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Contract Number MDWPC 76-10(3)

ENVIRONMENTAL ENGINEERING PROGRAM
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ABSTRACT

Three activated sludge reactors of different sizes and designs, one in pilot-plant-scale of 7,000 gallon liquid volume, and two in bench-scales of 7.5 gallon and 1.2 gallon liquid volumes, respectively, were used to evaluate the scale-up effects on the process at outdoor temperatures. The reactors were operated under equivalent conditions by regulating the mean cell residence time, mean hydraulic detention time, and biomass concentrations in their reactors so that these operating factors were respectively of the same values. The reactors were installed outdoors, so that they were exposed to the same ambient temperatures ranging from 5°C to 13°C through the winter months, 1976-1977.

In Phase I of the study, the flow regimes prevailing in the reactors were not of the same degree of mixing. While Biochemical Oxygen Demand (BOD) and Chemical Oxygen Demand (COD) removals were practically the same in the two bench-scale reactors, there was a small but statistically significant difference in BOD and COD removals between the bench-scale and pilot-scale reactors. However, the difference could be attributed to other than scale-up effects. As for NH_3-N removal, nitrification and total-phosphorus removal, there were no differences among the reactors. The data were further analyzed by using current kinetic models, and again there were practically no differences in the performance of the three reactors. Therefore, a bench-scale reactor, as small as 1.2 gallon, could be applied to generate treatability information as reliable as a pilot-scale reactor having a liquid volume of 7,000 gallons.

In Phase II of the study, the results indicate that the modifications in operation intended to bring about the complete-mixing flow regime in all reactors did not alter the similarity in performance of the reactors as depicted in Phase I. This phase of the study also revealed the fact that there were neither biochemical reactions involving nitrification and total-phosphorus removal, nor significant carbonaceous bio-oxidation taking place in the final clarifier. Seemingly the assumption of no substrate removal by biochemical reactions in the final clarifiers, commonly made in developing the kinetic models for activated sludge processes, is reasonable.

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I. INTRODUCTION

Bench-scale evaluations of the activated sludge process have been widely used for treatability tests and other laboratory experimentation. They are usually conducted indoors in reactors with various liquid volumes ranging from 5 liters to 50 liters. These small reactors are easy to control, flexible and inexpensive to operate, and sensitive in response to varied operating conditions. However, it is often argued that the data generated with such bench-scale studies are not reliable for the purpose of designing full-scale facilities, and pilot plant evaluations must be used to confirm or refine the information.

In 1972, a U. S. Environmental Protection Agency (USEPA) report (1) *concluded that pilot plants of adequate size are generally acceptable* to provide data for engineering full-scale facilities. Such pilot plant studies involve aeration tanks up to several thousand gallons, which are usually installed outdoors, and subject to seasonal and diurnal variations in climate, and wastewater flow and strength. It is obvious that the pilot plants demonstrate operating conditions and results closer to the full-scale facilities. But they are much less flexible and of high cost in operation. Because sufficient design fees and ample study time are commonly not available in engineering practice, they are not as often used as the bench-scale studies. Therefore, if the scale-up differences between pilot plant and bench apparatus can be established, the usefulness of the bench-scale studies can be extensively enhanced in providing reliable data for the design of full-scale facilities.

II. OBJECTIVES

The general objective of this research is to compare the performance of laboratory scale and pilot plant scale activated sludge processes. It is hoped that the results of the study would provide reliable interpretation of the data generated by bench-scale tests for the purpose of designing full-scale facilities.

Specifically, the study is aimed to investigate the scale-up effects in terms of

BOD removal
COD removal
Nitrification
Phospohrus removal

In addition, current kinetic models for carbonaceous bio-oxidation and nitrification will be tested; these models will be used to study the scale-up effects by calculating and evaluating appropriate kinetic parameters. In this investigation the reactors are to be installed outdoors to study the effect of seasonal and diurnal variations of temperature and other climatic factors on reactor performance.

III. PREVIOUS INVESTIGATIONS

A rational approach for investigating the scale-up effects on the performance of various systems of the activated-sludge process is to operate the experimental reactors of different sizes ranging from bench-scales commonly used in laboratory studies to adequate pilot-plant-scales under similar performance-controlling conditions. To relate the process performance to microbial growth, kinetic models (2,3,4,5,6) have been developed, which mathematically describe the biological and biochemical interactions involved in the complete-mix activated sludge process.

A simplified model proposed by Sherrad and Schroeder (6) which relates the substrate utilization to an observed biomass production will be used for this study.

A. Sherrad and Schroeder Model for Carbonaceous Bio-oxidation (6)

For an activated sludge process with cellular recycle, as shown in Figure 1, the rate of food utilization in the system can be evaluated from a material balance on substrate; under the assumption of steady state and no substrate removal in the final clarifier, the following equation results,

$$\frac{\Delta F}{\Delta t} = \frac{Q(S_o - S)}{V} \qquad (1)$$

where $\frac{\Delta F}{\Delta t}$ = rate of substrate utilization on a finite basis,mass/volume-time.

Q = influent flowrate, volume/time

V = volume of aeration basin, volume.

S_o = influent substrate concentration, mass/volume

S = effluent substrate concentration, mass/volume.

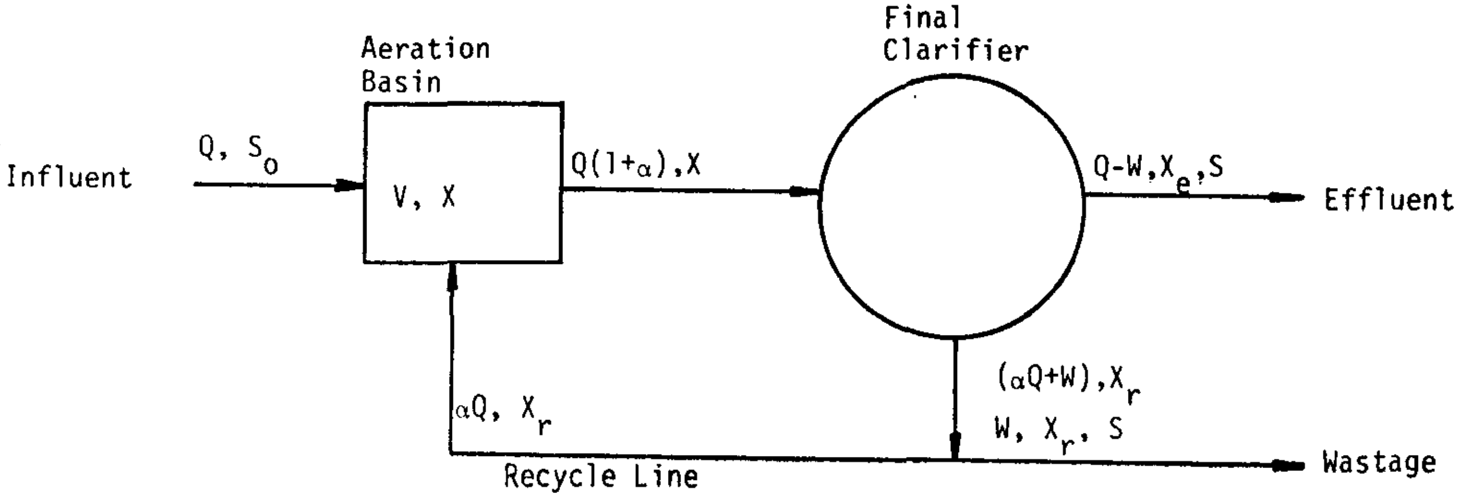


Figure 1. Schematic Diagram of a Complete-Mix Reactor with Cellular Recycle

The ratio of $\frac{\Delta F}{\Delta t}$ to X, known as F/M ratio and denoted by U, can be written as

$$U = \frac{\Delta F/\Delta t}{X} = \frac{Q(S_o-S)}{VX} = \frac{S_o-S}{\theta X} \tag{2}$$

where X = concentration of active microorganisms, mass/volume

U = F/M ratio, $\frac{\text{mass of substrate consumed}}{\text{mass of active cells-time}}$

$\theta = \frac{V}{Q}$ = mean hydraulic detention time

The expression for the specific growth rate, defined as the growth rate per unit mass of microorganisms, can also be obtained by striking a material balance of microorganisms in the system on the following assumptions (2,3):

- The concentration of dead organic matter (cell debris, dead cells) is negligible as compared with the concentration of live (active) cells.
- There is no significant growth of cells in the clarifier.
- Cell growth is exponential.

$$\mu = \frac{\Delta X/\Delta t}{X} = \frac{(Q-W)X_e + WX_r}{XV} \tag{3}$$

where μ = specific growth rate, 1/time

$\Delta X/\Delta t$ = net growth rate of microorganisms on a finite basis, mass/volume-time

W = wastage flowrate from recycle line, volume/time

X_e = microorganism concentration in the effluent, mass/volume

X_r = microorganisms concentration in the recycle line, mass/volume.

The reciprocal of μ, $X/\frac{\Delta X}{\Delta t}$, has often been referred to as the mean cell residence time, θ_c; therefore,

$$\theta_c = \frac{1}{\mu} \qquad (3a)$$

In a system with a properly operating settling unit, the quantity of cells in the effluent is very small, and Equation (3) can be simplified to yield

$$\mu = \frac{WX_r}{VX} \qquad (4)$$

If the cell wastage is drawn off from the aeration basin, Equation (4) becomes

$$\mu = \frac{W'X}{VX} = \frac{W'}{V} \qquad (5)$$

where W' = wasteage flowrate from aeration basin, volume/time

Sherrad and Schroeder combined Equation (2) and (3) and incorporated the concept of observed yield, as follows

$$\mu = \frac{1}{\theta_c} = Y_o U \qquad (6)$$

$$Y_o = \frac{(Q-W)\,X_e + WX_r}{Q(S_o-S)} \qquad (7)$$

where Y_o = observed yield, mass of cells/mass of substrate

The value of Y_o is a parameter that indicates how much biomass is produced per unit mass of substrate utilized in the treatment system and can be determined experimentally via Equation (7). It has the following properties:

- Y_o is an observed yield which results from many reactions in the system and accounts for factors such as oxidation-reduction of carbon source and nutrients, polymerization of substrate, growth rate, death rate, predation and cell maintenance reactions (8).

- Y_o is a function of the specific substrate utilization, and its highest value is equal to the true yield. This can be shown by using the well-known expressions (2,3):

$$\mu_N = \frac{(Q-W)X_e f + WX_r f}{VXf}$$

$$= \frac{[(Q-W)X_e + WX_r]f}{VXf}$$

$$= \frac{(Q-W)X_e + WX_r}{VX}$$

The last term of the above equation is the same as μ for the carbonaceous oxidation. So we can write

$$\mu_N = \mu \qquad (17)$$

By incorporating the mean cell residence time, and combining Equations (15) (16) and (17), the kinetic model for nitrification can be expressed as

$$\mu = Y_N U_N \qquad (18)$$

and

$$Y_N = \frac{(Q-W)X_e + WX_r}{Q(N_o-N)} \cdot f \qquad (19)$$

where Y_N = observed yield for nitrifiers, $\frac{\text{mass of nitrifiers produced}}{\text{mass of } NH_3\text{-N consumed}}$

Theoretically, Y_N can be computed from Equation (18) or (19) if the value of f is known. But no specific technique is available for direct determination of nitrifiers in a batch of activated sludge, and the value of f can only be estimated from the knowledge of theoretical yields of the nitrifiers (10).

Although Y_N is formulated similarly after the kinetic models for heterotrophic microorganisms, it has a different meaning from that of Y_o. Being autotrophic, nitrifiers use dissolved CO_2 as carbon source and small amounts of NH_3-N as nitrogen source for synthesis, and derive all the needed energy by converting NH_3-N to NO_3-N. In view of the different metabolism of heterotrophic microorganisms which utilize carbonaceous substrates for both energy and synthesis, Y_N can be less affected by the concentration of NH_3 than Y_o by the concentration of carbonaceous substrate. Furthermore, since the amount of NH_3 used for synthesis is insignificantly small, the reduction of NH_3 can be replaced by the formation of NO_3 in the calculations of the F/M ratio for nitrification and Y_N as follows:

$$U_N = \frac{\text{mass of } NO_3\text{-N formed}}{\text{mass of nitrifiers-time}}$$

$$Y_N = \frac{\text{mass of nitrifiers produced}}{\text{mass of } NO_3\text{-N formed}}$$

C. Operating Factors Affecting Process Efficiency

By the incorporation of Equations (2) and (3a), Equation (6) can be rearranged as

$$E = \frac{100\ X\ \theta}{\theta_c\ Y_o\ S_o} \tag{20}$$

where E is equal to $\frac{100\ (S_o - S)}{S_o}$, process efficiency in percent.

As shown in Equation (20), the process efficiency is a function of five variables. If the mean hydraulic detention time (θ) and the mean cell residence time (θ_c) are held constant for different sizes of aerators, and a same influent is fed to the aerators, that is, S_o is the same, the process efficiency becomes dependent only on two of the parameters, observed yield (Y_o), and biomass concentration (X). If biomass concentration is further maintained the same for all the reactors by varying the recycle flowrate of each unit, the efficiency will depend on Y_o only. The control of biomass concentration will be further discussed elsewhere.

Several operating factors affect the values of Y_o and biomass concentration, which include the following:

1) Dissolved oxygen concentration.
2) Recirculation ratio and settleability of biosolids.
3) Mixing characteristics.
4) Temperature.

1) Dissolved Oxygen (DO) Concentration

It has been established that the concentration of DO above 0.1 to 0.3 mg/ℓ has no effect on the metabolic rate of dispersed cells (9). In activated sludge systems, the BOD removal efficiency and nitrification rate have been found independent of DO concentrations as long as the DO level is above 0.5 mg/ℓ and 1 mg/ℓ respectively in the aeration tanks (5, 10).

Reports showed that extremely high DO levels tend to inhibit some microorganisms, especially nitrifiers, and to stimulate the growth of filamentous microorganisms (11, 12). But whether high DO levels are toxic or inhibitory otherwise has not been definitely established. Investigation by Okun and Lynn (11), concluded that DO levels from 2 to 25 mg/ℓ had no measurable effect on oxygen utilization rate in activated sludge processes. Thus a control of DO levels at 2 to 10 mg/ℓ in the aeration tanks should have little effect on the process performance.

2) Recirculation Ratio and Settleability of Biosolids

A material balance on biomass for the aerator only reveals the relationship between recirculation ratio and biomass concentration in the reactor. This material balance under steady-state is

$$\alpha Q\, X_r - Q(1+\alpha)X + \frac{dX}{dt}V = 0 \tag{21}$$

where $\frac{dX}{dt}$ = net growth rate of microorganisms, mass/volume-time

α = R/Q = recirculation ratio

R = recycle flowrate, volume/time

Q = influent flowrate, volume/time

Considering the equation above on a finite basis and incorporating the mean cell residence time, θ_c, and the mean hydraulic detention time, θ, it can be rearranged as

$$X = \frac{\alpha X_r\, \theta_c}{(1+\alpha)\theta_c - \theta} \tag{22}$$

X_r is the concentration of biomass in the recycle line and is a function of the settleability of biosolids in the final clarifier, thus it cannot be controlled experimentally. But X_r can be estimated by measuring the mixed liquor suspended solids and the sludge volume index (SVI). Thus, the concentration of biomass in the aerator, X, can be maintained at a steady level by operating at the same θ_c and θ, but varying recirculation ratios according to the value of X_r so that the product of α and X_r remains constant. In such an operation, the biomass concentration in the aerator can be kept steady, and the effect of recirculation ratio and settleability of biomass on the process performance is minimized and the process efficiency depends on Y_o only.

3) Mixing Characteristics

The kinetic models described previously are based on the ideal complete-mix flow regime. But in reality, the flow regime in a mixed aeration tank varies with the size and configuration of the tank, design of the inlet and outlet structures, and arrangement of the aeration systems. In most aeration tanks, the degree of mixing is somewhere between complete-mix and plug-flow. A tracer method can be used to evaluate the flow regime in aeration tanks. After instantaneous release of a tracer into the influent stream, the concentration of tracer in the effluent stream of the aerator is measured with respect to time. If the peak tracer concentration appears instantaneously in the effluent stream, a complete-mix condition prevails. If the time for the appearance of peak tracer concentration in the effluent equals to the theoretical detention time, an ideal plug-flow regime is assumed (13).

As to the effect of flow regime on the performance of an aeration tank, based on the current kinetic models (2,14), it could be stated that the plug-flow mode is theoretically more efficient in wastewater treatment than the complete-mix system. However, no observable differences in the Y_o value and in treatment performance were obtained in both laboratory (14) and field (1) experiments.

4) Temperature

The effect of temperature on the activated sludge process can be found in the literature (15,16,17,18,19,20,21). Temperature affects the true yield (Y), respiration coefficient of microorganisms (K_d), the dominant microorganism group, and the availability of nutrients (18,21).

However, the temperature response may vary, as the microorganisms are being acclimated (19). Temperature may also affect activated sludge settling rates, although reports on the effects are contradictory (18,19).

Sawyer (15) conducted a series of laboratory experiments and showed that equivalent BOD removals were obtained under conditions of temperature higher than 10°C and mixed liquor suspended solids above 1000 mg/ℓ. Ludzack _et al._ (16) found that BOD and COD removals were about 10 percent higher at 30°C than at 5°C at high solid loading rates. Hunter (17) and Friedman (18) showed little change in BOD removal as temperature increased from 4°C to 45°C. Laboratory experiments conducted by Benedict (19) under long-term temperature acclimation resulted in little difference in COD removal efficiency at temperatures of 4°C and 19°C, but showed that the removal efficiency decreased at a temperature of 32°C.

Also Keefer (20) reported that the BOD removal in a plant-scale operation increased from 84.5 percent at a sewage temperature of 11°C to 91.5 percent at a temperature of 24°C when the sewage flow varied between 18 and 20 MGD, and the BOD removal increased from 89.5 percent at 12°C to 91.5 percent at 23°C when flow varied between 12 and 14 MGD. Novak (21) generalized that the temperature response of biological processes was found to depend on the substrate concentration in the system. Aerobic processes were found to be nearly independent of temperature variations at low substrate levels (21). It may be surmised that no significant changes in BOD removal should be expected in an activated sludge system operated at relatively long mean cell residence times because of seasonal or diurnal variations of temperature.

Benedict and Carlson (19) found that the true yield reached a maximum at 20°C and reduced when temperature either increased or decreased, which was confirmed by Friedman (18).

Nitrification, unlike BOD removals, is markedly affected by temperature changes (10, 26). Maximum nitrification occurs at about 30°C. As the temperature decreases, the nitrification drops rapidly. The same trend has been observed in the growth rate of nitrifiers (26).

IV. METHODS OF APPROACH

PHASE I

A. Equipment

Three activated sludge reactors of different sizes and shapes were used for this phase of study, which are:

(a) A DAVCO aeration unit with 7000 gallons of aeration basin volume. It has been used for studies related to the treatment of Amherst wastewater in the past several years.

(b) A 7.5 gallon plexiglass continuous-flow reactor. It has been used for studying shock-load effects on activated sludge processes.

(c) A 1.2 gallon glass continuous-flow reactor designed by Ludzack (22). It has been widely used for laboratory treatability studies.

The general features of the three reactors are given in Table I and Figures 2, 3, and 4.

The two smaller units were installed outdoors near the DAVCO unit and were sheltered in a wooden box which can be closed to keep them from rain and snow during the winter. The inside of the wooden box was heated with two 60-watt electric light bulbs. The heating was controlled manually by switching on and off the light bulbs so that the temperature difference of the bench-scale units and the pilot plant-scale unit were within $\pm 1^{\circ}$C, and the connecting tubings to and from the bench-scale units were kept free from being frozen.

