the maximum draw-down level. For example, if the MLTSS level is 2000 mg/l and the reactor volume is 500 m^3, the volume occupied by the solids (total 1000 kg) would be 100 m^3 if the design SVI is 100 ml/g and 200 m^3 if the design SVI is 200 ml/g. The corresponding maximum draw-down level can be calculated with knowledge of the reactor

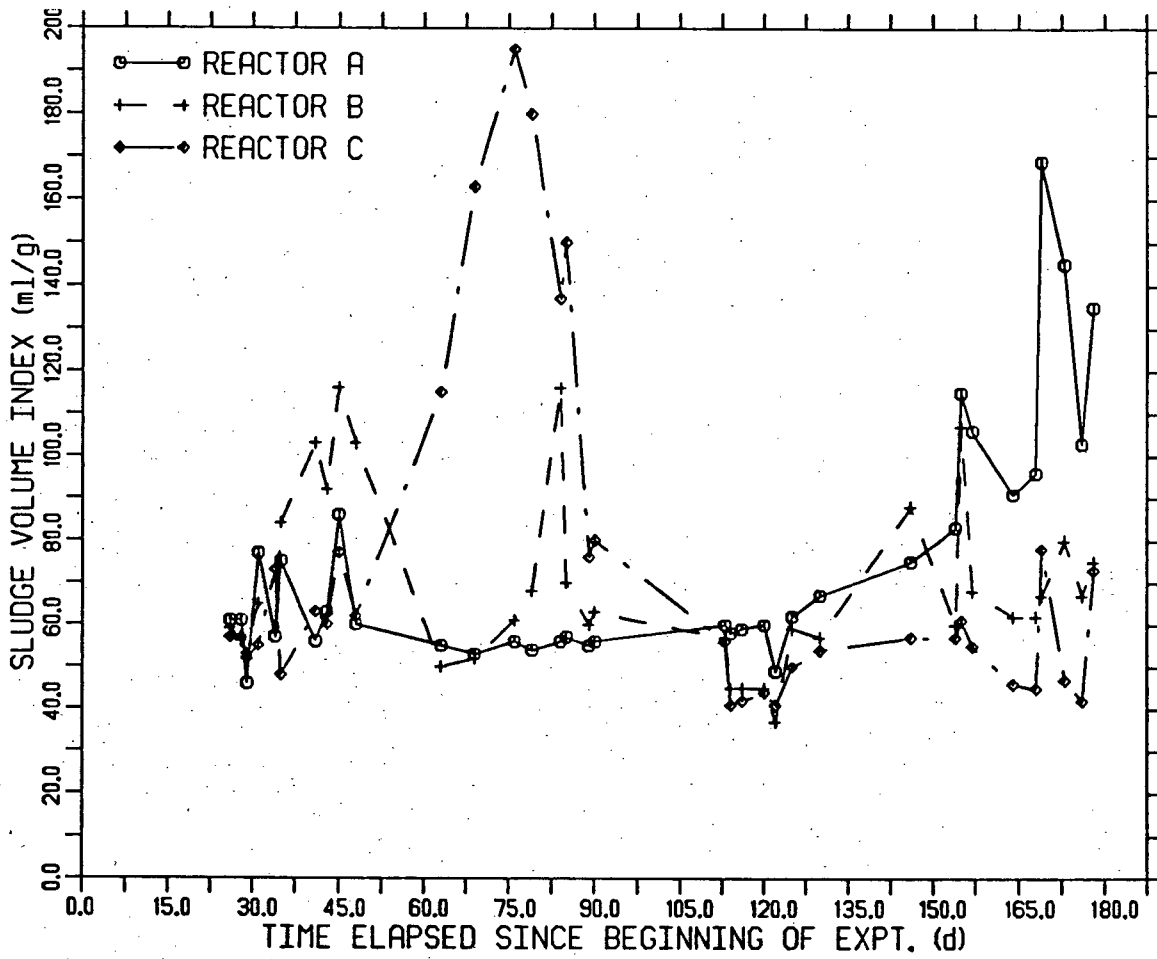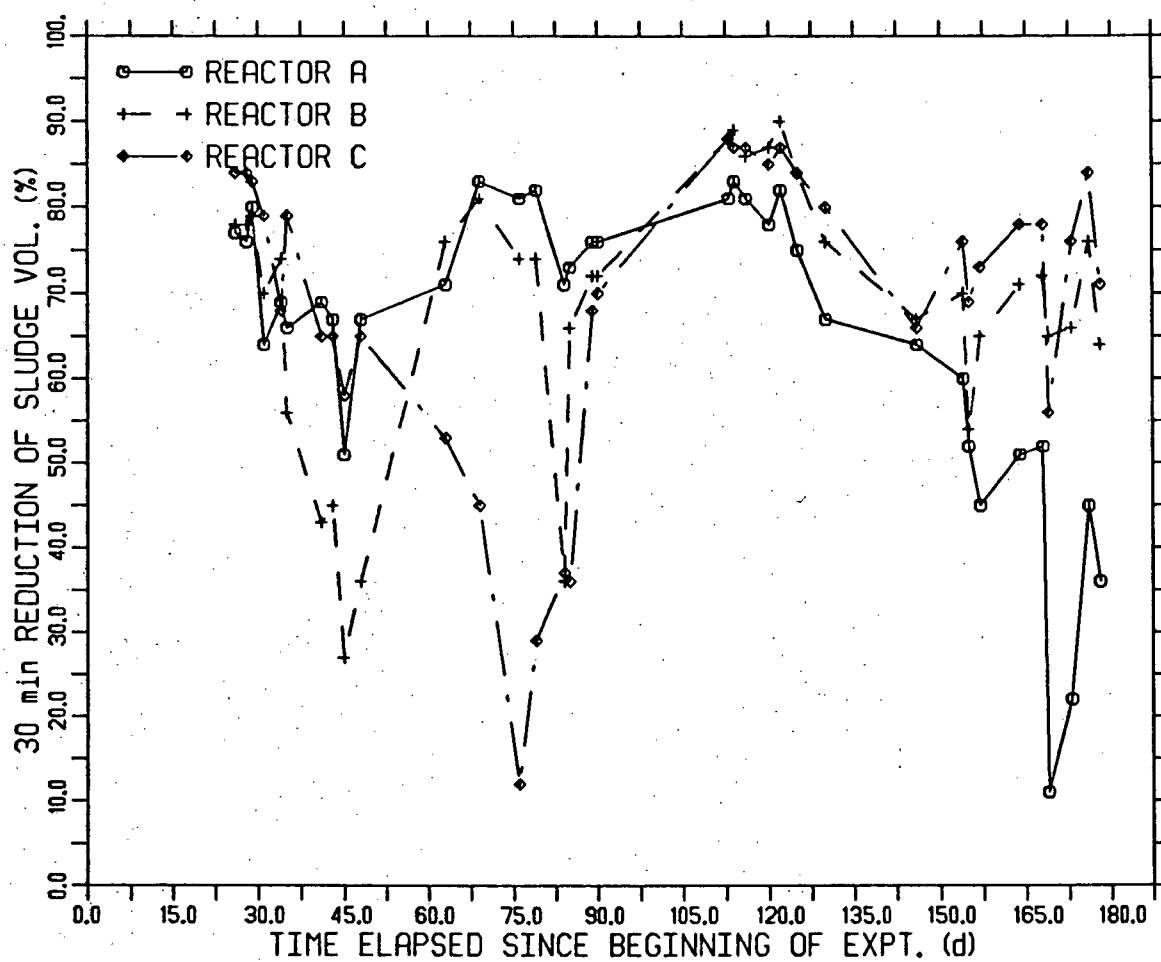FIG. 4.39 : SLUDGE VOLUME INDEX RECORD OF ALL EXPT. PERIODS

FIG. 4.40 : SLUDGE VOLUME REDUCTION RECORD OF ALL EXPT. PERIODS

geometry.

