

THE EFFECT OF MEDIA TYPE  
on the  
ANAEROBIC HYBRID REACTOR

by

Vijay B. Thadani

A thesis  
presented to the University of Manitoba  
in partial fulfillment of the  
requirements for the degree of  
Master of Science in Engineering

Department of Civil Engineering  
Winnipeg, Manitoba  
April 1987

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ISBN 0-315-44097-X

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## ABSTRACT

A parallel study was conducted with four laboratory-scale anaerobic hybrid (Anhybrid) reactors. Three different media configurations were investigated: random media (no orientation); media with vertical orientation; and no media (an UASB configuration). The reactors were operated at a temperature of 35°C and fed a synthetic sulphite evaporator condensate. The performance and operation of the reactors was monitored over a 100 day period. It was found that the reactors with random media were best able to retain biomass and that their performance, in terms of organic removal efficiency, was more stable than the reactors with the other two media configurations.

Tracer tests were conducted on the reactors to examine the effect of media type on the hydraulic regime in the reactors. Methylene blue and rhodamine B were used as tracers. The visual results obtained with the methylene blue indicated that the hydraulic regime in the reactors with random media gravitated towards plug flow. The other reactors dispersed the tracer to a much greater degree, tending towards a completely mixed hydraulic regime.

The dispersion number of each reactor configuration was calculated with tracer-response curves plotted from the tracer tests using rhodamine B. The reactors with random media consistently had lower dispersion numbers than the other two reactors. This suggests that there is a correlation between the ability of a media type to retain biomass and dispersion number.

## ACKNOWLEDGEMENTS

The author wishes to express his most sincere gratitude to Dr. Jan Oleszkiewicz for his support, understanding and patience throughout the duration of this study.

Thanks go to Judy Tingley who was never too busy to lend a helping hand in the laboratory.

The author is grateful to Prof. Arthur Sparling and Dr. Gregory Blank for serving on his Examining Committee.

Special thanks are extended to Wendy Seversen and Norma Lindsay who always had an encouraging word.

Thanks go to my good friend Mary-Ann Bralec who worked above and beyond the call of duty in completing this project.

Last, but certainly not least, the author wishes to acknowledge the assistance provided by his fellow graduate students, Daryl McCartney and Shahab Shafai.

Thank you all.

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## NOMENCLATURE

Anbiof	anaerobic filter, anaerobic biofilter
Ancont	anaerobic contact process
Anhybrid	anaerobic hybrid reactor
C	tracer concentration, $M L^{-3}$
COD <sub>si</sub>	soluble influent chemical oxygen demand, $M L^{-3}$
COD <sub>se</sub>	soluble effluent chemical oxygen demand, $M L^{-3}$
d	longitudinal dispersion coefficient, $L^2 T^{-1}$
F/M	food to microorganism ratio, $T^{-1}$
HRT, $\theta$	hydraulic retention time, T
Hyan	hybrid anerobic reactor
$k_d$	endogenous decay coefficient, $T^{-1}$
$K_s$	half velocity constant, $M L^{-3}$
$K_l$	specific substrate utilization constant, $T^{-1}$
l	length of path in reactor, L
Q	flow rate, $L^3 T^{-1}$
$r_g$	growth rate, $M L^{-3} T^{-1}$
$r_n$	net rate of growth, $M L^{-3} T^{-1}$
$r_s$	rate of substrate utilization, $M L^{-3} T^{-1}$
S	concentration of limiting substrate in solution, $M L^{-3}$
SEC	sulphite evaporator condensate
$S_o$	influent substrate concentration, $M L^{-3}$
SRT, $\theta_c$	solids retention time, T
t	time, T
$\mu_m$	maximum specific growth rate, $T^{-1}$

u	mean flow velocity along reactor, $L T^{-1}$
U	specific substrate utilization rate, $T^{-1}$
UASB	upflow sludge bed reactor
VFA	volatile fatty acids
X	concentration of microorganisms, $M L^{-3}$
Y	growth yield coefficient, $M M^{-1}$

## CHAPTER I

### INTRODUCTION TO ANAEROBIC PROCESSES

#### 1.1 Advantages of Anaerobic Processes

Anaerobic processes have been utilized sporadically for wastewater treatment for some time, but their full potential has yet to be realized. The reason anaerobic technology is in its infancy in the field of environmental engineering is not a lack of research; it is a result of the apprehension of design engineers towards the process and their unfamiliarity with the microbiological and biochemical fundamentals involved. This attitude can be attributed to the many misconceptions that prevail with regard to anaerobic treatment including the following:

- (a) the process is extremely sensitive, and susceptible to shocks;
- (b) it produces offensive odours; and
- (c) it is not adaptable to a variety of waste streams.

In fact, anaerobic treatment does have some drawbacks. The most prominent of these are the long start up times required (up to 12 weeks) and the general lack of practical experience with regard to the treatment of wastewater.

However, the advantages of anaerobic treatment are far more numerous and include the following:

- (a) a high degree of waste stabilization at high organic load rates;
- (b) the production of methane as a recyclable end product;
- (c) small quantities of excess sludge are produced and these are easily dewatered;
- (d) if properly acclimated, stabilization of toxic compounds is possible; and
- (e) a well acclimated sludge can be left unfed for long periods of time without noticeable deterioration.

Given these benefits, anaerobic treatment is sure to secure a greater place in environmental engineering, and will continue to be utilized and improved upon in the future.

## 1.2 Fundamentals of Anaerobic Digestion

The anaerobic treatment of complex organic matter is essentially a three stage process. As Figure 1.1 illustrates, the first stage involves a group of fermentative bacteria which hydrolyze long chain organics to soluble organics, including fatty acids and alcohols. In the second stage, the degradation of propionate and longer chain fatty acids to acetate, hydrogen gas and carbon dioxide is accomplished by a group of acetogenic bacteria. Finally, a group of strictly anaerobic, methanogenic bacteria utilize the acetate as a substrate (along with formate, methanol, and hydrogen gas) to produce methane and carbon dioxide.

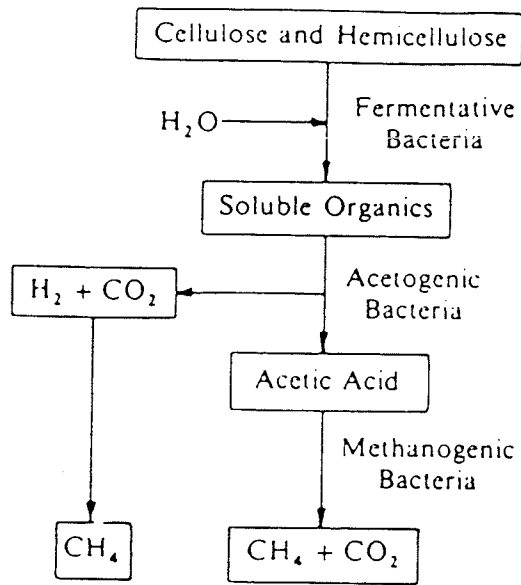


Figure 1.1 Schematic of the anaerobic process (from Pfeffer, 1980).

It was initially believed that the rate limiting step in the anaerobic treatment process was the conversion of organic acids to methane gas (1). Subsequent research, however, has shown that this is not always the case (2). In the digestion of sewage sludge at 35°C (with solids retention times of greater than 10 days), the rate limiting step has been found to be the hydrolysis of organic solids. Further, the digestion of municipal solid wastes is believed to be limited by the rate of cellulose hydrolysis.

The anaerobic process is governed by a number of system parameters. In order for the process to be efficient, these system parameters must be maintained within specific ranges. They include pH, alkalinity, and temperature. Other important factors in anaerobic treatment are methanogenic inhibition, nutrient requirements, and biokinetic relationships.



### 1.2.1 pH

The methanogenic bacteria in the anaerobic process have an optimum pH of approximately 6.6 to 7.6. The consequence of operating a system outside of these limits is a decrease in process efficiency.

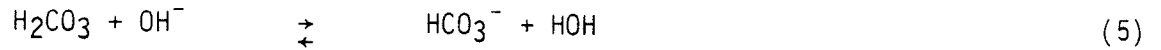
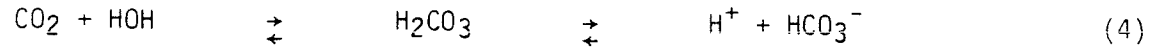
Below a pH of 6.6, the inhibition of the methanogenic bacteria takes place due to an increase in free hydrogen ion concentration. The result is a reduction in methane production. Production of volatile fatty acids continues, however, and when the system reaches a pH of approximately 4.5, methane production ceases. This situation can be averted by maintaining the pH within the optimum range through the addition of alkalinity.

At pH levels above 7.6, the volatile acids in the system are converted into their salts. The methanogenic bacteria are not capable of utilizing these salts and methanogenic inhibition occurs once again.

### 1.2.2 Alkalinity

The term alkalinity refers to the acid-neutralizing capacity of water (3). This parameter is particularly important in anaerobic treatment because the methanogenic bacteria, as described earlier, utilize acetate as a substrate. When the concentration of volatile acids in a system is low, the bicarbonate alkalinity generally represents the total alkalinity. In a system which is maintained and operating properly, the acidic intermediates are converted





When an imbalance occurs in the system, the methanogenic bacteria do not convert the substrate as rapidly as required, resulting in an increase in the volatile acids concentration. This, in turn, leads to a drop in pH, as mentioned in the previous section. To prevent this, the alkalinity in the system should be maintained between 2500 and 5000 mg/L (2).

### 1.2.3 Temperature

For every 10°C rise in temperature, the rate of microbiological reactions is approximately doubled (4). Therefore, higher temperatures will normally yield more efficient operation of anaerobic systems at the same hydraulic retention times.

Bacteria are generally classified into one of three temperature related categories: psychrophilic; mesophilic; or thermophilic (4). These categories and the temperature ranges to which they correspond are listed in Table 1.1.

TABLE 1.1  
Temperature Ranges for Various Bacteria

Type	Temperature (°C)	
	Range	Optimum
Psychrophilic	-2 to 30	12 to 18
Mesophilic	20 to 45	25 to 40
Thermophilic	45 to 75	55 to 65

Although it seems that thermophilic bacteria are most suited for anaerobic digestion because of their high optimum temperatures, this has been found not to be true. In fact, processes operating at thermophilic temperatures are generally uneconomical because of the large quantities of methane required to maintain the high temperatures. For this reason, most anaerobic processes operate in the mesophilic temperature range.

#### 1.2.4 Methanogenic Inhibition

Methanogenic bacteria are susceptible to a number of conditions which can limit or prevent their growth. One of these conditions is an excessively low or high pH level. Another is the presence of oxygen, even in minute quantities. This is because the methanogenic bacteria are obligate anaerobes and require a highly reduced environment for optimum growth. Thus, inhibition of the methanogenic bacteria occurs when they are exposed to any oxidized compounds such as nitrates and nitrites.

Toxicity may result in the anaerobic process when the methanogenic bacteria are exposed to certain materials in solution at threshold concentrations. These include sulfides, heavy metals, alkali and alkaline earth-metal salts, and toxic organics. A concentration of 1700 mg/L of  $\text{NH}_3\text{-N}$  has been found to be the threshold toxicity level for methane production. In the anaerobic process, the total ammonia nitrogen exists in two forms,  $\text{NH}_4^+$  ion and free  $\text{NH}_3$  form, the latter being toxic to methanogens at concentrations exceeding 150 mg/L (5).

Methanogenic bacteria can be acclimated (to a varying degree) to virtually any toxic or inhibitory material over a period of time. Other means of controlling toxicity or inhibition in a waste stream include:

- (a) adding a material that is antagonistic to the undesired material;
- (b) removing the undesired material by methods such as chemical precipitation; and
- (c) diluting the waste stream such that the concentration of the undesired material is below the toxic threshold.

### 1.2.5 Nutrient Requirements

The most important nutrients for anaerobic treatment are those that are required in the greatest concentrations - nitrogen and phosphorous. A variety of other elements including iron, magnesium, potassium, and calcium must also be present for the process to operate, but in smaller

concentrations. These nutrients are compulsory for biological growth and poor effluent quality is the result of their absence.

Based on the assumption that the average chemical composition of a biological cell is  $C_5H_9O_3N$  (2), the theoretical amount of nitrogen required for sludge growth in the anaerobic process can be calculated to be approximately 11% of the volatile solids weight of the cell. The phosphorous needed is about one fifth of this quantity or approximately 2% of the volatile solids weight of the cell.

Such an assumption, however, may result in inadequate nutrient quantities for efficient operation of the process. This is because under high organic loads, the first two stages of the process continue while the methanogenic bacteria stabilize the waste. The fermentative and acetogenic bacteria continually require the essential nutrients during this time. Nutrient requirements should, therefore, be based on the actual removal of the waste from the system as opposed to the load on the system. The minimum carbon to nitrogen to phosphorous ratio for the anaerobic process to proceed has been estimated at 100:6:1 (6).

### 1.2.6 Kinetic Models

An understanding of the basic kinetic models of microbial growth is essential for the efficient design and operation of any microbiological system. The majority of kinetic models are based on the assumption that the

growth rate of microorganisms is based upon some limiting substrate. This is illustrated in Figure 1.2.

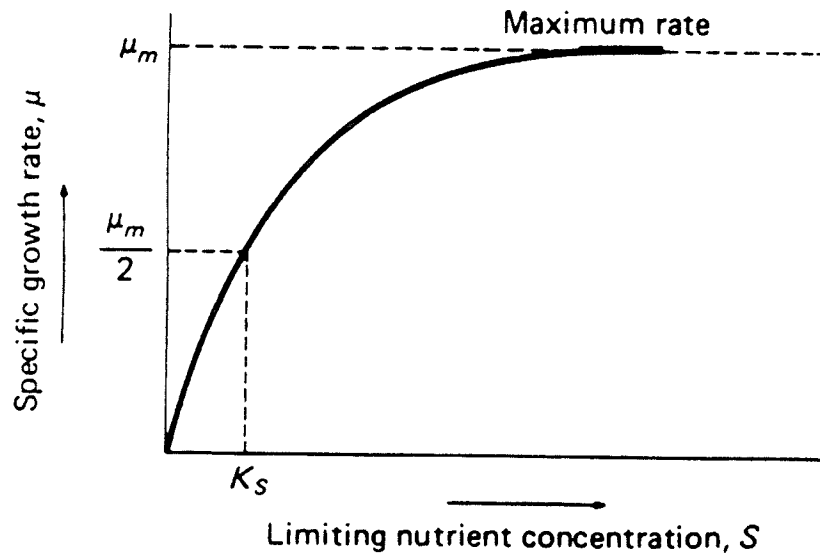


Figure 1.2 Effect of limiting substrate on growth rate (from Metcalf and Eddy, 1979).

From that empirical representation, Monod proposed the following expression for bacterial growth:

$$r_g = \frac{\mu_m X S}{K_s + S} \quad (6)$$

where  $r_g$  = growth rate,  $M T^{-1} L^{-3}$

$\mu_m$  = maximum specific growth rate,  $T^{-1}$

$X$  = concentration of microorganisms,  $M L^{-3}$

$S$  = concentration of limiting substrate in solution,  $M L^{-3}$

$K_S$  = substrate concentration at half of the maximum growth rate,  $M L^{-3}$

The net change in the concentration of microorganisms is a function of cell growth through substrate utilization and cell loss through endogenous decay. Thus, the net rate of bacterial growth may be described as follows:

$$r_n = \frac{\mu_m X S}{K_S + S} - k_d X \quad (7)$$

where  $k_d$  = endogenous decay coefficient,  $T^{-1}$ ,

or

$$r_n = -Y r_s - k_d X \quad (8)$$

where  $Y$  = growth yield coefficient,  $M M^{-1}$

$r_s$  = rate of substrate utilization,  $M T^{-1} L^{-3}$ .

Under steady state conditions it can be assumed that there is no change in the concentration of the microorganisms in the system. A mass balance can then be conducted on the system in Figure 1.3(a) and it can be found that:

$$\frac{Q}{V} = \frac{1}{\theta} = \frac{\mu_m S}{K_S + S} - k_d \quad (9)$$

where  $\theta = V/Q$  = hydraulic retention time,  $T$ .



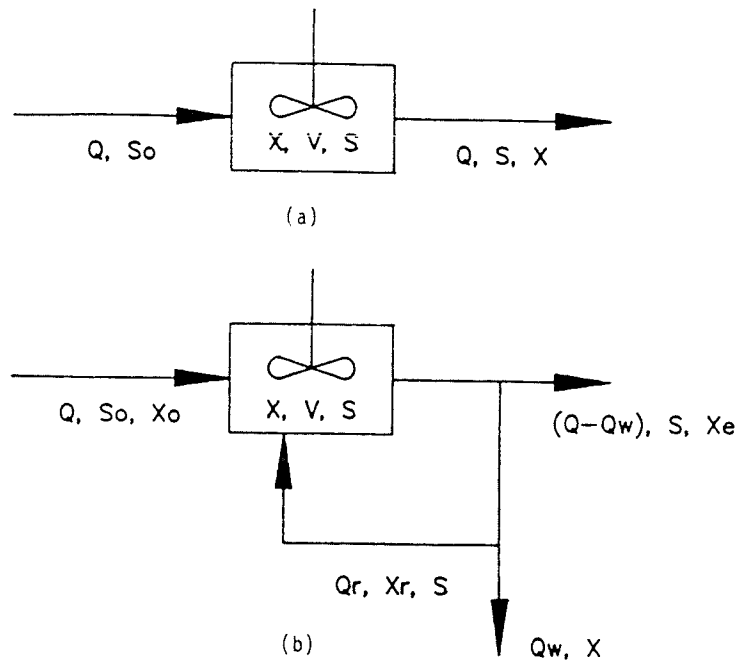


Figure 1.3 Schematic flow diagram of system (a) without recycle and (b) with recycle.

In a system with no recycle, the solids residence time,  $\theta_c$ , can be determined if both  $Q$  and  $V$  are multiplied by  $X$ , the concentration of microorganisms in the system:

$$\theta_c = \frac{VX \text{ (mass of cells in system)}}{QX \text{ (mass of cells wasted daily)}} \quad (10)$$

The term  $r_s$  can now be defined as:

$$r_s = -\frac{Q}{V} (S_o - S) = -\frac{S_o - S}{\theta} \quad (11)$$

where  $S_o - S$  = concentration of substrate utilized,  $M L^{-3}$ .

The specific substrate utilization rate can then be calculated as:

$$U = \frac{S_0 - S}{\theta X} = \frac{Q}{V} \frac{(S_0 - S)}{X} \quad (12)$$

and the solids retention time may be expressed as:

$$\frac{1}{\theta_c} = YU - k_d \quad (13)$$

The Food to Microorganism ratio, F/M, a term used to describe the organic loading on a system can be determined from the following expression:

$$F/M = \frac{S_0}{\theta X} \quad (14)$$

where  $S_0$  = the influent substrate concentration,  $M L^{-3}$ .

This ratio can be related to the specific utilization rate as follows:

$$U = F/M \frac{(S_0 - S)}{S_0} \quad (15)$$

A materials mass balance can also be conducted for the system with recycle shown in Figure 1.3(b). The hydraulic retention time for this system is also:

$$\theta = \frac{V}{\bar{Q}} \quad (16)$$

Recalling equation (10), the solids retention time can be expressed as:

$$\theta_c = \frac{VX}{Q_wX + (Q - Q_w)X_e} \quad (17)$$

where  $Q_w$  = flow rate of liquid stream with cells to be wasted,  $L^3 T^{-1}$

$X_e$  = effluent microorganism concentration,  $M L^{-3}$ .

Assuming steady-state conditions prevail and the influent concentration of microorganisms is equal to zero, the following expression may be derived for biological growth:

$$\frac{Q_wX + (Q - Q_w)X_e}{VX} = -\frac{Yr_s}{X} - k_d \quad (18)$$

This can be simplified using equation (17) to:

$$\frac{1}{\theta_c} = -\frac{Yr_s}{X} - k_d \quad (19)$$

or

$$\frac{1}{\theta_c} = YU - k_d \quad (20)$$

as in equation (13).

By varying the mass balance according to the parameters of the system, these relationships can be determined for any situation. Once established, they are instrumental in monitoring the performance and operation of the system.

### 1.3 Flow Characteristics

A reactor may be defined as any vessel, container or tank in which a biological or chemical reaction takes place. In terms of flow characteristics, there are three general classifications of reactors. These are:

- (a) the Plug Flow reactor in which each fluid particle that enters the reactor stays in the same longitudinal position with respect to the other fluid particles that enter the reactor for a period of time equivalent to the hydraulic retention time. There is no longitudinal dispersion of the fluid;
- (b) the Completely-Mixed Stirred-Tank reactor in which all the fluid particles that enter the reactor are immediately dissipated throughout it. They leave the reactor in accordance with their initial numbers; and
- (c) the Arbitrary Flow reactor in which the fluid particles are dispersed to any degree between the plug flow and the completely mixed reactors.

In practice, it is impossible to achieve true plug flow or complete mixing in a reactor. Generally, however, during operation the flow conditions are close enough to either situation for the purpose of design or analysis. The degree of mixing in a reactor may be estimated by developing a tracer-response curve and determining the liquid dispersion based upon a suitable model.

### 1.3.1 Tracer-Response Curves

Tracer-response curves are used to help establish the nature of the hydraulic regime in various vessels, including anaerobic reactors. A number of parameters can be established through these curves which can be of great use in the design and operation of the Anhybrid reactor. They include the following:

- (a) the time to first appearance (the time at which the tracer first appears in the reactor effluent);
- (b) the modal time (the time at which the greatest effluent tracer concentration is recorded); and
- (c) the dispersion number (a measure of the longitudinal dispersion of the tracer as it travels through the reactor).

The material used as a tracer for the purpose of hydraulic testing depends upon the ease with which it can be detected and the situation in which it is required (6). The tracer may be a dye, a fluorescent compound, a radioactive isotope, or any nonreactive, detectable chemical. The tracer may be applied as either a step or impulse disturbance.

In the case of the step disturbance, a tracer of constant concentration is injected continuously into the reactor and the effluent is monitored at regular intervals to determine the tracer concentration. The data collected is then used to plot a tracer-response curve. The curves obtained when a

step disturbance is applied to a plug flow, a completely mixed, and an arbitrary flow reactor are shown in Figure 1.4(a).

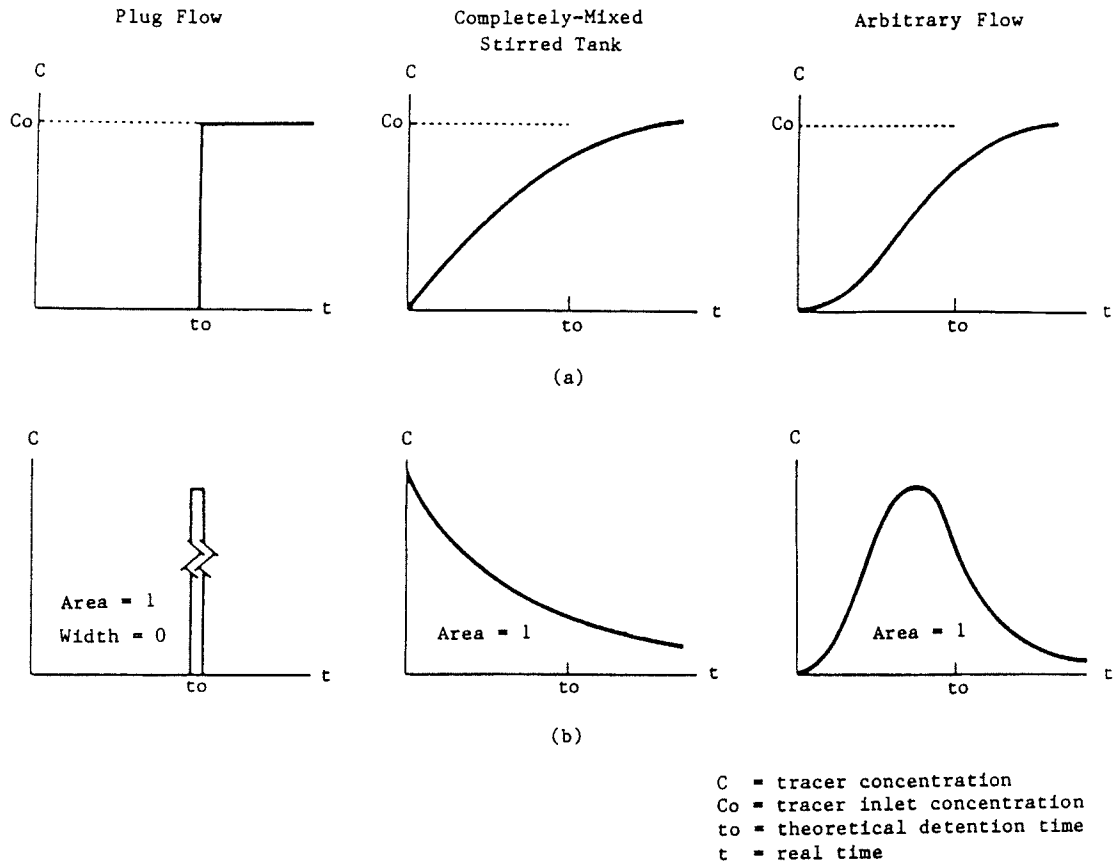


Figure 1.4 Tracer-response curves for (a) step disturbance and (b) impulse disturbance (from Levenspiel, 1972).

If a slug of tracer is injected, an impulse disturbance is created. The effluent is monitored at regular intervals and the tracer-response curve is then plotted. Figure 1.4(b) shows the curves obtained when such a disturbance is applied to a plug flow, a completely-mixed, and an arbitrary flow reactor.

The theoretical effluent concentrations for the plug flow and completely-mixed reactors may also be calculated as a function of time by conducting a materials mass balance around the system.

### 1.3.2 Dispersion Models

A very useful parameter that can be used to help delineate the characteristics of the fluid flow in a reactor is the dispersion number. The dispersion number may be estimated by comparing the shape of the tracer-response curve obtained through testing to the ideal curves shown in Figure 1.4. A far more precise method, however, is to determine the variance of the resultant curve and then relate it to the dispersion number.

Van der Laan has suggested that such a relationship exists for a closed vessel of finite length with no diffusion across its boundaries (6). For a test monitored at regular time intervals, "t", the concentration "C", of tracer in the effluent is determined. After the desired time has elapsed, (usually 4 to 7 hydraulic retention times), the mean time of the concentration curve can be calculated as follows:

$$t = \frac{\sum tC}{\sum C} \quad (21)$$

The standard deviation of the curve is:

$$\sigma_t^2 = \frac{\sum t^2 C}{\sum C} - \bar{t}^2 \quad (22)$$

The variance in terms of dimensionless time can then be calculated as:

$$\sigma^2 = \frac{\sigma_t^2}{\bar{t}^2} \quad (23)$$

The variance can then be related to the dispersion as shown in the following equation:

$$\sigma^2 = 2(d/ul) - 2(d/ul)^2(1 - e^{-ul/d}) \quad (24)$$

where  $d$  = longitudinal dispersion coefficient,  $L^2 T^{-1}$

$u$  = mean flow velocity along reactor,  $L T^{-1}$

$l$  = length of path of travel in reactor,  $L$ .

The term  $d/ul$  is collectively referred to as the dispersion number. For a plug flow reactor the theoretical dispersion number is zero. For a completely mixed reactor the theoretical dispersion number is infinity.



CHAPTER 2  
LITERATURE REVIEW PERTINENT TO  
ADVANCED ANAEROBIC REACTORS

2.1 Development of Advanced Anaerobic Systems

Anaerobic processes, as previously mentioned, have been used sporadically in wastewater treatment for decades. Successful application of anaerobic treatment has been documented in papers dated as early as 1939 by Buswell (9). The last decade, however, has seen a marked rise in interest in anaerobic digestion as a wastewater treatment alternative. According to Oleszkiewicz (10), this increase in interest is due to a number of environmental and economic factors. Among these are:

- (a) new requirements to conserve water resulting in more concentrated wastewaters which are well suited to anaerobic treatment;
- (b) rising energy costs that necessitate efficient methods of treatment;
- (c) the ease with which anaerobic treatment may be applied to complex organic compounds; and
- (d) the energy independence that may be achieved with anaerobic treatment through the recovery of biogas.