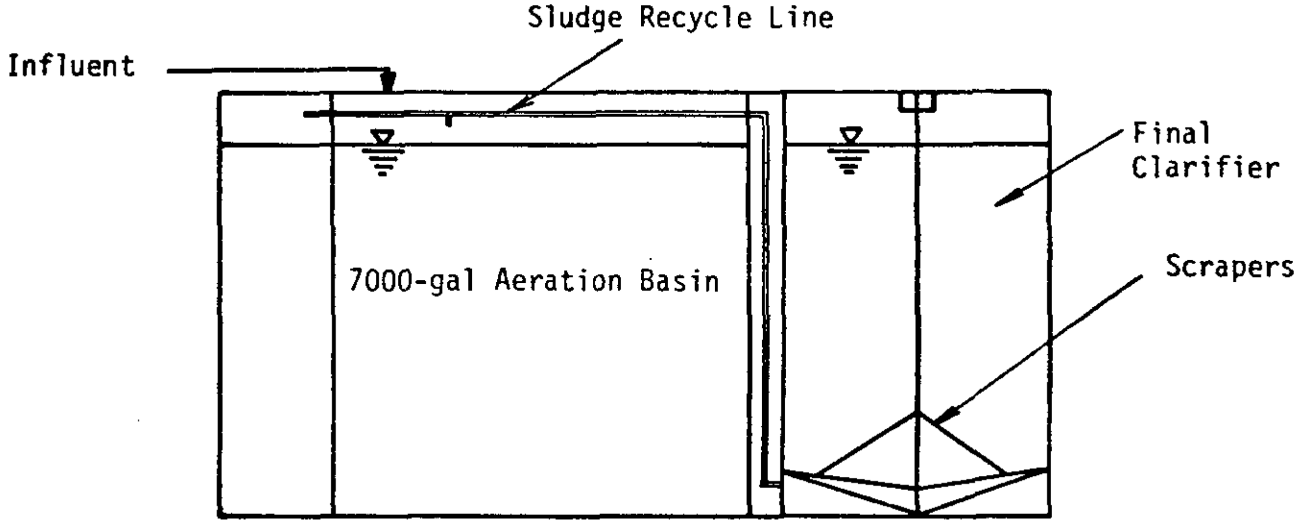
All the piping outside the wooden box was well insulated with asbestos and plastic tapes. The whole system was in operation without interruption throughout the winter months.

TABLE 1

General Features of Experimental Reactors

Reactor Number	Unit Classification	Reactor Shape	Reactor Volume gal (ℓ)	Final Clarifier Volume gal (ℓ)
A	Pilot-plant Scale	Rectangular	7,000 (26,490)	1,748 (6,616)
B	Bench-Scale	Rectangular	7.5 (28.2)	3.85 (14.6)
C	Bench-Scale	Circular	1.2 (4.6)	0.42 (1.6)

(a) Side View



(b) Top View

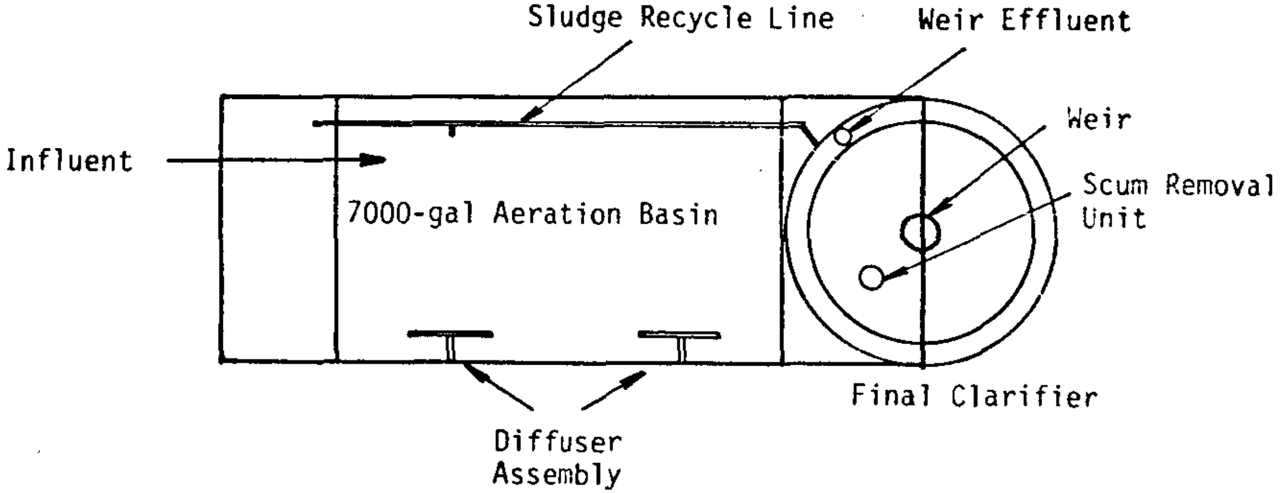
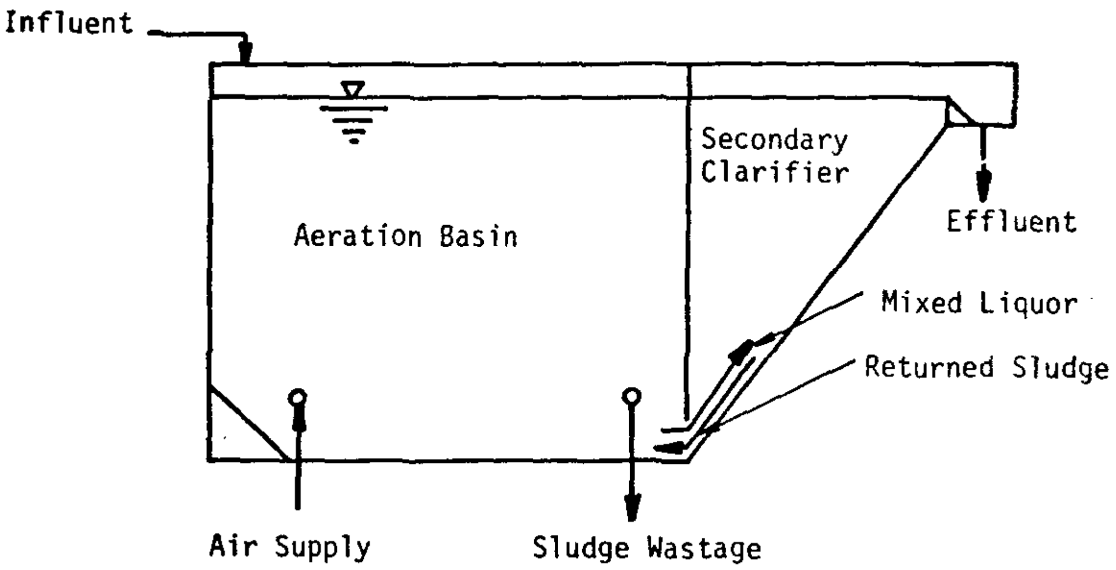


Figure 2. 7000-Gallon DAVCO Unit.

(a) Side View



(b) Top View

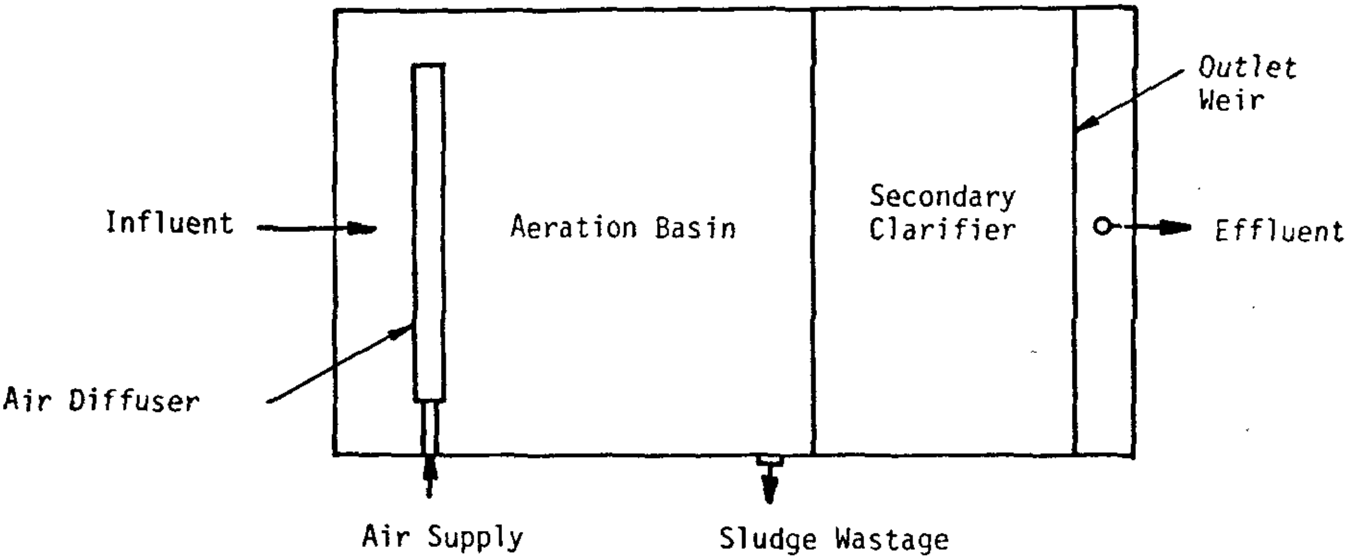
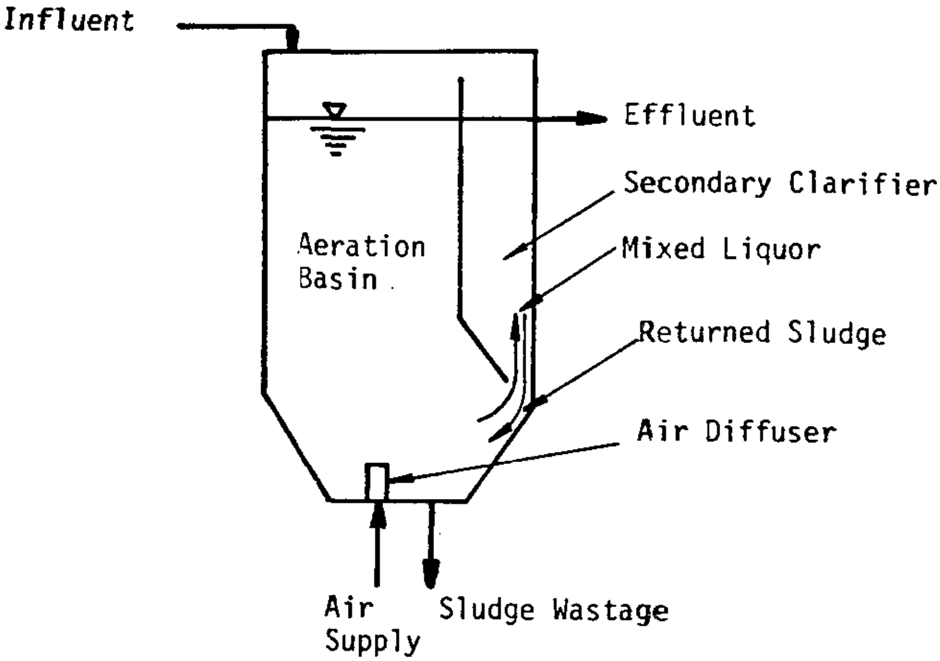


Figure 3. 7.5-Gallon Activated Sludge Unit.

(a) Side View



(b) Top View

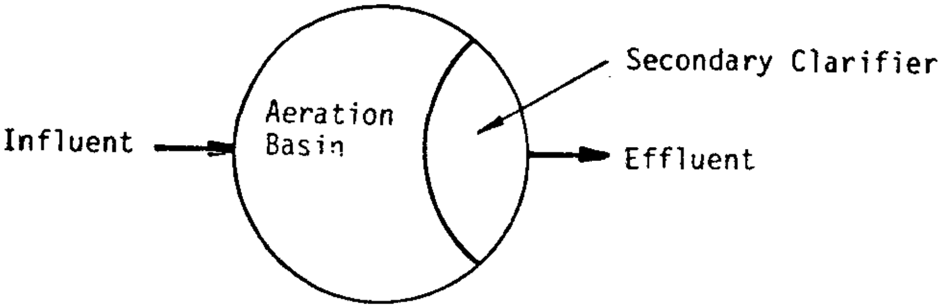


Figure 4. 1.2-Gallon Activated Sludge Unit.

B. Procedures

Raw wastewater from the Amherst Wastewater Treatment Plant was use as the feed substrate, which was first pumped to an elevated holding tank and screened by a hydroseive. The screened wastewater flowed by gravity to the DAVCO unit, and was pumped by Masterflex pumps to the other reactors. The schematic diagram of the whole system is shown in Figure 5.

The activated sludge units were controlled by mean cell residence time (4), a practical method that was found simple and required a minimum of laboratory work. A constant mean cell residence time was maintained by wasting a constant percent of sludge from the system each day. Wasting was completed once a day from the recycle line for the DAVCO unit, but from the aeration basins for the other two reactors. Wastage volumes were precalculated for each of the three reactors so that a mean cell residence time of 20 days was achieved.

Constant flowrates for the influents to the three units were maintained to keep a mean hydraulic detention time of 12 hours. The flowrate was checked regularly once a week. During the check-up, the solid deposits in the tubing and piping were removed and the flowrates were adjusted, if needed.

The DAVCO unit was aerated with a special diffuser system while the bench-scale reactors were operated with common air diffusing stones. The concentrations of DO were maintained above 2 mg/ℓ for all three units. Usually the DO in the aeration tanks was higher at night than in the day time due to the lower temperatures prevailing in the former case.

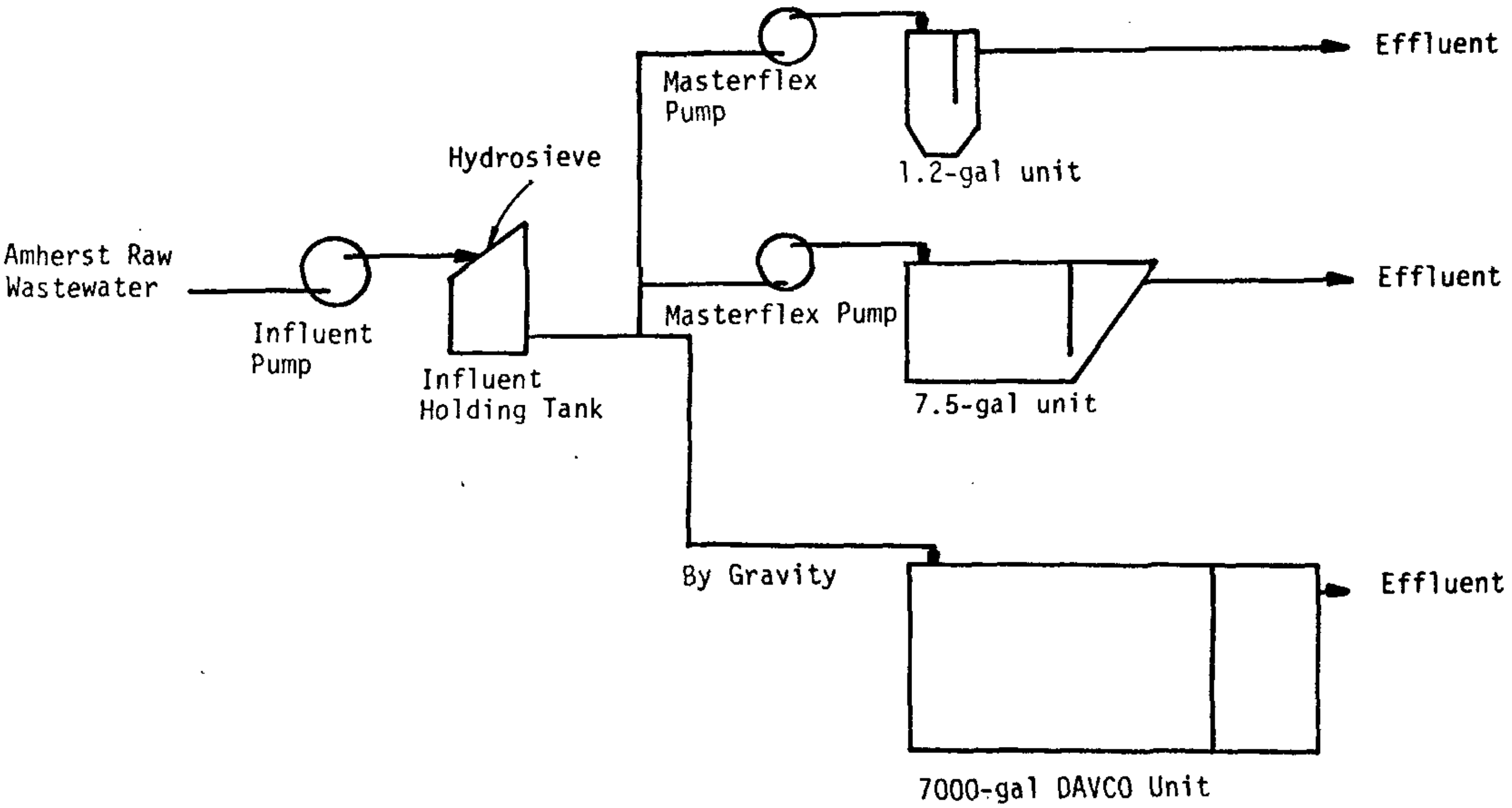


Figure 5. Schematic Diagram of the Experimental System.

Typical diurnal DO response curves of the three reactors are shown in Figure 6. Thus, an adequate DO concentration in the day time would ensure enough DO through the night.

The diurnal temperature variation of mixed-liquors was monitored and typical curves are shown in Figure 6. For that particular day the range of variation was from 5^0C to 8^0C.

The alkalinity of the Amherst wastewater is so low that it is not enough for converting all ammonia to nitrate. Thus, the alkalinity was supplemented.

The alkalinity-supplementing chemical was to be selected between lime and sodium bicarbonate. In separate experiments sufficient amounts of lime and sodium bicarbonate were added to the reactors near their influent ends so that the concentrations of CaO and $NaHCO_3$ were 25 mg/ℓ and 50 mg/ℓ, respectively. The pH and alkalinity were then measured against time. The data are presented in Figure 7. As shown, the addition of lime resulted in high, but quite different pH's in the units immediately after the addition; the smaller the unit, the higher the pH. Such a large initial pH discrepancy among the units should be avoided in order to minimize all the side effects which would influence the process. On the other hand, the addition of $NaHCO_3$ did not produce a wide difference in the initial pH among the units. In addition, being safe and easy to handle, sodium bicarbonate has been reported as a preferred source of supplemental alkalinity (24). As a result of the tests, sodium bicarbonate was chosen as the alkalinity supplement for the study.

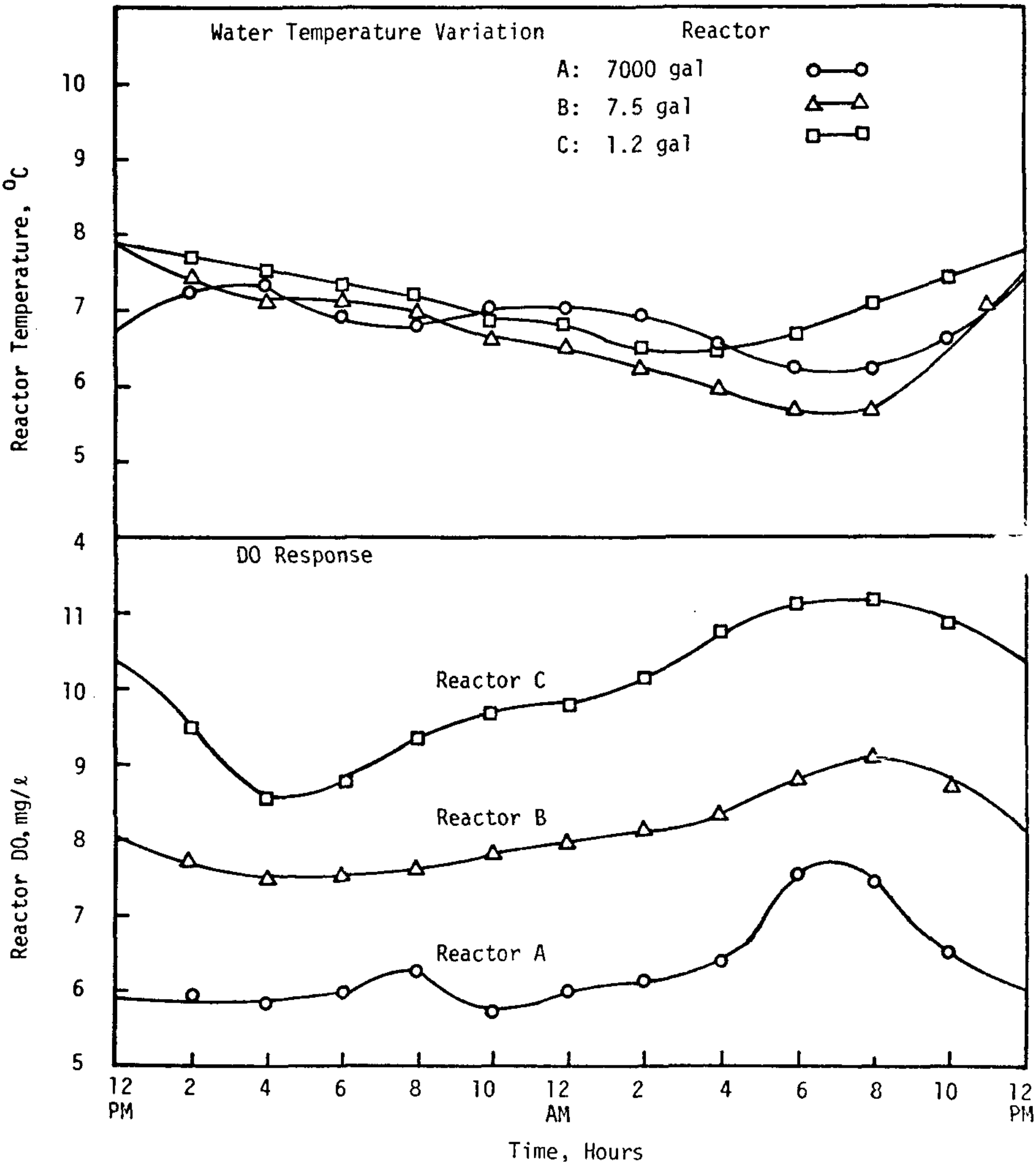


Figure 6. Diurnal Water Temperature Variations and Corresponding DO Response in Reactors.