However, in order to use the SVI for design purposes, the length of the SETTLE phase should be used as the time-span of the test instead of the 30 min standard time. Moreover, results from settling velocity tests should also be consulted.

4.6 OTHER ANALYSES

4.6.1 DISSOLVED OXYGEN LEVEL

The mixed-liquor dissolved oxygen (D.O.) level of the reactors was monitored during track analysis #4. The results are plotted in fig.4.41. During this experiment, the D.O. probe was placed at the bottom of the reactor before aeration began. From fig. 4.41, it can be seen that The D.O. level increased immediately as aeration commenced due to mixing with the supernatant, which D.O. level was in general above 3 mg/l. The ML D.O. then decreased steadily for approximately 10 min when substrate consumption was most active. After this period, the D.O. level would increase steadily for approximately 30 minutes until the saturation level was reached.

The D.O. profile of the reactor at the end of SETTLE was also monitored. In general, the supernatant D.O. concentration was maintained above 3 mg/l and the sludge D.O. was always below 1 mg/l.

4.6.2 PH

The influent and mixed-liquor pH were monitored occasionally and its course of change during aeration was traced during track analysis #2. No significant difference in pH was noted between the

FIG. 4.41 : PLOT OF MIXED-LIQUOR D.O. LEVEL VS. AERATION TIME IN TRACK ANALYSIS #4 (OP.PER.II, 6h CYC., 1.5 1/CYC.)
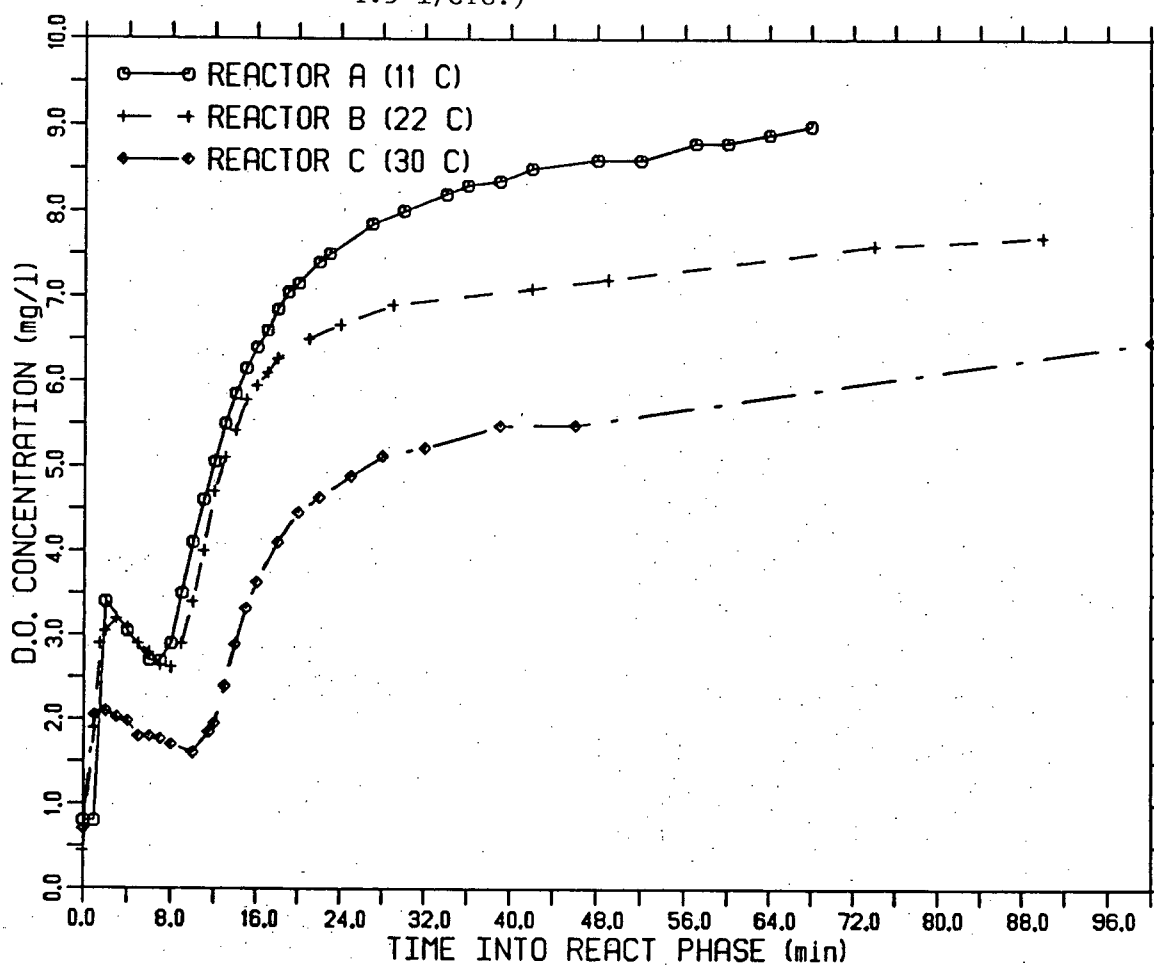
three reactors — the pH of the reactors stayed between 7.3 and 7.8 throughout all the experimental periods. The influent milking-waste pH, on the other hand, fluctuated between 7.3 and 8.2.