Recognition of the benefits of anaerobic treatment has prompted a great deal of research with regard to reactor configuration. As reported by van den Berg and Kennedy (11) and Olthof et al. (12, 13), this research has led to the development of a number of advanced anaerobic reactors in an attempt to

optimize the treatment of a large variety of wastewaters. Among the more notable of these are the anaerobic contact reactor (Ancont), the anaerobic filter (Anbiof), the upflow anaerobic sludge-bed reactor (UASB), the fluidized bed biofilter (Fanbiof), and the anaerobic hybrid reactor (Anhybrid). Each of these is a retained biomass reactor, the difference between reactors being the manner in which they retain the biomass. Figure 2.1 illustrates the reactors schematically.

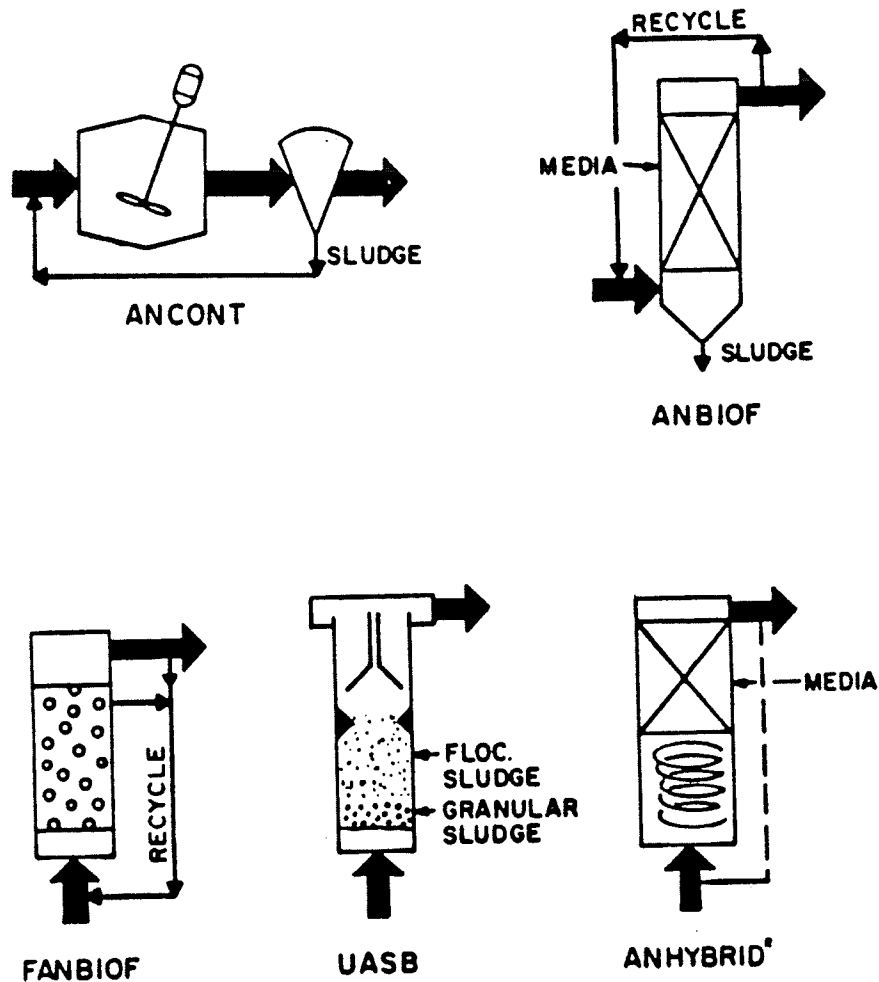


Figure 2.1 Advanced anaerobic reactors (from Olthof et al., 1984).

## 2.2 The Anaerobic Filter

The anaerobic filter is a treatment vessel filled with a fixed, or stationary, bed of porous media. The reactor operates fully submerged, usually in a plug flow hydraulic regime. Waste is passed through the media and is stabilized by biomass within the reactor. The biomass is trapped in the interstitial voids in the media and is also attached to the media itself. For this reason the anaerobic filter is referred to as a "fixed-film" process.

The attachment of biomass to the media enables the anaerobic filter to maintain long solids retention times. A high treatment efficiency is achieved without settling and recycling effluent solids and solids separation is not required. There is a low synthesis of biomass resulting in low nutrient requirements and the minimization of sludge disposal problems.

### 2.2.1 Studies on Media Effects

The anaerobic filter was first investigated in a pilot-scale study by Coulter et al. (14) and subsequently developed by Young and McCarty (15) in systematic laboratory studies. In their research, Young and McCarty employed anaerobic filters constructed of plexiglass 1.83 m in height with a diameter of 15.2 cm. Smooth, quartzite stone 2.5 to 3.8 cm in diameter was used as a media to retain biomass and to provide a surface for the biomass

to attach itself. The filters, when completed, had a porosity of 0.42 and a liquid volume of 12 L.

The effect of the media on the performance of the filters was evident immediately. Biomass developed on the surface of the media and was also trapped in the interstitial voids in the media. In addition, the media provided a mechanism that separated the solids from the gas produced in the system.

The filters were fed a volatile acid and a protein-carbohydrate waste. They exhibited removal efficiencies of up to 80% at organic load rates as high as 3.4 kg COD/m<sup>3</sup>.day with low effluent suspended solids concentrations. However, there was an increase in solids in the filters due to biological synthesis. When the filters became filled with this highly concentrated biomass, a sudden rise in the effluent suspended solids was observed. Thus, while the ability of media to retain biomass in the anaerobic filter is a great benefit, the accumulation of biomass is a major drawback in terms of the hydraulic short-circuiting it can cause and the solids carryover that results.

Since the pioneering work by Young and McCarty, numerous laboratory and pilot-scale studies have been conducted on the anaerobic filter to determine its applicability to various waste streams (16). The effects of temperature, retention time, and more recently, the effects of packing media on filter performance have also been examined (17,18).

A study by Mueller and Mancini (19) directed toward the biokinetics in anaerobic filters had interesting results with regard to media type. Their experimental apparatus consisted of two deep cast acrylic filters with a height of 1.98 m and a diameter of 12.7 cm. Polypropylene Pall rings with a 1.6 cm diameter were placed in the units to a height of 1.2 m. The filters had a final liquid volume of 13.1 L and a porosity of 0.85.

Mueller and Mancini concluded that lightweight plastic media has a distinct advantage over rock media because it has a greater specific surface area (allowing more biological growth per unit volume) and the ability to change shape. The maximum substrate removal for the filters in the study was 17.2 kg COD/m<sup>3</sup>.day. Although this is a relatively high removal rate, it was noted that plugging and solids carryover at high organic loads is a problem that requires further consideration.

In their research, Chian and DeWalle (20) recognized the fact that in a plug flow reactor, such as an anaerobic filter, there is an initial decrease in pH in the direction of flow as a result of acid fermentation. A gradual increase in pH then takes place as the biomass degrades the fatty acids generated. Generally, large quantities of buffer solution are added to the influent to combat this decrease in pH. Chian and DeWalle reasoned that a completely mixed reactor does not experience such a shift in pH because the hydraulic regime maintains a uniform pH level throughout the reactor. They proposed to demonstrate this by providing an anaerobic filter with an effluent recycle which would dilute the incoming waste stream and raise its pH.

The reactors used in their study were constructed of plexiglass with a height of 2.46 m and an interior diameter of 18.7 cm. A recirculation surge vessel separated the effluent and recycled flow streams. The media used in the filter consisted of plastic "Surfpac" (Dow Chemical) slabs with plastic strips between sheets. The specific surface area of the material was 206 m<sup>2</sup>/m<sup>3</sup>. The porosity of the filter was 0.94 with a liquid volume of 13.2 L.

Landfill leachate, a high strength wastewater which provided an influent with reasonably high alkalinity, was used as feed. The filter was operated at a hydraulic retention time of 42 days with a recycle ratio of 20:1. It was assumed that this recycle rate resulted in the unit being completely mixed every 1.8 days. The time required for mixing was much shorter than the hydraulic retention time (1.8 days as compared to 42 days) and the hydraulic regime in the filter was, therefore, considered completely mixed.

Chian and DeWalle found that it was possible to operate an anaerobic filter as a completely mixed unit, given the proper media and recycle ratio. Organic loads of 7.0 kg COD/m<sup>3</sup>.day were treated with 89% removal efficiency. Buffer solutions were not required to neutralize the influent and the filter was also capable of dealing with increases in organic loads and shock loads.

Hudson et al. (21) used two bench-scale anaerobic filters to compare granite stones to oyster shells as a media. Their columns were 1.53 m in height with a 15.2 cm diameter. The granite stone media had a porosity of 0.53

while that of the oyster shell media was 0.82. Both units were operated at similar organic loads and hydraulic retention times.

Their results indicated that the filter packed with the oyster shell media was capable of achieving COD removals 20% to 50% higher than the filter packed with granite stone. It was believed that this was due to the high specific surface area (estimated to be twice that of the granite stone) of the oyster shell media.

Van den Berg and Lentz (22) tested anaerobic filters in the upflow and downflow mode. It is their contention that upflow anaerobic filters operate as a combination of fixed-film and upflow sludge-bed reactors because there is a great deal of biological activity in the interstitial voids in the media. This is particularly true in the lower portion of the column. Downflow filters, on the other hand, tend to function exclusively as fixed-film reactors.

A number of researchers have examined the effects of implementing surface active material, such as activated carbon, in anaerobic filters (23-25). Khan et al. (25) used two anaerobic filters to compare the performance of granular activated carbon to anthracite as packing material. Their reactors were 0.61 m in height with a 5.1 cm diameter. The porosities and specific areas of the media were not recorded. A high recycle ratio of 25:1 was utilized to ensure a completely mixed hydraulic regime.

In all phases of their study, they found that the granular activated carbon surpassed the anthracite in terms of substrate removal, methane production, and biomass retention. The specific reasons for the superior performance of the activated carbon were not identified in the study. It is suspected, however, that it was due to fluctuations in substrate concentration and the ability of the activated carbon to retain biomass.

The question of media-related design criteria has been addressed by Dahab and Young (26) in a series of investigations beginning in 1980. Their studies involved a number of different types and sizes of media with laboratory-scale reactors 2 m in height and 0.5 m in diameter. The reactors were fed a synthetic alcohol stillage at organic load rates of 0.5 to 16 kg COD/m<sup>3</sup>.day. The media and their characteristics are tabulated in Table 2.1.

From an analysis of COD and suspended solids profiles throughout the height of the filters, Dahab and Young found that most of the COD removal took place in the lower one-third of the reactor. There was a very high biomass concentration (up to 60 g/L) in this portion of the filters, one-half to two-thirds of which was not attached to the media. Virtually all of the biomass in the upper one-third of the filters constituted attached growth.



TABLE 2.1  
 Characteristics of Media Tested by Dahab and Young (26)

Media Type	Porosity	Specific Surface Area (m <sup>2</sup> /m <sup>3</sup> )	Average Pore Diameter (mm)
25-37 mm Quartzite Stone	0.47	20-30	12
90 x 90 mm Pall Rings (Norton Actifil No. 90E)	0.95	102	20
50 x 80 mm large openings Corrugated Sheets (Munters No. 27060)	0.95	98	46
20 x 40 mm small openings Corrugated Sheets (Munters No. 19060)	0.95	138	32

As explained by Dahab and Young, gas scouring in the higher regions of the media results in some sloughing of the attached matter. These solids may be lost in the effluent or in-bed flocculation may occur and they will settle downward. Eventually, the reactor can be expected to fill with biomass and plug up. The reactor must, therefore, be designed, and the media selected, to ensure that these solids can be removed by flushing or gravity drainage.

All of the filters in the study had similar removal efficiencies indicating that unit surface area does not play a large role in COD removal or solids distribution. This is believed to be because the greatest biological activity takes place in the interstitial voids in the media.

Porosity did not appear to alter the reactor performance. Pore size, however, did seem to affect the removal efficiencies, particularly in the case of the corrugated plastic media. The large, corrugated plastic media with the greatest pore size had the highest removal efficiency, followed by small, corrugated plastic media and then the Pall rings. The two factors most probably responsible for the differences in performance are channeling and solids accumulation.

The anaerobic filter has been tested extensively in Galway, Ireland (27-29). Barry and Colleran (27) compared the performance of four media types (fired clay, coral, mussel shell, and plastic) in different reactors, each with a total void volume of 21.3 L. The characteristics of the media in their reactors are listed in Table 2.2.

TABLE 2.2  
Characteristics of Media Tested by Barry and Colleran (27)

Media Type	Porosity	Specific Surface Area (m <sup>2</sup> /m <sup>3</sup> )	Active Filter Liquid Volume (L)
3.8 x 2.5 cm Fired Clay	0.69	119	12.42
1.5 cm length 0.2 cm diameter Coral	0.71	490	12.78
6 x 2 cm Mussel Shells	0.80	161	14.40
3.8 - 5 cm diameter Plastic	0.94	179	16.92

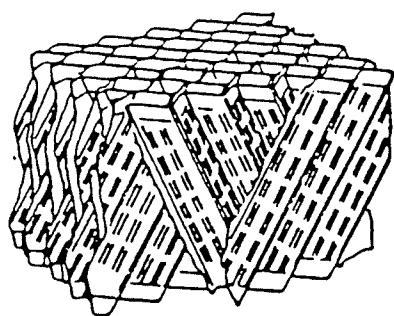
While all four filters displayed similar performance, the clay media appeared to be marginally superior in terms of COD removal, methane production, and COD yield. The mussel shell media yielded the poorest treatment efficiency.

These results are in agreement with Dahab and Young (26), who reported that the specific surface area of the media does not affect reactor performance. In their study however, Barry and Colleran, however, could not correlate porosity to treatment efficiency.

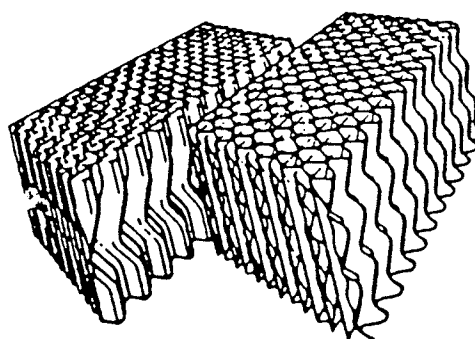
Wilkie, et al. (29) state that the superior performance of the clay media in the previous study was not related to high porosity or specific surface area. Pore size and media alignment may affect reactor performance, but in the case of the clay media, surface roughness and physicochemical interactions such as electrostatic attraction or leaching of essential inorganic nutrients may be more important factors.

In one of the most thorough studies conducted, Song and Young (30) assessed the importance of horizontal alignment and specific surface area of the media in anaerobic filters. The reactors were large, laboratory-scale reactors with a height of 1.83 m and a diameter of 0.5 m resulting in a total liquid volume of 37 L. A synthetic wastewater was used as a substrate.

The research was carried out in three phases. In the first phase, the effect of the specific area of the media on reactor performance was examined. The second phase attempted to find a correlation between the horizontal alignment of the media and reactor performance. The third phase was identical to the second, but the specific surface area of the media was increased. In all of the phases, three of the reactors were packed with cross-flow, corrugated, plastic media and one with plastic, tubular media as a further comparison. The packing media are shown in Figure 2.2. Their characteristics are listed in Table 2.3.



Cross-Flow



Tubular

Figure 2.2 Illustration of plastic, modular media (Song and Young, 1982).

TABLE 2.3

Characteristics of Media Tested by Song and Young (30)

Reactor Number	Media	Porosity	Specific Surface Area (m <sup>2</sup> /m <sup>3</sup> )	Channel Slope (degrees)
Phase 1				
1	Cross-flow	0.97	98	60
2	Cross-flow	0.95	138	60
3	Cross-flow	0.93	223	60
4	Tubular	0.97	98	45
Phase 2				
1	Tubular	0.97	98	90
2	Cross-flow	0.97	98	67.5
3	Cross-flow	0.97	98	45
4	Cross-flow	0.97	98	22.5
Phase 3				
1	Tubular	0.93	223	90
2	Cross-flow	0.93	223	67.5
3	Cross-flow	0.93	223	45
4	Cross-flow	0.93	223	22.5

The results of the first phase showed that the performance of the anaerobic filters was only slightly affected by the specific surface area of the media. The cross-flow media had a much better treatment efficiency than the tubular media with the same specific surface area. This led Song and Young to postulate that the ability of a media to redistribute flow when plugging occurs is one of the most important media design factors.

The horizontal alignment of the media had a significant effect on treatment efficiency. The cross-flow, plastic media with the interstitial channels placed at a slope of  $22.5^\circ$  provided the best COD removal. The difference, however, between that and the removal efficiency of the media oriented at  $45^\circ$  would probably be negated by long term plugging. Song and Young believed that when plugging and structural integrity of the media are considered, the optimum channel slope most likely lies between  $45^\circ$  and  $60^\circ$ .

The results show that pore size, or size of the interstitial openings in the interstitial channels, is also related to the potential for plugging in the cross-flow media. The number of intersections, or contact points, appears to affect performance, but only to a small degree.

It is apparent that the choice of media in an anaerobic filter is of great importance because of the effect it can have on reactor performance and stability. This factor must be taken into consideration during the design phase. However, the drawbacks that accompany the use of media must also be recognized. The most serious of these are (31-33):

- (a) channeling and hydraulic short-circuiting occurring as a result of gas bubbles rising through a limited number of channels;
- (b) mixing of the sludge is not possible or, at best, hampered;
- (c) a large portion of the total reactor volume may be lost to the volume of the media, depending on the media used;
- (d) treating wastes with high concentrations of suspended solids is difficult because of rapid plugging of the media; and

- (e) packing media can be very expensive and an anaerobic filter must be able to treat wastes economically in order to offset the initial investment on media.

## 2.3 The Upflow Anaerobic Sludge-Bed (UASB) Reactor

The UASB reactor operates as a suspended growth process and, therefore, no media is utilized. It is composed of a treatment vessel with a dense sludge bed at the bottom and a sludge blanket that extends up to a gas-solids separator at the top. Waste is stabilized as it passes upward through the sludge-bed and blanket.

The digestion process results in the production of gas bubbles that have a tendency to carry biomass particles with them as they rise. In order to maintain the long solids retention times necessary for effective waste treatment, these particles must be kept in the reactor. This is accomplished by the gas-solids separator at the top of the reactor which allows the effluent to escape, but retains the solids in the reactor.

### 2.3.1 Development of the UASB Reactor

A variation of the UASB concept was first implemented in South Africa where a Dorr-Oliver "Clarigester" was modified to meet the treatment standards for specific industrial effluents (34). The modification involved reversing the flow of the clarigester and pumping the influent into the lower digester compartment. The organic matter in the influent was stabilized by a dense

bed of anaerobic sludge. Sludge displaced by the influent settled in a clarifier within the reactor and was returned to the lower digester compartment.

Since 1971 the UASB concept has been investigated and developed extensively in the Netherlands by Lettinga and his co-workers (31, 33, 35-37). This work has been summarized in a number of papers which provide a great deal of information with regard to the design, performance and operation of UASB reactors.

One of the most interesting features of the UASB reactor discovered by Lettinga et al. (37) is the formation of granules that takes place in the sludge bed. This granulation takes place in three phases (37-39):

- (a) Phase 1 (organic loads of up to 2 kg COD/m<sup>3</sup>.day). In this phase the sludge bed expands because of the increasing hydraulic load and gas production. The solids that are lost in the effluent are primarily colloidal particles;
- (b) Phase 2 (organic loads of between 2 and 5 kg COD/m<sup>3</sup>.day). This phase is distinguished by an increase in the washout of solids, particularly flocculent sludge. The organic load rate rises rapidly, but the volumetric load rate remains unchanged because of the loss of biomass. Granules begin to form; and
- (c) Phase 3 (organic loads above 5 kg COD/m<sup>3</sup>.day). The growth of sludge granules prevails over the flocculent sludge. The sludge concentration



increases in the reactor, thereby allowing an increase in the volumetric load rate.

Most researchers agree that these three phases occur in the process of granulation but the specific conditions required for granulation remain unknown (40-43). Lettinga et al. (35) have suggested several factors that they believe to play a role in the process:

- (a) the presence of sufficient nutrients in the system for bacterial growth and the formation of bonding agents between the bacteria;
- (b) the continuous washout of flocculating sludge from the system under high hydraulic load rates;
- (c) the presence of bivalent cations that aid in flocculation;
- (d) the gentle, vertical agitation caused by gas production that results in the gravity compression of sludges; and
- (e) the specific activity of the seed sludge and the concentration of inert particles in it.

Numerous species of methanogenic bacteria have been found in the sludge granules of UASB reactors. The predominant species depends upon the origin of the seed sludge, the substrate utilized, and the process conditions (38). The granules that develop with acetic acid substrates are composed mainly of the Methanosarcina species. When a reactor that has been seeded with a digested sewage sludge is given a volatile fatty acid mixture as a substrate, the Methanothrix soehngenii species of bacterium prevails. These granules can be either rod-like or filamentous in nature (37).

Callander and Barford (44,45) used polyelectrolytes as flocculating agents in their studies with Upflow Floc digesters (a variation on the UASB concept). They found that the polyelectrolytes enhanced the accumulation of biomass and suspended solids within the reactor. This accelerated the granulation process and allowed the volumetric loads to be increased at a more rapid pace than is normal.

Cail and Barford (46) had similar results when they compared the granulation process in an Upflow Floc digester and a UASB reactor. The addition of polyelectrolytes did allow for more rapid increases in volumetric load, but, once granules were present in both reactors, their performance was virtually indistinguishable.

A number of polymers and elements are being tested for their effect on the development of granules (43). Operating conditions, such as thermophilic temperatures, are also being investigated to increase the applicability of the UASB reactor.

Granular sludge is conducive to the application of high hydraulic loads in the UASB reactor, but flocculent sludge is preferred when dealing with complex wastes with high suspended solids concentrations (41). Insufficient removal of these solids may necessitate the installation of an external filter as well as a separate sludge digester. Thus, flocculent sludge is desirable in a variety of situations because of its inherent ability to remove suspended solids. And, in order for a waste to make proper contact

with a granular sludge, a more sophisticated influent distribution system is required.

The influent distribution must be designed such that the risks of channeling in the reactor are minimized. These risks are greatest when low volumetric loads are applied to the reactor (the gas production is too low for mixing to take place); the height of the sludge bed is too low; the feed is not distributed evenly across the bottom of the reactor; and the sludge has a high settleability (35-37).

The primary difficulty encountered in the operation of the UASB reactor is the escape of biomass from the reactor. The gas-solids separator that Lettinga et al. (35) have developed in response to this problem serves four main purposes:

- (a) the separation of biogas from the effluent and from floating sludge particles;
- (b) the separation of biomass by a combination of settling, flocculation, and entrapment in the sludge bed;
- (c) retention of the sludge in the digester compartment; and
- (d) prevention of expansion of the sludge blanket.

Van der Meer and de Vletter (47) have studied the design and operation of the gas-solids separator in depth. They report that, for the settler to function most efficiently, the gas must be separated from the particulate matter before entering the actual settling compartment. The retention time

in this compartment should be kept to a minimum to avoid the production of biogas which can reduce settling efficiency. A sludge blanket is desirable in the settler because it helps prevent sludge bed expansion, entraps solid particles, and provides a reduction in effluent COD.

A settler that fulfills all of these conditions is shown in Figure 2.3. The gas is separated from the solids which then enter a compartment that allows the escape of gas released by expansion. It also dampens turbulence caused by gas generated in the reactor. The solids are separated from the effluent in the settling compartment where the flow is predominantly laminar. This allows the thickened solids to slide back into the reactor on the bottom plates of the settler. The optimum angle for these plates has been found to be between  $45^{\circ}$  and  $50^{\circ}$  (33, 35-37, 47).

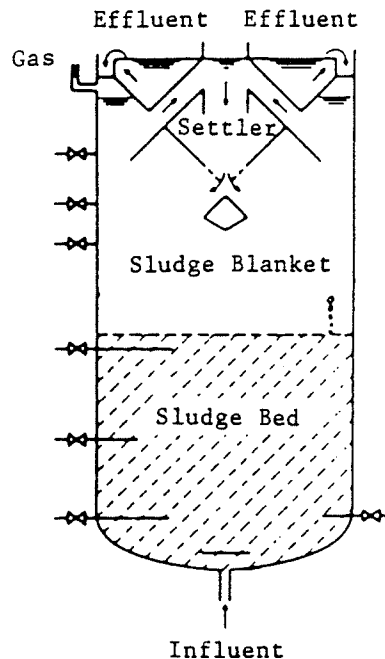


Figure 2.3 UASB reactor with settler (from van der Meer and Vletter, 1982).

When the system is subjected to large organic loads, there is a strong tendency towards flotation in the sludge. Mechanical mixing may then be required at the top of the settler. If the flotation is caused by lipids or fatty acids, a skimmer should be installed in the settler to remove them (37).

### 2.3.2 Hydraulic Characteristics of the UASB Reactor

The reason for modelling the hydraulic characteristics of a reactor are twofold: it allows the prediction of the reaction of the system to changes in operating conditions without performing full-scale experiments; and phenomenon that cannot be measured directly, such as hydraulic short-circuiting and dead spaces, can be estimated (48,49). A number of hydraulic studies have been undertaken to determine the mixing efficiencies and sludge distributions in anaerobic digesters (48-50). However, relatively little work has been done on the hydraulic characteristics of the newer, advanced anaerobic reactors.

In a study by Heertjes and van der Meer (51), three mathematical models were developed to predict the dynamics of the liquid flow in a 6 m<sup>3</sup>, laboratory-scale UASB reactor under various operating conditions. For analytical purposes, the reactor was divided into three parts: the settler, a plug flow region; the sludge blanket, a completely mixed region; and the sludge bed, in which the flow is predominantly completely mixed, but dead zones exist and bypassing may also occur.

The reactor was operated under three different process and reactor conditions. A step disturbance was applied to the reactor during each of the operating conditions. The tracer used was LiCl. The tracer-response curves obtained from these hydraulic tests were then compared to the curves generated by the computer models for the same operating conditions.

Heertjes and van der Meer concluded that the models presented were effective in predicting the dynamic behaviour of the fluid in the reactor and, therefore, their division of the reactor into three parts was reasonable. They also noted that an increase in the gas production of the reactor from 2.7 to 4.5 m<sup>3</sup>/h caused no significant change in the fluid flow pattern. An increase in the height of the sludge bed, however, from 1.2 to 2.2 m resulted in increased bypassing of the sludge by the influent flow stream. It is believed that this was due to compression settling of the sludge and an influent distribution system which was not capable of introducing the influent evenly.

Bolle et al. (52) used a similar mathematical model to predict the fluid flow in a 30 m<sup>3</sup>, pilot-scale UASB reactor. Their model, as in the previous study, divided the reactor into three parts to ease the analysis. They discovered that the short circuiting of flow over the sludge bed increases linearly with height and with increasing gas velocity. There is a possibility that the maximum short circuiting flow is a function of the gas velocity in the UASB reactor.

Hydraulic tests have been conducted at Environment Canada's Wastewater Technology Centre on an UASB reactor and an anaerobic filter (53). The tests were designed to monitor the changes in the internal hydraulics of the reactors due to the accumulation of biomass after long term operation. Specifically, the biomass-associated dead volumes and the flow channeling in the reactors were assessed using tracer-response techniques coupled to non-linear modelling of the fluid flow. Estimates of the active flow, the portion of the influent flow that does not bypass the treatment zone of the reactor, and the active volume, the portion of the reactor that constitutes the treatment zone, were made for both reactors.

In the anaerobic filter, both the active flow and the active volume decreased significantly because of the accumulation of biomass over a three year period. The active volume was reduced by 45%, and 30% of the influent flow bypassed the treatment zone. These hydraulic difficulties were corrected on a regular basis by shutting down the reactor and removing the excess biomass. In full-scale applications, this remedial action is both costly and time-consuming.

During the same period, an equal amount of biomass was measured in the UASB reactor. The active volume in the UASB reactor was usually greater than 90% of the total void volume and the bypass flow averaged only 3% of the influent flow.

The results of these hydraulic tests indicate that the UASB process is far more efficient than the anaerobic filter. When the performance of the UASB

reactor was monitored, however, it was observed that the loss of biomass in the reactor was very rapid, in spite of the presence of a gas-solids separator. This problem has been detailed in a number of other studies and indicates the potential need for external clarification and recycling of solids with the UASB reactor (32, 54).

#### 2.4 The Anaerobic Hybrid (Anhybrid) Reactor

Many researchers have examined the anaerobic filter and the UASB reactor and found that, while both processes are viable, they possess drawbacks inherent in their designs. To minimize these weaknesses, a number of reactors have been designed that incorporate the basic elements of each process (54-56). The resulting hybrid reactors are composed of a treatment vessel with an upflow sludge bed at the bottom and a zone of support media at the top. In theory, this suspended-growth, fixed-film configuration has many benefits because it combines the most desirable features of the UASB reactor and the anaerobic filter.