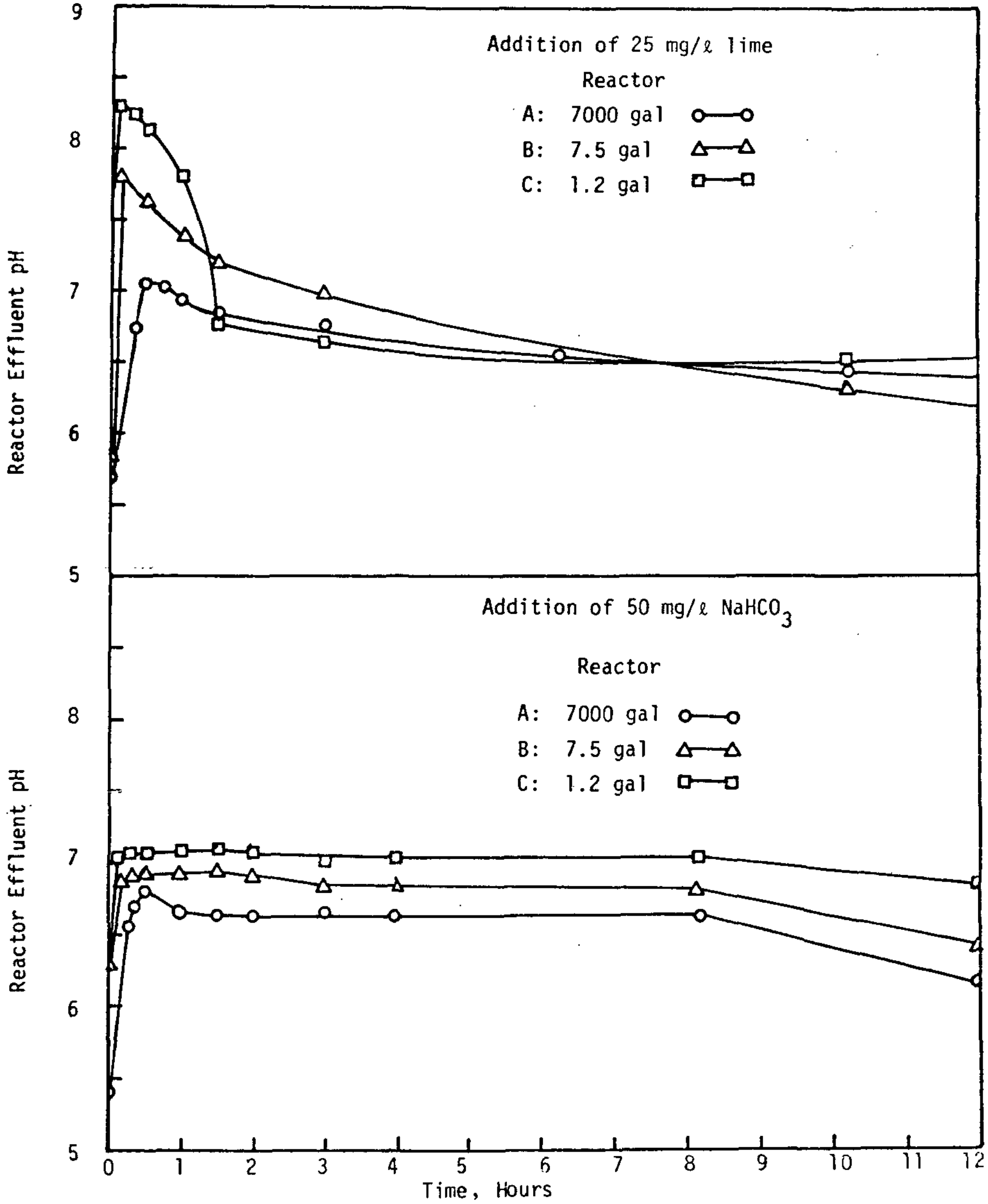


Figure 7. pH Response of Lime Addition and Sodium Bicarbonate Addition in Reactors.

As described previously, the control of biomass concentration can be accomplished by measuring the biomass concentration in the recycle line and varying the recirculation ratio according to Equation (22), until a desired level is reached in the reactor. Because there was no way to control the recycle flowrates of both bench-scale reactors, the recycle flowrate of the DAVCO unit was adjusted so that its biomass concentration was matched as closely as possible to the biomass concentration of bench-scale reactors. The needed recirculation ratios of the DAVCO unit were found to be between 1.5 to 1.8 which were applied throughout the experimental period.

In Phase I studies, no specific provisions were applied to induce an equivalent degree of complete-mixing in all three reactors. Without specific tests, the prevailing flow regimes of the reactors can be compared by viewing the pH response after the addition of lime as shown in Figure 7. It seems that the smaller the unit, the closer the flow regime to the complete-mixing mode, as indicated by the fact that the smaller the unit, the shorter the elapsing time before the peak pH appears at the effluent end.

Phase I studies started in the middle of December 1976 and ended at the end of February 1977. Winter temperatures prevailed, which were closely monitored to determine the effect of lower temperatures on the process.

C. Analyses

Samples were taken simultaneously from the units to determine the process performance measurements and the reactor operating factors which are listed below.

Performance Measurements*	Reactor Operating Factors
BOD_5^{20}	MLSS
COD	MLVSS
NH_3-N	DO
NO_3-N	Temperature
Total-P	SVI
Alkalinity	
pH	

*Unfiltered samples were used.

All the measurements of process performance were conducted in accordance with Standard Methods (24), except the analyses of NH_3-N and NO_3-N. NH_3-N and NO_3-N were measured by the Orion Ion Meter, Model 407A, using respectively NH_3-N and NO_3-N sensing probes. Calibration of the probes was performed prior to each analysis. The NH_3-N determination was performed to monitor the progress of nitrification.

In the measurement of BOD, the azide modification of Winkler's Method was employed. Total phosphorus was measured colorimetrically, after persulfate digestion, by the standard ascorbic acid method (24), using a Bausch and Lomb Spec-20 for measurement of light absorbance.

Samples of mixed liquor were taken from the aeration tanks. The Gelman Type A-E glassfiber filters were used to determine the concentration of mixed liquor suspended solids (MLSS) and mixed liquor volatile suspended solids (MLVSS). Average concentrations of MLSS and MLVSS of the DAVCO unit were obtained by analyzing samples taken at various spots along the aeration tank.

Temperature and DO concentration were measured twice a day at 10 AM and 2 PM, respectively, with a precalibrated YSI model 54A DO meter. The average of these two measurements was recorded as the representative daily temperature and DO.

Sludge Volume Index of the DAVCO unit was determined daily to compute the wastage sludge volume. Occasionally, SVI of the bench-scale reactors were determined to check if any bulking conditions had occurred.

PHASE II

A. Equipment

In order to study the scale-up effect on the activated sludge process under a flow regime as close to the completely-mixing mode as possible, the activated sludge units used in Phase I were modified either in operation only, or in both operation and reactor-structure as follows:

1. Bench-scale units. The air flowrate of the 7.5 gallon unit was increased so that its flow regime became as completely mixed as the smaller 1.2 gallon unit.
2. DAVCO unit. A new inlet structure was installed to distribute the influent along the length of the aeration tank, and the recycled sludge was returned to the aeration tank at its one-third and two-third points down stream from the inlet end. Such an arrangement enhanced the degree of mixing in the rectangular DAVCO unit.

B. Procedures

Because the raw wastewater feed pump was broken, it was decided to use, instead, the primary effluent from the Amherst Wastewater Treatment

Plant as the feed to the experimental reactors. A period of about two weeks was allowed for the acclimation of the activated sludge to the change of feed substrate.

The test of pH response to lime addition conducted in Phase I was again used to evaluate the mixing condition of the modified units. As shown in Figure 8, in all three reactors, their respective pH's appeared almost immediately after the lime addition and therefore the condition of complete-mixing prevailed.

The procedures, described in Phase I study, to maintain all three reactors under similar operating conditions, with respect to mean cell residence time, mean hydraulic detention time, and concentration of mixed liquor suspended solids, were used in this phase of the study.

This phase of operation spanned from March to April 1977. In addition to investigation of the scale-up effect, the role of the settling tank in the over-all performance of an activated sludge system was also evaluated.

C. Analyses

Samples were taken from the aeration tanks and collected after final clarification. As presented previously, the kinetic models for both carbonaceous bio-oxidation and nitrification were developed on the assumption that the biochemical reactions occurred predominantly in the aerator. Therefore, θ_c, U and U_N were computed on the basis of the aeration tank volume. However, it is desirable to clarify the question whether there is significant biochemical reactions taking place in the secondary clarifier, which was investigated in this phase of the study. Thus, samples of mixed liquor and effluent from the settling compartment of each experimental system were taken for analyses.

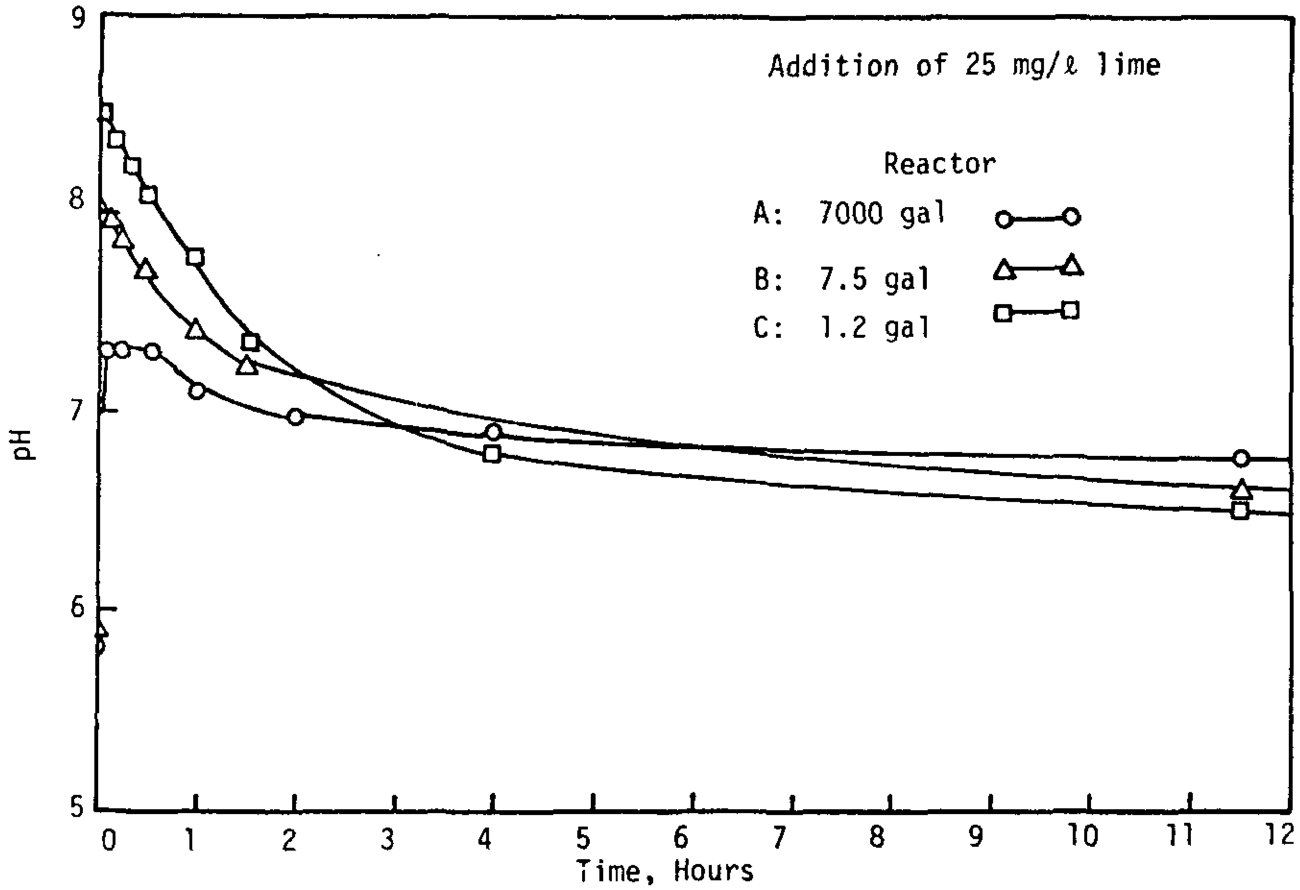


Figure 8. The pH Response Curves of Lime Addition for the Three Reactors.

All samples were filtered through Gelman Type A-E glassfiber filters before chemical analysis, in order to exclude the interference of the biosolids carryover. Therefore the results would indicate whether the biochemical reactions took place in the secondary clarifiers.

The following table shows the analyses performed:

Performance Measurements	Reactor Operating Factors
TOC	MLSS
NH_3-N	MLVSS
NO_3-N	DO
	Temperature
	SVI

Total organic carbon (TOC) was measured with a Beckman Model 915 Total Organic Carbon Analyzer. The procedures for the other measurements were described in the Phase I study.

V. RESULTS AND DISCUSSIONS

PHASE I

This phase of the study started in December, 1976 and ended in February of 1977, which covered the months of severe cold. The data obtained, therefore, depict the performance of the experimental activated sludge system at possibly the lowest ambient temperatures prevailing in this Northeastern region.

The average characteristics of the hydrosieved Amherst raw wastewater during the experimental period are shown in Table II. Similar results were obtained by Martel and others (27).

TABLE II

Characteristics of Amherst Raw Wastewater after Hydrosieving*

Parameter	BOD mg/ℓ	COD mg/ℓ	NH_3-N mg/ℓ	Total-P mg/ℓ	Alkalinity mg/ℓ as $CaCO_3$	pH
Mean	217	401	25	7.3	127	7.6
Standard Deviation	45	127	4.6	1.0	18	0.3

*Average of 22 samples

According to the following chemical reactions involved in nitrification,

$$NH_4HCO_3 + 2O_2 \rightarrow H^+ + NO_3^- + CO_2 + 2H_2O$$

$$H^+ + HCO_3^- \rightarrow CO_2 + H_2O$$

7.14 mg/ℓ of alkalinity as $CaCO_3$ are needed for the conversion of 1 mg/ℓ of NH_3-N to NO_3-N. Therefore, for the 25 mg/ℓ of NH_3-N present, at least 178.5 mg/ℓ of alkalinity as $CaCO_3$ were needed for the nitrification reaction. The required supplementation of alkalinity was provided by adding sodium bicarbonate to the influent end of the aerators.

The results are discussed in the following three parts:

- Operating Conditions
- Process Performance
- Summary

A. Operating Conditions

The methods of controlling the operating conditions, temperature, DO, MLSS, MLVSS, and others, were described previously in the section on Methods of Approach. During the experimentation, the operating conditions were monitored. The following are the results:

1. Temperature

Usually, the larger the water volume, the less its temperature varies as the air temperature fluctuates. Because the bench-scale reactors were housed in a wooden box heated by light bulbs, their temperature variations were somehow mitigated. As a result, although the temperature varied from 5°C to 13°C in the three winter months, the differences among the reactors were within ± 1°C, which are shown in Figure 9. The water temperature in all reactors changed gradually at a maximum rate of about 1°C per day. It may be assumed that the differences in temperature among reactors were so small that it would not have significant effect on the process performance.

Temperatures in the final clarifiers were almost the same as those of their respective reactors. The effect of temperature on the settling of activated sludge was not included in this study.

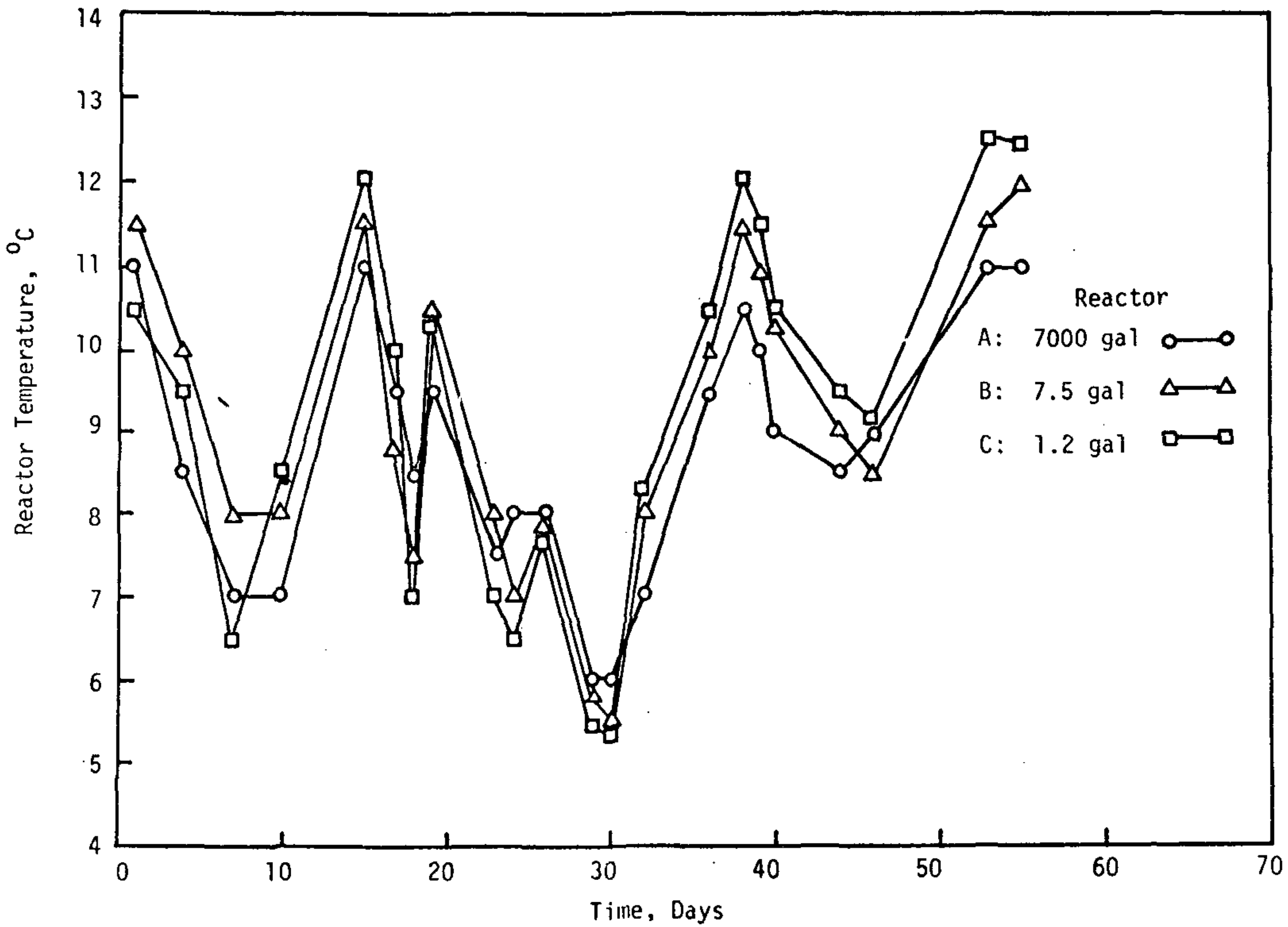


Figure 9. Daily Reactor Temperature

2. DO

As stated earlier, it has been known that the biochemical oxidation is independent of DO concentration as long as DO is above 2 mg/ℓ. Figure 10 shows the DO concentrations in the experimental reactors. They were all maintained above 2 mg/ℓ except during a breakdown of one of the air compressors, the DO of the DAVCO unit dropped below 0.5 mg/ℓ for a period of less than two days. The effect of low DO on nitrification will be discussed elsewhere.