4.6.3 20-DAY BOD

Eight BOD_{20} tests were carried out during the experimental periods — two for the screened milking-centre wastewater and two for each of the SBR effluents. The mean BOD_5/BOD_{20} ratio were as follows : Influent = 0.81, reactor A effluent = 0.67, reactor B effluent = 0.63, reactor C effluent = 0.79.

4.6.4 FILTRATE BOD AND COD

Two filtrate BOD_5 and COD analyses were carried out. One on April 03 and one on May 28. The results are as follows :

TEST	INF.	FILT. INF.	EFF. BOD (mg/l) A	B	C	FILT. BOD (mg/l) A	B	C
4.03	---	---	23	6	29	4	4	0
5.28	223	149	29	22	14	13	10	14

TEST	INF.	FILT. INF.	EFF. COD (mg/l) A	B	C	FILT. COD (mg/l) A	B	C
4.03	---	---	360	252	284	254	198	168
5.28	836	643	213	129	102	167	116	68
5.28	836	643	213	129	102	167	116	68

The April 03 effluent TSS concentrations were 244, 112 and 240 mg/l for reactors A, B and C respectively; and the May 28 effluent TSS concentration were 19, 4 and 4 mg/l for reactors A, B and C respectively. While the effluent TSS level was expected to play a definite role in the parametric differentiation of "raw" and filtered

effluents, a correlation was not made because of the limited amount of filtrate data collected in this study. Throughout the experimental periods of this project, the focus was on supernatant parameters which was deemed more realistic then filtrate parameters in comparing the *overall* treatment efficiency of the reactors under different operation conditions.

4.7 OPERATION PARAMETERS IN SBR SYSTEMS

The foundamental difference between a continuous-flow system and a sequencing batch operation is that the former operates under steady state assumption and the latter does not. Consequently, a lot of the operation parameters and terminologies familiar to continuous-system designers cannot be applied indiscriminately to a sequencing batch system. New terminologies will also have to be invented in order to clearly describe a sequencing batch operation. The term "specific cycle" used in this study is one such parameter.

Other familiar terms such as sludge age (or mean cell residence time) and food to microorganism ratio have been adopted by some researchers; however, the subtle difference should be clearly noted. Since both substrate and microorganism concentrations are time dependent variables in a SBR, all parameters dependent on these variables are therefore also functions of time. Usually, the initial or extreme (maximum or minimum) values are used to provide some indication of the parameter interested.