The hydraulic regime in the hybrid reactor is similar to that of the UASB reactor. The sludge bed is the main treatment zone while the media acts as a gas-solids separator. The entire sludge bed volume of the reactor can, consequently, be considered active and channeling or bypassing of the treatment zone (a major problem in the anaerobic filter) is virtually eliminated. The absence of media also allows for the system to be easily monitored and sludge to be wasted, when necessary.



The potential for the granulation of biomass exists in the hybrid reactor when the physical, chemical, and operational conditions are favourable. In cases where the granulation process is slow (due to unfavourable operating conditions, wastewater characteristics, or poor sludge settleability), the media serves to retain the unattached biomass in the reactor. This is most beneficial during start up and other periods of reactor instability such as transient operating conditions. As a result, the hybrid reactor, like the anaerobic filter, has shorter start up times than the UASB reactor (57).

The biomass that occupies the media zone, both attached and unattached, polishes the wastewater that leaves the sludge bed zone and increases the overall treatment capacity of the hybrid reactor. If a system failure occurs, this biomass is a potential source of seed sludge when the reactor is restarted. As the biomass accumulates in the media zone, channeling and plugging will begin to affect the hydraulic efficiency of the reactor. The extent of the reduction will depend on the type of media used and its volume in relation to the total volume of the reactor.

In the event that the media zone of a hybrid reactor with a well acclimated sludge becomes plugged, the performance of the reactor may be only slightly altered. This situation is temporary because the excess biomass can be removed from the media by draining the reactor until the media is no longer submerged. The process is less time-consuming than with an anaerobic filter because of the smaller volume of media involved.

Commercially available media contributes between 100 and 300 dollars/m<sup>3</sup> (Canadian, 1985) to the capital cost of a high rate anaerobic reactor (58). The capital cost of the hybrid configuration is substantially lower than that of the anaerobic filter because of the 50 to 75% reduction in media volume. This fact alone makes the hybrid reactor a very attractive treatment alternative.

#### 2.4.1 Development of the Anaerobic Hybrid Reactor

The hybrid reactor was conceptualized by Maxham in a study that compared four parallel upflow reactors (57). The reactors were 35.6 cm in height with 3.8 cm diameters. Glass beads with a 6 mm diameter were used as media. Three of the columns were operated as hybrid reactors and one was filled entirely with glass beads to determine the effect of the additional media on reactor performance. The reactors were monitored for their treatment of biomass gasification wastewaters.

The hybrid reactors and the anaerobic filter exhibited similar treatment efficiencies in the study. However, the comparison between reactor types lasted only 7 weeks because the anaerobic filter was taken out of operation. After 12 weeks of operation, all of the reactors were shut down and dismantled. No attached growth was observed in any of the reactors, presumably because of the brief duration of the study. In spite of this, the feasibility of the hybrid configuration was demonstrated as they provided COD removals of up to 80% at organic load rates of 2.3 kg COD/m<sup>3</sup>.day.

The Anhybrid reactor was conceived in 1981 at Duncan, Lagnese and Associates (DLA) in Pittsburgh, Pennsylvania. Pilot studies were conducted with various waste streams and the Anhybrid reactor was subsequently introduced as part of a modular, packaged biological treatment plant for high strength organic wastes (over 1500 mg/L COD). A schematic flow diagram of a treatment train for a chemical wastewater is shown in Figure 2.4.

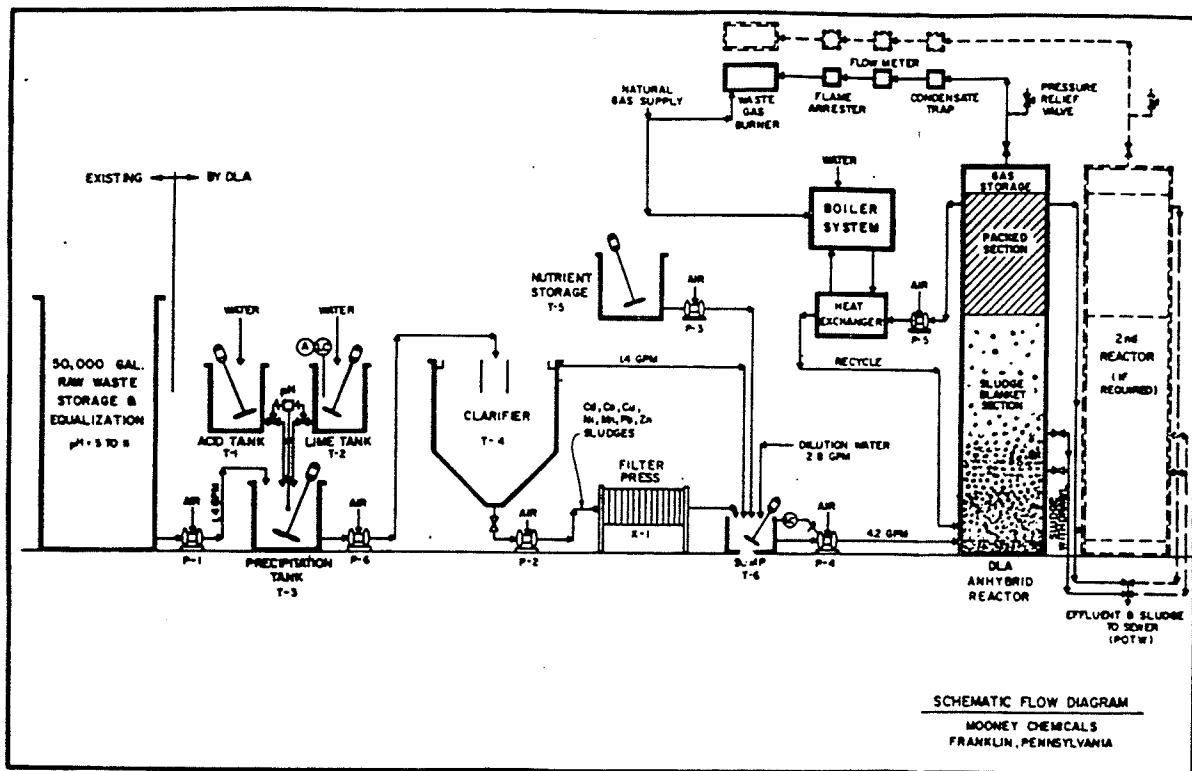


Figure 2.4 Schematic flow diagram of packaged, Anhybrid treatment plant (from DLA, 1982).

Guiot and van den Berg (60-62) used four identical reactors with hybrid configurations to test their effectiveness in treating a soluble sucrose wastewater of varying concentrations. The reactors had a liquid height of 62 cm and an interior diameter of 9.6 cm. The top third of the reactors was filled with 1.5 cm plastic Flexirings (Koch, Inc.). The seed sludge was a combination of flocculent sludge from a UASB reactor treating sugar waste at 35°C and granular sludge from a UASB treatment synthetic acetate waste at 27°C. The characteristics of the reactors and the operating conditions of the study are shown in Table 2.4.

TABLE 2.4

Physical Characteristics and Operational Conditions  
of Hybrid Reactor used by Guiot and van den Berg (60)

Total Reactor Volume (L)	4.25
Media Type	Flexirings (Koch, Inc.)
Number of Rings	250
Ring Density (kg/L)	0.85
Specific Surface Area (m <sup>2</sup> /m <sup>3</sup> )	235
Media Zone Volume (L)	1.41
Media Dead Volume (L)	0.14
Sludge Blanket Volume (L)	2.75
Recycle Ratio	5:4:1
Operating Temperature	27°C
Influent	Synthetic, soluble sucrose waste

An impulse disturbance was applied to the reactors to determine the mixing regime within them. The tracer-response curves obtained from these hydraulic tests were those of a completely-mixed system without dead spaces or hydraulic short circuiting.

Granulation of the sludge in the reactors was rapidly achieved due to the nature of the seed sludge. The granules that formed increased in size with operating time as well as organic load rate. The majority of biomass present in the reactors was granular, but filamentous biomass was found on, and near, the media.

Biomass yields as high as 0.024 g VSS/g COD removed were recorded during the study. Yields observed by other researchers treating the same waste with a different process were significantly lower than this (63).

As expected, the presence of media greatly enhanced reactor performance and the retention of biomass. Volumetric loadings of up to 25 kg COD/m<sup>3</sup>.day were treated with 90% COD removal efficiency. Biomass concentrations of 25 g VSS/L were observed in the media. Volatile suspended solids profiles of the reactors revealed that the media retained the biomass independently of the sludge bed. No major change in the effluent suspended solids concentrations took place in spite of an increase in upflow velocity in the reactors from 1 to 4 m/h, further evidence of the positive effects of the media.

Oziemblo et al. (64) ran a study with four Anhybrid reactors to assess the importance of the media zone volume to total volume ratio in the process. The reactors were constructed of plexiglass tubing with an 11 cm diameter. Each reactor had a sludge bed zone volume of 8.5 L. The media zone volume was varied. Unglazed, 2.5 cm, ceramic Rashig rings were used as media and placed randomly in the media zone. The dimensions of the reactors and the operating conditions of the study are tabulated in Table 2.5.

**TABLE 2.5**  
Physical Characteristics and Operating Conditions  
of Anhybrid Reactors used by Oziemblo (64)

Parameter	Reactor			
	R1	R2	R3	R4
Reactor Height (m)	1.8	1.5	1.2	0.95
Total Reactor Volume (L)	16.1	13.4	10.7	8.5
Media Zone Volume (L)	8.05	5.35	2.68	0.40
Media Zone Vol./Total Vol.	0.50	0.40	0.25	0.05
Recycle Ratio	4.3 to 164.2			
Operating Temperature	35°C			
Influent	Synthetic materials and industrial wastewater			

At organic loads of between 2 and 4 kg COD/m<sup>3</sup>.day all of the reactors provided treatment efficiencies greater than 90%. Increases in the organic load rate had no significant effect on the performance of R1 and R2. They maintained high treatment efficiencies throughout the study. Reactors R3 and R4, in contrast, displayed a progressive deterioration in performance and, by the end of the study, their COD removal efficiencies were below 50%.

This disparity in performance was attributed to the volume of the media zone in the reactors. The reactor with the highest media zone volume to total volume ratio, R1, lost the least amount of sludge during the course of the study. The other reactors lost larger volumes of sludge in proportion to the volume of their media zones.

The F/M ratios in the reactors with lower media zone volume to total volume ratios increased because of the extensive loss of biomass. Near the end of the study, R4 was operating at an F/M ratio of more than 2 kg COD/kg VSS.day. The maximum substrate removal rate in a mixed anaerobic culture at 35°C has been reported to be only 1 kg COD/kg VSS.day (65).

The retention of biomass was cited as a contributing factor in the start up of the reactors. At the end of the study, the quantities of volatile suspended solids in the reactors were proportional to the volumes of the media zones. Visual examination of the Rashig rings in the reactors revealed no attachment of biomass. Biomass was discovered however, in the interstitial voids of the media in all of the reactors. While this biomass

was observed to cause some channeling in the media, its impact on reactor performance was minimal, as attested to by the superior performance of R1.

Among his conclusions, Oziemblo noted that the optimum ratio of volume occupied by the media to the total volume of the Anhybrid reactor lies between 0.25 and 0.40. An analysis of the biomass in the reactors indicated that solids loss from the Anhybrid is proportional to the volume of the media zone.

In an ongoing study by Canviro Consultants Limited (66) at Portage La Prairie, Manitoba, three parallel, 1000L, pilot-scale reactors are being loaded with a carbohydrate influent to examine the effects of different media on their performance. The reactors are operating at a temperature of 35°C under the following conditions:

- (a) an Anhybrid configuration with cross-flow oriented media (Munters);
- (b) an UASB configuration with no media; and
- (c) an Anhybrid configuration with randomly placed 90 mm Norton Pall rings.

The reactor with Munters media has just recently been started, while the other two have been in operation for several months.

The reactor with Munters media was initially seeded with sludge from the existing Portage La Prairie treatment plant. The mass of the total solids in the reactor at the time was 15.7 kg. After 4 weeks, a reduction in the



volatile solids content of the reactor was noted and the reactor was again seeded, this time with a granular sludge previously acclimated to a brewery waste. There has been a continual decrease in the organic content of the reactor in spite of the introduction of the new sludge.

The same difficulty has been encountered in the UASB configuration; a decrease in the total volatile solids content of the reactor has been noted each week and reseeded has been necessary on several occasions. This reactor has also been seeded with granular sludge recently. The results of this action have yet to be realized. It is suspected, however, that a further decrease in the organic content of both reactors will continue because the sludge is not acclimated to the present substrate. This will result in some disintegration of the granules after which, if the required conditions are met, some improvement will take place and the granules will redevelop.

To date, the Anhybrid reactor with the random media has provided the best overall performance. Organic load rates of up to 9 kg COD/m<sup>3</sup>.day have been treated with removal efficiencies of 90%. Biomass retention has not been a problem and the reactor has remained relatively stable throughout the study. The UASB reactor, in contrast, was shut down due to an excessive loss of sludge.

## 2.4.2 Full-Scale Applications

At present, there are very few full-scale hybrid reactors in existence in Canada because the technology is still relatively new. Those that are in operation, however, have met with successful results. At Hardee Farms International in Lambeth, Ontario, a vegetable processing wastewater is being anaerobically treated by two parallel, hybrid reactors (53). Plastic, modular, Munters media has been installed in the media zone at different depths to examine the effect of bed height on reactor performance. The system has been limited by wastewater availability but the system has operated as close to design conditions as possible.

A hybrid facility has been constructed at Lakeview Water Pollution Control Plant in Mississauga, Ontario by Gore & Storrie Limited to anaerobically treat thermal sludge conditioning liquor (67). The construction of the reactors (referred to as the "Hyan" process) involved the modification of two existing digesters by incorporating a 2 m media zone with random, plastic media. Figure 2.5 is a diagram of the Hyan design for the Lakeview Plant.

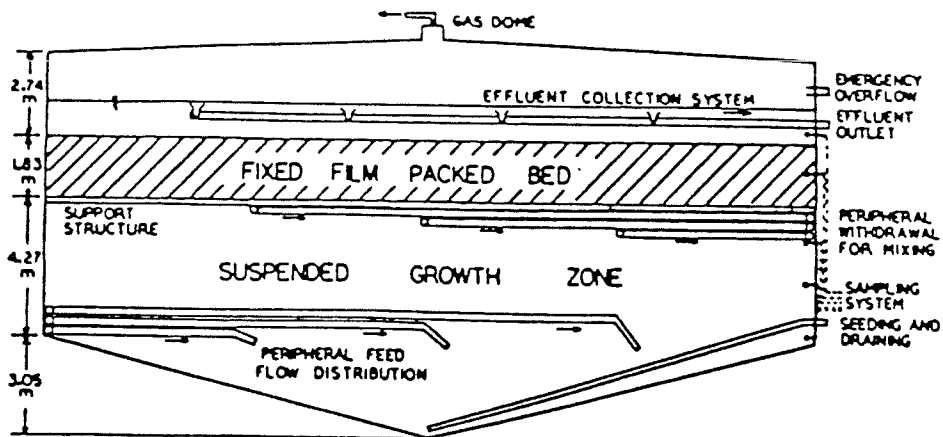


Figure 2.5 Diagram of Hyan reactor at Lakeview Water Pollution Control Plant (from Crawford and Teletzke, 1986).

A number of startup techniques were investigated prior to the actual operation of the reactors. Among these was the use of a target reactor effluent acid level of 2000 mg/L as a means of feed control. The performance of the reactors has exceeded the expected levels of treatment efficiency. At loads of over 6 kg COD/m<sup>3</sup>.day and hydraulic retention times of between 32 and 48 hours, the reactors have consistently achieved COD removals of 72%.

A third hybrid facility is being designed for the treatment of thermal conditioning liquor by Proctor and Redfern for the Highland Creek Treatment Plant in Scarborough, Ontario (68). The reactors will be a converted primary and a modified secondary digester with 15.9 cm Propak biorings placed at random in the media zone. The sludge blanket will occupy approximately 50% of the total reactor height.

## CHAPTER 3

### SCOPE OF STUDY

#### 3.1 Background

The hybrid reactor configuration is a recent development in anaerobic wastewater treatment. Because of this, design information and performance data are limited and design standards have not yet been developed. Extensive laboratory and pilot-scale studies are essential, therefore, prior to any full-scale application.

This research is a continuation of the work previously conducted on the Anhybrid reactor by Oziemblo (64) at the University of Manitoba. In the earlier study, the volume of the media zone in relation to the total volume of the reactor was examined. The results of that research indicated that an optimum ratio between the media zone volume and the total volume of the Anhybrid reactor did exist, and that further examination of media effects on the reactor were warranted.

#### 3.2 Objectives

The general purpose of this research was to provide information on the design and scale-up of the Anhybrid reactor. The specific objective of this study was to quantitatively determine the effect of media type on the performance and operation of the Anhybrid reactor.

### 3.3 Study Approach

A study with four parallel Anhybrid reactors was conducted over a 100 day period. Three different reactor configurations were examined:

- (a) random media (no orientation);
- (b) vertically oriented media; and
- (c) no media.

The study was broken into two stages; an operational stage and a hydraulic stage. During the operational stage, the performance of the reactors equipped with the various media types under increasing organic load rates was monitored. This was followed by an examination of the hydraulic effects of the different media through the application of tracer tests.

## CHAPTER 4

### MATERIALS AND METHODS

#### 4.1 Apparatus

##### 4.1.1 Reactors

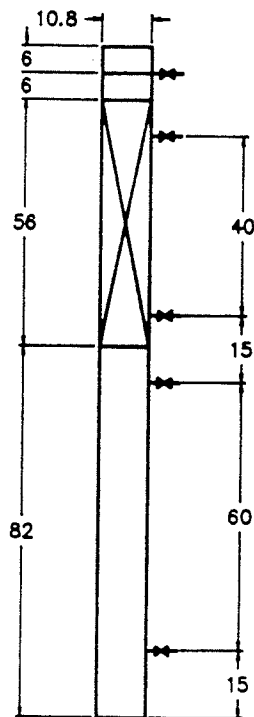
The Anhybrid reactors utilized in this study were constructed of plexiglass columns with an interior diameter of 0.11 m and a height of 1.5 m. Sampling taps were installed in each reactor at heights of 15 cm, 75 cm, 90 cm, and 130 cm.

Reactor 1 (R1) and Reactor 2 (R2) contained randomly oriented, unglazed, 2.5 cm, ceramic Rashig rings as media. Reactor 3 (R3) contained rigid, vertically oriented PVC tubing with an interior diameter of 2.5 cm as media. Reactor 4 (R4) was operated as a UASB reactor and contained no media. Figure 4.1 illustrates the dimensions of the reactors. Table 4.1 lists the physical characteristics of each reactor.

TABLE 4.1

Physical Characteristics of Laboratory-Scale Anhybrid Reactors

Parameter	Reactor			
	R1	R2	R3	R4
Height (m)	1.5	1.5	1.5	1.5
Volume (L)	13.4	13.4	13.4	13.4
Media Zone Volume (L)	5.35	5.35	5.35	-
Media Support	Rashig Rings	Rashig Rings	PVC Tubing	None
Specific Surface Area (m <sup>2</sup> /m <sup>3</sup> )	190	190	65	-
Porosity	0.70	0.70	0.94	1.00



All dimensions in centimeters.

Figure 4.1 Dimensions of laboratory-scale Anhybrid reactors.

The reactors were housed in a walk-in environmental chamber maintained at a temperature of 35°C. The influent, synthetic feed used in the study was prepared and stored in a similar, adjoining environmental chamber maintained at a temperature of 5°C. The second chamber was set at this temperature to minimize the decomposition of the feed during storage. Figure 4.2 shows the reactors in the walk-in environmental chamber.

The influent was pumped from the feed tank into a mixing chamber which combined the raw influent and recycle flows. This combined flow was then pumped into the reactor through a distributor located at the bottom of the reactor.

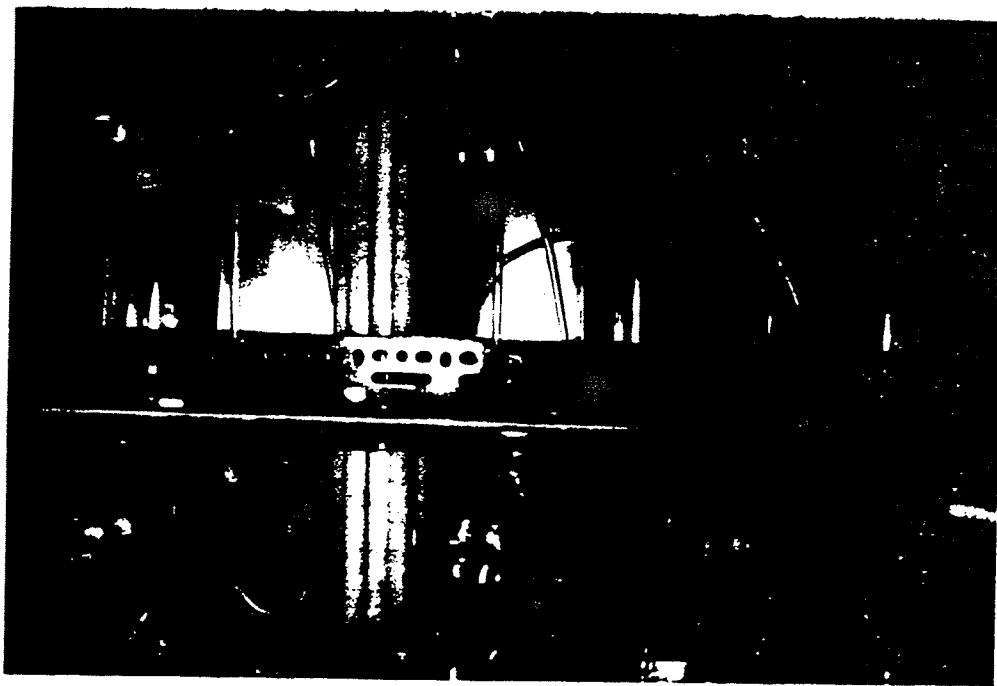


Figure 4.2 Anybrid reactors in walk-in environmental chamber.



The effluent flow was separated from the biogas produced in the reactors by using a splitter box. A thermal shock solids separator was installed in the second environmental chamber to enhance solids removal from the effluent and to measure lost solids. The effluent was stored in a container for later measurement and analyses.

Pumping of the influent and recycle flows was accomplished with Masterflex variable speed pumps (Cole Parmer Instrument Co., Model No. 7553-00). Masterflex Neoprene tubing was used in the pumps. Tygon tubing was used for all other gas and liquid transport lines in the system.

The biogas was directed through a water seal, after which it was measured on a gas meter and released to the atmosphere. Figure 4.3 is a schematic of the experimental apparatus in the laboratory.

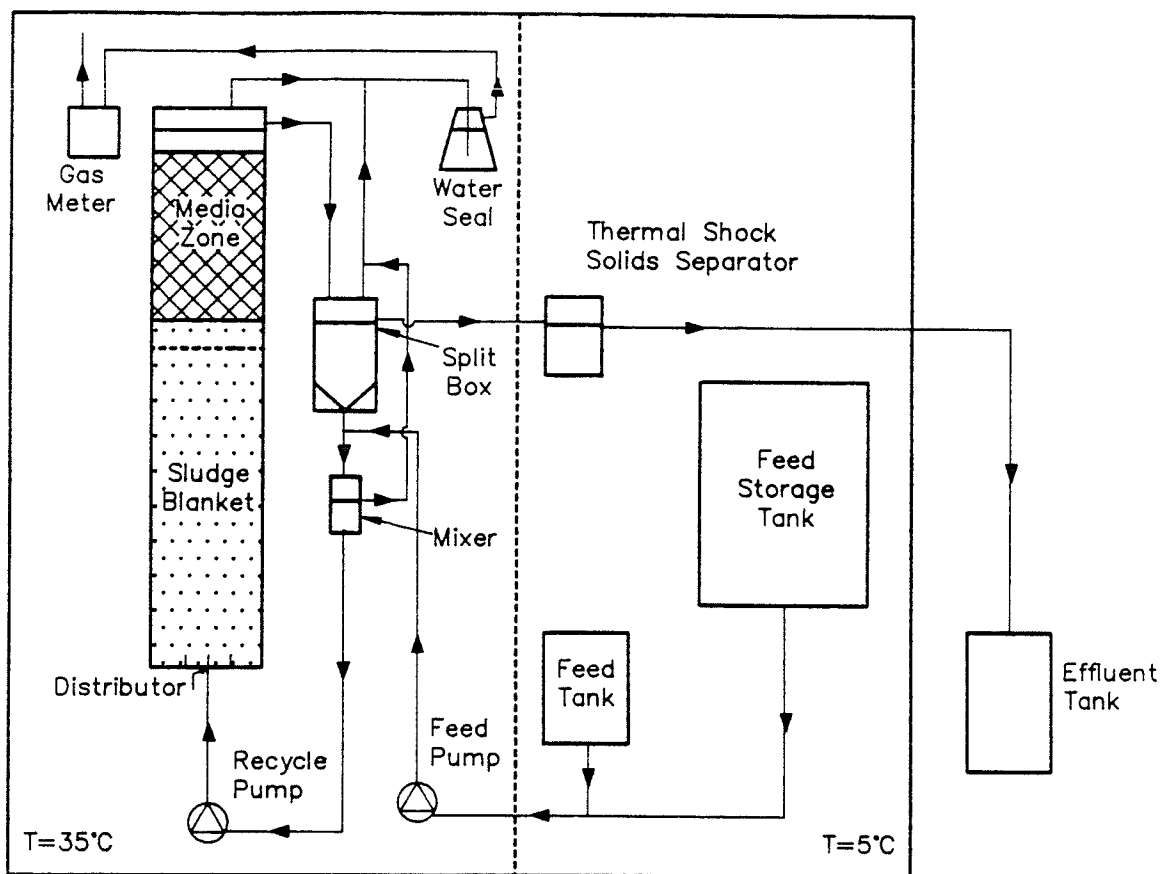


Figure 4.3 Schematic flow diagram of experimental apparatus.

#### 4.1.2 Synthetic Feed

A synthetic sulphite evaporator condensate (SEC) was employed as the feed (67). The feed was prepared in two 100 L containers, as required. The constituents of the feed are listed in Table 4.2. The influent feed also contained a trace metal solution that provided 1 mg/L each of nickel, cobalt, and iron. Sodium carbonate was used to supplement alkalinity.

TABLE 4.2

Composition of Synthetic Sulphite  
Evaporator Condensate (SEC) Feed

Constituent	Concentration (mg/L)
Acetic Acid	2400
Methanol	850
Ethanol	450
Phenol	100
Acetone	100
Ammonium Phosphate (Dibasic)	10
Ammonium Chloride	10
Potassium Phosphate	10
Magnesium Chloride	10
Furfural	10-25*
Sodium Sulphide	10-25*

\* The concentrations of Furfural and Sodium Sulphide were increased during the course of the study to concentrations that could reasonably be expected in actual SEC. This was done to allow the biomass to acclimate to the compounds.

The SEC feed was chosen to simulate an actual industrial waste, and also to ensure that the nutrient ratio of C:N:P (100:6:1) required for anaerobic treatment was maintained. The influent feed concentration was approximately 5500 mg COD/L.

## 4.2 Operational Procedure

### 4.2.1 Start-up and Acclimation

The seed sludge for each reactor was obtained from the Winnipeg North End Treatment Plant. Each reactor was charged with 4.14 L of thickened, actively digesting sludge containing a volatile solids content of approximately 20 g/L. The initial organic load rate was 1 kg COD/m<sup>3</sup>.day at an F/M ratio of approximately 0.1 COD/kg COD/kg VSS.day.

The reactors were operated at a solids retention time of approximately 50 days. Generally, steady state conditions are assumed when consistent operation of a reactor is recorded over a period of two or three solids retention times. However, due to the time constraints involved in this study, it was assumed that pseudo-steady state conditions were achieved after a period of two weeks. That is, the removal data were obtained in four consecutive measurements after which the organic load rate was increased step-wise as shown in Figure 4.4.