Also shown in Figure 10, the DO concentrations were not the same in the three units, probably due to the fact that the aeration rate applied to the smaller units was higher, the smaller the unit, the higher the DO level. The average DO concentrations were 4.2 mg/ℓ for the DAVCO unit, 7.0 for the larger bench-scale unit, and 9.4 for the smaller one.

3. MLSS and MLVSS

The biomass concentrations in the three reactors were controlled at the same level. The difficulties lie in the fact that the MLSS measurement is rather time-consuming since it takes several hours. Therefore, during this study the adjustment of recycle flowrate, if needed, was often done on the next day. But still the results seemed satisfactory, as shown in Figure 11. Although there were large differences among the reactors in some periods, the average values of MLSS were as follows:

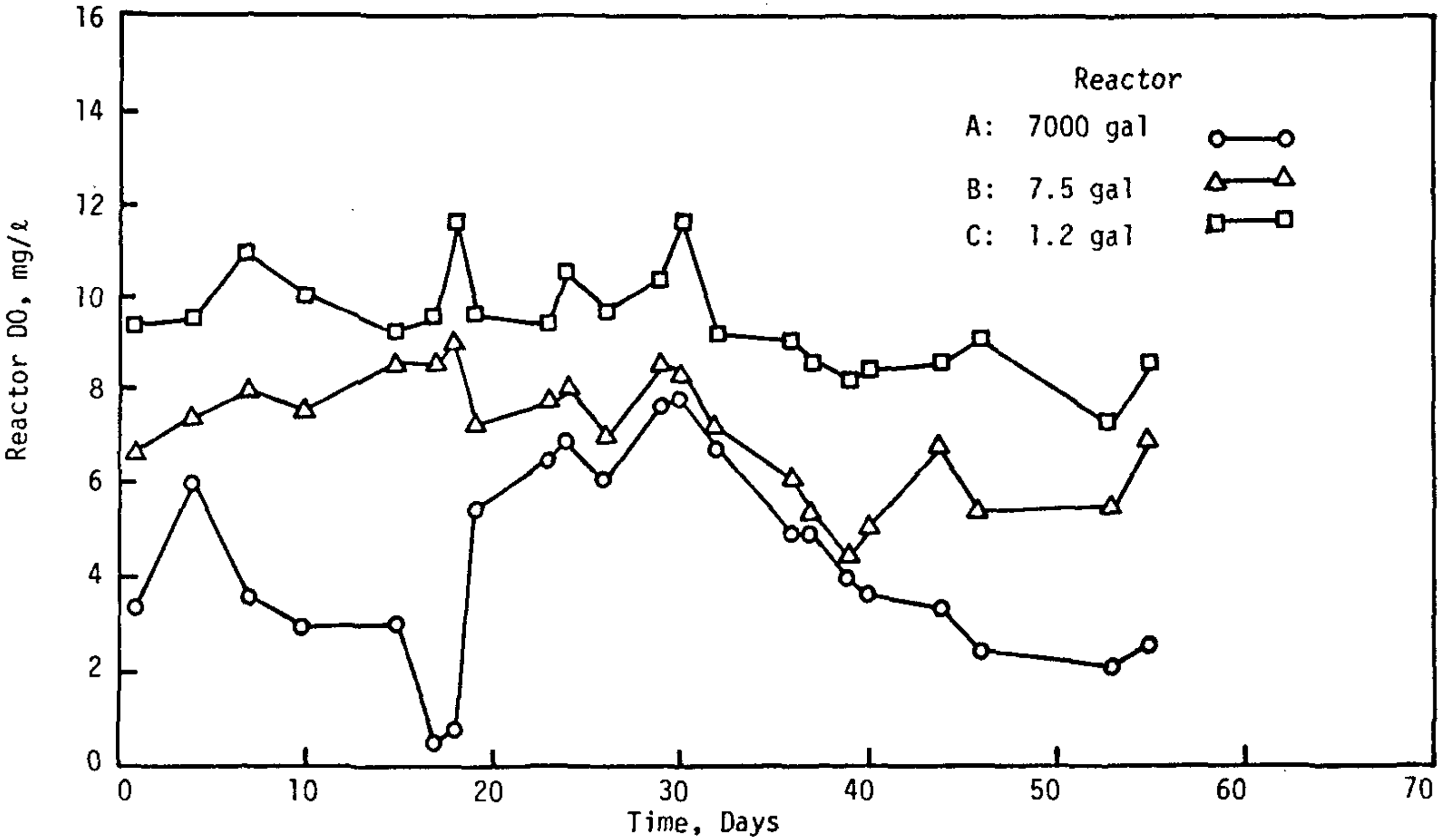


Figure 10. Daily Reactor DO

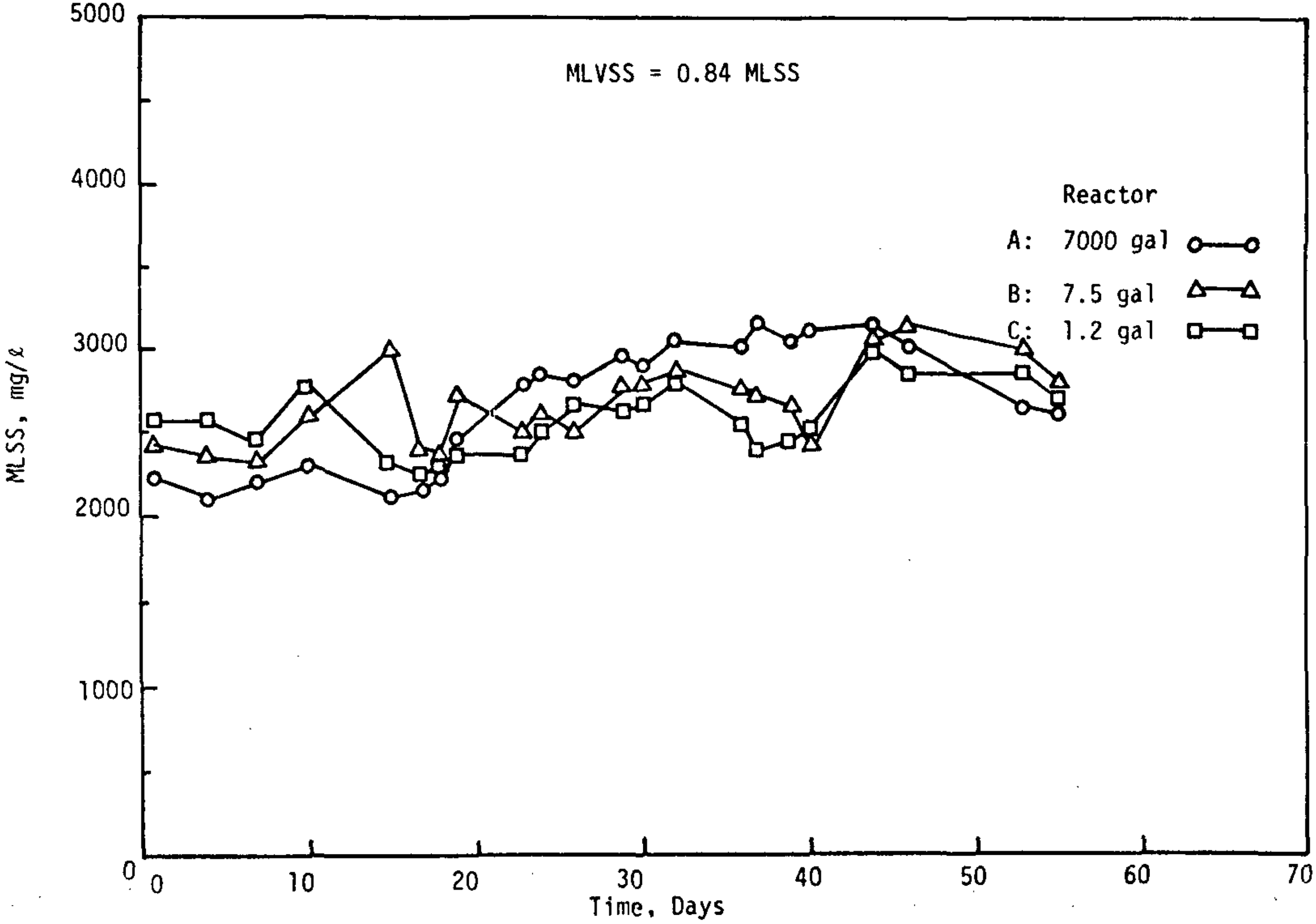


Figure 11. Daily MLSS Concentrations.

Reactor	Average MLSS mg/ℓ	Standard Deviation
A (7000 gal)	2690	391
B (7.5 gal)	2675	245
C (1.2 gal)	2570	212

The MLVSS was found to be about 84 percent of the MLSS for all the reactors.

4. Others

Sludge volume index is an indicator for biosolids settleability and can be used to estimate the biosolids concentration in the recycle line. During the study SVI was measured for calculating wastage rate of sludge in order to control the MLSS concentration. Fortunately, throughout the experimental period, no bulking conditions occurred.

B. Process Performance

In this section, the results of process performance such as BOD and COD removals, NH_3-N removal and nitrification, total phosphorus removal, and responses of pH and alkalinity are discussed. Kinetic coefficients are calculated on the basis of biochemical reactions involving BOD, COD, NH_3-N and NO_3-N. Temperature effects on the process are also presented and discussed under each reaction.

1. BOD Removal

Figure 12 shows the BOD concentrations of influent to and effluent from the three reactors. It can be seen that the pilot plant-scale reactor (Reactor A) seemed consistently to have an effluent of higher concentrations in BOD, and therefore, had lower removal efficiencies.

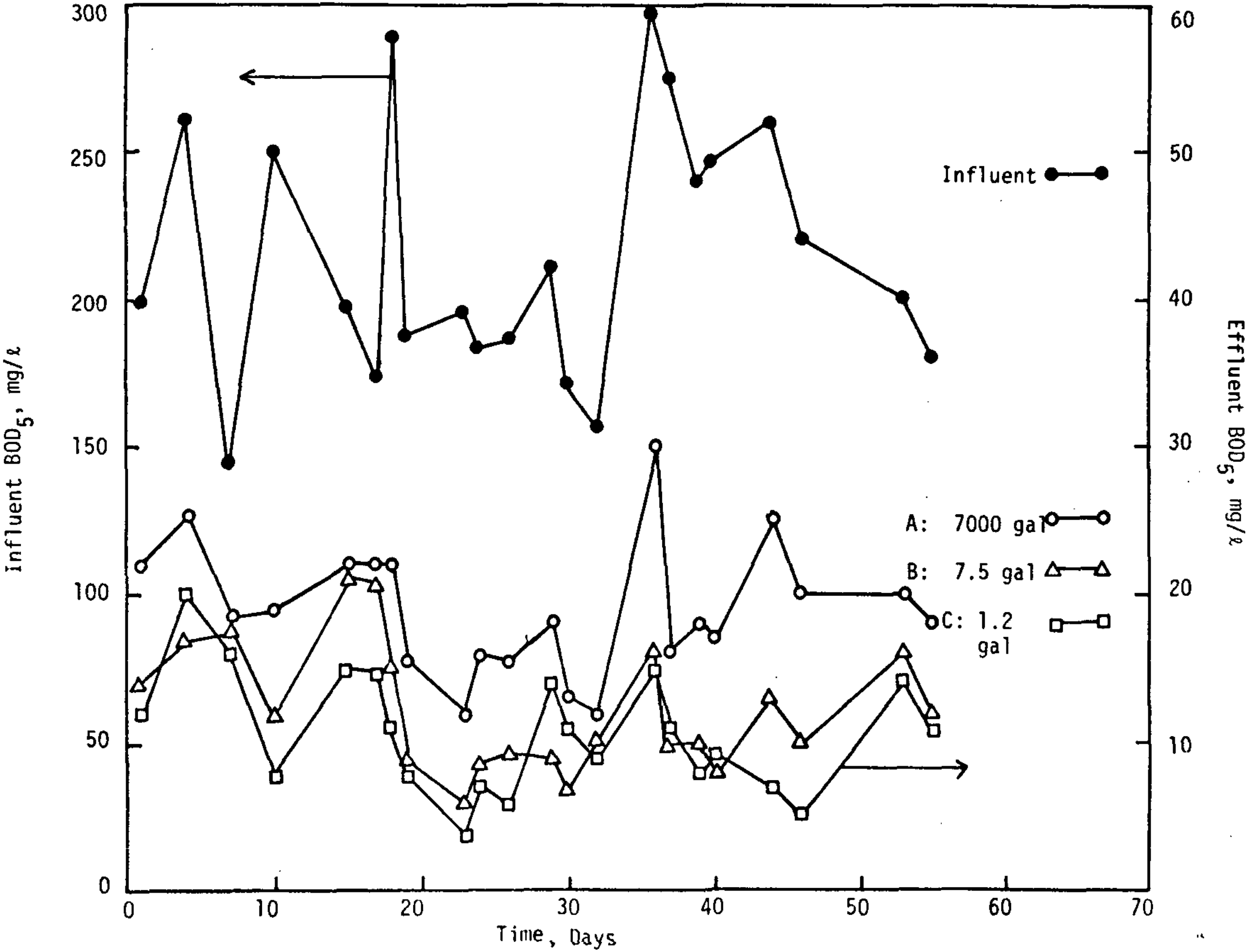


Figure 12. Influent and Effluent BOD Concentration.

The average effluent concentrations, removal efficiencies, and their respective standard deviations of the three reactors are listed in Table III.

TABLE III

Average Effluent Concentrations and Corresponding Removal Efficiencies*

Reactor	Effluent BOD		BOD Removal	
	Concentration, mg/ℓ	S.D.	Efficiency, %	S.D.
A	19	4.5	91	1.7
B	12	4.4	94	2.7
C	11	4.0	95	2.4

*Average of 22 runs

A statistical analysis using the Student t-test (See Appendix II) shows that there was a significant difference in BOD removal efficiency between Reactor A and the other two reactors at the 5 percent significance level, but no difference between Reactor B and Reactor C. That is to say that the bench-scale reactors yielded better BOD removals than the pilot-scale reactor. However, in a practical sense, this difference was small in terms of total BOD removal.

The results are further scrutinized in the respect of possible temperature effect. A plot of BOD removal vs temperature for each reactor is shown in Figure 13. The straight horizontal lines in the figure are graphically the best fit curves for the experimental data points, which indicates that the removal of BOD remained the same in each unit regardless of the variation of temperature ranging from 5°C to 13° C. It apparently

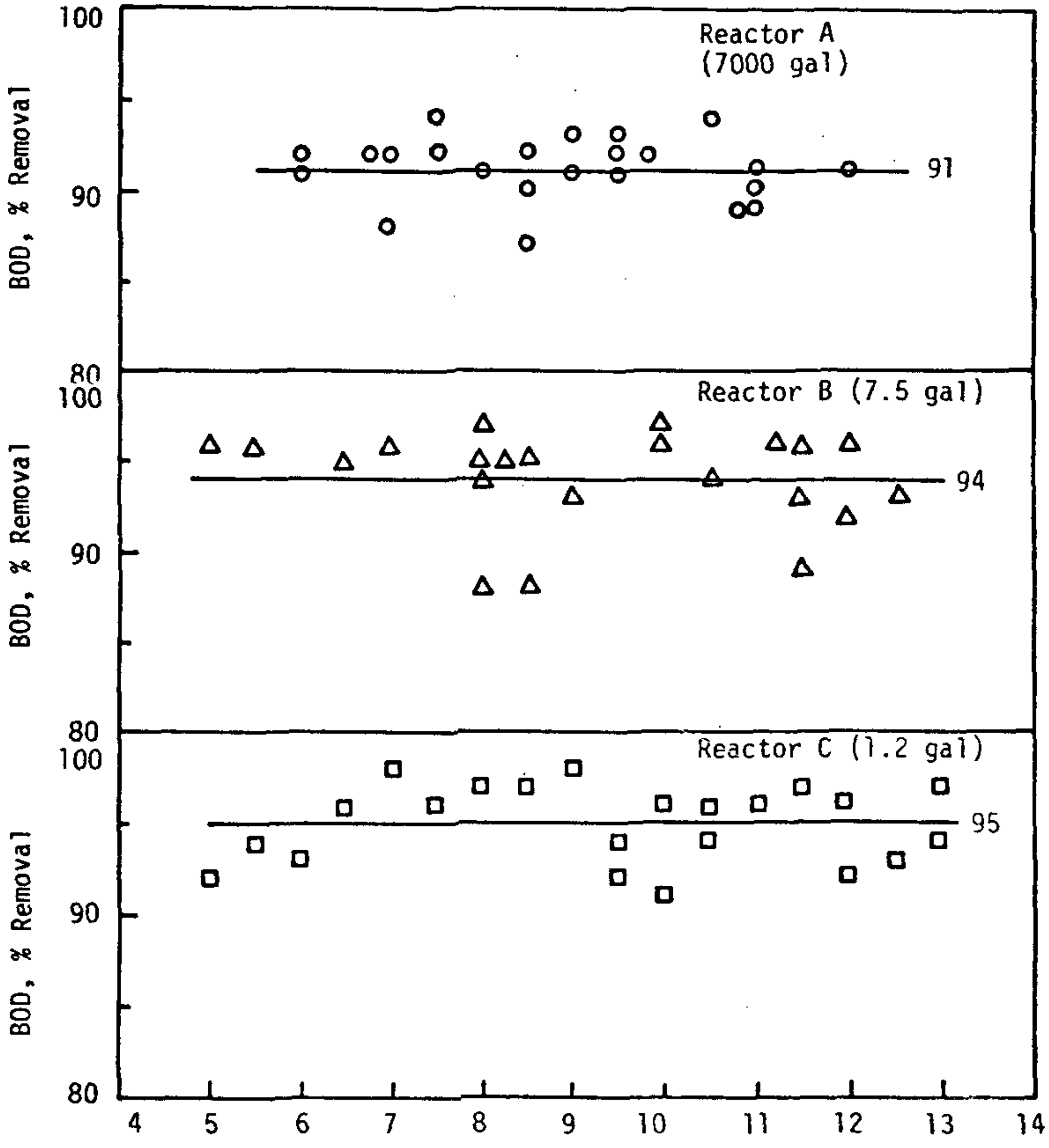


Figure 13. Response of BOD Removal to Temperature.

agreed with the findings of others to the effect that under low substrate conditions BOD removal is independent of temperature, since the experimental processes were operated at a mean cell residence time of 20 days and a mean hydraulic detention time of 12 hours, that is, a low-substrate operation.

Furthermore, the observed yield, Y_o, and F/M ratio, U, as formulated in Equations (6) and (7) are calculated. The Y_o values are plotted against temperature as shown in Figure 14. The average F/M ratios, average Y_o values, and their respective standard deviations are listed in Table IV.

TABLE IV*

Average F/M Ratios, Y_o Values, and Standard Deviations

Reactor	F/M Ratio, $\frac{\text{lb BOD removed}}{\text{lb MLVSS}}$	S.D.	Y_o, $\frac{\text{lb MLVSS produced}}{\text{lb BOD removed}}$	S.D.
A	0.18	0.04	0.28	0.07
B	0.19	0.04	0.26	0.06
C	0.19	0.04	0.26	0.06

*Average of 22 runs.