Specific Cycle is defined in this study as the total number of cycles employed to treat the same amount of wastewater per day. In another words, it is the number of cycles employed per unit flow :

Specific Cycle   = # of cycles/ volume of wastewater treated/day

= # of cycles/ daily flow rate

By this definition, the specific cycle of operating periods II, V and VI was 0.67 d/l, of period III was 1.0 d/l and period IV was 1.33 d/l. The average hydraulic retention time, calculated as the filled reactor volume divided by the daily flow rate, was 0.833 days for operating periods II to IV. During operating periods V, the average hydraulic retention time for A and B remained 0.833 days while that of C was decreased to 0.750 days due to a 10 % reduction of the filled reactor volume. In operating period VI, the average hydraulic retention time were decreased to 0.278 days for A and B and 0.250 days for C. It should be noted that although the specific cycles for reactors A and B were the same in periods II and VI, their average hydraulic retention times were differed by a factor of three.

For the same average hydraulic retention time, the larger the value of specific cycle (that is, the more cycles employed per unit flow), the smaller would be the ratio of treatment/total volume per cycle, and the closer the kinetics of the system would resemble that of a continuous system. This is because semi-batch operation is in effect, over a long period of time, a continuous process with finite periods. It is therefore desirable to minimize the value of specific cycle (minimize the number of cycles per unit flow) in order to fully realize the kinetic advantage of a batch reactor.

In this study, the effect of changing the specific cycle was studied during operating periods II to IV. During these operating periods, the average hydraulic retention time was kept constant while the specific cycle was varied from 0.76 to 1.33 d/l. The purpose of investigating the effect of specific cycle is that in practical application of SBR systems to treating

periodic flows, it is useful to know how flexible the SBR is in accomodating changes in the incoming flow-pattern (while the total daily flow remains the same). For example, if there is an operational re-scheduling in a milking centre which effluent is treated by a SBR, and the number of milking cycle is increased from 2 to 3 (the daily flow remains the same), it would be useful to know whether the SBR operation can be adjusted proportionately from 2 cycles to 3 cycles without sacrificing any treatment efficiency.

The results of this experimental study indicated that under the operational conditions of this project, changing the specific cycle from 0.67 to 1.33 d/l did not have any significant effect on the BOD_5, COD, and NH_3-N removal efficiency of the reactors.

One possible reason for this lack of response to specific cycle variation maybe that the reactors were overdesigned. The possibility of overdesigning cannot be totally rejected as the reactors were sized semi-arbitrarily based on other researchers' experience. However, the fact that the treatment efficiency of reactor C was decreased by 2.4 % when the average retention time was reduced by only 10 % (see section 4.1.1) suggests that gross overdesigning was unlikely to be the case.

It is therefore suspected that the range of specific cycles studied was not large enough to induce noticeable difference in the reactor's treatment efficiencies. However, the range studied (4 to 8 cycles per day) was considered sufficiently large for practical systems.

Since almost all of the SBR researches done to-date utilized longer cycles of six hours or more, the longest cycle chosen for this study was kept to six hours. The lower limit of cycle length for the purpose of understanding the effect of specific cycles was three hours. In real

systems, a large number of short-length cycles would present operational problems such as inadequate sedimentation time and quicker equipment wear-out and energy wastage due to frequent switching of ON/OFF modes. Furthermore, as the number of cycles needed increases, the influent flow pattern is approaching continuous flow. In this case, either a multiple tank system or a continuous system should be considered.

To summarize, within the range of this study (4 to 8 cycles/day), varying the specific cycle has no effect on the treatment efficiencies of the reactors. This allows more flexibility in the operation strategy planning of the Sequencing Batch System.

5. SUMMARY AND CONCLUSIONS

Based on the literature review and the experimental work of this study, the major results, together with the conclusions drawn from them, are summarized as follows :

1. Very high and reliable BOD_5, COD, NH_3-N and Suspended Solids removal from milking parlour effluent can be achieved by using a Sequencing Batch Biological Reactor. At 21.8 and 29.8 °C, over 90% BOD_5 removal, 80% COD removal and 90% NH_3-N removal was attained by the bench-scale reactors used in this study.