Figure 4.4

Load Rate vs Time

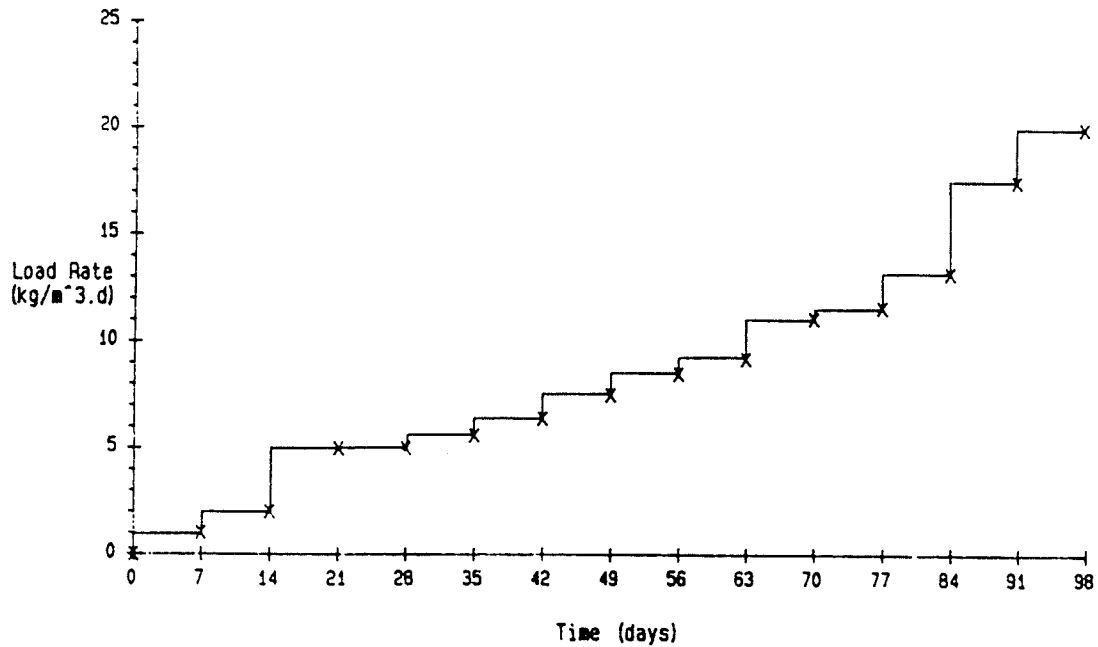


Figure 4.4 Increase in organic load rate with respect to time.

#### 4.2.2 Performance Study

The duration of this study was 100 days. The reactors were monitored throughout the study in order to compare the effects of the media on their performance and operation. The parameters chosen as the basis for comparison were as follows:

- (a) soluble COD removal;
- (b) soluble organic carbon (SOC) removal;
- (c) cumulative gas production; and
- (d) volatile fatty acids concentrations.

The reactors were operated at organic load rates from 1 to 20 kg COD/m<sup>3</sup>.day at varying recycle rates.

#### 4.2.3 Hydraulic Study

Hydraulic testing of the reactors involved a slug injection of known tracer concentration into a "T" connection between the recycle pump and the distributor for the reactors. The effluent flow was not recycled.

The tracers, methylene blue and rhodamine B, were chosen for their visual and fluorescent properties, respectively.

In the first phase of the study, the reactors were cleaned thoroughly and filled only with water. A slug of methylene blue was injected in each reactor. The progress of the tracer through each reactor was then photographed at regular time intervals in order to visually distinguish the effects of the different media.

In the second phase, each reactor was charged with 4.14 L of sludge, acclimated during the performance study. Rhodamine B was injected into the reactors in a similar manner; the concentration of the rhodamine B in the

effluent was monitored every 5 minutes for 30 minutes and then every 30 minutes for a total elapsed time of 360 minutes (6 hydraulic retention times).

The reactors were subsequently started again at an organic load rate of 0.5 kg COD/m<sup>3</sup>.day. When gas production reached 5 L/d, the procedure outlined above was repeated with rhodamine B to examine the combined effect of gas production and media type on the flow. The parameters of the hydraulic tests are listed in Table 4.4.

TABLE 4.3  
Parameters of Hydraulic Tests

Parameter	methylene blue	rhodamine B
Hydraulic Retention Time (min.)	60	60
Gas Production (L/d)	0	5
Tracer Concentration (mg/L)	750	100
Slug Volume (ml)	20	30

#### 4.3 Sampling and Analysis Schedule

The sampling schedule designed to monitor the reactors during the performance study is shown in Table 4.5. Those parameters that were vital to the daily operation of the system were monitored most frequently.

TABLE 4.4

Analysis Schedule for Anhybrid Reactors

Parameter	Days Monitored *
Gas Production	1 2 3 4 5 6 7
Influent pH	1 2 3 4 5 6 7
Effluent pH	1 2 3 4 5 6 7
Sludge Bed Height	1 2 3 4 5 6 7
Flow Rate	1 2 3 4 5 6 7
Recycle Rate	1 2 3 4 5 6 7
Effluent Volume	1 2 3 4 5 6 7
Effluent Suspended Solids (Total and Volatile)	2 4
COD (soluble)	2 4
SOC	2 4
Volatile Fatty Acids	3
Volatile Solids Content in Reactors	6
Volatile Solids Lost from System	6

- \* 1 - Monday  
 2 - Tuesday  
 3 - Wednesday  
 4 - Thursday  
 5 - Friday  
 6 - Saturday  
 7 - Sunday



The organic load rate was increased on day 1 (Monday) after pseudo-steady state conditions were reached. Samples from the reactor effluents were taken on days 2 (Tuesday) and 4 (Thursday) and analyzed for suspended solids, soluble COD and soluble organic carbon.

The volatile fatty acids concentration and the alkalinity in each reactor were each determined once per week, on days 3 and 5, respectively.

The volatile solids content in the reactors and the thermal solids shock separators were measured once per week in order to calculate the operating F/M ratio.

#### 4.4 Monitoring Techniques

##### 4.4.1 Gas Production

The daily gas production was measured by low volume gas meters from Triton Electronics (Model P.180). The meters required no maintenance other than changing the power supply and an occasional visual inspection.

##### 4.4.2 pH

All pH measurements were made with a Radiometer pH meter (Model No. PHM 29 b) using a glass pH electrode and a calomel reverse-sleeve reference electrode. The pH meter was calibrated daily.

#### 4.4.3 Flow and Recycle Rates

Pump speeds and flow rates were correlated at the beginning of the study. The feed and recycle pumps were monitored twice daily with a Shimpo hand digital tachometer (Model No. DT-105).

#### 4.4.4 Total and Volatile Suspended Solids

The total suspended solids in each reactor effluent were determined according to the procedure outlined in Standard Methods (3) Section 209C, "Total Suspended Solids Dried at 103-105°C". The volatile portion of this was then calculated by following the procedure in Section 209D, "Fixed Volatile Solids Ignited at 550°C".

The total volatile solids in both the reactor and the thermal shock solids separator were determined by following the procedures in Sections 209A, "Total Solids Dried at 103-105°C", and 209D.

#### 4.4.5 Soluble Chemical Oxygen Demand

The soluble COD of the influent and effluent samples was determined after filtration (Whatman grade 934AH glass-fiber filters) as outlined in Standard Methods (3) Section 508C, "Closed Reflux, Colorimetric Method" using a Baush and Lomb Spectronic 20 spectrophotometer ( $\lambda = 600 \mu\text{m}$ ).

#### 4.4.6 Soluble Organic Carbon

Samples analyzed for soluble organic carbon were filtered as described in Section 4.4.5 and then acidified with phosphoric acid to a pH 2. Samples were analyzed on a Dohrmann Carbon Analyzer (Model No. DC-80) with an Auto Sampler (Model No. ASM-1).

#### 4.4.7 Gas Analysis

The constituents of the gas produced by each reactor were determined according to Standard Methods (3) Section 511 B, "Gas Chromatograph Method". A Gow Mac Thermal Conductivity Detector (Model 550) with a Porepak Q column was used.

#### 4.4.8 Volatile Fatty Acids

Volatile fatty acids concentrations in the reactors were determined using a Gow Mac Flame Ionization Detector (Model 750) with a Chromasorb 101 column.

#### 4.4.9 Hydraulic Tests

The fluorescence emitted by the rhodamine B in the hydraulic tests was measured on a Turner Model 110 fluorometer. Baush and Lomb Spectronic 20 glass cuvettes (Model No. 33-29-27) were used to contain the samples in the fluorometer.

## CHAPTER 5

### RESULTS

#### 5.1 Summary of Results

The data accumulated from the analyses conducted during the performance portion of the study are shown in Appendix A. These results have been tabulated in Tables 5.1(a) to 5.1(1) along with the corresponding operating parameters for the reactors.

Tables 5.2(a) and 5.2(b) list the values obtained for the parameters determined in the hydraulic portion of the study. Appendix B contains the raw data and the tracer-response curves that were used to evaluate the hydraulic characteristics of the reactors.

TABLE 5.1(a)  
Performance Comparison (OLR<sup>\*</sup> = 1 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	140	133	167	173
COD <sub>5</sub> Removal (%)	50	68	51	59
VFA (mg/L)	3150	2460	2980	3340
Average Daily Gas Production (L)	8.4	13.4	10.6	10.1
L CH <sub>4</sub> /g COD Removed **	0.30	0.35	0.37	0.31
HRT (d)	4.9	4.9	4.9	4.9

\* OLR = Organic Load Rate

\*\* At 35°C.

TABLE 5.1(b)

Performance Comparison (OLR = 2 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	140	100	140	167
COD <sub>S</sub> Removal (%)	55	64	55	54
SOC Removal (%)	33	84	45	52
Average Daily Gas Production (L)	16.2	23.0	17.8	17.5
L CH <sub>4</sub> /g COD Removed	0.19	0.27	0.21	0.21
HRT (d)	2.5	2.5	2.5	2.5

TABLE 5.1(c)  
Performance Comparison (OLR = 5 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	113	80	153	180
COD <sub>5</sub> Removal (%)	79	87	52	75
SOC Removal (%)	86	91	65	82
VFA (mg/L)	380	320	710	640
Average Daily Gas Production (L)	32.3	33.5	27.7	27.9
L CH <sub>4</sub> /g COD Removed	0.62	0.58	0.80	0.56
HRT (d)	1.0	1.0	1.0	1.0
SRT (d)	60.2	64.3	50.9	47.9
F/M (kg COD/kg VSS.day)	1.4	1.4	2.0	2.2

TABLE 5.1(d)  
Performance Comparison (OLR = 6.4 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	260	160	233	213
COD <sub>5</sub> Removal (%)	84	85	81	79
VFA (mg/L)	320	320	540	430
Average Daily Gas Production (L)	43.3	44.6	42.9	35.0
L CH <sub>4</sub> /g COD Removed	0.45	0.45	0.46	0.38
HRT (d)	0.79	0.79	0.79	0.79
SRT (d)	61.3	57.2	44.0	47.0
F/M (kg COD/kg VSS.day)	0.85	0.86	1.0	1.4



**TABLE 5.1(e)**  
Performance Comparison (OLR = 7.5 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	133	114	93	100
COD <sub>S</sub> Removal (%)	78	89	80	79
SOC Removal (%)	79	89	82	82
VFA (mg/L)	3150	2460	2980	3340
Average Daily Gas Production (L)	8.4	13.4	10.6	10.1
L CH <sub>4</sub> /g COD Removed	0.30	0.35	0.37	0.31
HRT (d)	4.9	4.9	4.9	4.9
SRT (d)	48.5	45.3	40.7	44.2
F/M (kg COD/kg VSS.day)	1.5	1.4	1.6	1.9

TABLE 5.1(f)  
Performance Comparison (OLR = 8.5 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	167	160	287	320
COD <sub>S</sub> Removal (%)	63	82	69	62
SOC Removal (%)	61	80	63	32
VFA (mg/L)	1800	540	440	1380
Average Daily Gas Production (L)	30.3	54.1	48.1	-
L CH <sub>4</sub> /g COD Removed	0.22	0.43	0.36	
HRT (d)	0.59	0.59	0.59	0.59
SRT (d)	61.8	78.6	50.3	47.6
F/M (kg COD/kg VSS.day)	1.3	1.2	1.4	1.4

TABLE 5.1(g)

Performance Comparison (OLR = 9.2 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	187	100	187	160
COD <sub>S</sub> Removal (%)	60	72	66	76
SOC Removal (%)	50	82	83	48
VFA (mg/L)	2060	520	500	1150
Average Daily Gas Production (L)	53.1	44.2	45.7	41.3
L CH <sub>4</sub> /g COD Removed	0.55	0.41	0.46	0.39
HRT (d)	0.55	0.55	0.55	0.55
SRT (d)	58.8	59.6	43.5	55.1
F/M (kg COD/kg VSS.day)	1.6	1.5	1.7	1.4

TABLE 5.1(h)  
Performance Comparison (OLR = 11.1 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	53	73	107	107
COD <sub>S</sub> Removal (%)	68	66	67	61
SOC Removal (%)	69	69	67	42
VFA (mg/L)	1860	1110	1950	1575
Average Daily Gas Production (L)	39.6	49.2	30.9	37.7
L CH <sub>4</sub> /g COD Removed	0.29	0.39	0.23	0.30
HRT (d)	0.45	0.45	0.45	0.45
SRT (d)	63.6	47.7	35.1	53.3
F/M (kg COD/kg VSS.day)	1.9	1.8	2.1	1.6

TABLE 5.1(i)

Performance Comparison (OLR = 11.6 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	167	203	353	313
COD <sub>5</sub> Removal (%)	76	72	65	63
SOC Removal (%)	73	68	56	45
VFA (mg/L)	1210	860	940	1380
Average Daily Gas Production (L)	47.6	52.8	42.0	43.3
L CH <sub>4</sub> /g COD Removed	0.28	0.35	0.25	0.33
HRT (d)	0.43	0.43	0.43	0.43
SRT (d)	41.8	46.4	33.1	41.3
F/M (kg COD/kg VSS.day)	3.3	3.2	3.9	3.0

TABLE 5.1(j)

Performance Comparison (OLR = 13.2 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	203	267	323	387
COD <sub>S</sub> Removal (%)	61	57	44	54
SOC Removal (%)	66	62	46	51
VFA (mg/L)	1430	1580	1540	1710
Average Daily Gas Production (L)	62.3	57.8	62.5	40.8
L CH <sub>4</sub> /g COD Removed	0.35	0.38	0.43	0.30
HRT (d)	0.38	0.38	0.38	0.38
SRT (d)	61.5	32.9	30.1	32.3
F/M (kg COD/kg VSS.day)	4.6	4.4	5.7	4.6

TABLE 5.1(k)

Performance Comparison (OLR = 17.5 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	203	333	387	467
COD <sub>S</sub> Removal (%)	33	24	20	22
SOC Removal (%)	23	14	12	8.5
VFA (mg/L)	1950	2300	2460	2380
Average Daily Gas Production (L)	37.6	37.8	37.9	31.3
L CH <sub>4</sub> /g COD Removed	0.30	0.45	0.52	0.40
HRT (d)	0.29	0.29	0.29	0.29
SRT (d)	42.8	25.8	16.4	20.5
F/M (kg COD/kg VSS.day)	7.5	7.1	8.7	7.6

TABLE 5.1(I)

Performance Comparison (OLR = 20.0 kg COD/m<sup>3</sup>.day)

Parameter	Reactor			
	R1	R2	R3	R4
VSS (mg/L)	333	387	433	467
COD <sub>S</sub> Removal (%)	13	13	9	11
SOC Removal (%)	12	12	8	7
VFA (mg/L)	2120	2310	2590	2600
Average Daily Gas Production (L)	28.8	27.3	24.1	23.6
L CH <sub>4</sub> /g COD Removed	0.85	0.81	1.02	0.83
HRT (d)	0.25	0.25	0.25	0.25
SRT (d)	23.5	25.8	16.4	20.5
F/M (kg COD/kg VSS.day)	7.2	7.0	8.8	7.7



TABLE 5.2(a)

Hydraulic Comparison (Gas Production = 0 L/d, HRT = 1 hr.)

Parameter	Reactor		
	R1	R3	R4
	<u>Run #1</u>		
Time of first tracer appearance (min.)	10	5	5
Modal Time (min.)	55	50	50
Mean Time (min.)	73.1	75.1	79.9
Variance	0.26	0.31	0.32
d/u1	0.15	0.19	0.30
	<u>Run #2</u>		
Time of first tracer appearance (min.)	10	10	5
Modal Time (min.)	45	35	35
Mean Time (min.)	70.0	60.96	79.92
Variance	0.30	0.35	0.33
d/u1	0.19	0.23	0.21

TABLE 5.2(b)

Hydraulic Comparison (Gas Production = 5 L/d, HRT = 1 hr.)

Parameter	Reactor		
	R1	R3	R4
	<u>Run #1</u>		
Time of first tracer appearance (min.)	15	5	5
Modal Time (min.)	60	55	99
Mean Time (min.)	80.8	78.2	80.9
Variance	0.25	0.31	0.28
d/u1	0.15	0.19	0.17
	<u>Run #2</u>		
Time of first tracer appearance (min.)	10	5	5
Modal Time (min.)	50	55	65
Mean Time (min.)	78.3	80.1	82.5
Variance	0.29	0.30	0.32
d/u1	0.18	0.19	0.20

## 5.2 Performance Results

To ensure that all four Anhydrid reactors were operating under identical conditions, the volatile suspended solids content of each reactor and its effluent were measured on a regular basis. This data was used to monitor the solids retention times of the reactor and their respective substrate utilization rates.

A suitable model was chosen which characterizes the effects of a variable substrate concentration on substrate utilization. It has been suggested by Grau et al. (68) that:

$$\frac{S_o - S}{\theta X} = \frac{K_1 S}{S_o} \quad (25)$$

where  $S_o$  = initial substrate concentration,  $M L^{-3}$

$S$  = substrate concentration surrounding the biomass  
at any time,  $M L^{-3}$

$K_1$  = specific substrate utilization constant,  $T^{-1}$

$X$  = active biomass concentration,  $M L^{-3}$

If a plot is drawn of  $(S_o - S)/X\theta$  versus  $S/S_o$ , the slope of the resulting curve will give the specific substrate utilization constant,  $K_1$ . Using this equation, the value of  $K_1$  of each reactor was determined (Figure 5.1) and is listed in Table 5.3. It should be noted that the values of  $K_1$  for all the reactors are comparable, indicating that their operating conditions were similar.

TABLE 5.3

Comparison of Reactor Performance

Parameter	Reactor			
	R1	R2	R3	R4
K1 (days <sup>-1</sup> )	6.28	3.92	5.98	3.85
Average COD Removal (%)	58.9	66.4	55.5	56.2
Average SOC Removal (%)	58.9	66.0	55.5	52.0
Cumulative Gas Production (L)	3589	3756	3381	2857
VFA at Failure (mg/L)	2120	2310	2590	2600

Determination of Biokinetic Constant, K1

Trends of all Reactors

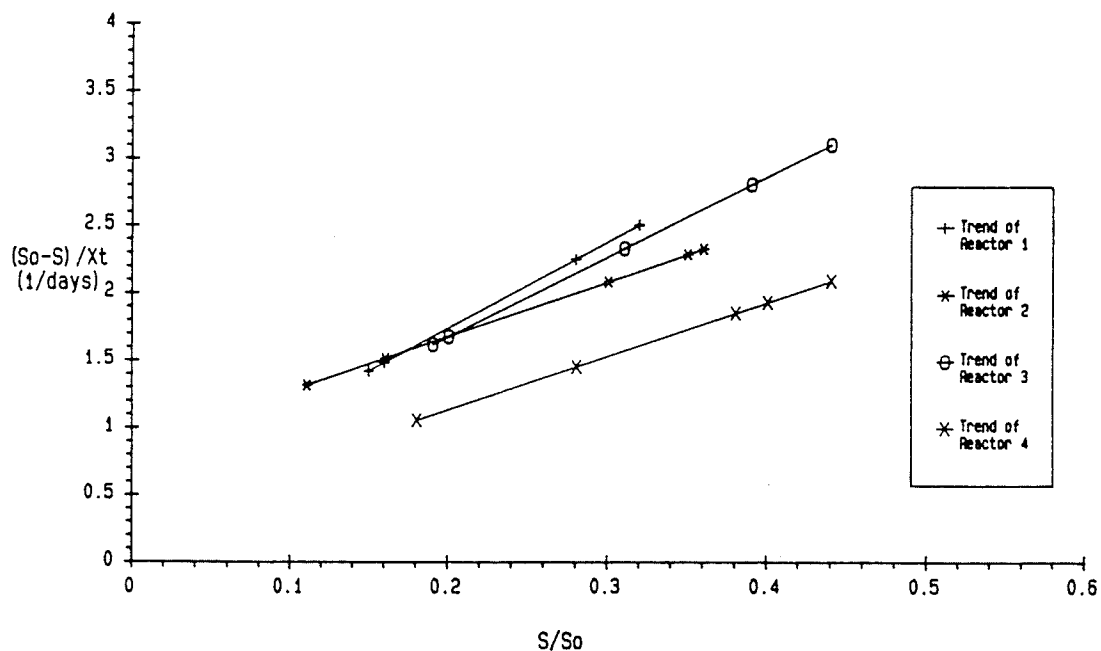


Figure 5.1 Determination of biokinetic constant, K1.

The soluble influent and effluent COD and SOC concentrations were monitored to calculate the organic removal efficiencies of the reactors under the various operating conditions. As shown in Figures 5.2 and 5.3, the removal efficiencies in the reactors increased as they acclimated to the synthetic feed until day 20 at which point the organic load rate was increased step-wise. This was continued until failure was reached in the reactors.

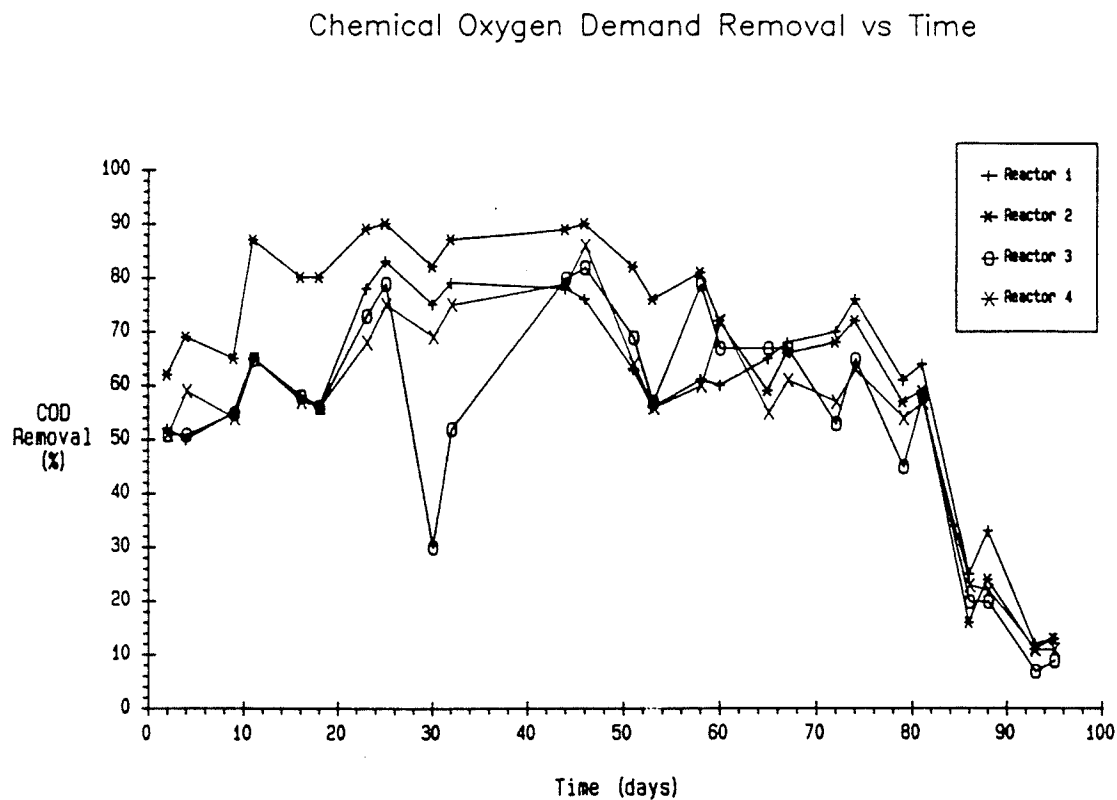


Figure 5.2 Illustration of COD removal with respect to time.

The reactors maintained consistent removals until day 77 when the load was increased to 13.2 kg COD/m<sup>3</sup>.day. At this point in time, a deterioration in the performance of all four reactors was noted with the greatest reductions taking place in R3 and R4.

While all four reactors failed at approximately 100 days, it is apparent from Figure 5.2 that R1 and R2 were providing greater COD removals than R3 and R4 prior to failure. This is reflected in the average COD removals over the entire study as listed in Table 5.3. The average values for R1 and R2 are both higher than those of R3 and R4. The same holds true for the SOC removals recorded during the study.

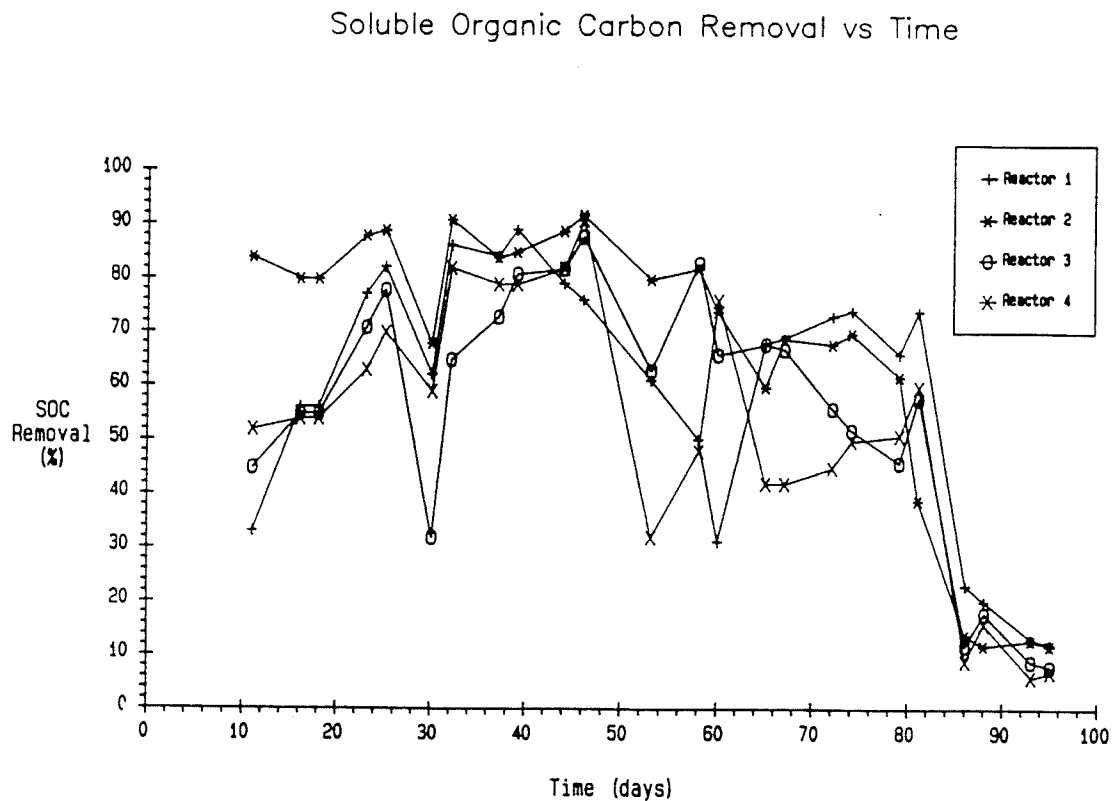


Figure 5.3 Illustration of SOC removal with respect to time.

It may be observed that, although R1 and R2 possessed identical media and were operating under comparable conditions, R2 had higher average COD and SOC removals than R1. This discrepancy can be attributed to an upset that occurred in R1 early in the study due to a mechanical difficulty. Similarly, R3 experienced an upset that required a small period of time for recovery, as seen in Figure 5.2.

Figure 5.4 illustrates the cumulative production of biogas with respect to time for the reactors. By day 21 both R1 and R2 were producing larger quantities of biogas than R3 and R4. This trend continued until the end of the study. In all cases, the production of biogas decreased with a corresponding decrease in COD removal efficiency - most notably in R4. The values for the cumulative gas production of the reactors are shown in Table 5.3.