Figure 14 and Table IV indicate the following:

- It seemed that a temperature variation ranging from 5°C to 13°C exerted no significant effect on the observed yield, Y_o.
- The F/M ratio values of the three units were effectively maintained at nearly the same level of 0.19 by regulating the mean cell residence time.
- The value of Y_o varied from 0.28 for Reactor A to 0.26 for both Reactor B and Reactor C. The lack of distinct disparity in Y_o-value among the three units indicates that the mode of biochemical reactions involved was not altered as a result of the differences in size and geometry of the reactors.

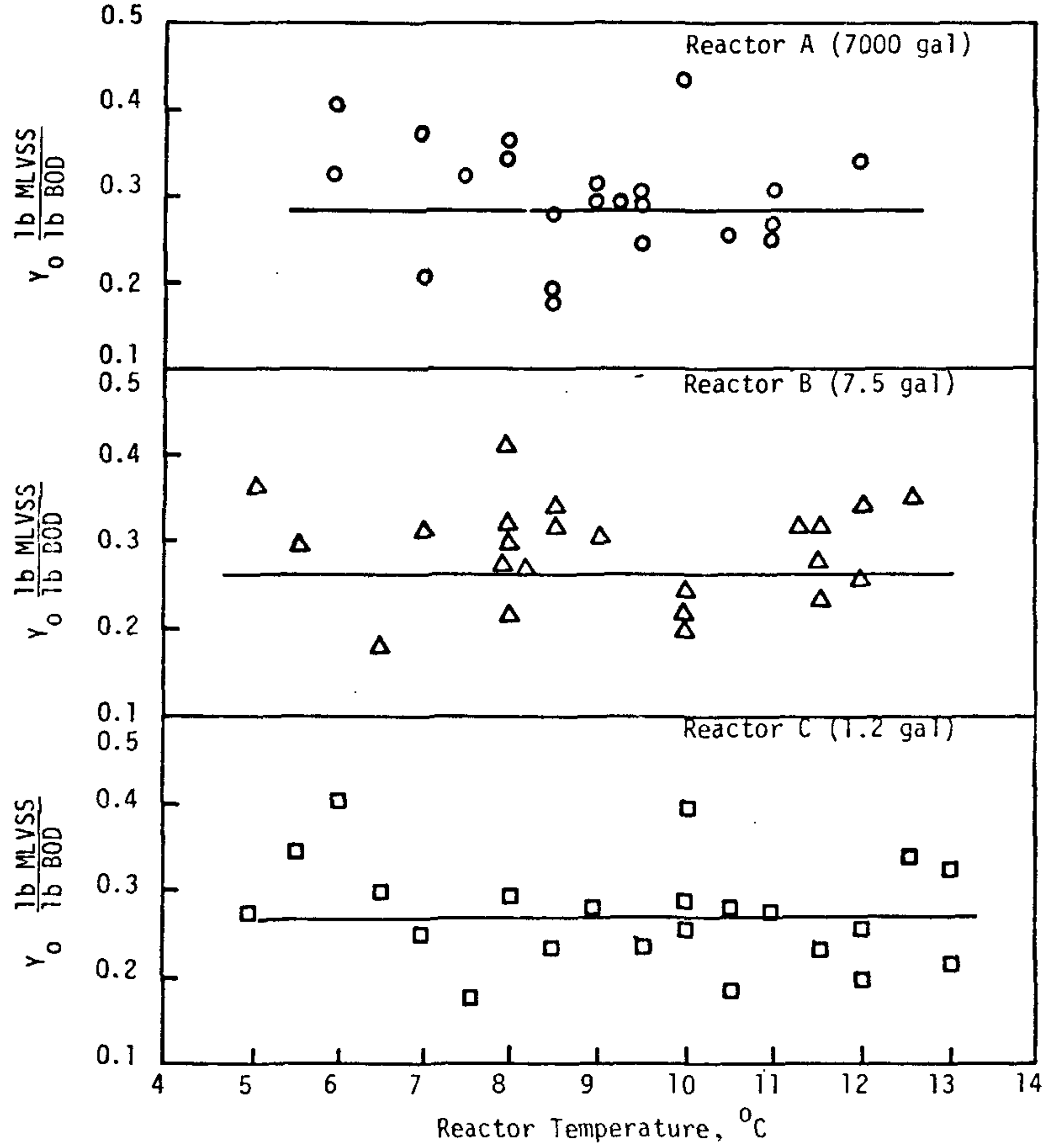


Figure 14. Response of Y_o Values to Temperature.

Therefore, the small, but significant difference in BOD removal among the units, as presented previously were apparently not due to any difference in biochemical reactions, but other causes which are speculated as follows:

a. Difference in the arrangement of piping for delivering influent and collecting effluent.

b. Difference in the provision of facilities for settling and then recycling sludge.

c. Difference in sludge settleability as a result of different degree of mixing by aeration.

d. Different extent of biochemical reaction taking place in the settling compartment of each unit.

While a portion of the Phase II study was to be devoted to scrutiniz whether there existed a significant extent of biochemical oxidation taking place in the secondary settling facilities, the possible BOD discrepancy in influent to the aeration compartment due to the difference in influent piping was investigated. Samples were taken of the influent just before its entrance to each aerator and analyzed. The results are presented in Table V.

As shown in Table V, the BOD and COD of the influent to Reactor A were higher than the BOD and COD of the influent to Reactor B and C. Whether the apparent discrepancy was due to the slow flowing of the influent through the long tubing during its transmission to Reactor B and C, and furthermore whether the discrepancy attributed to the higher BOD removal in Reactor B and C, are not conclusive because of the limited information. However, if the data given in Table V were reliable, the

TABLE V*

Analyses for the Influent Samples Taken at Aeration Inlets

Reactor	BOD mg/ℓ	COD mg/ℓ	NH_3-N mg/ℓ	Total-P mg/ℓ	Alkalinity mg/ℓ as $CaCO_3$	pH Units
A	170	408	28.5	7.6	140	7.8
B	165	396	29.0	7.6	141	7.8
C	162	391	29.5	7.6	140	7.8

*Average of three samples.

influent delivery arrangement could have increased the BOD removal in Reactor B and C.

2. COD Removal

Figure 15 shows the influent and effluent COD concentrations. Figure 16 shows the COD removal vs temperature. In general, COD removal followed a similar mode as BOD removal. The following are the observations:

a. The COD removal was independent of the temperature variation varying from 5°C to 13°C.

b. The COD removal percent, observed yield based on COD, ratio of effluent COD to BOD, and ratio of Y_0 based on COD to Y_0 based on BOD, are calculated and their average values are listed in Table VI.

3. Ammonium Removal and Nitrification

Figures 17 and 18 show the NH_3-N concentrations in the influent and effluent, and the NO_3-N concentrations in the effluent. The results are averaged and listed in Table VII.

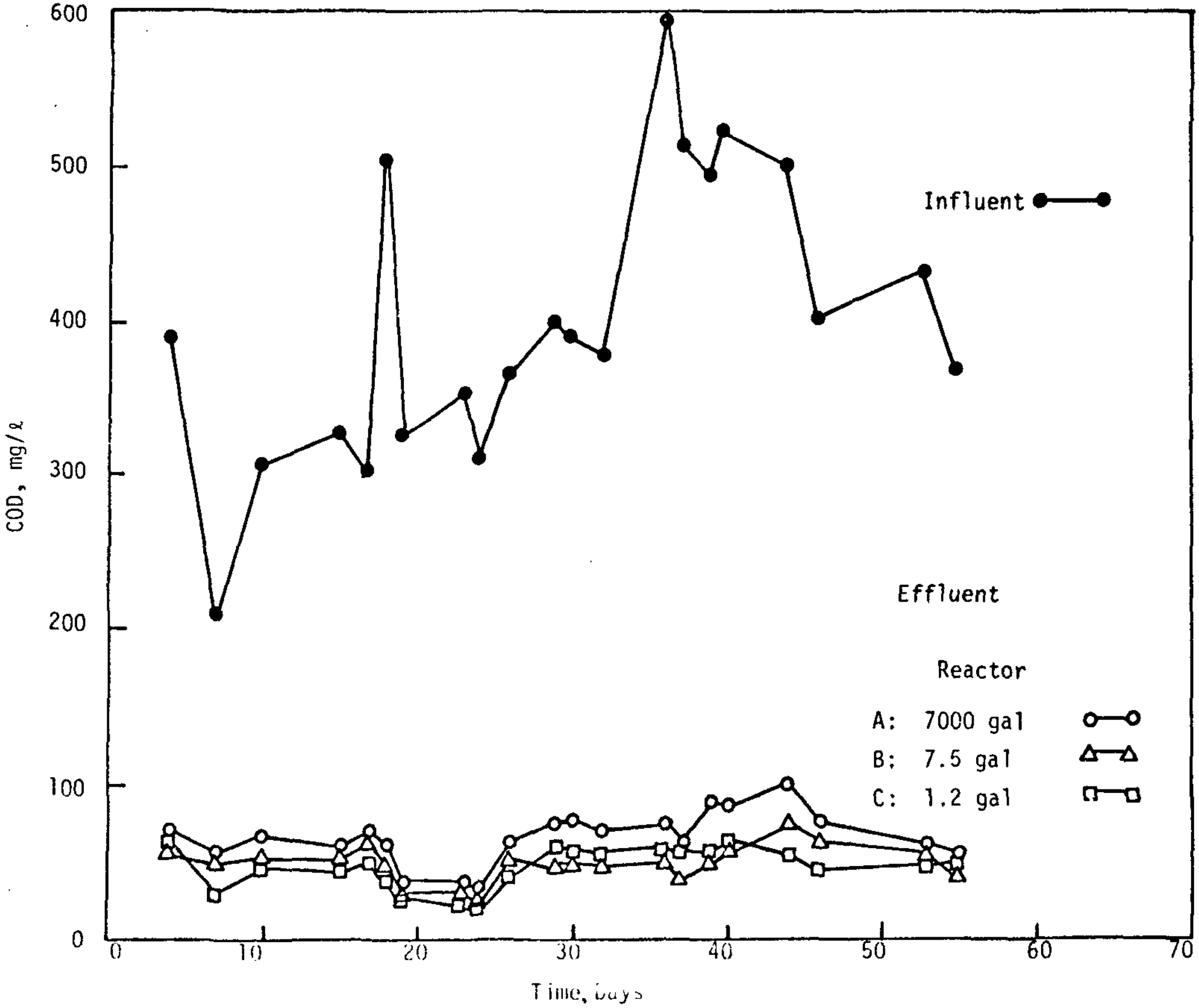


Figure 15. Influent and Effluent COD Concentrations.

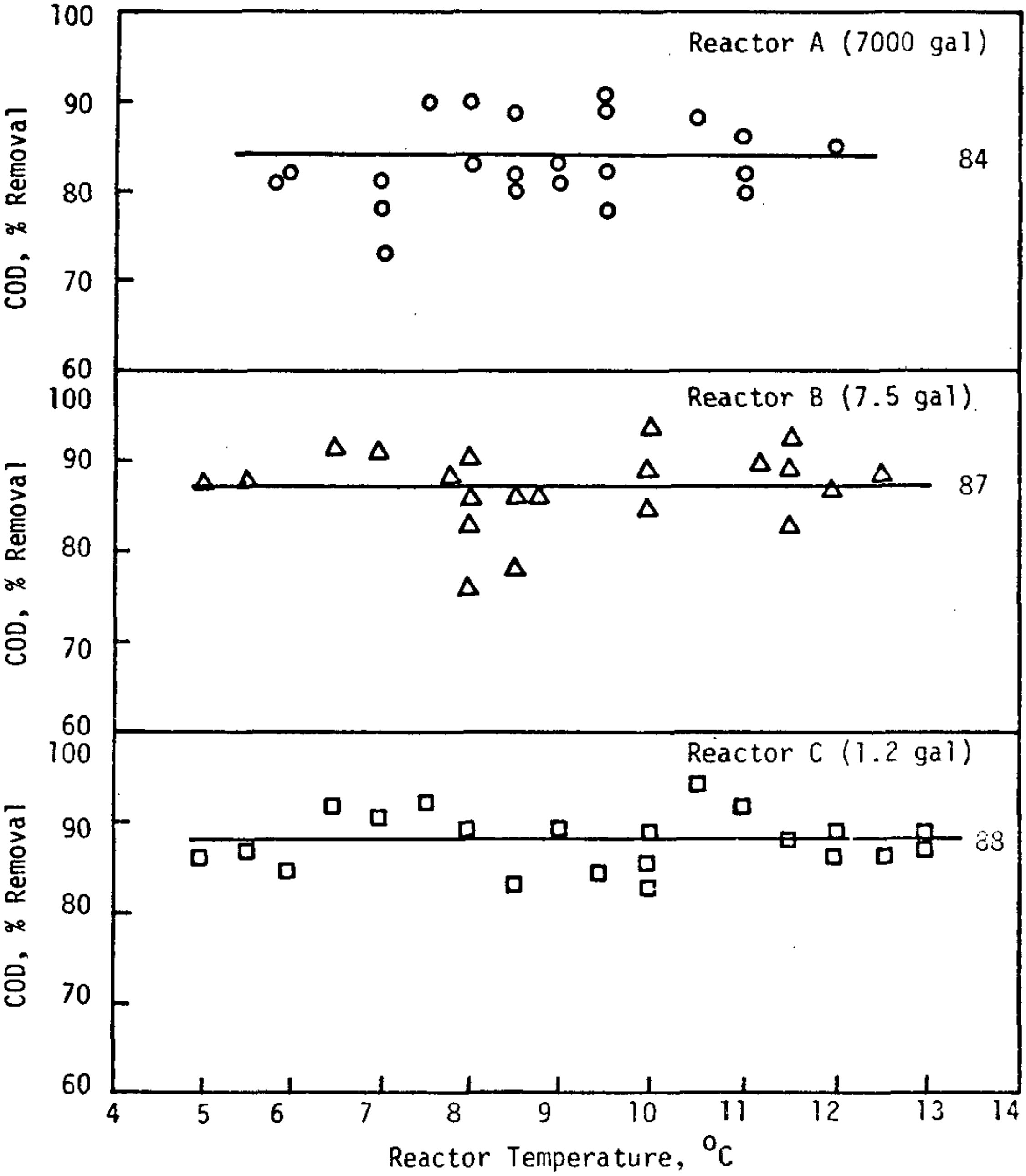


Figure 16. Response of COD Removal to Temperature.

TABLE VI

Analysis of COD Removal Data

Reactor	COD Removal %	Y_o (COD)* $\frac{\text{lb MLVSS produced}}{\text{lb COD removed}}$	$\frac{\text{COD}}{\text{BOD}}$	$\frac{Y_o(\text{BOD})^*}{Y_o(\text{COD})}$
A	84	0.18	3.8	1.6
B	88	0.17	4.2	1.6
C	88	0.17	4.3	1.6

*Y_o(COD) = Y_o based on COD, Y_o(BOD) = Y_o based on BOD.

TABLE VII

Effluent NH_3-N and NO_3-N Concentrations and NH_3-N Removal Efficiency

(Average Influent NH_3-N = 25 mg/ℓ)

Reactor	Effluent NH_3-N		NH_3-N		Effluent NO_3-N	
	Conc. mg/ℓ	S.D.	Removal, %	S.D.	Conc. mg/ℓ	S.D.
A*	4.0	2.1	84	7	20	3.8
B**	4.1	1.5	84	5	20	3.3
C**	3.7	1.7	85	8	21	6.0

*Average of 19 runs (3 runs were excluded because of low DO level and resulted abnormally low nitrification).

**Average of 22 runs.

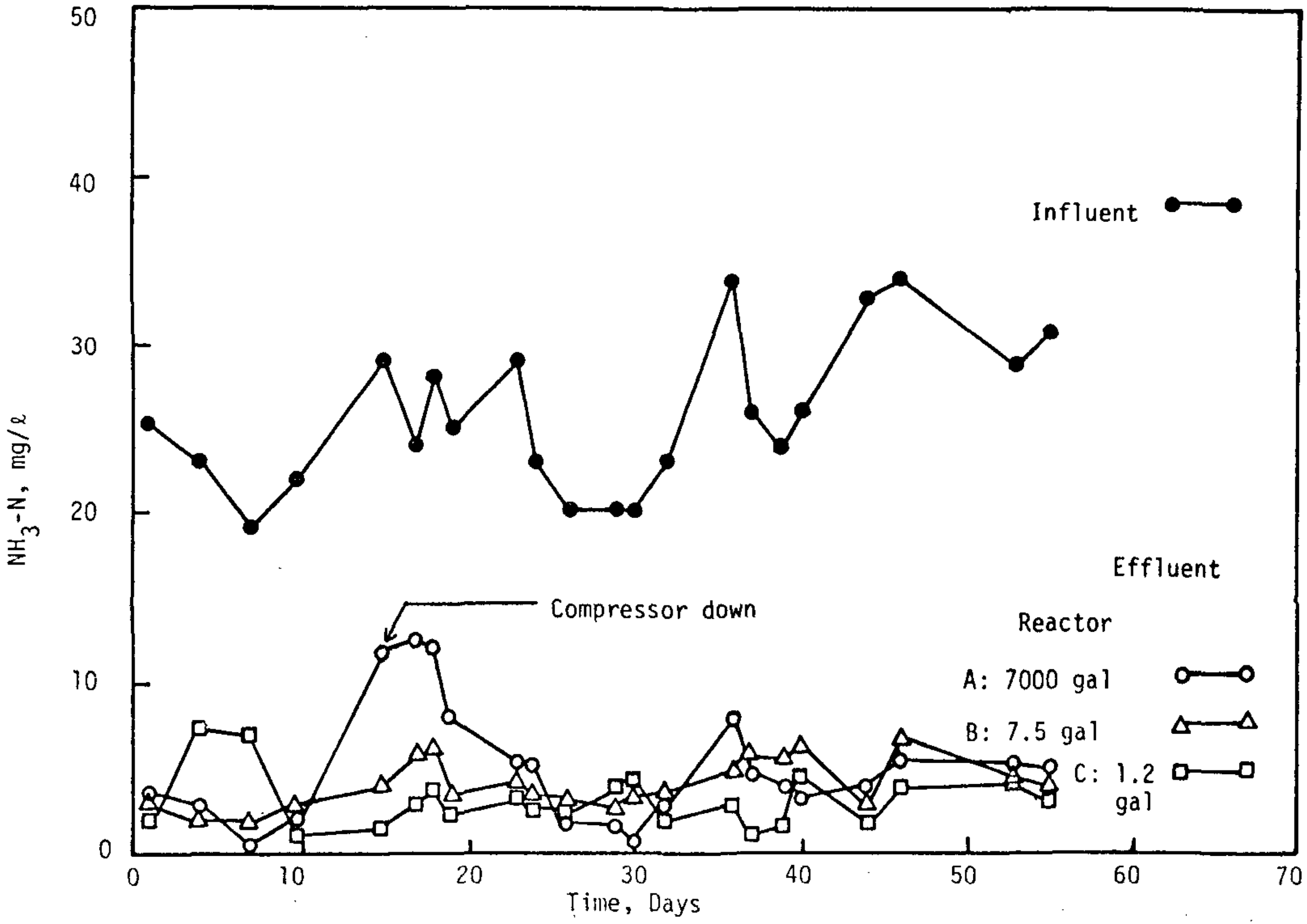


Figure 17. Daily Influent and Effluent NH_3-N Concentrations.