2. There was no noticeable difference in the treatment efficiency of the SBR operating at 21.8 and 29.8°C. The treatment efficiency of the low temperature reactor (10.5 and 3.7°C) was lower; however, an average of 71 to 92 % BOD_5 and COD removal could still be achieved. Ammonia removal at these low temperatures was more fluctuating; the percentage removal of NH_3-N by the low temperature reactors ranged from 49 to 85%.

3. The experimental results showed that changing the specific cycle within the range 0.67 to 1.33 d/l (4 to 8 cycles per day) had no noticeable effect on the treatment efficiency of the Sequencing Batch Reactors.

4. Compare to the BOD_5 data, the COD data on treatment efficiency showed more sensitivity to the effect of low operating temperatures.

5. Instantaneous and uncontrolled denitrification occured at the beginning of aeration. Substrate introduction was required for this denitrification process; but the extent of denitrification was limited.

6. The experimental results showed that the nitrification and denitrification processes were most efficient in the 29.8°C reactor and least efficient in the low temperature reactor (10.5 and 3.7°C operations).

7. The treatment efficiency of the reactors did not seem to be directly affected by the sludge age of the biomass population. Both settling characteristics and nitrification power of the activated sludge were satisfactory within the sludge age studied (5.3 to 20 days).

8. BOD_5 and COD removal during the first 20 to 30 minutes of aeration in the SBR can be approximated as a pseudo first-order reaction with respect to substrate concentration. The reaction kinetics assumed a much lower rate constant after this initial period.

9. Aerobic conditions generally prevailed in the supernatant faction at the end of the SETTLE phase. In this study, the supernatant D.O. was generally above 3.0 mg/l at the end of sedimentation. Anaerobic conditions developed quickly at the bottom of the sludge mass after sedimentation begins; however, this transient anaerobic stage did not have any adverse effect on the settling characteristics of the biomass population.

10. The BOD_5 to BOD_{20} ratio was found to be 0.81 for the seived milking centr effluent and 0.63 to 0.79 for the SBR treated effluents.

11. The strength of the UBC milking centre varied greatly in terms of BOD_5, COD, NH_3-N and TSS. However, the percentage removal treatment efficiency of the reactors remained relatively constant.

6. RECOMMENDATIONS

Based on the experience gained from this study, some recommendations for the design and research on Sequencing Batch Reactors are made as follows :

1. The high efficiency of Sequencing Batch Reactors permit shorter hydraulic retention times and smaller reactor volumes as compared with continuous systems. However, experience gained from operating period II of this study showed that a sudden change in waste characteristics can upset the sludge mass. Presumably, this effect is more prominant in SBRs because of the quick change of the reactor environment as a result of short retention time. A pre-reactor equalization tank properly sized to take this factor into account is therefore strongly recommended for the Sequencing Batch Biological Systems.

2. The discrepancy between the unseeded BOD_5 data and COD data in temperature effect on treatment efficiency suggests further investigations with seeded BOD_5 test as an additional indicator.

3. A survey of the literature showed that presently available data on the flow rate and characteristics of milking-centre effluents show great variation in both quantity and quality. There appears to be a need for more information in this area to facilitate the design and management of both water and wastewater handling facilities in the milking-centres.

4. The denitrification that takes place at the beginning of aeration is suspected to be a result of the consumption of denitrifying enzymes accumulated in the sludge mass during SETTLE. However, further research works, probably involving some extensive biochemical experiments, will be needed before this denitrification process at the beginning of an aerobic phase can be better understood and utilized.

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5. The using of SBR to treat milking-centre wastewater has been very successful in bench-scale and pilot-scale operations. Even the worried problem of diffuser clogging did not occur. The applicability of SBR to treating other types of wastes should now be examined.

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