Cumulative Gas Production vs Time

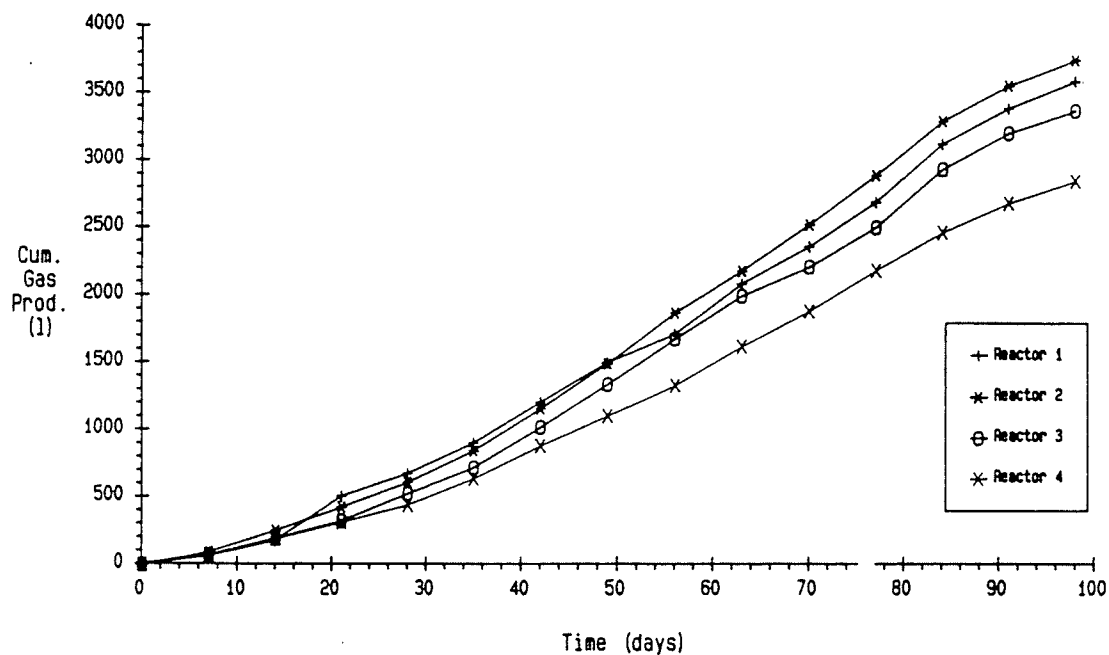


Figure 5.4 Illustration of cumulative gas production with respect to time.

The variation in the concentration of volatile fatty acids in all four reactors with respect to organic load rate is shown in Figure 5.5. The upsets that took place early in the study in R1 and R3 are visible, as are the subsequent periods of stabilization. The volatile fatty acids in all four reactors generally increased with each increase in the organic load rate. These increases in volatile fatty acids also correspond with the decreases in COD removal seen in Figure 5.2. By day 80, the volatile fatty acids concentrations in the reactors had reached a level whereby the operating pH of 7 to 7.5 could no longer be maintained. As shown in Figure 5.5, the volatile fatty acids concentrations in all the reactors increased dramatically at an organic load rate of approximately 13.2 kg COD/m<sup>2</sup>.day. The pH decreased consistently after that point until failure. The volatile fatty acids concentrations at failure are listed in Table 5.3.

Volatile Fatty Acids vs Organic Load Rate

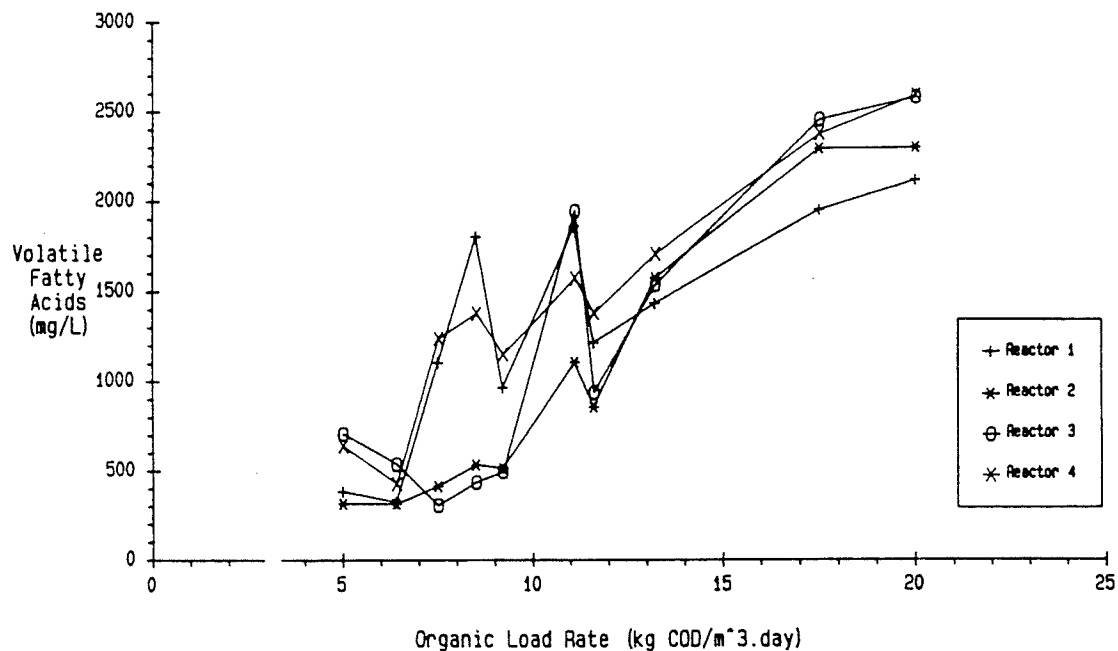


Figure 5.5 Illustration of volatile fatty acids with respect to organic load rate.



The effluent volatile suspended solids concentrations from each reactor were monitored throughout the study. Figure 5.6 shows the change in volatile suspended solids concentrations in each reactor with respect to time. As can be seen, while the concentration from each reactor increased toward the end of the study, the concentrations in the effluent from R3 and R4 were greater than R1 and R2.

Volatile Suspended Solids vs Time

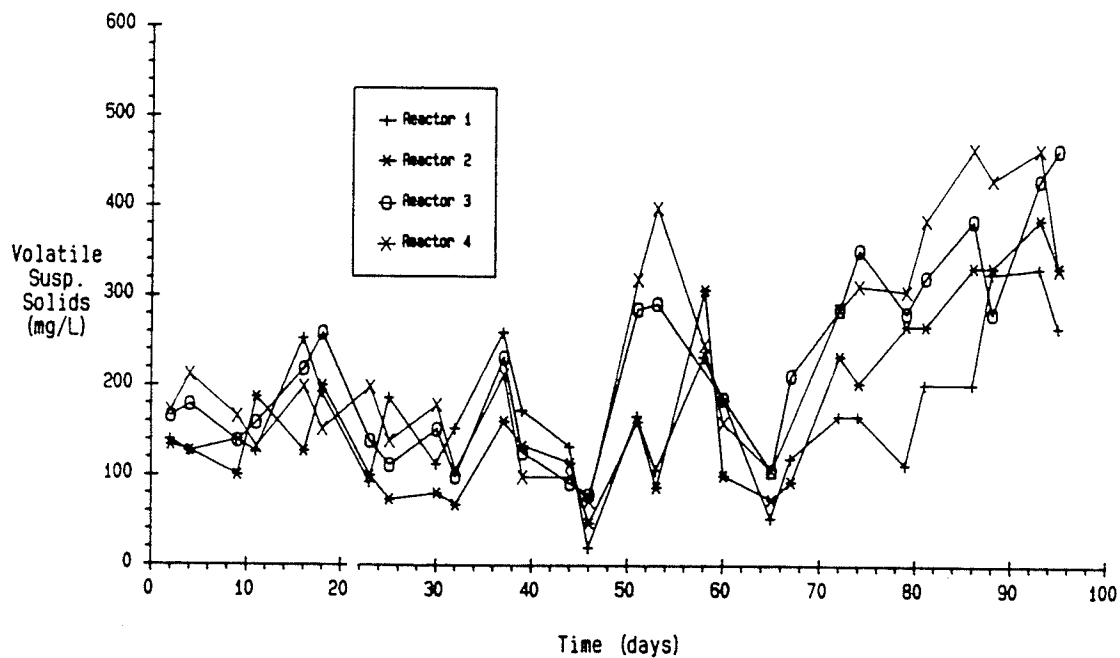


Figure 5.6 Illustration of volatile suspended solids in effluent with respect to time.

The concentrations of the volatile suspended solids and their distributions along the height of the reactors were also monitored during the study. Figure 5.7 is an example of one of the volatile suspended solids profiles obtained for the reactors. It is clear from this visual representation that R1 and R2 retained a greater concentration of solids both in the sludge bed zone and in the media volume zone.

### Reactor Height vs Volatile Suspended Solids

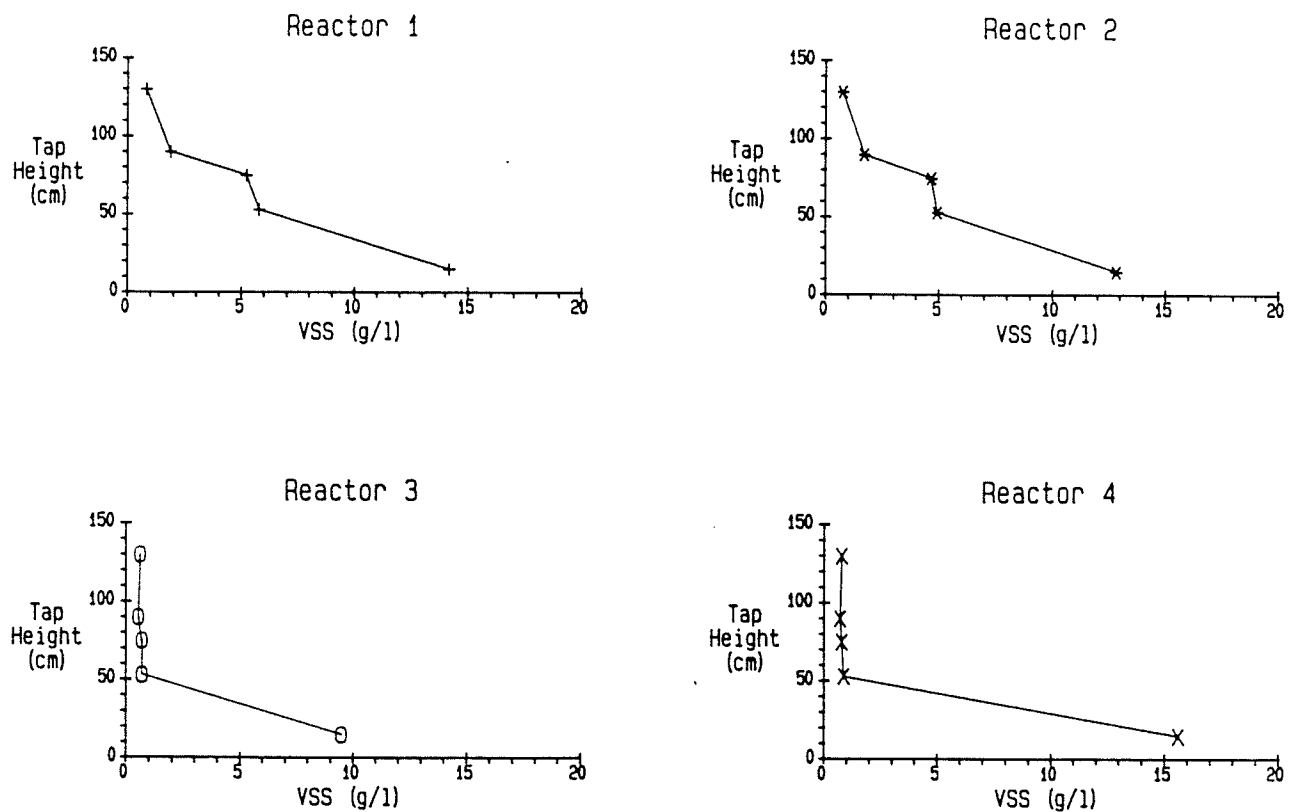


Figure 5.7 Illustration of volatile suspended solids profile along reactor height.

### 5.3 Tracer Test Results

The visual comparison of media type using methylene blue as a tracer is shown in Figures 5.8(a) to 5.8(g). The difference in the hydraulic regime of the reactors was apparent (Figure 5.8(c), 10 minutes after the introduction of the tracer into the reactors. This illustration shows the greater difference of the dispersion of the slug of methylene blue in R4 as compared to R2. The dispersion of R3 was between these two reactors.

In Figure 5.8(d), 15 minutes after the input of the tracer, an even larger dispersion in R3 and R4 can be seen. While these reactors displayed evidence of hydraulic short-circuiting and mixing, R2 continued to operate essentially as a plug flow reactor.

After 20 minutes, the methylene blue was dispersed completely throughout the contents of R4. The slug of tracer in R2, however, had not yet reached the level of the media. It may be observed that the tracer in R1 continued to behave as a plug while the tracer in the other two reactors was mixed to a large degree. These visual observations indicate that the hydraulic regime in R2 was essentially plug flow while R3 and R4 gravitated more towards completely mixed flow.



Figure 5.8(a) Reactors prior to introduction of tracer.



Figure 5.8(b) Five minutes after introduction of tracer.



Figure 5.8(c) Ten minutes after introduction of tracer.

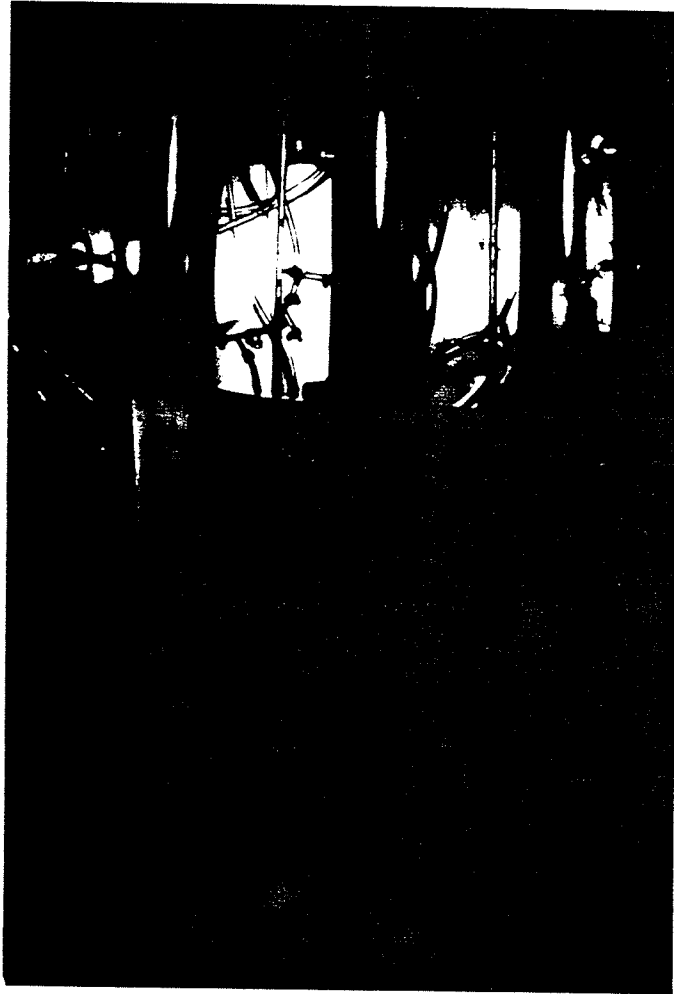


Figure 5.8(d) Fifteen minutes after introduction of tracer.

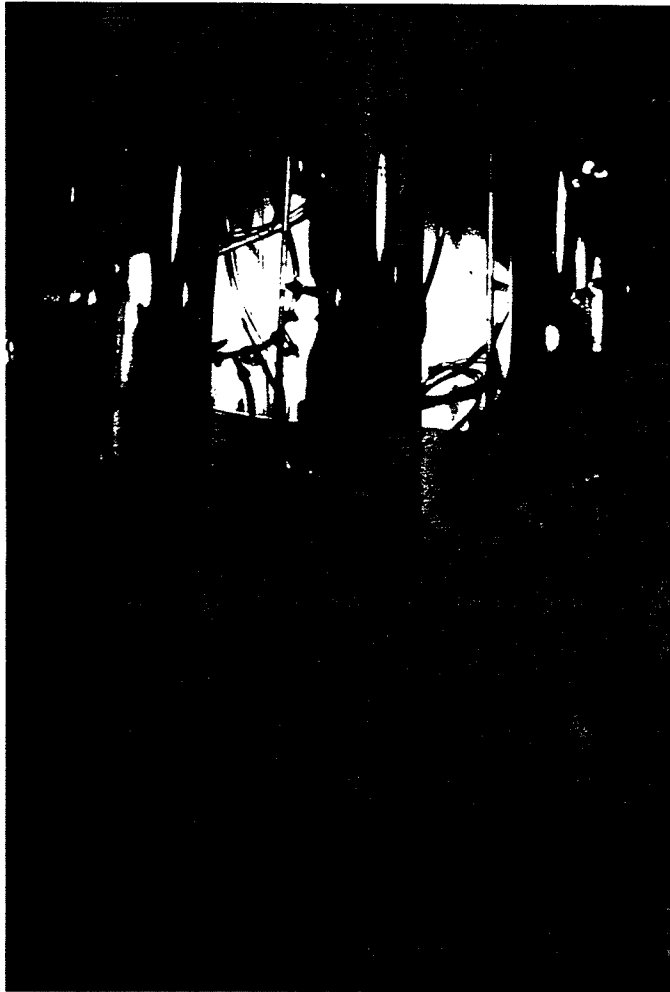


Figure 5.8(e) Twenty minutes after introduction of tracer.



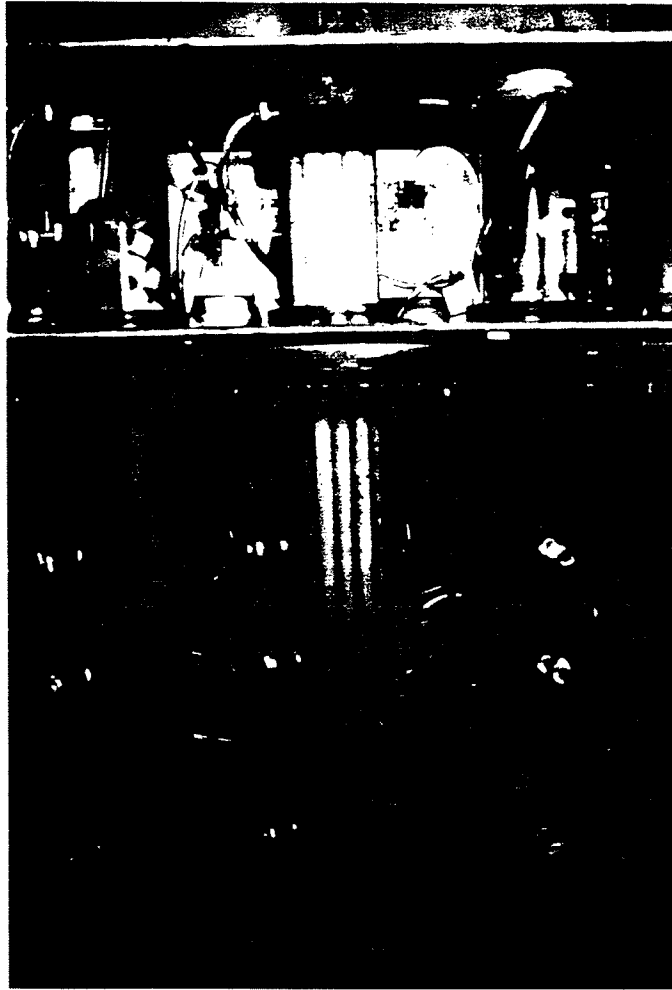


Figure 5.8(f) Twenty five minutes after introduction of tracer.

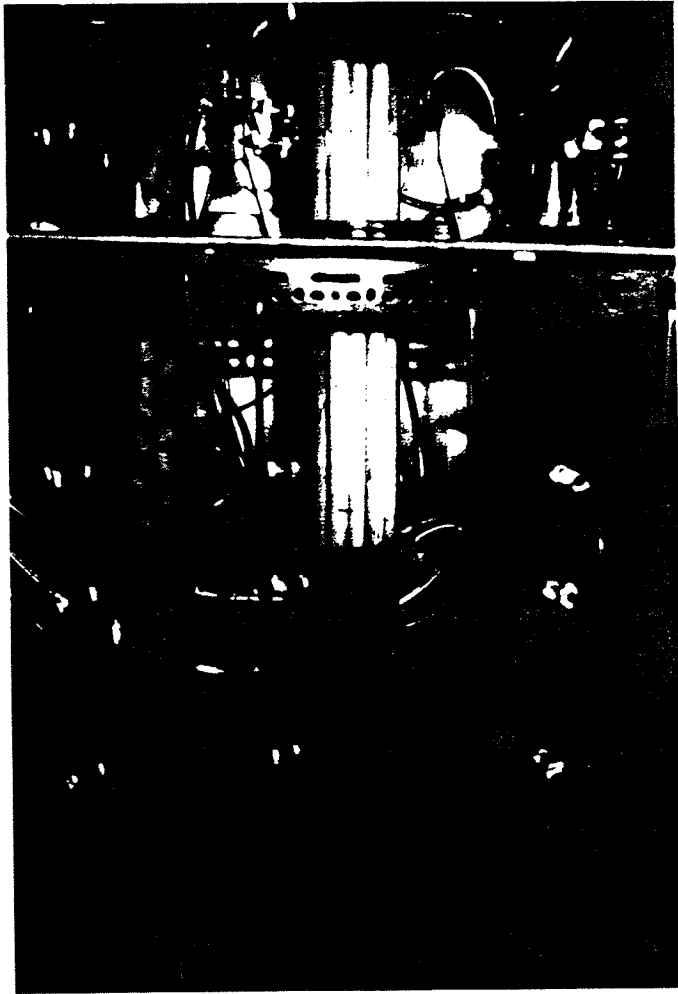


Figure 5.8(g) Thirty minutes after introduction of tracer.

Figures 5.9(a) to 5.9(d) illustrate the tracer-response curves resulting from the tests using rhodamine B as a tracer. These corroborate the visual observations mentioned above. In each curve, the modal time increases with the porosity of the media. That is, R1 had the greatest modal time, followed by R3, and then R4.

All of the reactors had low dispersion numbers, which was expected because the tests were run without a recycle flow stream. The reactors were, therefore, operating with a plug flow hydraulic regime. In all of the tests, the dispersion number of R1 was lower than that of R3 and R4.

### Tracer Concentration vs Time

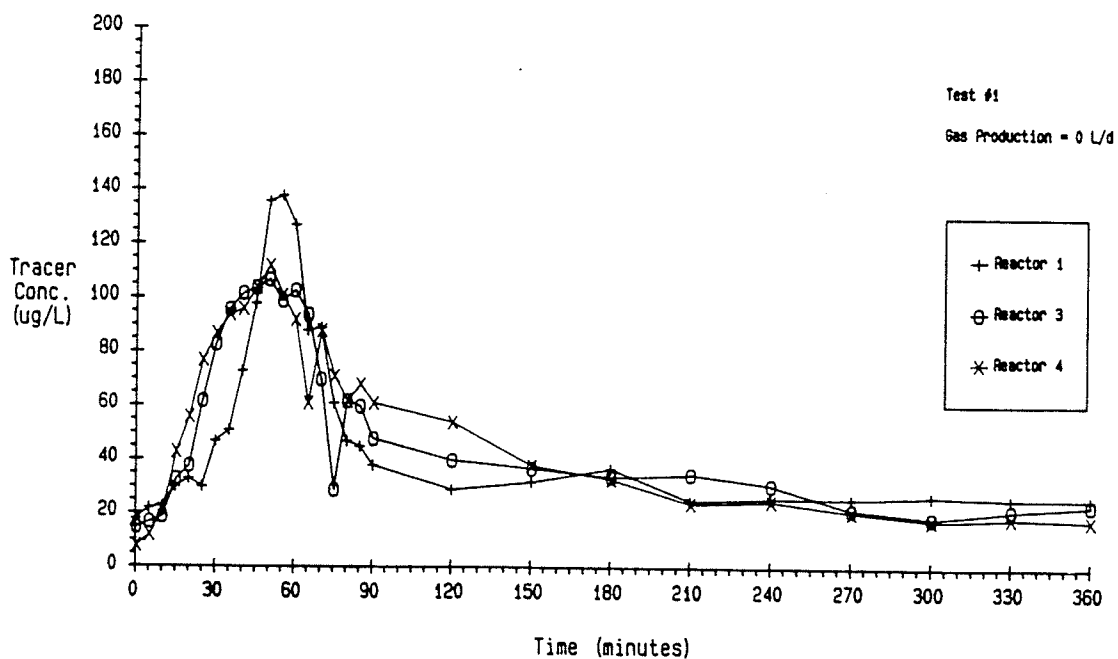


Figure 5.9(a) Tracer test with rhodamine B  
(Test #1 - Gas Production = 0 L/d).

### Tracer Concentration vs Time

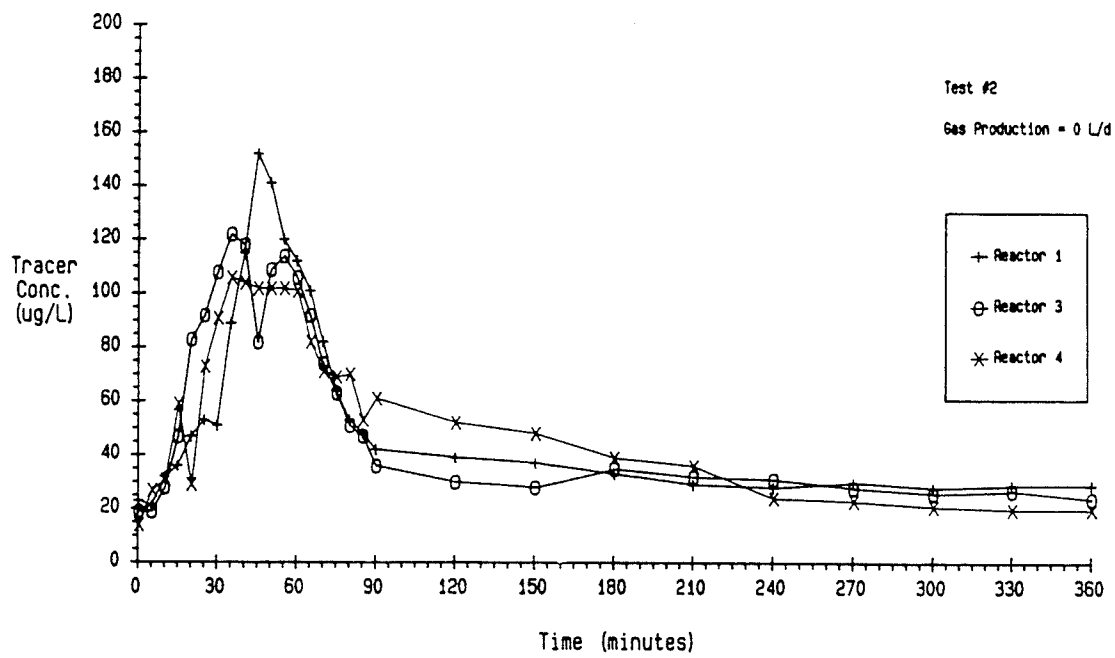


Figure 5.9(b) Tracer test with rhodamine B  
(Test #2 - Gas Production = 0 L/d).