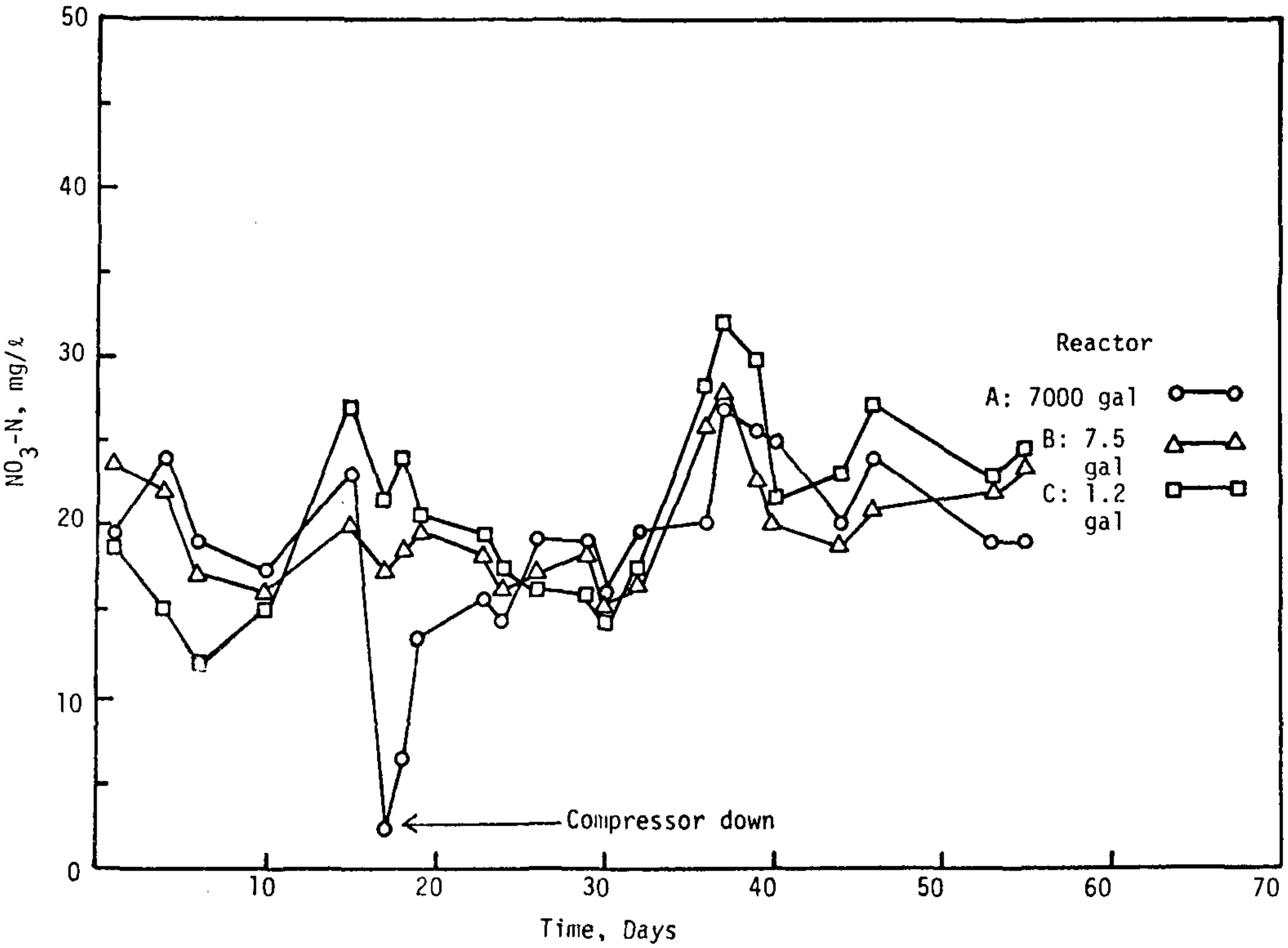


Figure 18. Daily Effluent NO_3-N Concentrations.

There was practically no difference in the nitrification performance of the three reactors. As shown in Figure 17, the high NH_3-N in the effluent and low nitrification during the period of day 15 to day 17 in the DAVCO unit was caused by the low DO level because of the breakdown of one of the compressors.

Influent NO_3-N concentrations were usually below 0.5 mg/ℓ and close to zero for most of the time.

NH_3-N removal and nitrification were reduced at low temperatures in all three reactors. Figure 19 and 20 show the responses of NH_3-N removal and nitrification, respectively, to temperature. As shown in Figure 19, the slopes of the best-fit straight lines for NH_3-N removal are nearly equal. The results show that the experimental information on NH_3-N removal from the bench-scale reactors as small as Reactor C was as reliable as that from a pilot-scale reactor as large as Reactor A.

As shown in Figure 20, generally nitrification decreased as temperature lowered, similarly to NH_3-N removal. However, as temperatures rose to above and beyond 11^{o}C, higher concentration of NO_3-N prevailed in the bench-scale reactors than in the pilot-scale reactor.

According to the kinetic models for nitrification previously described, the observed yield of nitrifiers, Y_N, could be calculated if it were possible to determine the nitrifier fraction, f; however, this would be involved and difficult. Nevertheless, if f would remain constant in two activated sludge systems, the value of Y_N/f could be used instead of Y_N to demonstrate the relative difference in their nitrifier yield. For this study, the only difference among the experimental activated sludge

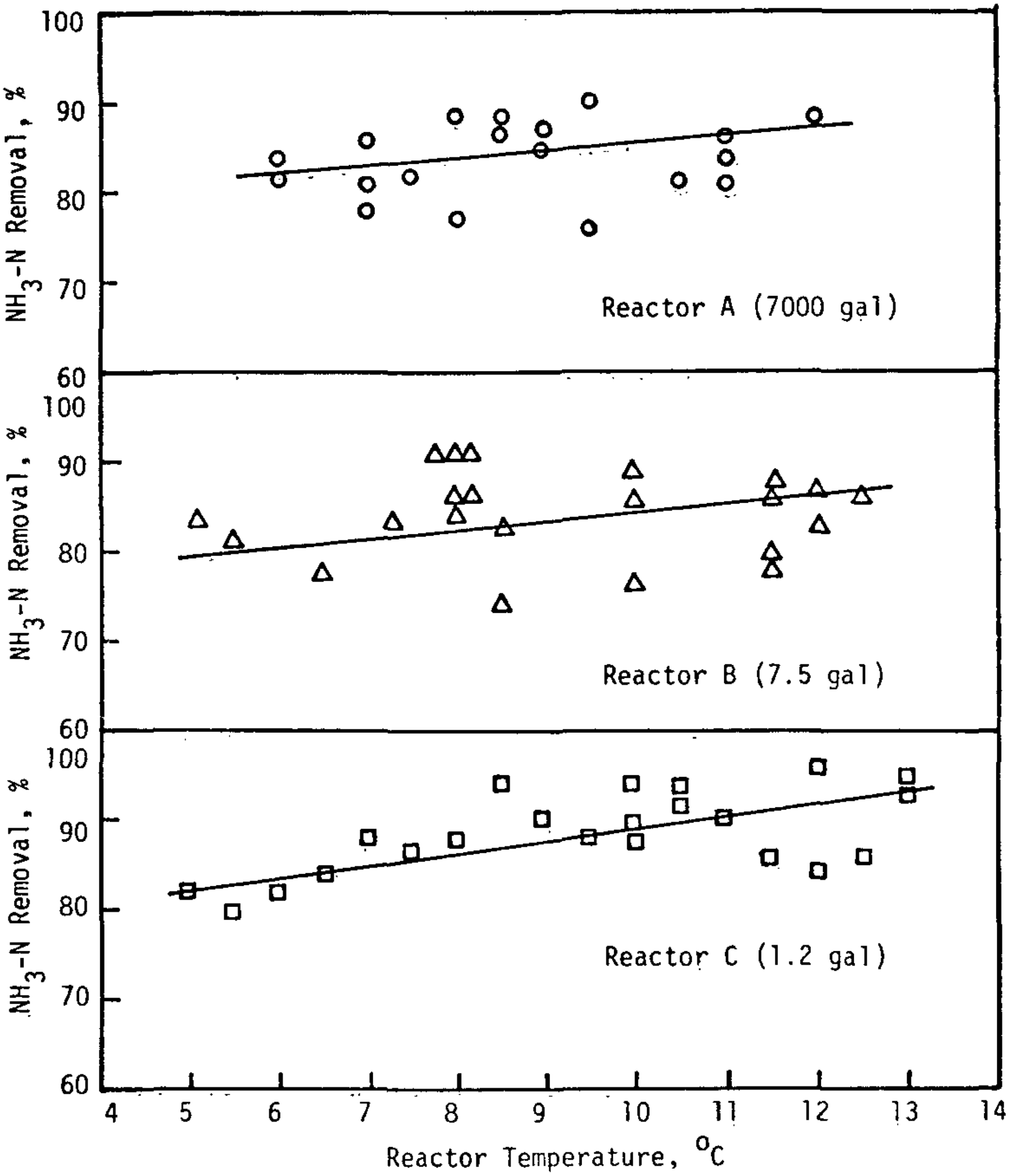


Figure 19. Response of NH_3-N Removal to Temperature.

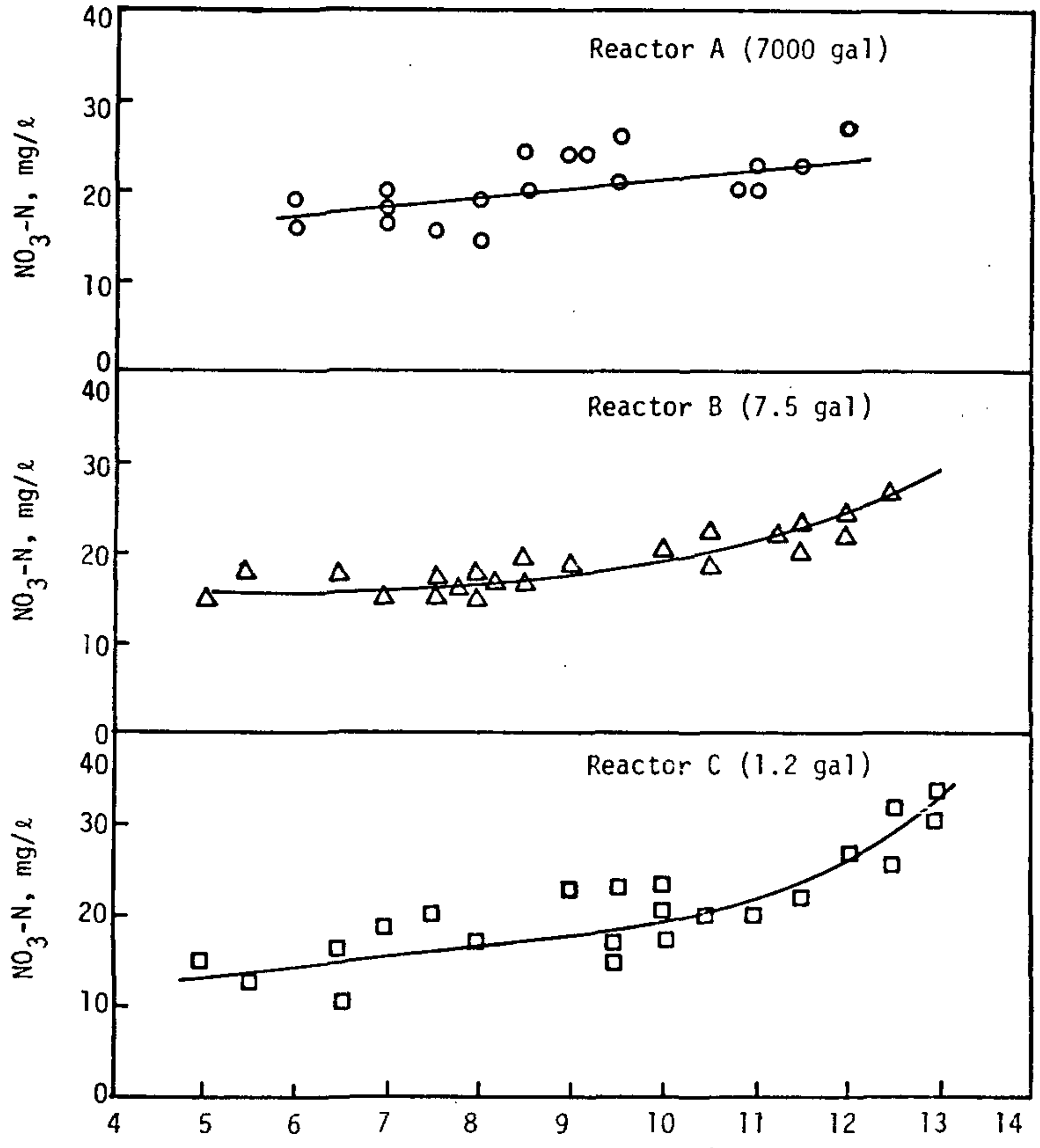


Figure 20. Response of Effluent NO_3-H Concentration to Temperature.

systems was the size and geometry of the reactors, and it is reasonable to assume the nitrifier fraction, f, was the same for all three units. Therefore, for the comparison of their nitrifier yields, the values of Y_N/f are calculated via Equation 19 and the averages are presented in Table VIII. As shown, there was no significant difference in nitrifier yield among the three reactors.

TABLE VIII

Y_N/f Values for NH_3-N Removal and NO_3-N Production

Reactor	$\frac{Y_N}{f}$ $\frac{\text{lb MLVSS produced}}{\text{lb } NH_3\text{-N removed}}$	S.D.	$\frac{Y_N}{f}$ $\frac{\text{lb MLVSS produced}}{\text{lb } NO_3\text{-N formed}}$	S.D.
A*	5.5	1.0	6.0	1.3
B**	5.2	0.9	5.7	1.0
C**	5.0	1.3	5.5	1.7

*Average of 19 runs

**Average of 22 runs

Furthermore the response of Y_N/f to temperature is shown in Figure 21. The best fit straight lines show that Y_N/f decreased linearly as temperature increased, that is, nitrifier yield per unit mass of NH_3-N utilized increased as temperature decreased. The results agreed with the finding of others (10) which stated that the F/M ratio for nitrification decreases with declining of temperature.

It is significant to note the nearly same slope of decline with temperature increase for all three best-fit straight lines, which again indicates that there was no scale-up effect of the experimental reactors on nitrification.

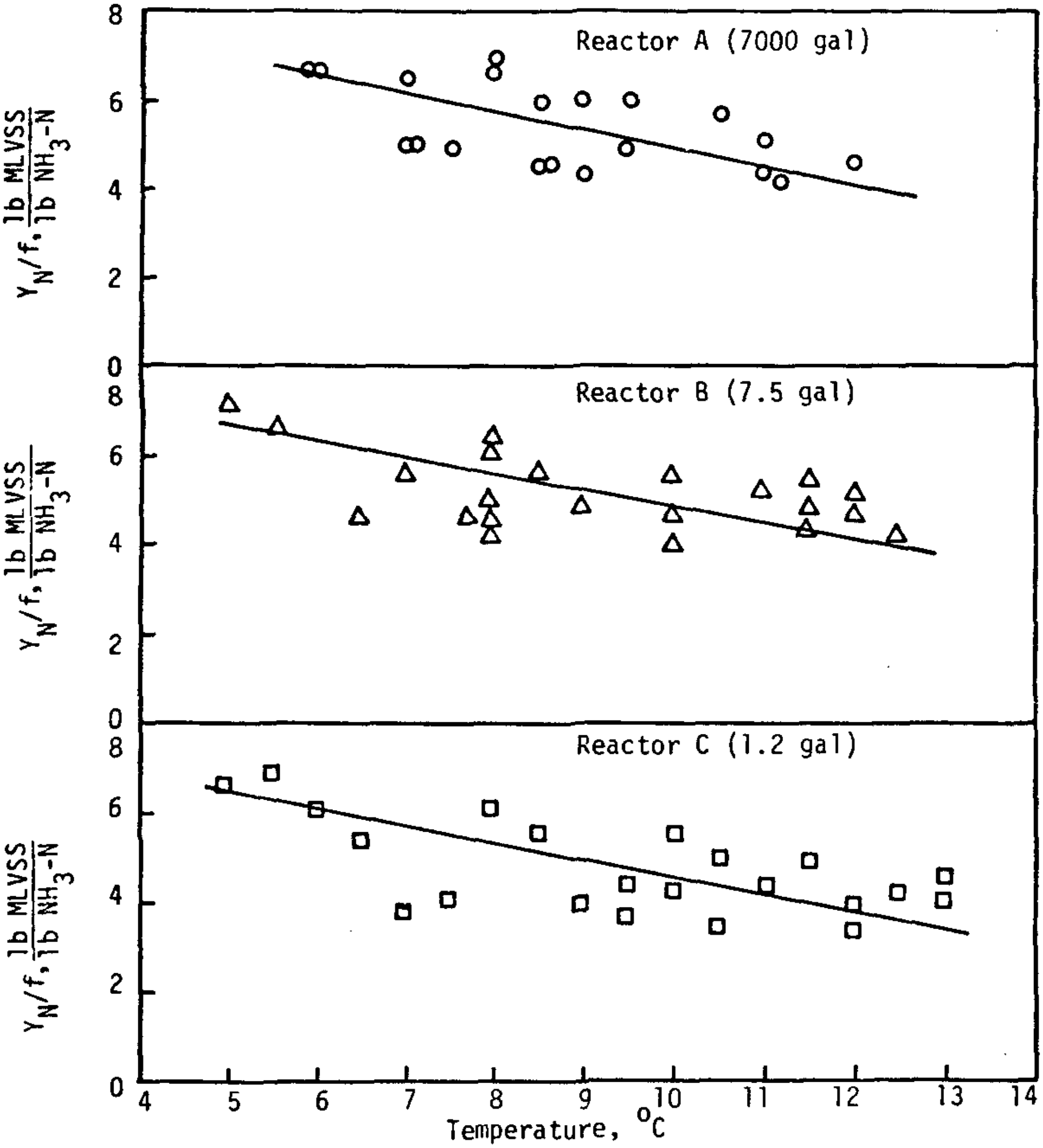


Figure 21. Response of Y_N/f to Temperature.

4. Total Phosphorus Removal

Table IX summarizes the information on effluent Total-P concentration and Total-P removed by the three reactors. The calculations are based on the data presented in Figure 22.

TABLE IX

Effluent Total-P Concentration and Total-P Removal*

Reactor	Effluent Total-P		Total-P Removal	
	conc. mg/ℓ	S.D.	Removal, %	S.D.
A	5.1	1.0	30	9.3
B	5.1	0.9	30	7.2
C	5.1	1.0	30	7.6

*Average of 22 runs.

The calculated average values of effluent Total-P concentrations and removal rates were respectively of the same values for all three reactors.

It is further shown in Figure 23 that the response of total phosphorus removal to temperature change was nearly identical in the three reactors. Therefore it seems that there was no scale-up effect on phosphorus removal in the experiments.

In Figure 23, it is also shown that the total phosphorus removal decreased as the temperature increased. Any attempt to interpret the phenomenum must be referred to the phosphorus removal mechanisms in activated sludge which have been studied and reported elsewhere (28).