### Tracer Concentration vs Time

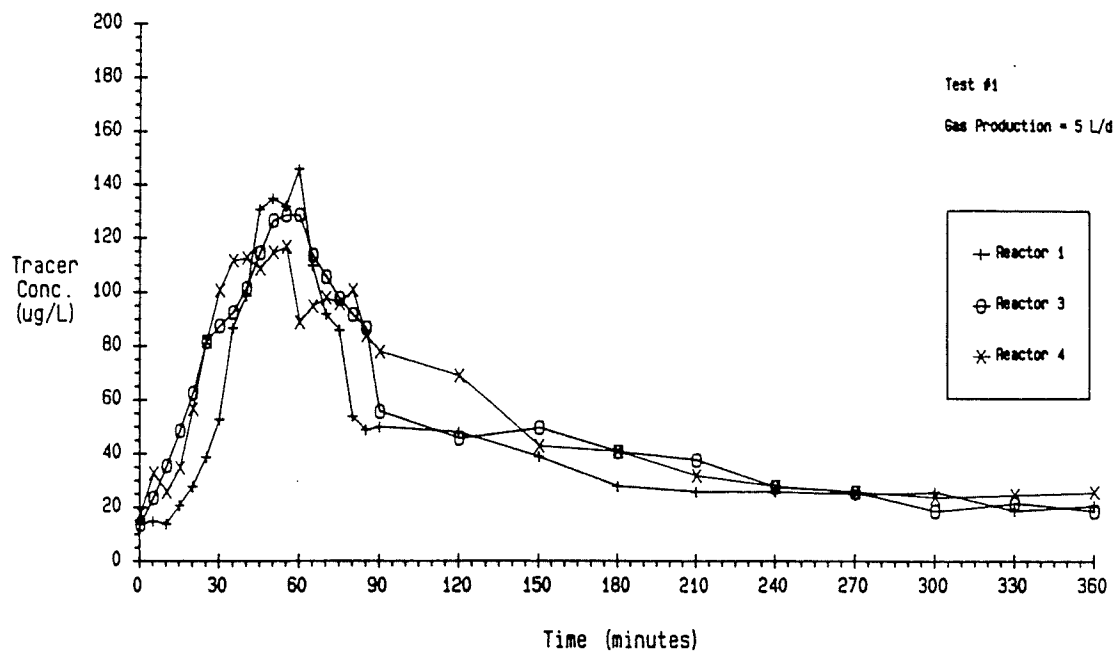


Figure 5.9(c) Tracer test with rhodamine B  
(Test #1 - Gas Production = 5 L/d).

### Tracer Concentration vs Time

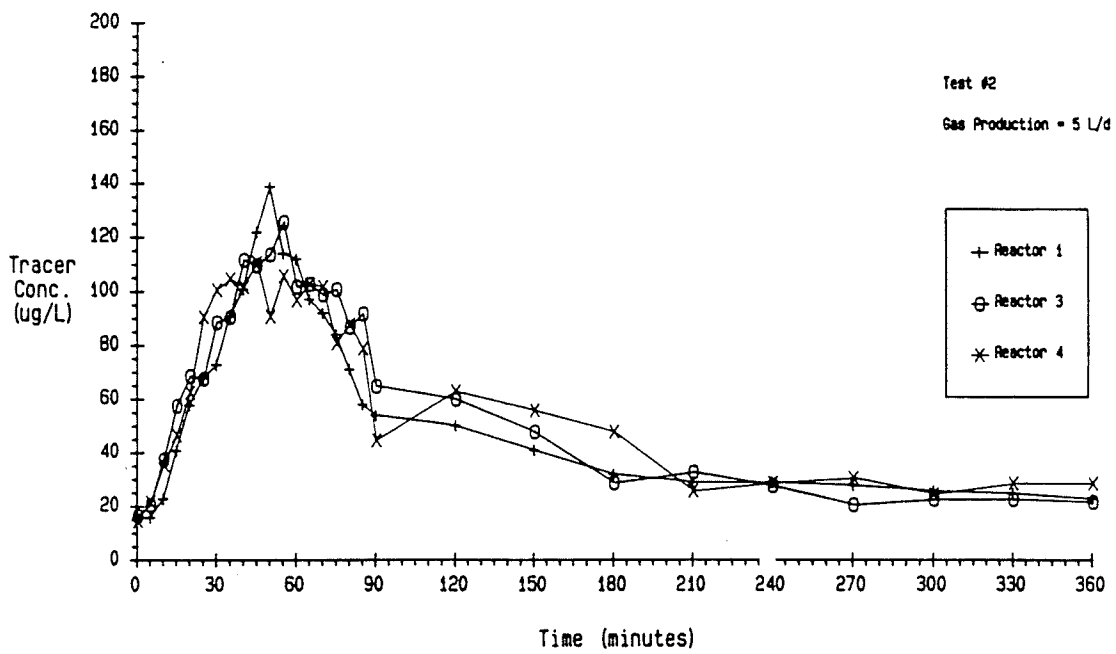


Figure 5.9(d) Tracer test with rhodamine B  
(Test #2 - Gas Production = 5 L/d).

## CHAPTER 6

### DISCUSSION

The results of this study indicate that the type of media utilized in the Anhybrid reactor does, in fact, have a significant effect on reactor performance. Throughout the operational stage of the study, the performance of the two reactors with random media was characterized by more consistent operation than the reactor with oriented media and the reactor with no media. The reason for this stability was the ability of the random media to retain biomass more effectively than the other two configurations.

In their research, Maxham and Wakamiya (57) reported the presence of media in the hybrid reactor was most beneficial during start-up and reactor instability. This was also the case in this study. The reactors with the Anhybrid configuration were more easily acclimated to the synthetic sulphite evaporator condensate substrate than the other reactors. When the organic load rate was increased after 14 days, all of the reactors were providing similar COD removal efficiencies. After 28 days, R4 began falling behind the other reactors in terms of treatment efficiency - a trend that continued until the end of the study. Similarly, R3 did not provide as consistent COD removal as R1 and R2 during the entire study. This was partially due to an upset in the reactor. The average values for COD removal reflected their performance during the study (R1 and R2 had the highest averages followed by R3 and then R4).

During other transient conditions throughout the study, it was observed that the reactors with the random media were least susceptible to biomass loss. The reactors were operated at an SRT of approximately 50 days throughout the study. As time progressed, it became increasingly difficult to maintain the solids in R3 and R4. R1 and R2, however, were able to operate at a relatively constant SRT until the end of the study. The volatile fatty acids concentrations in R1 and R2 were, consequently, lower than in R3 and R4 at failure because the F/M ratios in the reactors were higher.

Dahab and Young (26) reported that most of the biomass was in the lower portion of the anaerobic filters in their study. Less than half of this constituted attached growth. Because there was very little biomass in the interstitial voids in the upper region, it can be reasoned that the media in this region serves to retain biomass. Thus, the need for media in the lower two-thirds of the reactor must be questioned.

The volatile suspended solids profiles taken during the course of this study showed that this was the case. The greatest quantity of biomass was found to be in the bottom one-third of the reactors. Although the amount of biomass in the reactors was similar at the beginning of the study, after failure R1 and R2 had the greatest quantities of biomass remaining followed by R3 and R4. A large part of this biomass was retained in the interstitial voids of the media. No attached growth was observed on the media, most likely because sufficient time was not allowed for a layer to develop.



Granulation of the sludge did not occur in any of the reactors during the study. There are several possible reasons for this:

- (a) the hydraulic flow rate was not high enough to promote granulation;
- (b) the synthetic feed lacked the nutrients required for granulation; and
- (c) the seed sludge was not conducive to the formation of granules.

The most plausible reason for the absence of granulation in the reactors is the fact that the seed sludge was of a flocculent nature. Granulation may have eventually taken place, but the study was not of sufficient duration to examine the development of this phenomenon.

In agreement with the findings of Song and Young (30), it appears that pore size and horizontal alignment of the media were the properties which had the greatest affect on media performance. It was assumed that the pore size of the Rashig rings in the Anhybrid reactors ranged from 0 to 2.5 cm because they were placed at random. The horizontal alignment, therefore, also varied from 0 to 90°. The PVC tubing had a pore size of 2.5 cm and a vertical alignment. The pore size of the reactor without media was 11 cm, the interior diameter of the reactor itself.

The significance of these properties on fluid flow are readily apparent from the illustrations in the previous section. The Rashig rings served to dampen the fluid flow through the reactor and thereby reduce the longitudinal dispersion. While the methylene blue tracer behaved as a plug in R2, it was dispersed in R3 and even more so in R4.

The difference in the results of the tracer tests with rhodamine B were not as marked. A mass balance of the tracer showed that a portion of the rhodamine B remained in the reactors and was assumed to have been absorbed by the sludge. It is believed that this affected the tracer-response curves somewhat and a true representation of the dispersion in the reactors was not obtained.

The calculated dispersion numbers indicated that a plug flow hydraulic regime prevailed in the reactors. This was expected because no recycle flow was applied during the tracer tests. The dispersion number for R1 in each of the tests was smaller than R3 and R4, indicating that the hydraulic regime in it most closely approximated plug flow. The dispersion increased when the gas production was increased to 5 L/d. The gas production, however, was not high enough to corroborate the research by Bolle et al. (52) which suggested that the maximum short-circuiting flow is a function of gas velocity in the UASB reactor.

## CHAPTER 7

### CONCLUSIONS

The following conclusions are drawn from the results of this study:

- (a) the type of media utilized in the reactors affected their performance (the reactors with random media displayed superior performance in terms of organic removal efficiency);
- (b) porosity and horizontal alignment of the media play a major role in biomass retention in the reactors;
- (c) the operation of the reactors with random media was more stable than the reactor with oriented media and the reactor with the UASB configuration during start up and periods of increased organic loading;
- (d) the quantity of biomass retained in the reactors varied in accordance with the type of media utilized (the reactors with random media were better able to maintain biomass than the reactor with oriented media and the reactor with the UASB configuration);
- (e) the type of media utilized affected the hydraulic regime in the reactors (the reactors with random media has less longitudinal dispersion than the reactor with oriented media and the UASB configured reactor).

## CHAPTER 8

### SUGGESTIONS FOR FURTHER STUDY

The Anhybrid reactor has proven itself both in the laboratory and in full scale facilities. Further research, however, is still required on the configuration to optimize its design. It is suggested that the following topics receive priority:

- (a) a laboratory study should be conducted in which tracer tests are run on the Anhybrid with a variety of media types. To prevent absorption of the tracer, glass beads may be used in the sludge bed zone instead of actual sludge. An artificial source of gas can be introduced to examine the combined effects of media type and gas velocity on the hydraulic regime in the reactor; and
- (b) tracer tests should be run on existing full-scale installations without recycle to quantify the dispersion of the flow. These results can then be compared to the performance data that has already been obtained from the facilities to determine the effects of media type on full-scale Anhybrid reactors.

## CHAPTER 9

### ENGINEERING SIGNIFICANCE

Based upon the results of this study, it appears that the pore size and horizontal alignment of the media utilized are critical to the performance of the Anhybrid reactor. Various media may also be better suited to different waste streams. Great care, therefore, should be taken when choosing a media for a particular influent.

The effect of the media type in the reactor can be quantified through the use of tracer tests. By running a tracer through the reactor without recycle flow, the deviation from plug flow can be quantified for any combination of media type and waste stream. The dispersion number obtained can then be used to choose the optimum media type.

# APPENDIX A

## PERFORMANCE AND OPERATION DATA

A-2

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
03/06/85	1	1	4.6	4.9	9.4	5.8	23
		2	4.6	6.7	8.7	13.4	26
		3	4.6	5.1	9.4	7.6	28
		4	4.6	5.0	9.6	8.1	23
04/06/85	2	1	4.2	6.4	11.6	8.5	23
		2	4.3	7.3	11.2	10.3	23
		3	4.3	6.5	11.9	9.8	24
		4	4.3	6.6	14.1	11.3	22
05/06/85	3	1	6.4	7.0	11.8	9.8	25
		2	6.4	7.5	11.5	10.0	25
		3	6.4	6.9	11.2	10.8	24
		4	6.4	7.1	11.9	11.6	22
06/06/85	4	1	7.4	7.6	10.2	7.6	23
		2	7.2	7.7	10.3	-	25
		3	7.2	7.2	10.0	8.4	24
		4	7.2	7.4	10.0	9.4	20
07/06/85	5	1	7.9	7.6	13.0	-	22
		2	7.9	7.8	13.1	16.7	23
		3	7.9	7.6	12.8	10.1	23
		4	7.7	7.7	12.4	11.3	21
08/06/85	6	1	7.2	7.8	11.8	-	22
		2	7.2	8.0	12.1	16.0	26
		3	7.2	7.6	12.1	8.5	23
		4	7.2	7.7	12.0	9.4	21
09/06/85	7	1	6.0	7.0	10.0	-	80+
		2	6.0	7.6	9.8	16.8	80+
		3	6.0	7.5	10.4	18.6	80+
		4	6.0	7.3	9.8	9.5	80+

Date (D/M/Y)	Day	Reactor	pHi	pHe	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
10/06/85	8	1	6.0	6.9	10.7	10.2	23
		2	6.0	7.3	10.4	17.8	25
		3	6.0	7.3	10.4	12.0	22
		4	6.0	7.1	10.0	12.2	20
11/06/85	9	1	5.7	6.9	11.9	12.0	24
		2	5.7	7.4	11.6	20.0	25
		3	5.7	7.2	11.8	12.0	23
		4	5.7	17.0	11.6	-	20
12/06/85	10	1	5.4	6.6	11.4	12.7	80+
		2	5.4	7.2	11.3	19.9	80+
		3	5.4	7.1	11.9	15.2	80+
		4	5.4	7.1	11.3	-	80+
A-3 13/06/85	11	1	5.9	6.7	10.2	14.7	80+
		2	5.9	8.9	2.2	22.8	80+
		3	5.9	7.2	9.7	17.8	23
		4	5.9	7.3	9.9	16.2	80+
14/06/85	12	1	5.7	6.7	10.8	18.9	22
		2	5.7	7.8	10.2	26.6	25
		3	5.7	7.3	10.7	20.6	20
		4	5.7	7.0	10.1	19.5	19
15/06/85	13	1	5.7	6.8	12.7	22.0	21
		2	5.7	7.7	12.6	27.3	-
		3	5.7	7.1	12.2	24.3	19
		4	5.7	7.3	12.0	20.8	17
16/06/85	14	1	5.8	7.0	12.1	22.8	22
		2	5.8	7.7	12.1	26.5	24
		3	5.8	7.2	12.0	24.4	20
		4	5.8	-	11.8	19.1	17



A-4

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
17/06/85	15	1	5.3	7.1	9.5	18.7	22
		2	5.3	7.7	9.1	21.8	27
		3	5.3	7.2	7.2	19.4	16
		4	5.3	7.3	9.3	14.6	15
18/06/85	16	1	5.2	7.1	11.3	23.5	-
		2	5.2	7.7	11.0	27.3	-
		3	5.2	7.0	11.0	23.3	13
		4	5.2	7.1	11.0	17.2	12
19/06/85	17	1	5.2	7.2	11.1	22.7	18
		2	5.2	7.5	11.0	23.8	24
		3	5.2	7.2	11.2	20.1	13
		4	5.2	6.9	11.3	17.9	12
20/06/85	18	1	5.3	7.1	11.6	24.7	19
		2	5.4	7.6	11.4	28.0	25
		3	5.4	7.2	11.4	22.1	12
		4	5.4	6.9	11.6	17.0	12
21/06/85	19	1	5.1	7.0	10.3	99.4	18
		2	5.1	7.6	10.1	25.5	-
		3	5.1	7.0	10.6	22.6	12
		4	5.1	6.8	10.2	15.8	-
22/06/85	20	1	5.0	7.4	14.2	.4	23
		2	-	-	-	5.6	21
		3	5.0	7.1	14.2	32.1	10
		4	5.0	6.0	14.0	22.4	10
23/06/85	21	1	5.2	7.1	12.1	.2	17
		2	5.2	7.4	11.4	21.3	23
		3	5.2	7.1	11.8	23.7	15
		4	5.2	7.4	12.6	16.9	17

A-5

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
24/06/85	22	1	5.1	7.1	10.2	-	15
		2	5.1	7.5	10.0	19.4	20
		3	5.1	7.1	10.4	21.9	15
		4	5.1	7.3	10.2	15.5	15
25/06/85	23	1	5.3	7.6	14.1	22.3	21
		2	5.3	7.7	12.1	21.1	26
		3	5.3	7.4	14.2	22.2	15
		4	5.3	7.6	13.9	15.4	17
26/06/85	24	1	5.1	7.3	12.1	27.0	22
		2	5.1	7.8	12.2	25.5	25
		3	5.1	7.3	12.1	27.1	16
		4	5.1	7.3	11.9	19.7	
27/06/85	25	1	5.2	7.5	9.2	22.3	22
		2	5.2	7.6	9.4	21.5	26
		3	5.2	7.3	9.3	22.6	14
		4	5.2	7.2	9.1	18.0	20
28/06/85	26	1	5.3	7.5	11.3	25.8	21
		2	5.3	7.7	11.6	25.9	26
		3	5.3	7.4	10.5	26.4	12
		4	5.3	7.4	11.2	21.6	20
29/06/85	27	1	5.3	7.6	13.1	27.2	23
		2	5.3	7.5	12.9	26.0	25
		3	5.3	7.5	13.5	26.1	10
		4	5.3	7.5	13.1	22.0	10
30/06/85	28	1	5.3	7.4	10.1	21.6	21
		2	5.3	7.3	10.6	20.2	25
		3	5.3	7.6	11.7	21.3	16
		4	5.3	7.3	9.5	17.7	18

A-6

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
01/06/85	29	Reactors shut off for 24 hours to measure sludge content.					
02/07/85	30	1	5.0	7.7	7.8	18.6	22
		2	5.0	7.7	8.7	16.3	25
		3	5.0	7.4	8.6	18.1	23
		4	5.0	7.5	9.0	17.7	-
03/07/85	31	1	4.9	7.6	9.1	23.8	23
		2	4.9	7.6	9.5	24.1	23
		3	4.9	6.5	9.2	15.1	23
		4	4.9	7.4	9.6	20.8	17
04/07/85	32	1	4.9	7.6	9.4	26.5	23
		2	4.9	7.6	10.0	25.4	23
		3	4.9	7.0	10.6	22.6	16
		4	4.9	7.4	10.1	22.4	19
05/07/85	33	1	4.7	7.6	-	30.5	23
		2	4.7	7.6	-	17.7	24
		3	4.7	7.4	-	29.7	23
		4	4.7	7.5	-	25.2	20
06/07/85	34	1	4.6	7.1	15.9	48.9	22
		2	4.6	7.5	13.2	-	24
		3	4.6	6.5	14.2	33.9	26
		4	4.6	7.4	14.4	42.9	20
06/07/85	35	1	4.7	7.1	11.4	45.6	26
		2	4.7	7.3	11.3	-	23
		3	4.7	7.3	11.1	46.5	23
		4	4.7	7.5	11.0	38.3	20

A-7

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
08/07/85	36	1	4.7	7.3	6.3	26.4	19
		2	4.7	7.7	6.2	-	20
		3	4.7	7.4	6.3	28.0	15
		4	4.7	7.3	23.0	0.0	
09/07/85	37	1	4.8	7.4	12.8	49.0	22
		2	4.8	7.3	11.5	6.0	22
		3	4.8	7.3	12.3	50.8	17
		4	4.8	7.3	11.7	37.7	16
10/07/85	38	1	4.6	7.5	13.4	52.4	23
		2	4.6	7.2	12.8	52.6	23
		3	4.6	7.4	13.8	52.6	16
		4	4.6	2.3	13.2	42.3	14
11/07/85	39	1	4.6	8.0	6.2	22.4	18
		2	4.6	7.1	14.2	49.5	23
		3	4.6	7.1	14.5	45.7	13
		4	4.6	7.2	14.7	39.3	13
12/07/85	40	1	4.6	7.6	14.2	47.4	23
		2	4.6	7.3	13.4	51.7	23
		3	4.6	7.2	14.1	47.9	12
		4	4.6	7.2	14.0	40.0	13
13/07/85	41	1	4.6	7.2	16.9	51.7	22
		2	4.6	7.0	15.2	56.5	22
		3	4.6	7.0	16.9	50.6	11
		4	4.6	6.9	16.9	42.1	12
14/07/85	42	1	4.6	7.7	16.5	53.6	21
		2	4.6	7.5	14.5	57.2	22
		3	4.6	7.5	15.7	52.8	11
		4	4.6	7.4	1.6	43.6	13

A-8

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
15/07/85	43	1	4.6	7.3	10.7	36.6	22
		2	4.6	7.4	10.0	37.5	22
		3	4.6	7.3	10.5	34.6	13
		4	4.6	7.3	10.0	29.8	16
16/07/85	44	1	4.4	7.3	14.8	38.4	18
		2	4.4	7.5	12.5	83.4	21
		3	4.4	7.3	14.8	45.7	9
		4	4.4	7.2	14.7	37.3	16
17/07/85	45	1	4.5	7.2	15.5	40.5	22
		2	4.5	7.5	13.2	46.5	23
		3	4.5	7.3	15.3	45.4	11
		4	4.5	2.2	15.1	28.5	14
18/07/85	46	1	4.5	7.0	14.3	42.5	
		2	4.5	7.5	11.9	50.1	
		3	4.5	7.3	13.7	47.1	
		4	4.5	7.6	4.0	0	
19/07/85	47	1	4.5	7.0	1.60	51.5	17
		2	4.5	7.4	13.5	58.5	16
		3	4.5	7.2	15.2	54.1	9
		4	4.5	7.2	1.5	0	11
20/07/85	48	1	4.5	7.0	15.6	51.8	18
		2	4.5	7.3	14.4	57.4	14
		3	4.5	7.2	15.1	56.2	10
		4	4.5	7.1	15.3	48.1	13
21/07/85	49	1	4.6	7.1	15.7	53.2	19
		2	4.6	7.2	14.3	60.1	16
		3	4.6	7.1	15.4	58.2	13
		4	4.6	7.0	15.1	49.9	14

A-9

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
22/07/85	50	1	4.6	7.0	20.9	22.8	23
		2	4.6	7.2	22.1	39.7	23
		3	4.6	7.1	18.6	36.6	17
		4	4.6	6.8	19.2	-	20
23/07/85	51	1	4.6	6.8	18.5	20.0	22
		2	4.6	7.2	15.0	41.5	23
		3	4.6	7.1	17.6	34.3	16
		4	4.6	6.5	20.8	-	16
24/07/85	52	1		5.1	1.70	24.7	19
		2		6.9	15.2	55.9	21
		3		6.2	16.0	45.5	23
		4		4.6	4.0	-	22
25/07/85	53	1	4.6	5.1	14.7	21.7	22
		2	4.6	7.1	13.7	42.0	22
		3	4.6	6.7	15.7	38.3	17
		4	4.6	4.9	.3	.1	22
26/07/85	54	1	4.6	4.9	19.7	30.6	22
		2	4.6	7.0	18.3	53.6	22
		3	4.6	6.8	19.4	51.3	12
		4	4.6	4.8	3.6	.1	18
27/07/85	55	1	4.6	5.0	18.2	45.4	22
		2	4.6	7.0	19.5	68.6	21
		3	4.6	6.8	18.3	61.4	12
		4	4.6	4.9	19.2	12.0	24
28/07/85	56	1	4.6	4.9	20.2	39.2	22
		2	4.6	6.9	19.8	62.8	23
		3	4.6	6.8	20.1	58.0	12
		4	4.6	4.6	20.2	16.5	-

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)	
29/07/85	57	Reactors shut off for 24 hours to measure sludge content.						
30/07/85	58	1	5.1	6.1	15.1	29.8	22	
		2	5.1	7.2	15.1	45.3	22	
		3	5.1	7.6	6.0	39.4	17	
		4	5.1	6.5	12.8	3.1	23	
31/07/85	59	1	5.1	6.0	-	38.2	22	
		2	5.1	6.9	-	48.4	23	
		3	5.1	7.1	-	44.5	28	
		4	5.1	6.6	-	31.0	27	
A-10 01/08/85	60	1	5.0	5.5	-	37.1	23	
		2	5.0	7.1	-	55.9	22	
		3	5.0	6.8	-	41.1	18	
		4	5.0	7.6	-	9.1	22	
02/08/85	61	1	5.1	6.7	18.6	65.2	22	
		2	5.1	7.2	17.2	26.9	23	
		3	5.1	7.1	17.9	53.1	17	
		4	5.1	7.4	18.3	8.6	22	
03/08/85	62	1	5.1	6.8	16.4	56.2	23	
		2	5.1	6.9	16.8	44.4	23	
		3	5.1	7.1	16.3	50.1	18	
		4	5.1	7.4	16.7	51.3	22	
04/08/85	63	Reactors shut off for 24 hours to measure sludge content.						
05/08/85	64	1	5.1	7.0	19.2	54.1	21	
		2	5.0	7.2	19.4	47.3	23	
		3	5.0	7.1	18.9	52.8	22	
		4	5.0	7.0	19.6	49.8	21	

A-11

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
06/08/85	65	1	5.1	6.8	20.1	30.9	
		2	5.1	6.7	19.9	55.1	
		3	5.1	6.8	20.0	44.2	
		4	5.1	5.9	20.0	53.6	
07/08/85	66	1	5.0	5.6	20.2	32.1	
		2	5.0	6.7	20.1	54.6	
		3	5.0	5.5	19.9	20.0	
		4	5.0	5.2	20.2	34.2	
08/08/85	67	1	5.8	6.2	18	30.7	19
		2	5.8	7.2	17	58.8	22
		3	5.8	5.9	19	16.2	17
		4	5.8	6.4	18	27.9	22
09/08/85	68	1	5.6	6.4	20.5	34.9	-
		2	5.8	6.4	19.8	29.4	22
		3	5.6	6.1	20	14.9	-
		4	5.6	6.3	20.2	15.2	-
10/08/85	69	1	5.6	6.6	18	35.5	15
		2	5.6	6.8	18.2	50.1	-
		3	5.6	6.7	17.8	37.3	12
		4	5.6	6.7	18	45.3	20
11/08/85	70	1	6.4	6.7	20.2	59.2	15
		2	6.4	7.2	20.5	105.3	-
		3	6.4	7.1	19.8	86.4	16
		4	6.4	7.2		78.5	-
12/08/85	71	1	6.6	6.8	20.1	26.6	16
		2	6.6	7.4	19.8	53.7	22
		3	6.6	7.1	19.6	46.5	16
		4	6.6	7.3	19.6	10.4	-



A-12

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
13/08/85	72	1	5.4	6.8	35.2	56.9	14
		2	5.4	6.9	35.0	69.0	20
		3	5.4	7.1	35.4	38.4	12
		4	5.4	7.0	35.8	42.8	19
14/08/85	73	1	5.4	6.9	29.8	52.8	15
		2	5.4	6.9	29.6	54.6	21
		3	5.4	7.0	29.8	42.2	14
		4	5.4	7.0	29.4	43.5	19
15/08/85	74	1	5.4	7.2	19.8	45.8	13
		2	5.4	7.3	20.1	43.2	21
		3	5.4	7.0	19.6	38.8	13
		4	5.4	7.4	19.7	40.3	-
16/08/85	75	1	5.5	7.1	21.8	48.2	14
		2	5.4	7.1	21.4	47.2	20
		3	5.4	7.0	20.9	40.3	16
		4	5.4	7.3	20.6	44.4	-
17/08/85	76	1	5.4	7.0	22.1	49.6	-
		2	5.2	7.1	21.4	49.8	20
		3	5.4	7.1	20.8	43.2	15
		4	5.4	7.2	21.4	44.3	-
18/08/85	77	1	5.4	7.0	21.6	52.9	-
		2	5.4	7.0	21.4	51.8	19
		3	5.4	7.1	21.0	44.4	14
		4	5.4	7.1	21.6	44.5	-
19/08/85	78	1	4.8	6.6	14.8	27.4	16
		2	4.9	6.8	13.6	30.3	20
		3	4.9	6.7	13.2	24.8	-
		4	4.9	6.3	13.8	28.9	15

A-13

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
20/08/85	79	1	4.9	6.4	26.8	57.2	14
		2	4.9	6.7	27.1	60.1	20
		3	5.0	6.6	27.6	63.2	-
		4	4.5	6.0	27.8	40.3	-
21/08/85	80	1	4.9	6.2	26.4	61.2	12
		2	4.9	6.6	27.1	55.6	21
		3	5.0	6.5	28.0	74.1	-
		4	4.9	5.8	27.2	35.2	14
22/08/85	81	1	4.9	6.3	26.8	64.2	-
		2	4.9	6.5	27.1	57.9	-
		3	4.8	6.4	26.9	69.9	-
		4	4.9	5.2	27.0	23.9	14
23/08/85	82	1	5.4	6.6	21	46.2	12
		2	5.4	6.8	21	43.3	12
		3	5.4	7.0	23	58.4	13
		4	5.5	5.0	22	41.3	14
24/08/85	83	1	4.9	6.5	?	56.5	-
		2	5.0	6.6	?	51.3	-
		3	5.0	6.6	?	73.2	11
		4	4.9	5.3	?	36.1	13
25/08/85	84	1	5.0	6.6	?	61.4	-
		2	5.0	6.6	?	48.5	-
		3	5.0	6.6	?	71.1	10
		4	5.0	5.3	?	38.8	13
26/08/85	85	1	5.0	6.6	9.2	23.2	-
		2	5.0	6.7	9.4	21.5	21
		3	5.0	6.7	9.2	26.3	-
		4	5.0	5.3	9.3	19.4	15

A-14

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)
27/08/85	86	1	5.0	6.7	31.2	50.6	-
		2	5.0	6.8	30.8	49.3	21
		3	5.1	6.7	32.2	41.3	14
		4	5.0	6.8	30.9	37.9	17
28/08/85	87	1	4.5	6.6	24.1	54.7	10
		2	4.6	6.8	23.7	54.9	18
		3	4.6	6.8	24.2	60.6	-
		4	4.5	6.5	23.6	47.1	12
29/08/85	88	1	4.5	6.7	30.9	52.9	10
		2	4.5	6.8	31.8	55.6	19
		3	4.6	6.8	31.9	61.3	-
		4	4.5	6.8	32.3	48.4	-
30/08/85	89	1	4.5	6.9	31.2	53.9	10
		2	4.5	7.0	31.4	56.6	-
		3	4.5	7.1	30.9	50.9	-
		4	4.5	7.0	30.8	45.4	14
31/08/85	90	1	4.5	6.9	16.2	27.8	14
		2	4.5	7.1	16.4	26.1	16
		3	4.5	7.0	16.4	25.1	11
		4	4.5	7.0	16.3	21.2	12
01/09/85	91	1	4.5	7.0	28.1	46.3	-
		2	4.5	7.0	28.2	44.2	21
		3	4.5	6.9	28.2	34.3	-
		4	4.5	7.0	28.1	37.2	-
02/09/85	92	1	4.8	6.9	41.2	44.4	-
		2	4.8	6.8	41.4	43.3	20
		3	4.7	6.8	41.5	31.3	-
		4	4.8	6.8	41.0	40.0	-

Date (D/M/Y)	Day	Reactor	pH <sub>i</sub>	pH <sub>e</sub>	Effluent Volume (L)	Gas Volume (L)	Sludge Bed Bed Height (cm)	
03/09/85	93	1	4.8	6.4	38.2	43.3	-	
		2	4.8	6.6	38.4	40.2	19	
		3	4.7	6.3	38.2	34.2	-	
		4	4.8	6.3	38.1	38.2	-	
04/09/85	94	Reactors shut off.						
05/09/85	95	Reactors shut off.						

Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>Sj</sub> (mg/L)	COD <sub>Se</sub> (mg/L)	SOC <sub>j</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
03/06/85	1	1											
		2											
		3											
		4											
04/06/85	2	1	180	140	3800	1860							
		2	167	133	3800	1460							
		3	247	133	3800	1880							
		4	280	173	3800	1880							
05/06/85	3	1											
		2											
		3											
		4											
06/06/85	4	1	200	127	3800	1900			5/75/15				
		2	153	127	3800	1200			-/75/15				
		3	220	180	3800	1880			-/75/20				
		4	233	213	3800	1560			-/75/20				
07/06/85	5	1										3150	
		2										2460	
		3										2980	
		4										3340	
08/06/85	6	1											
		2											
		3											
		4											
09/06/85	7	1								1990	0.30		40.2
		2								2350	0.35		50.4
		3								1860	0.37		45.1
		4								2020	0.31		38.3

Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>si</sub> (mg/L)	COD <sub>se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
10/06/85	8	1 2 3 4											
11/06/85	9	1 2 3 4	233 187 213 260	140 100 140 167	4400 4400 4400 4400	2000 1570 2000 2010							
12/06/85	10	1 2 3 4							5/75/15 -/75/15 -/75/15 -/75/15				
13/06/85	11	1 2 3 4	207 273 267 260	127 187 160 133	5750 5750 5750 5750	2010 760 2000 2010	1772 1772 1772 1772	1184 290 978 857					
14/06/85	12	1 2 3 4											
15/06/85	13	1 2 3 4								1350 1650 1425 1525			
16/06/85	14	1 2 3 4									0.19 0.27 0.21 0.21		

Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>si</sub> (mg/L)	COD <sub>se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
17/06/85	15	1							-/60/10				
		2							-/75/15				
		3							-/75/15				
		4							-/70/15				
18/06/85	16	1	313	253	4800	2000	1708	745					
		2	200	127	4800	940	1708	337					
		3	340	220	4800	2000	1708	768					
		4	300	200	4800	2080	1708	780					
19/06/85	17	1										160	
		2										120	
		3										170	
		4										190	
20/06/85	18	1	247	193	4500	1960	1705	749					
		2	207	200	4500	880	1705	335					
		3	293	260	4500	2000	1705	767					
		4	213	153	4500	1980	1705	777					
21/06/85	19	1								1375			
		2								1700			
		3								1450			
		4								1575			
22/06/85	20	1											
		2											
		3											
		4											
23/06/85	21	1									.45	26.0	
		2									.35	45.7	
		3									.33	30.3	
		4									.23	23.5	

Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>si</sub> (mg/L)	COD <sub>se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
24/06/85	22	1							5/85/15 5/80/20 -/85/15 -/80/15				
		2											
		3											
		4											
25/06/85	23	1	207	93	4700	1020	1601	362					
		2	153	100	4700	540	1601	191					
		3	200	140	4700	1280	1601	597					
		4	353	200	4700	1500	1601	597					
26/06/85	24	1										100	
		2										120	
		3										160	
		4										170	
27/06/85	25	1	247	187	5100	860	1627	299					
		2	113	73	5100	500	1627	182					
		3	160	113	5100	1060	1627	360					
		4	187	140	5100	1280	1627	488					
28/06/85	26	1								1350			
		2								1725			
		3								1400			
		4								1525			
29/06/85	27	1											
		2											
		3											
		4											
30/06/85	28	1									0.45	46.1	
		2									0.48	47.4	
		3									0.47	32.7	
		4									0.38	29.9	



Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>sj</sub> (mg/L)	COD <sub>se</sub> (mg/L)	SOC <sub>j</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD <sub>R</sub>	VFA (mg/L)	VSS in Reactor (g)	
01/06/85	29	Reactors shut off for 24 hours to measure sludge content												
02/07/85	30	1	173	113	2950	720	1459	558						
		2	133	80	2950	520	1459	473						
		3	240	153	2950	2060	1459	991						
		4	253	180	2950	900	1459	595						
03/07/85	31	1										380		
		2										320		
		3											710	
		4											640	
04/07/85	32	1	187	153	3100	660	1435	204						
		2	80	67	3100	400	1435	130						
		3	147	100	3100	1500	1435	503						
		4	133	107	3100	730	1435	258						
05/07/85	33	1								980				
		2								970				
		3									1260			
		4									900			
06/07/85	34	1												
		2												
		3												
		4												
07/07/85	35	1												
		2									.62			
		3									.58			
		4									.80			
										.56				

Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>si</sub> (mg/L)	COD <sub>se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
08/07/85	36	1							-/70/25				
		2							-/65/30				
		3							-/65/30				
		4							-/65/30				
09/07/85	37	1	267	260	-	-	2015	332					
		2	167	160	-	-	2015	327					
		3	240	233	-	-	2015	387					
		4	253	133	-	-	2015	366					
10/07/85	38	1										320	
		2										320	
		3										540	
		4										480	
11/07/85	39	1	207	173	-	-	2060	223					
		2	153	133	-	-	2060	301					
		3	167	127	-	-	2060	393					
		4	133	100	-	-	2060	433					
12/07/85	40	1								1100			
		2								1000			
		3								1400			
		4								1100			
13/07/85	41	1											51.5
		2											50.8
		3											43.1
		4											31.6
14/07/85	42	1									.45		
		2									.45		
		3										.46	
		4										.38	

Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>Si</sub> (mg/L)	COD <sub>Se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
15/07/85	43	1								-/65/30			
		2								-/70/30			
		3								-/65/35			
		4								-/75/25			
16/07/85	44	1	147	133	4200	940	1622	340					
		2	147	114	4200	480	1622	181					
		3	107	93	4200	850	1622	290					
		4	107	100	4200	890	1622	297					
17/07/85	45	1										1100	
		2										420	
		3										310	
		4										1240	
18/07/85	46	1	53	20	4100	980	2001	486					
		2	107	47	4100	410	2001	152					
		3	160	80	4100	740	2001	245					
		4	113	73	4100	590	2001	186					
19/07/85	47	1								800			
		2								1000			
		3								800			
		4								600			
20/07/85	48	1											42.2
		2											44.2
		3											39.7
		4											32.8
21/07/85	49	1									.43		
		2									.53		
		3									.47		
		4									.38		

Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>Si</sub> (mg/L)	COD <sub>Se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
22/07/85	50	1								-/60/30			
		2								-/65/30			
		3								-/60/35			
		4								-/75/20			
23/07/85	51	1	200	167	4900	1800	-	650					
		2	207	160	4900	860	-	326					
		3	387	287	4900	1500	-	-					
		4	527	320	4900	1840	-	-					
24/07/85	52	1										1800	
		2										540	
		3										440	
		4										1380	
25/07/85	53	1	220	107	4300	1900	1912	749					
		2	160	87	4300	1020	1912	387					
		3	513	293	4300	1840	1912	712					
		4	540	400	4300	1900	1912	1301					
26/07/85	54	1								600			
		2								900			
		3								800			
		4								500			
27/07/85	55	1											62.8
		2											65.7
		3											58.4
		4											59.9
28/07/85	56	1							-/70/30		.22		
		2							-/75/25		.43		
		3							-/70/25		.36		
		4							-/70/25		-		

Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>Si</sub> (mg/L)	COD <sub>Se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> /CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
29/07/85	57	Reactors shut off for 24 hours to measure sludge content											
30/07/85	58	1	353	233	4800	1900	1613	812					
		2	487	307	4800	940	1613	283					
		3	1340	780	4800	1020	1613	276					
		4	353	247	4800	1920	1613	835					
31/07/85	59	1										2060	
		2										520	
		3										500	
		4										1150	
01/08/85	60	1	333	187	5200	2080	1755	1208					
		2	213	100	5200	1460	1735	449					
		3	320	187	5200	1740	1755	597					
		4	280	160	5200	1440	1755	414					
02/08/85	61	1								1320			
		2								1500			
		3								1500			
		4								1700			
03/08/85	62	1									.55		63.0
		2									.41		65.4
		3									.46		57.9
		4									.39		73.4
04/08/85	63	Reactors shut off for 24 hours to measure sludge content											
05/08/85	64	1							-/70/20				
		2							-/75/20				
		3							-/70/35				
		4							-/70/25				

Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>Sj</sub> (mg/L)	COD <sub>Se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
06/08/85	65	1	127	53	4750	1680	1756	565					
		2	140	73	4750	1940	1756	705					
		3	173	107	4750	1570	1756	564					
		4	173	107	4750	2130	1756	1021					
07/08/85	66	1										1860	
		2										1110	
		3										1950	
		4										1575	
08/08/85	67	1	160	120	5100	1620	1759	547					
		2	107	93	5100	1740	1759	543					
		3	267	213	5100	1700	1759	579					
		4	-	-	5100	1980	1759	1024					
09/08/85	68	1								1200			
		2								1600			
		3								1100			
		4								1300			
10/08/85	69	1											54.2
		2											55.7
		3											47.6
		4											62.2
11/08/85	70	1									.29		
		2									.39		
		3									.23		
		4									.30		
12/08/85	71	1											
		2											
		3											
		4											

Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>Si</sub> (mg/L)	COD <sub>Se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
13/08/85	72	1	200	167	4700	1400	1682	462					
		2	267	233	4700	1520	1682	538					
		3	343	287	4700	2210	1682	744					
		4	353	287	4700	2010	1682	921					
14/08/85	73	1										1210	
		2										860	
		3										940	
		4										1380	
15/08/85	74	1	223	167	5220	1250	1721	446					
		2	287	203	5220	1480	1721	521					
		3	413	353	5220	1850	1721	822					
		4	453	313	5220	1910	1721	867					
16/08/85	75	1								2700			
		2								2200			
		3								2900			
		4								3200			
17/08/85	76	1											
		2											
		3											
		4											
18/08/85	77	1									.28		46.8
		2									.35		48.9
		3									.25		40.3
		4									.33		51.7
19/08/85	78	1							-/65/30				
		2							-/70/20				
		3							-/70/25				
		4							-/75/20				

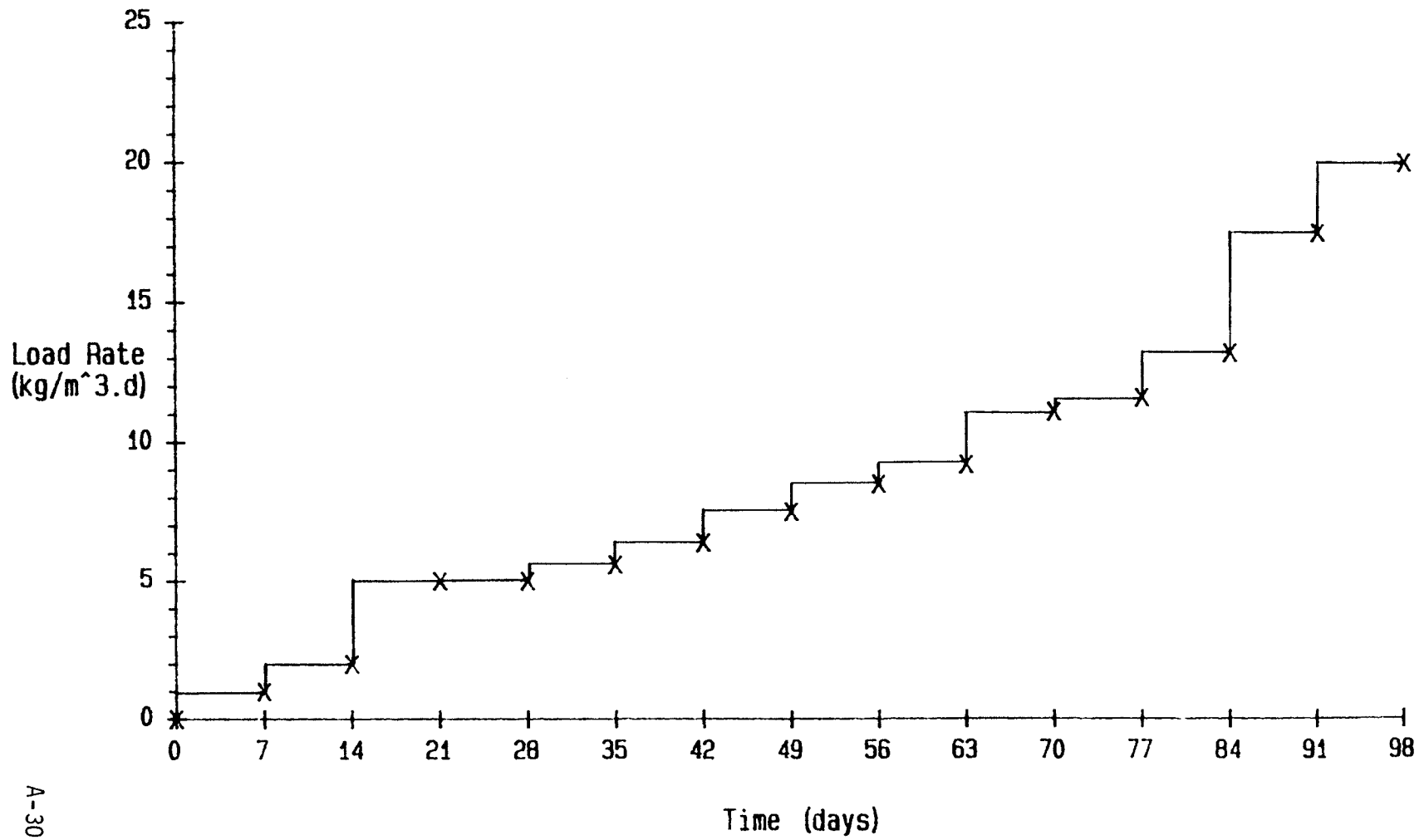
Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>Si</sub> (mg/L)	COD <sub>Se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD <sub>R</sub>	VFA (mg/L)	VSS in Reactor (g)
20/08/85	79	1	167	113	4220	1630	1546	528					
		2	327	267	4220	1810	1546	586					
		3	367	283	4220	2340	1546	841					
		4	413	307	4220	1960	1546	753					
21/08/85	80	1										1430	
		2										1580	
		3										1540	
		4										1710	
22/08/85	81	1	283	203	5050	1810	1738	458					
		2	343	267	5050	2050	1738	1054					
		3	367	323	5050	2130	1738	735					
		4	427	387	5050	2170	1738	701					
23/08/85	82	1								2400			
		2								2100			
		3								2600			
		4								2900			
24/08/85	83	1											
		2											
		3											
		4											
25/08/85	84	1									.35		39.2
		2									.38		40.7
		3									.43		31.6
		4									.30		38.8
26/08/85	85	1							-/60/20				
		2							-/65/30				
		3							-/65/30				
		4							-/65/25				



Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>Sj</sub> (mg/L)	COD <sub>Se</sub> (mg/L)	SOC <sub>j</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
27/08/85	86	1	283	203	5100	3820	1731	1337					
		2	367	333	5100	4260	1731	1496					
		3	433	387	5100	4100	1731	1528					
		4	527	467	5100	3950	1731	1583					
28/08/85	87	1								1950			
		2								2300			
		3								2460			
		4								2380			
29/08/85	88	1	383	327	5450	3650	1767	1410					
		2	413	333	5450	4130	1767	1562					
		3	427	283	5450	4380	1767	1443					
		4	467	433	5450	4250	1767	1489					
30/08/85	89	1								1840			
		2								1610			
		3								2130			
		4								2250			
31/08/85	90	1									.30		36.2
		2									.45		37.3
		3									.52		29.4
		4									.40		33.6
01/09/85	91	1							-/60/25				
		2							-/60/30				
		3							-/65/30				
		4							-/60/25				
02/09/85	92	1	367	333	5150	4530	1748						
		2	467	387	5150	4530	1748						
		3	533	433	5150	4530	1748						
		4	567	467	5150	4530	1748						

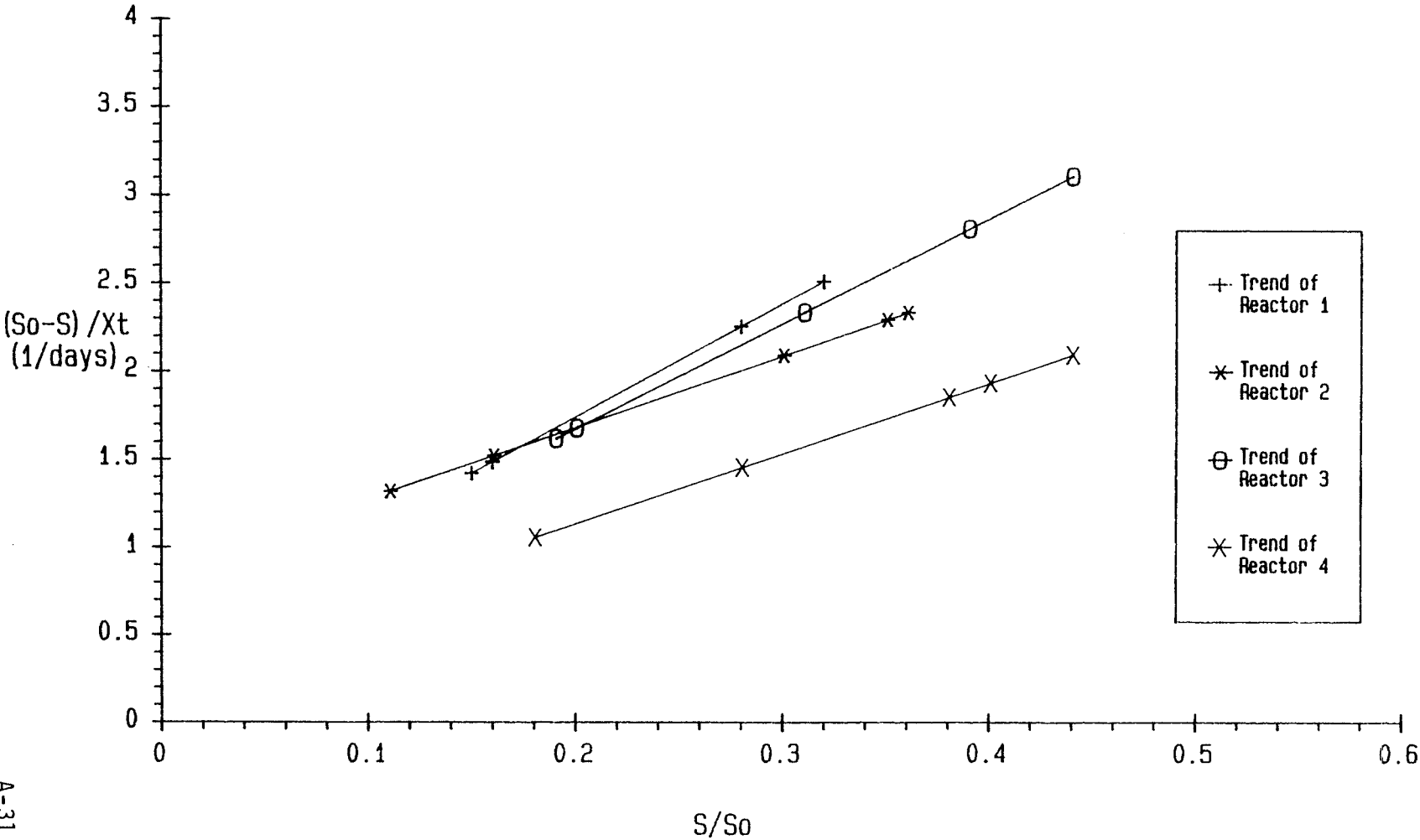
Date (D/M/Y)	Day	Reactor	TSS <sub>e</sub> (mg/L)	VSS <sub>e</sub> (mg/L)	COD <sub>Si</sub> (mg/L)	COD <sub>Se</sub> (mg/L)	SOC <sub>i</sub> (mg/L)	SOC <sub>e</sub> (mg/L)	Gas (%air/CH <sub>4</sub> / CO <sub>2</sub> )	Alkalinity (mg/L as CaCO <sub>3</sub> )	L CH <sub>4</sub> per g COD R	VFA (mg/L)	VSS in Reactor (g)
03/09/85	93	1										2120	
		2										2310	
		3										2590	
		4										2600	
04/09/85	94	1	343	267	5000	4350	1803	1590			.85		
		2	427	333	5000	4350	1803	1589			.81		
		3	533	467	5000	4550	1803	1659			1.02		
		4	413	333	5000	4450	1803	1676			.83		
05/09/85	95	1								1620			
		2								1560			
		3								1410			
		4								1130			
06/09/85	96	1											30.7
		2											32.4
		3											26.3
		4											30.1

# Load Rate vs Time



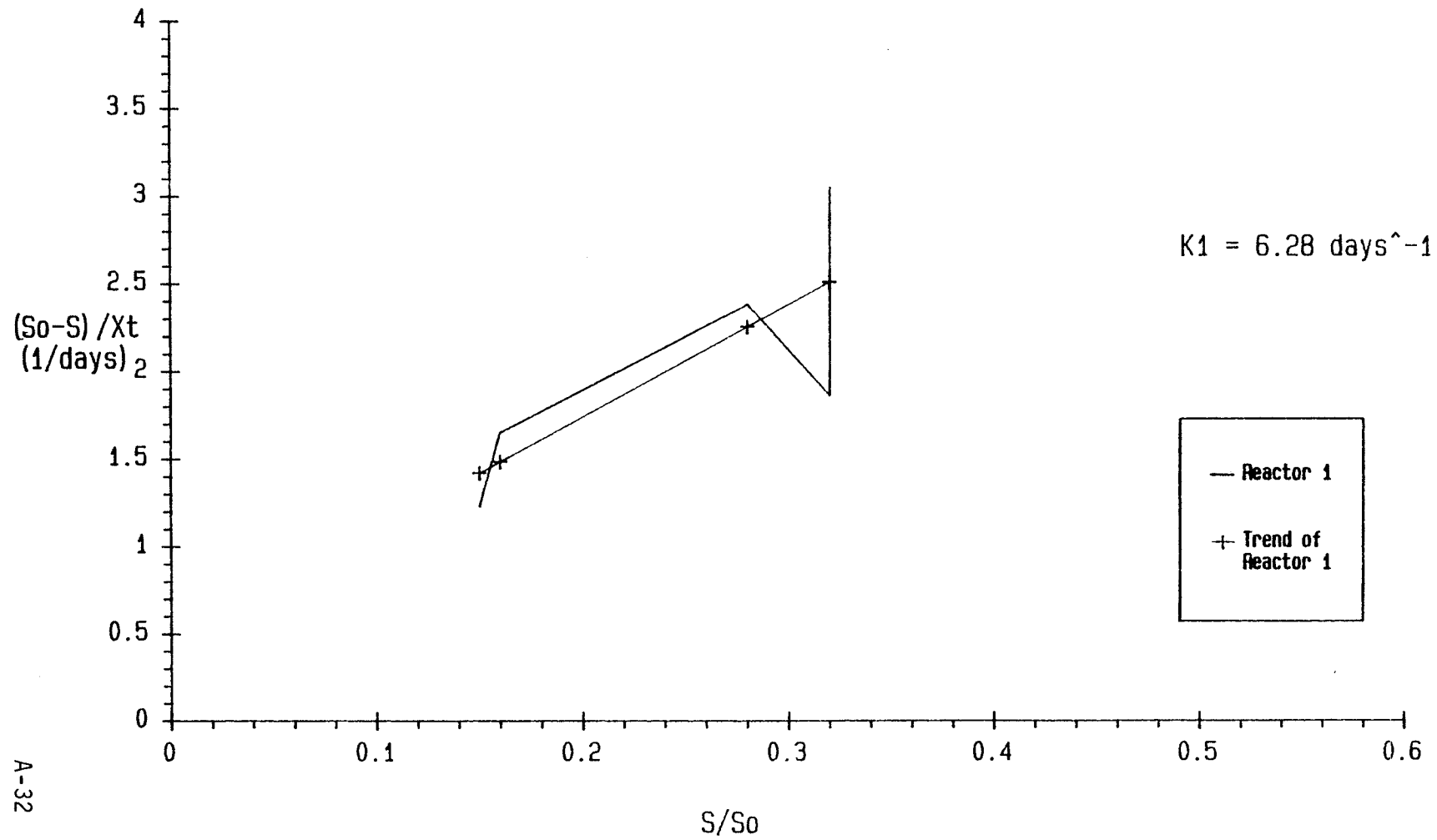
# Determination of Biokinetic Constant, K1

## Trends of all Reactors



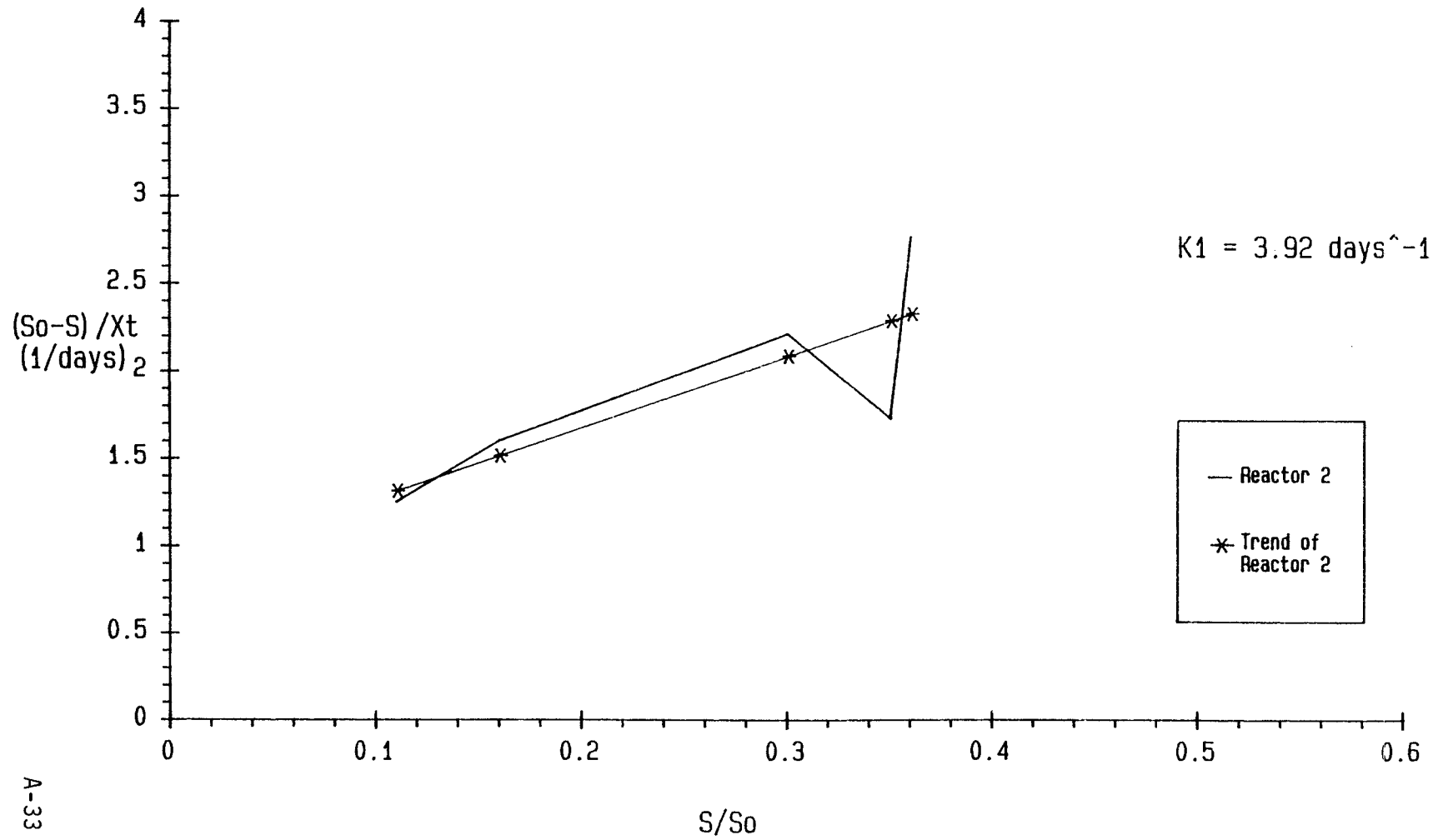
# Determination of Biokinetic Constant, K1