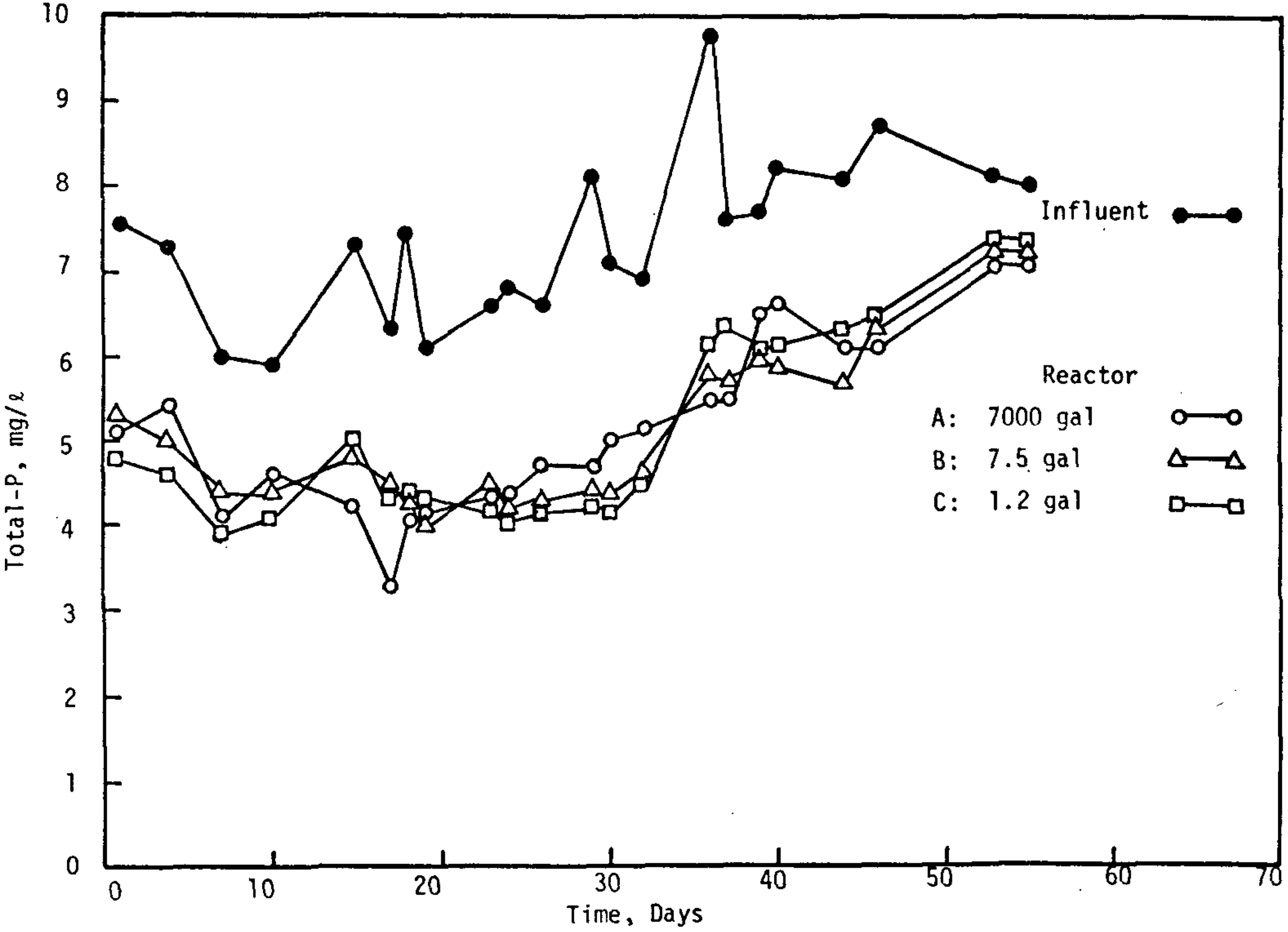


Figure 22. Daily Influent and Effluent Total-P Concentrations.

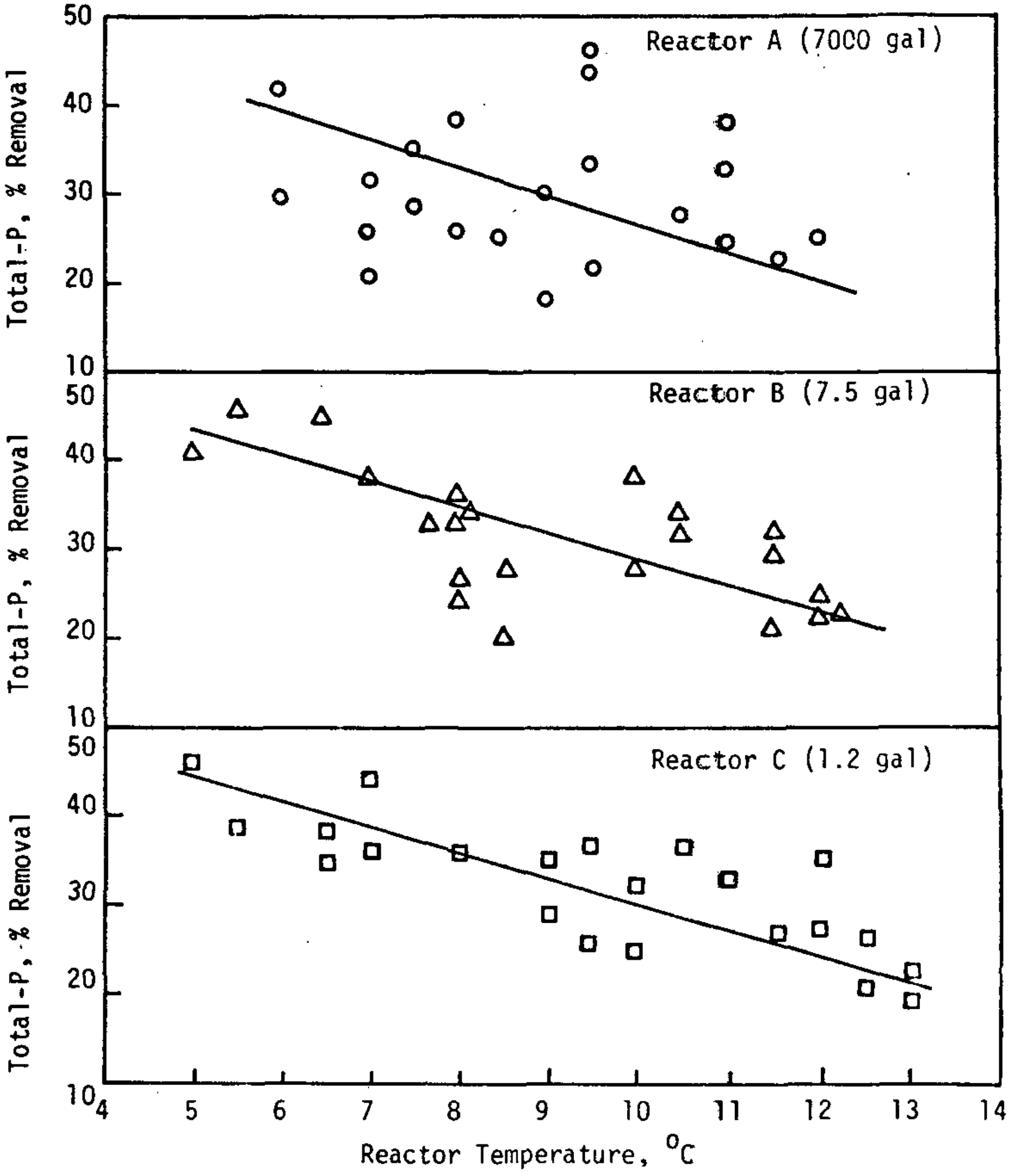


Figure 23. Response of Total-P Removal to Temperature.

Metabolic uptake by microorganisms (synthesis of bacterial cells) and chemical precipitation by calcium ions account for the phosphorus removal in activated sludge systems. Therefore it may be surmised that at low temperatures, more bacterial cells were synthesized, which somehow agreed with the higher nitrifier yield at lower temperatures. Many factors such as influent COD/P ratio, presence of fluoride ion, pH, and alkalinity all affect the phosphorus removal mechanisms.

5. Response to pH and Alkalinity Changes

pH and alkalinity greatly affect the performance of biochemical processes especially nitrification (2, 10). From the past experience, effluent pH above 5.5 would indicate adequate alkalinity present in the reactors for both carbonaceous oxidation and nitrification to occur. Figure 24 shows the influent and effluent pH and alkalinity prevailing in the reactors. The effluent pH of all units was above 5.5 throughout the experimentation, which indicates that adequate alkalinity prevailed. The sudden increase of alkalinity in the effluent of the pilot-scale unit in day 15 was due to the practical cease of nitrification due to the low DO concentrations. The chemical analyses show that the average pH was about 6.3 and average alkalinity about 14 mg/ℓ in the effluents of all the units.

C. Summary of Results

The average values of experimental results are summarized in Table X and XI. The kinetic coefficients for carbonaceous bio-oxidation and nitrification are listed in Table X, and the influent and effluent characteristics are presented in Table XI with the respective removal efficiencies.

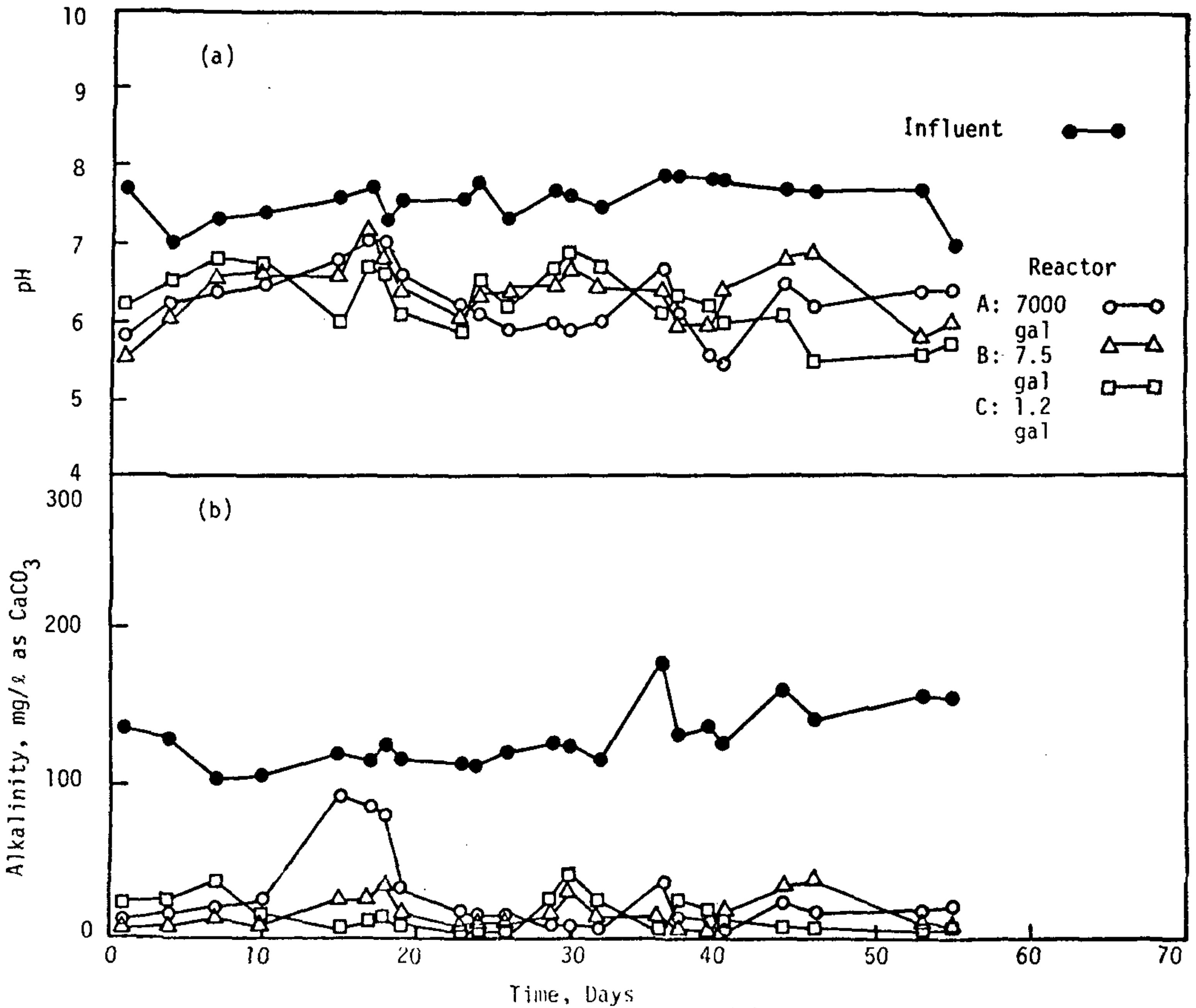


Figure 24 (a) Daily Influent and Effluent pH and (b) Daily Influent and Effluent Alkalinity.

TABLE X

Kinetic Coefficients for Carbonaceous Bio-oxidation and for Nitrification

Parameter	Reactor		
	A	B	C
$Y_o \frac{\text{lb MLVSS}}{\text{lb BOD}}$	0.28	0.26	0.26
$Y_o \frac{\text{lb MLVSS}}{\text{lb COD}}$	0.18	0.17	0.17
$Y_N/f \frac{\text{lb MLVSS}}{\text{lb NH}_3\text{-N}}$	5.5	5.2	5.0
$Y_N/f \frac{\text{lb MLVSS}}{\text{lb NO}_3\text{-N}}$	6.0	5.7	5.5

TABLE XI

Average Influent and Effluent Characteristics and the Respective Removal Efficiency

Reactor	Parameter	Influent Conc. mg/ℓ	Effluent Cond. mg/ℓ	Removal Eff. %
A (Pilot-Scale)	BOD	217	19	91
	COD	401	65	84
	NH_3-N	25	4.0	84
	NO_3-N	-	20	-
	Total-P	7.3	5.1	30
	Alkalinity (as $CaCO_3$)	127	14	-
	pH	7.6	6.2	-
B (Bench-Scale)	BOD	217	12	94
	COD	401	50	88
	NH_3-N	25	4.1	84
	NO_3-N	-	20	-
	Total-P	7.3	5.1	30
	Alkalinity (as $CaCO_3$)	127	15	-
	pH	7.6	6.4	-
C (Bench-Scale)	BOD	217	11	95
	COD	401	47	88
	NH_3-N	25	3.7	85
	NO_3-N	-	21	-
	Total-P	7.3	5.1	30
	Alkalinity (as $CaCO_3$)	127	13	-
	pH	7.6	6.3	-

PHASE II

Phase II studies were designed to complement the Phase I experiments in the following ways:

1. Operation of the reactors was modified so that their flow regimes were brought further to approach the complete-mixing mode.
2. All the samples of mixed liquor and final effluent were filtered prior to analyses in order to preclude the influence of residual flocs after settling in the calculations related to biochemical reactions taking place in the reactors.
3. Samples of mixed liquor were compared with samples of final effluent in terms of TOC, NO_3-N, and Total-P for the purpose of investigating whether there were biochemical reactions taking place during the retention time in the secondary clarifier.

The results are presented and discussed in the following three parts:

- Operating Conditions
- Effect of Final Clarifier on Process Efficiency
- Process Performance

A. Operating Conditions

Reactor temperatures were monitored and are shown in Figure 25(a). There was only a small difference in temperature among the reactors. The range of temperature was between 7 to 12°C and the average temperatures in the reactors were:

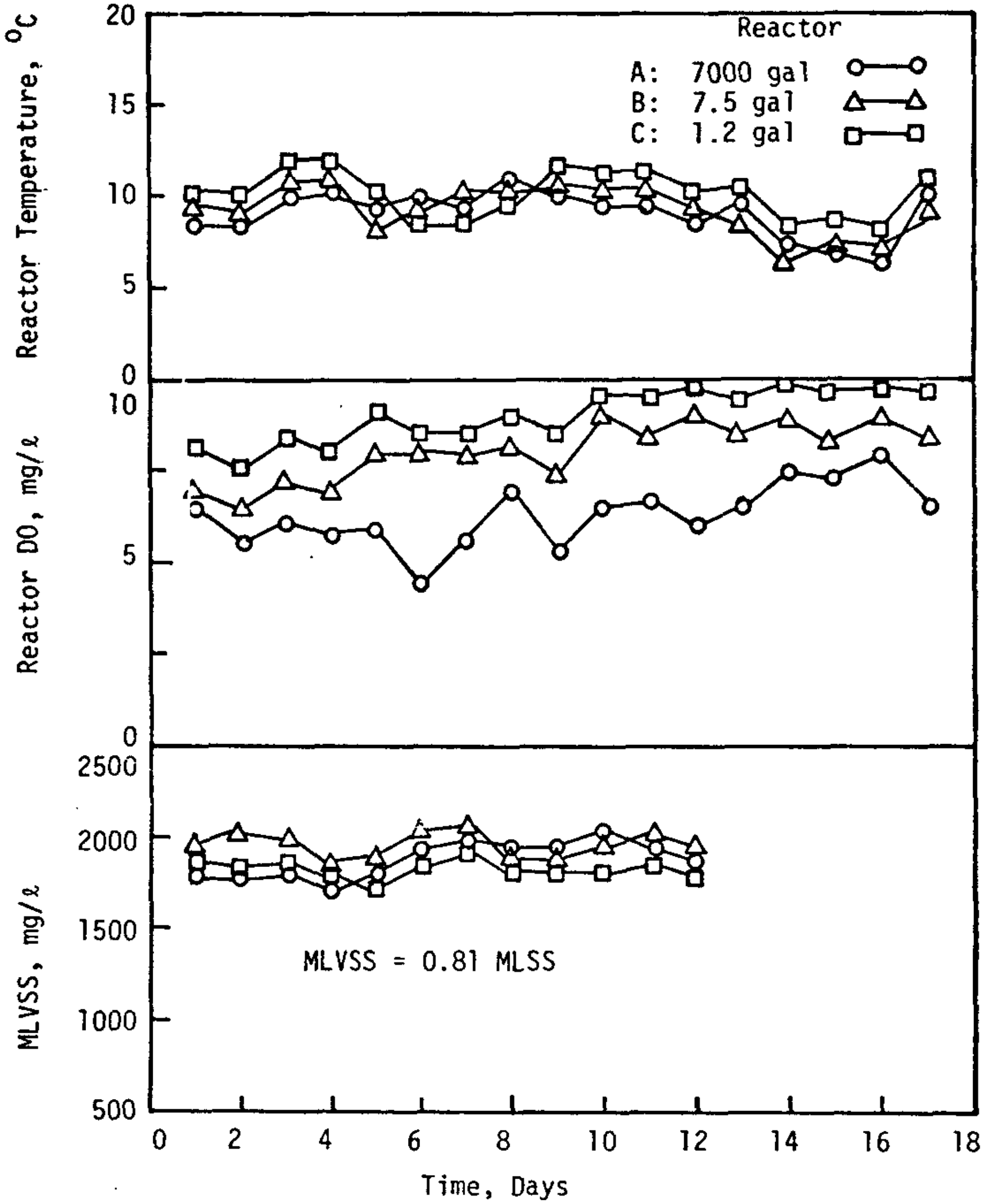


Figure 25. Daily Temperature, DO, and MLVSS.

Reactor A 9°C
Reactor B 9.5°C
Reactor C 10°C.

DO concentrations in the reactors were well above 2 mg/ℓ as shown in Figure 25(b). They were in the range of 4 to 10 mg/ℓ.

MLVSS concentrations were maintained within 1800 to 2000 mg/ℓ in the aerators as shown in Figure 25(c). The average values were 1810, 1930, and 1827 mg/ℓ , in Reactor A, B, and C, respectively, which were lower than the average MLVSS concentration of 2200 mg/ℓ maintained in the reactors in Phase I study.

B. Effect of Final Clarifier on Process Efficiency

The final clarifier could affect the process efficiency in two ways: (1) its efficiency in removing biosolids by settling and (2) biochemical reactions taking place during the detention time in the clarifier. A comparison of substrate concentrations in the mixed liquor and the final effluent would evaluate the effect of final clarification on process efficiency.

Table XII lists the average concentrations of TOC, NO_3-N, and Total-P of filtered samples of mixed liquor and final effluent and the percent removal of each component. The results of chemical analysis are given in Appendix III.

As shown in Table XII, while there was neither nitrate production nor phosphorus removal in the clarifiers of the units, there was small but persistent 1 to 3 percent additional TOC removal in the clarifiers.

TABLE XII

Average Concentrations of TOC, NO_3-N, and Total-P of Filtered Samples of Mixed Liquor and Effluent

Reactor	Parameter	TOC* mg/ℓ	TOC Removal %	NO_3-N** mg/ℓ	Total-P mg/ℓ	Total-P*** Removal,%
A	Mixed Liquor	8.4	82	18.1	3.8	27
	Effluent	7.2	85	17.8	3.8	27
B	Mixed Liquor	7.3	85	19.3	3.9	26
	Effluent	6.8	86	19.1	3.9	26
C	Mixed Liquor	7.7	84	20.2	4.1	24
	Effluent	6.5	86	20.2	4.0	24

*Average of 10 runs.

**Average of 17 runs.

***Average of 6 runs.

However, the assumption of no substrate removal by biochemical reactions, made for developing the kinetic models, is reasonable for all practical purposes.

C. Process Performance

1. TOC Removal

The influent and effluent TOC concentrations, TOC removals, and observed yields based on TOC are shown in Figure 26. The average values are summarized in Table XII. In general, there was no significant difference in biochemical reactions measured by TOC among the reactors.

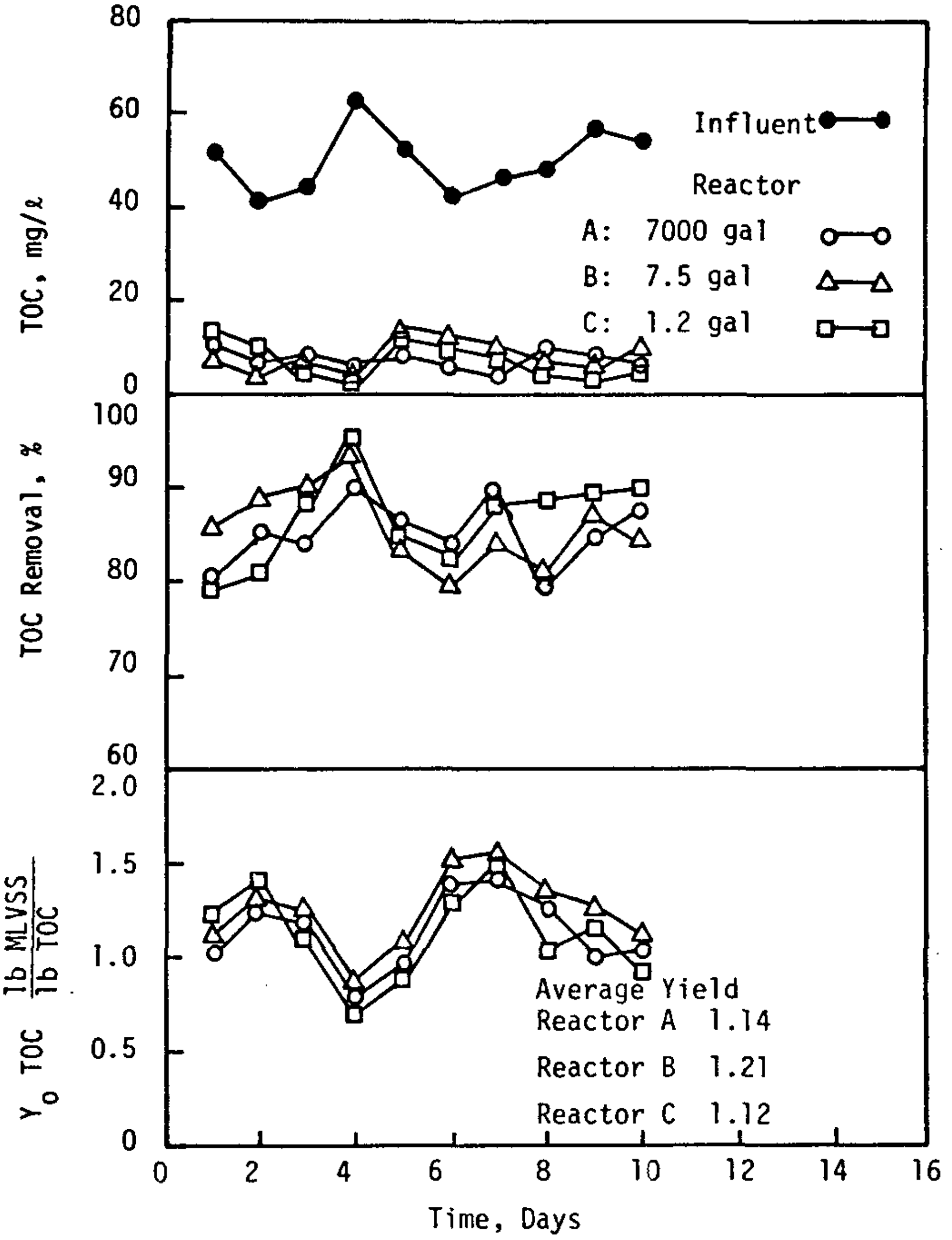


Figure 26. Daily Variations of Influent and Effluent TOC Concentration, TOC Removals, and Y_o TOC.

2. Nitrification and Phosphorus Removal

Figure 27 shows the effluent NO_3-N concentrations, influent and effluent total phosphorus concentrations, and phosphorus removals. The average values are tabulated in Table XII.

As shown, a slightly higher formation of NO_3-N occurred in the bench-scale units. However, in general there were no significant differences in biochemical reactions involving nitrogen and phosphorus among the reactors.

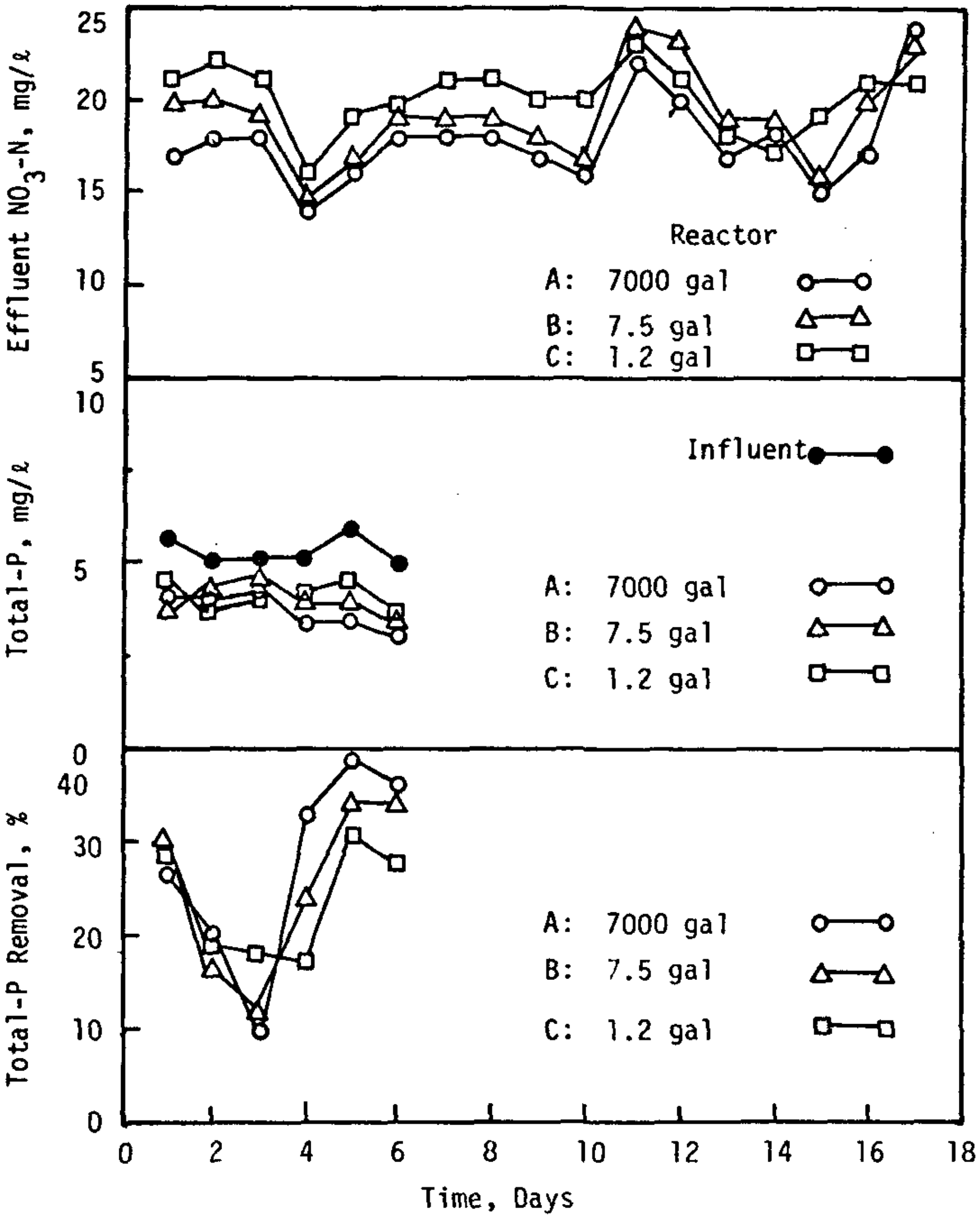


Figure 27. Daily Variations of Effluent NO_3-N Concentration, Influent and Effluent Total-P Concentration, and Total-P Removal

VI. CONCLUSIONS

As a result of this study, the following are the conclusions:

PHASE I:

1. While BOD and COD removals were practically the same in the two bench-scale reactors, there was a small, but statistically significant difference in BOD and COD removals between bench-scale and pilot-scale reactors. However, the difference could be attributed to other than the scale-up effects.
2. There were no differences in NH_3-N removal, nitrification, and total phosphorus removal among the three reactors.
3. Kinetic coefficients for carbonaceous bio-oxidation and nitrification are calculated, based on the kinetic models developed. The calculated values are practically the same for all three reactors.
4. The biochemical reactions in the three reactors responded similarly to the temperature variation ranging from 5°C to 13°C as follows:
    a. BOD and COD removals were independent of the temperature variation.
    b. NH_3-N removal and nitrification increased linearly as temperature rose from 5°C to 13°C.
    c. Total phosphorus removal reduced linearly as temperature increased from 5°C to 13°C.
5. In general, a bench-scale reactor, as small as 1.2 gallon, could be applied to generate treatability information as reliable as a pilot-scale reactor having a liquid volume of 7000 gallons.

PHASE II:

1. There were no nitrification and total phosphorus removal taking place in the final clarifiers of the three reactors. However, an additional 1-3 percent of TOC removal in the overall 85 percent removal was exerted in the final clarifiers.
2. The assumptions of no substrate removal by biochemical reactions and no microorganism growth in the final clarifiers commonly made for developing the kinetic models are reasonable for all practical purposes.

VII. RECOMMENDATIONS

The following additional investigations are recommended:

1. The effect of mean cell residence time on the treatment performance has been widely studied. Generally, a gradual reduction in treatment efficiency is expected with a decrease in mean cell residence time until a critical mean cell residence time is reached where the treatment efficiency drops rapidly. This study was operated at a relatively long mean cell residence time of 20 days. Therefore, it is recommended that the scale-up effect be studied at relatively short mean cell residence times, e.g. 5 days.
2. *It is recommended to extend the study on the performance of a* process in bench-scale reactors under the conditions of complete-mix flow as compared with that in pilot-scale reactors under conditions of arbitrary flows and plug flow.

VIII. REFERENCES

1. Toerber, E. D., "Full Scale Parallel Activated Sludge Process Evaluation." Report for Office of Research and Monitoring, USEPA, Environmental Protection Technol. Series (1972).

2. Metcalf and Eddy, Inc., Wastewater Engineering, Collection, Treatment, Disposal . McGraw-Hill, New York, New York (1972).

3. Lawrence, A. W., and McCarty, P. L., "Unified Basis for Biological Treatment Design and Operation." Jour. Sanitary Eng. Div., ASCE, 96, SA3, 757 (1970).

4. Jenkins, D., and Garrison, W. E., "Control of Activated Sludge by Mean Cell Residence Time." Jour. Water Poll. Control Fed., 40, 1905 (1968).

5. Eckenfelder, W. W., and O'Connor, D. J., Biological Waste Treatment, Pergamon Press, London (1964).

6. Sherrard, J. H., and Schroeder, E. D., "Relationship Between the Observed Cell Yield Coefficient and Mean Cell Residence Time in the Completely Mixed Activated Sludge Process." Water Research, (G.B.) 6, 1039 (1972).

7. Walker, L. F., "Hydraulically Controlling Solids Retention Time in the Activated Sludge Process." Jour. Water Poll. Control Fed., 43, 30 (1971).

8. Ribbons, D. W., "Quantitative Relationships Between Growth Media Constituents and Cellular Yields and Composition." Methods in Microbiology, (G.B.), 3A (1970).

9. Kalinske, A. A., "Effect of Dissolved Oxygen and Substrate Concentration on the Uptake Rate of Microbial Suspensions." Jour. Water Poll. Control Fed., 43, 73 (1971).

10. "Process Design Manual in Nitrogen Control". USEPA, Office of Technol. Transfer, Washington, D. C. (1975).

11. Okun, D. A., and Lynn, W. R., "Preliminary Investigations into the Effect of Oxygen Tension on Biological Sewage Treatment." McCabe and Eckenfelder (Ed.), Reinhold Publ. Corp., New York, New York (1956).

12. Benefield, L. D., Randall, E. R., "Stimulation of Filamentous Microorganisms in Activated Sludge by High Oxygen Concentration." Water Air Soil Poll., 5, 1 (1975).

13. Levenspiel, O., Chemical Reaction Engineering, Chap. 9, 2nd Ed., John Wiley and Sons, New York, New York (1972).

14 Schroeder, E. D., "Relationship Between Process Configuration and Process Performance." Jour. Water Poll. Control Fed., 47, 5 (1975).

15. Sawyer, C. N., "Results of Feeding Experiments to Determine the Effect of the Variables Temperature and Sludge Concentration." Sewage Works Jour., 12, 2,44 (1940).

16. Ludzack, F. J., et al., "Temperature and Feed as Variables in Activated Sludge Performance." Jour. Water Poll. Control Fed., 33, 141 (1961).

17. Hunter, J. V., Genetelli, E. J. and Gilwood, M. E., "Temperature and Retention Time Relationships in Activated Sludge Process." Proc. 21st Ind. Waste Conf., Purdue Univ. Ext. Ser. 121, 953 (1966).

18. Friedman, A. A., and Schroeder, E. D., "Temperature Effects on Growth Yield in Activated Sludge." Jour. Water Poll. Control Fed., 44, 1433 (1972).

19. Benedict, A. H., and Carlson, D. A., "Temperature Acclimation in Aerobic Bio-oxidation Systems." Jour. Water Poll. Control Fed., 45, 10 (1973).

20. Keefer, C. E., "Temperature and Efficiency of the Activated Sludge Proc. ." Jour Water Poll. Control Fed., 34, 1186 (1962).

21. Novak, O. T., "Temperature-substrate Interactions in Biological Treatment." Jour. Water Poll. Control Fed., 46, 1984 (1974).

22. Ludzack, E. J., "Laboratory Model Activated Sludge Unit." Jour. Water Poll. Control Fed., 32, 605 (1960).

23. Burchelt, M. E., and Tchobanoglous, G., "Facilities for Controlling the Activated Sludge Process by Mean Cell Residence Time." Jour. Water Poll. Control Fed., 46, 5 (1974).

24. Salvador, M., and O'Brien, J., "Alkalinity and Extended Aeration." Jour. New England Water Poll. Control Assoc., 10, 182 (1976).

25. Standard Methods for the Examination of Water and Wastewater, 13th Edition, APHA, AWWA, WPCF, New York, New York (1971).

26. "Nitrification and Denitrification Facilities". EPA Technol. Transfer Seminar Publication. (1973).

27. Martel, C. J. Pilot Plant Studies at University of Massachusetts. Personal Communication. (1976).

28. Morgan, W. E., and Fruh, E. G., "An Investigation of Phosphorus Removal Mechanisms in Activated Sludge Systems." Report for Office of Research & Monitoring, USEPA, Environ. Prot. Technol. Series (1972).

IX. APPENDIX

I. Notation

$\frac{\Delta F}{\Delta t}$	rate of substrate utilization on a finite basis, Mass/Volume-Time
Q	influent flowrate, Volume/Time
V	volume of the reactor, Volume
S_o	influent substrate concentration, Mass/Volume
S	effluent substrate concentration, Mass/Volume
X	concentration of the active microorganisms, Mass/Volume
U	food to microorganisms ratio, $\frac{\text{Mass of substrate consumed}}{\text{Mass of active cells-Time}}$
$\frac{\Delta X}{\Delta t}$	net growth rate of microorganisms on a finite basis, $\frac{\text{Mass}}{\text{Volume-Time}}$
W	wasteage flowrate, Volume/Time
X_e	microorganism concentration in the effluent, Mass/Volume
X_r	microorganism concentration in the recycle line, Mass/Volume
μ	specific growth rate, Mass/Mass-Time
Y_o	observed yield, $\frac{\text{Mass of cell growth}}{\text{Mass of substrate utilized}}$
Y	true yield, $\frac{\text{Mass of cell growth}}{\text{Mass of substrate utilized}}$
K_d	respiration coefficient, 1/Time
E	treatment efficiency, %
θ_c	mean cell residence time, Time
θ	mean hydraulic detention time, Time

Notation

R	recycle flowrate, Volume/Time
α	recirculation ratio
$(\frac{\Delta F}{\Delta t})_N$	rate of NH_3-N utilization, $\frac{\text{Mass of } NH_3\text{-N}}{\text{Volume-Time}}$
U_N	F/M ratio for nitrification, 1/Time
X_N	nitrifier concentration, Mass/Volume
X_{eN}	nitrifier concentration in the effluent, Mass/Volume
X_{rN}	nitrifier concentration in the recycle line, Mass/Volume
N_o	influent NH_3-N concentration, Mass/Volume
N	effluent NH_3-N concentration, Mass/Volume
f	nitrifier fraction of the total biomass.
μ_N	specific growth rate for nitrifiers, 1/Time
$(\frac{\Delta X}{\Delta t})_N$	net growth rate of nitrifiers, Mass/Volume-Time
Y_N	observed *yield* for nitrifiers, $\frac{\text{Mass of cells produced}}{\text{Mass of } NH_3\text{-N consumed}}$

II. THE STUDENT t DISTRIBUTION

Two population means may be compared by estimating their difference:

$$\mu_1 - \mu_2$$

A reasonable estimate of this difference in population means is the difference in sample means:

$$\overline{X}_1 - \overline{X}_2$$

Because of the sampling fluctuation in point estimates, interval estimate is desirable. Thus, the distribution of estimator $(\overline{X}_1-\overline{X}_2)$ can be used to make an inference about the population parameter $(\mu_1-\mu_2)$.

In practice the Student t distribution is only for small samples (usually less than 25) when the population variance, σ^2, is unknown. For independent samples, the population variance can be estimated by the pooled sample variance defined below:

$$S_p{}^2 = \frac{1}{(n_1+n_2-2)} \left(\sum_{i=1}^{n_1} (X_{1i}-\overline{X}_1) + \sum_{i=1}^{n_2} (X_{2i}-\overline{X}_2) \right)$$

where $S_p{}^2$ = pooled sample variance

n_1 = first sample size

n_2 = second sample size

X_{1i} = ith observation in the first sample

X_{2i} = ith observation in the second sample

Thus, the 95 percent confidence interval for the difference in means can be written in the following:

$$(\mu_1-\mu_2) = (X_1-X_2) \pm t_{0.025}\, S_p \sqrt{\frac{1}{n_1} + \frac{1}{n_2}}$$

where $t_{0.025}$ is the critical t value with degree of freedom $n_1 + n_2 - 2$.

If the difference in two population means is estimated between the 95 percent confidence interval, we say it is acceptable at the 95 percent confidence level or 5 percent significance level. All other values outside the confidence interval are rejected, and we call our results statistically significant at 95 percent confidence level or 5 percent significance level.

III. DATA ON PROCESS PERFORMANCE FROM PHASE II STUDIES

A. TOC (all in mg/ℓ)

Reactor	A		B		C	
Run No.	Reactor Substrate	Effluent Conc.	Reactor Substrate	Effluent Conc.	Reactor Substrate	Effluent Conc.
1	10	10	8	7.5	13	11
2	6	6.5	5	4.5	10	8
3	7	7	4	4.5	6	4.5
4	6. 5	5.5	4.5	3	4	3.5
5	10	7.5	8.5	8.5	8	8
6	10	6.5	10	8	10	7
7	7	4.5	5	7	5	4.5
8	11	9.5	10.5	9	6	5
9	9	8.5	8	7.5	8	7
10	7	6	9	8	7	6
Mean	8.4	7.2	7.3	6.8	7.7	6.5
S.D.	1.8	1.8	2.4	2.0	2.7	2.2

B. NO_3-N (all in mg/ℓ)

Reactor	A		B		C	
Run No.	Reactor Conc.	Effluent Conc.	Reactor Conc.	Effluent Conc.	Reactor Conc.	Effluent Conc.
1	18	17	21	20	23	21
2	18	18	21	20	22	22
3	18	18	20	19	21	21
4	14	14	15	15	16	16
5	16	16	16	16	19	19
6	18	18	18	18	19	19
7	18	18	20	19	21	21
8	18	18	19	19	21	21
9	17	17	18	18	20	20
10	16	16	16	16	20	22
11	23	22	24	24	23	24
12	20	20	24	23	21	21
13	19	17	20	19	18	18
14	18	18	19	18	18	18
15	15	15	16	16	19	18
16	17	17	18	20	21	21
17	25	24	23	24	20	21
Mean	18.1	17.8	19.3	19.1	20.2	20.2
S.D.	2.67	2.44	2.76	2.67	1.85	1.94