## Reactor 1



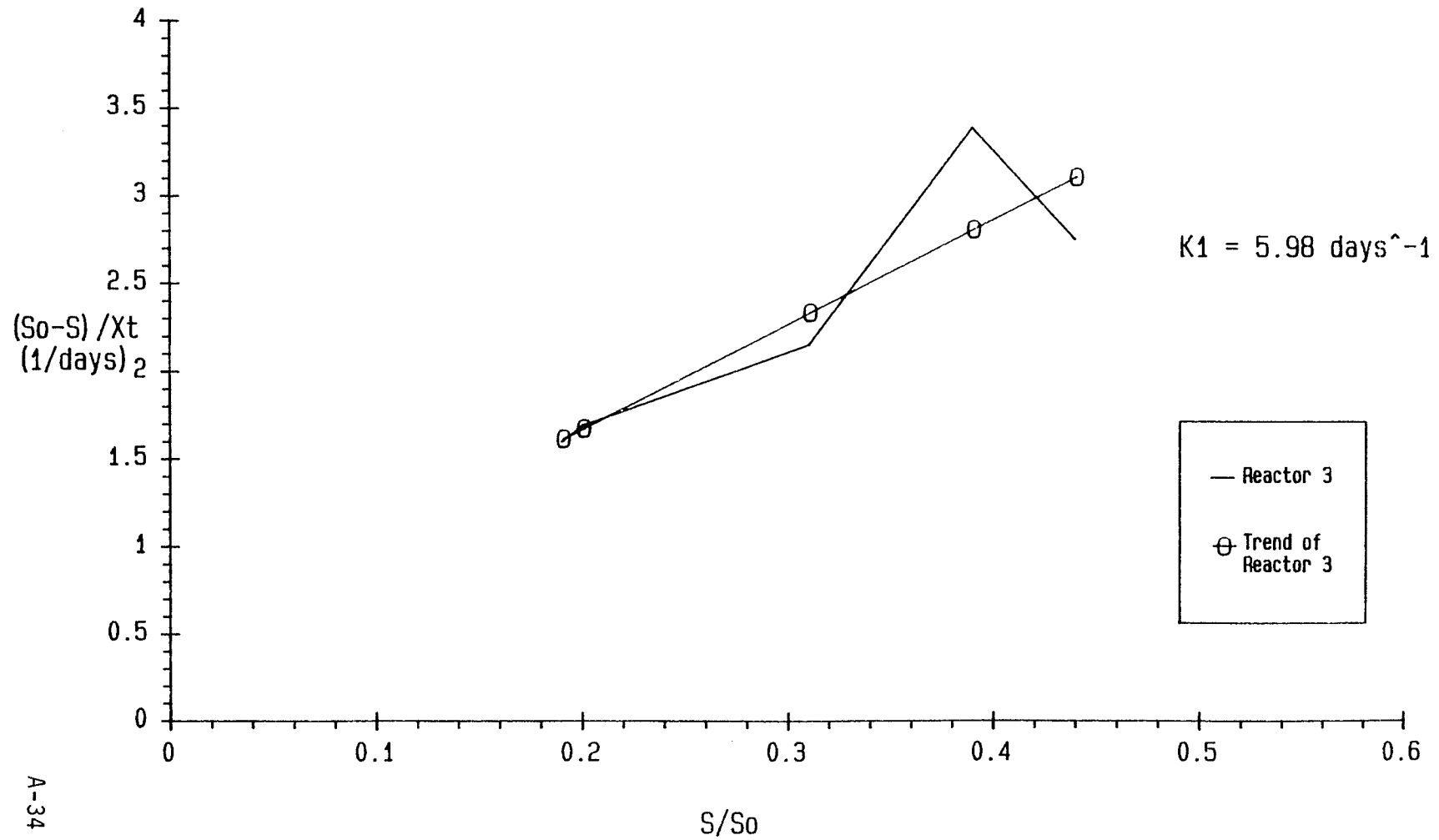
# Determination of Biokinetic Constant, K1

## Reactor 2



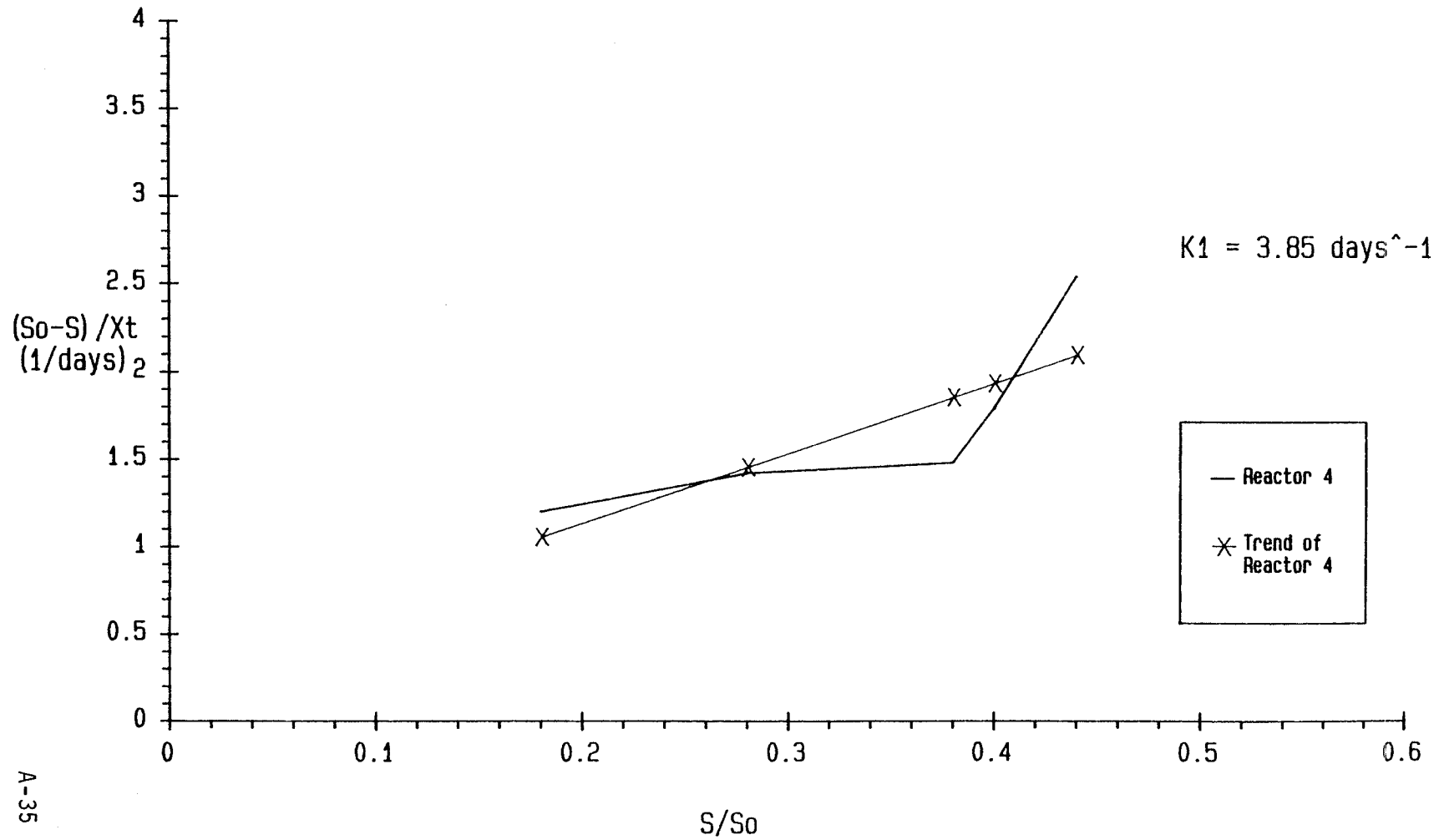
# Determination of Biokinetic Constant, K1

## Reactor 3



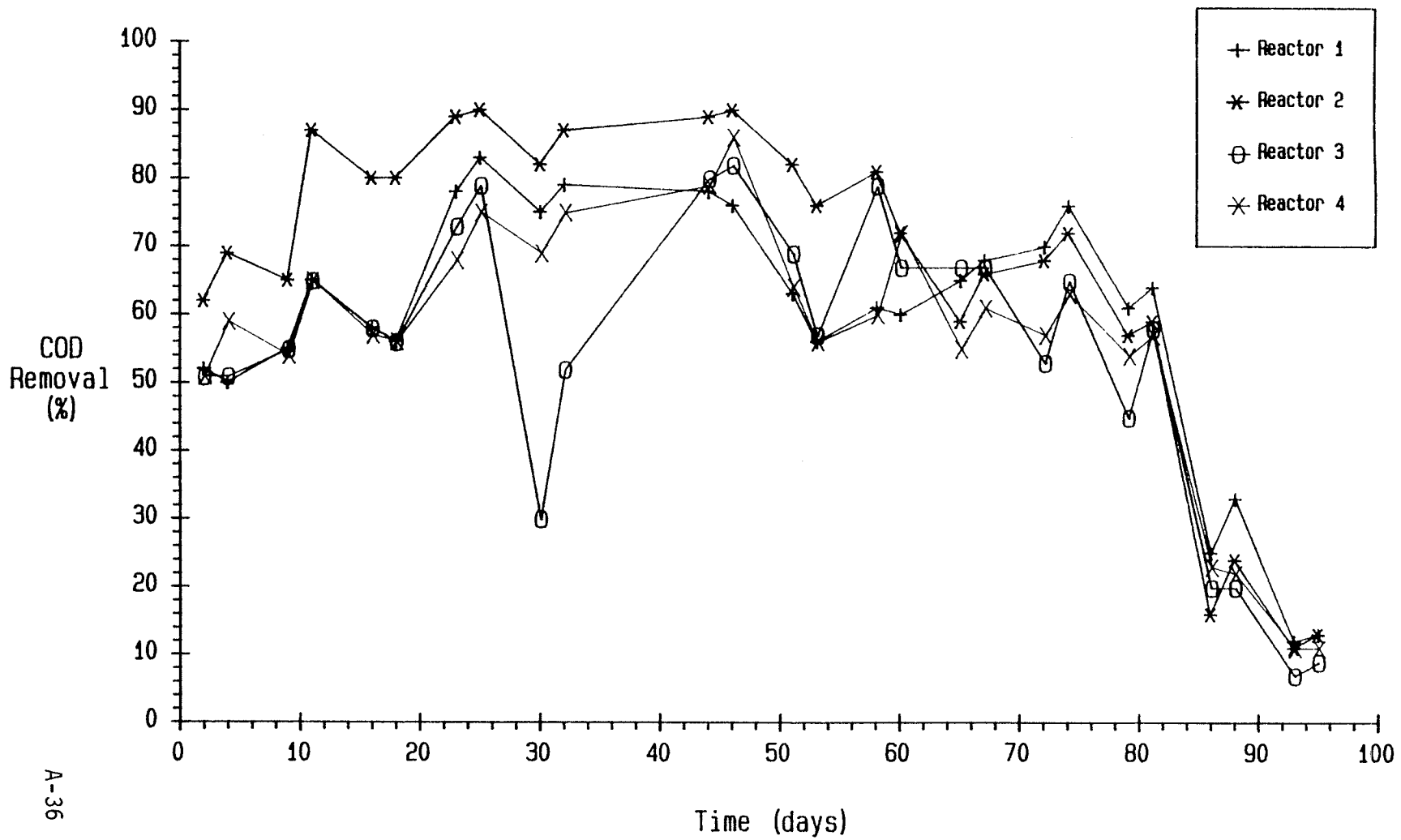
# Determination of Biokinetic Constant, K1

## Reactor 4



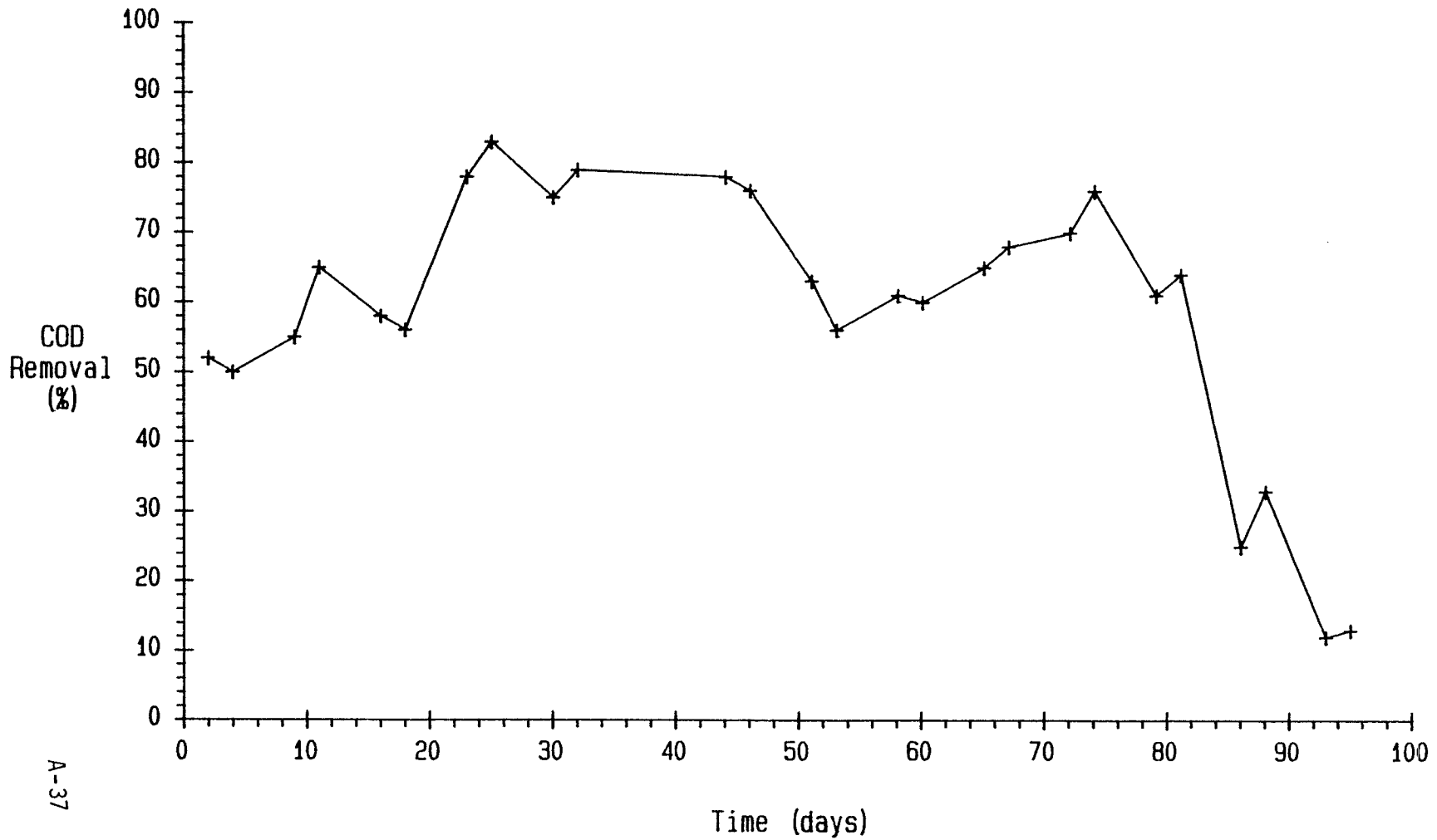


# Chemical Oxygen Demand Removal vs Time



# Chemical Oxygen Demand Removal vs Time

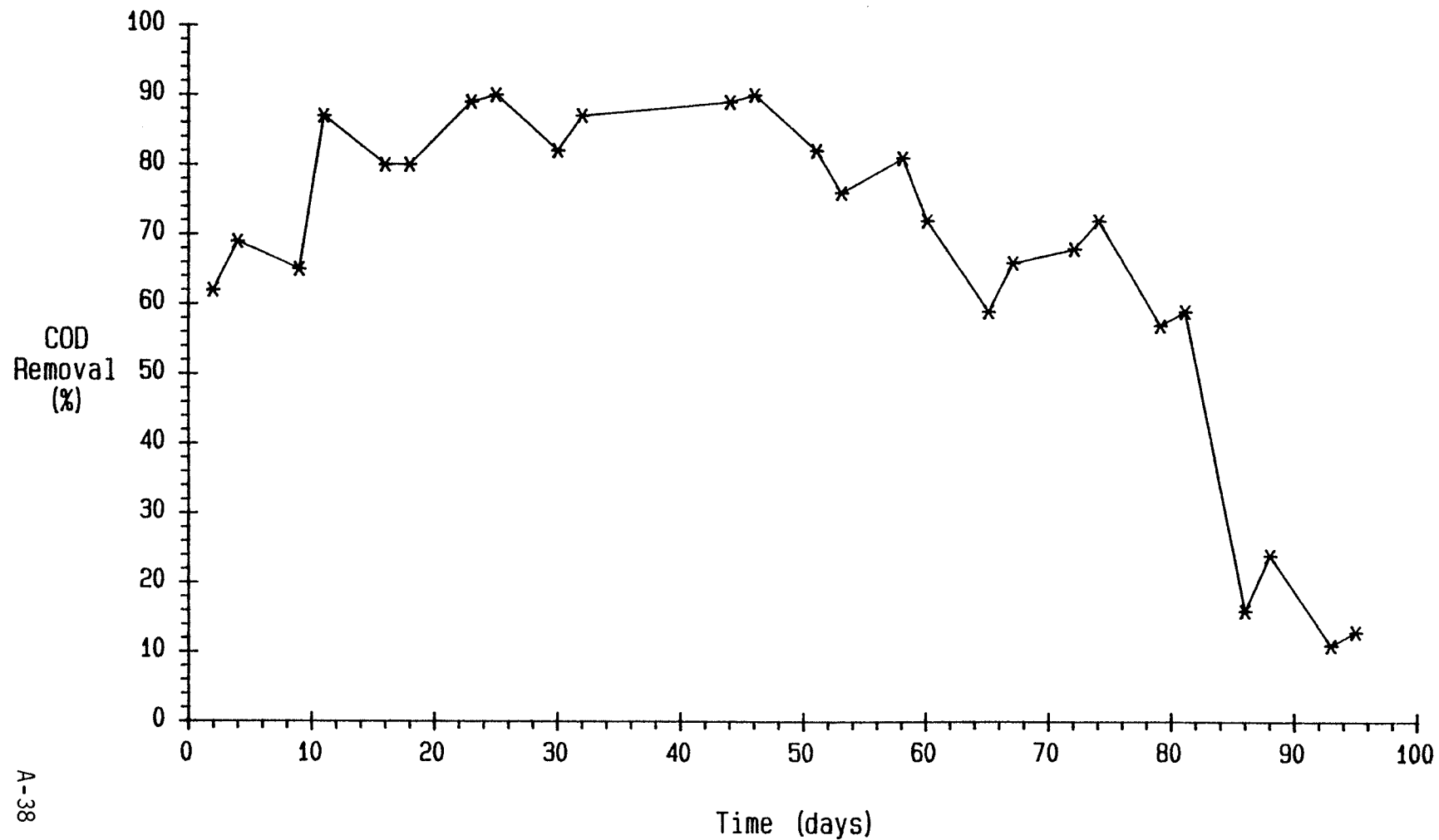
## Reactor 1



A-37

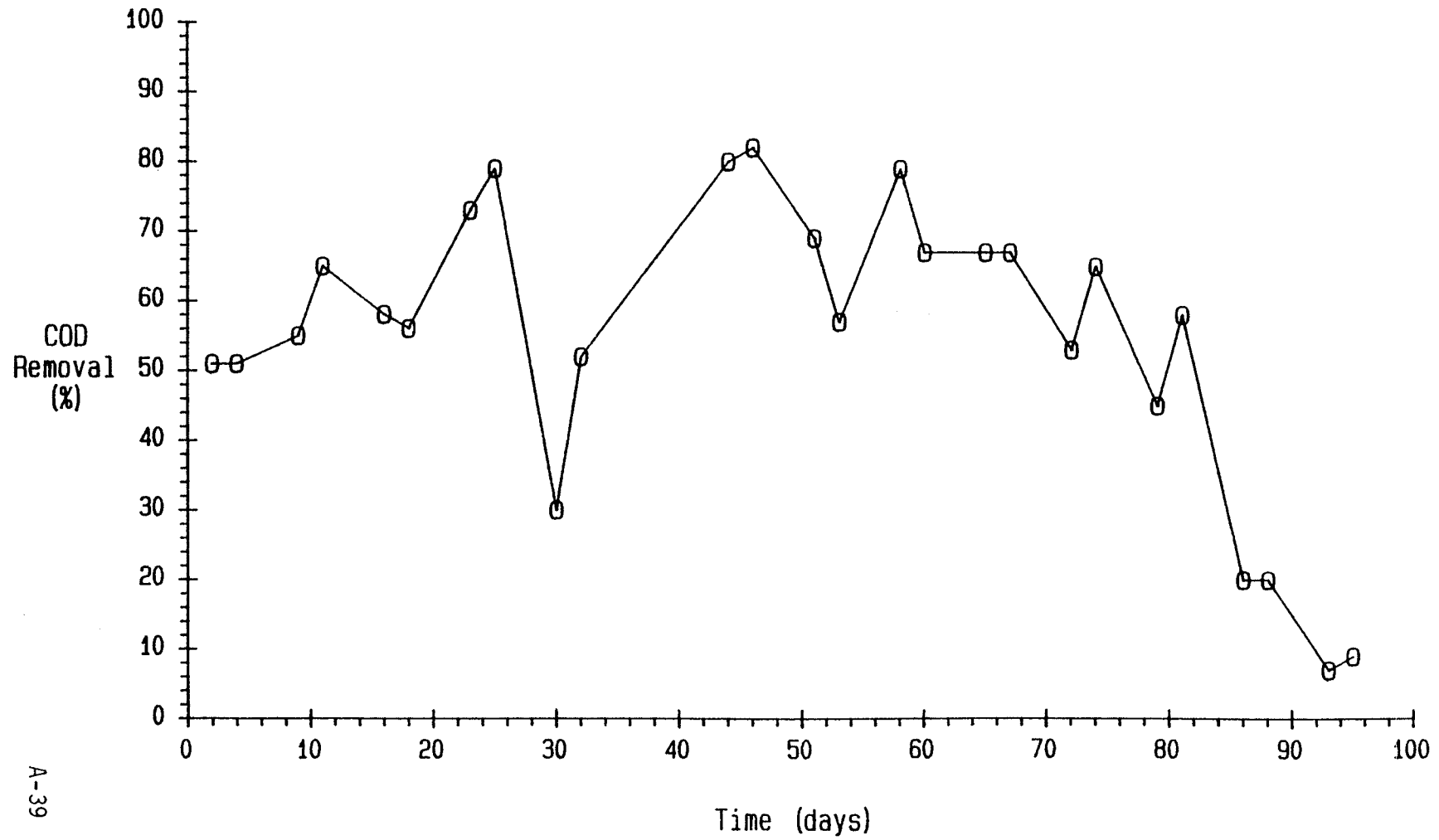
# Chemical Oxygen Demand Removal vs Time

## Reactor 2



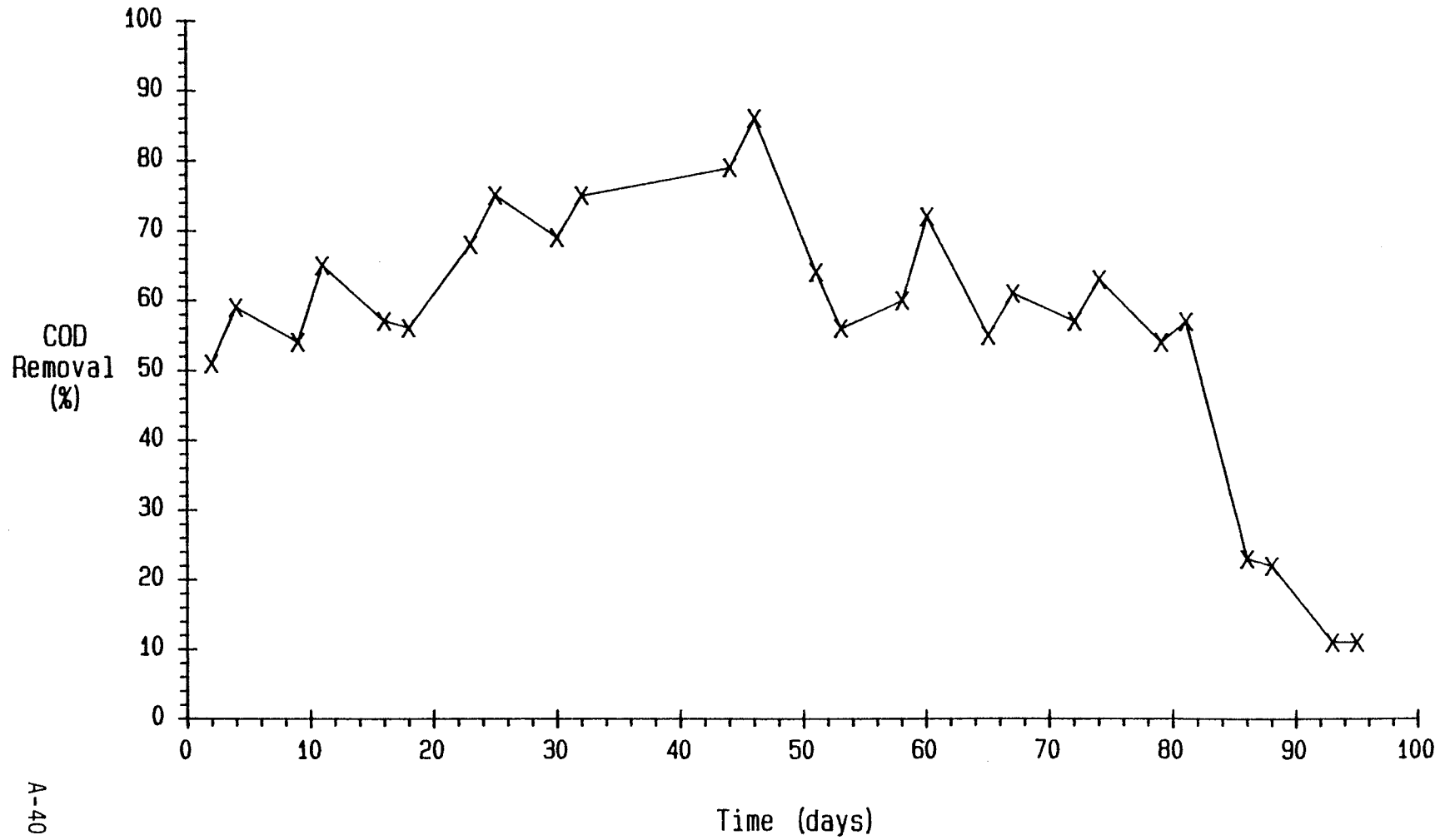
# Chemical Oxygen Demand Removal vs Time

## Reactor 3



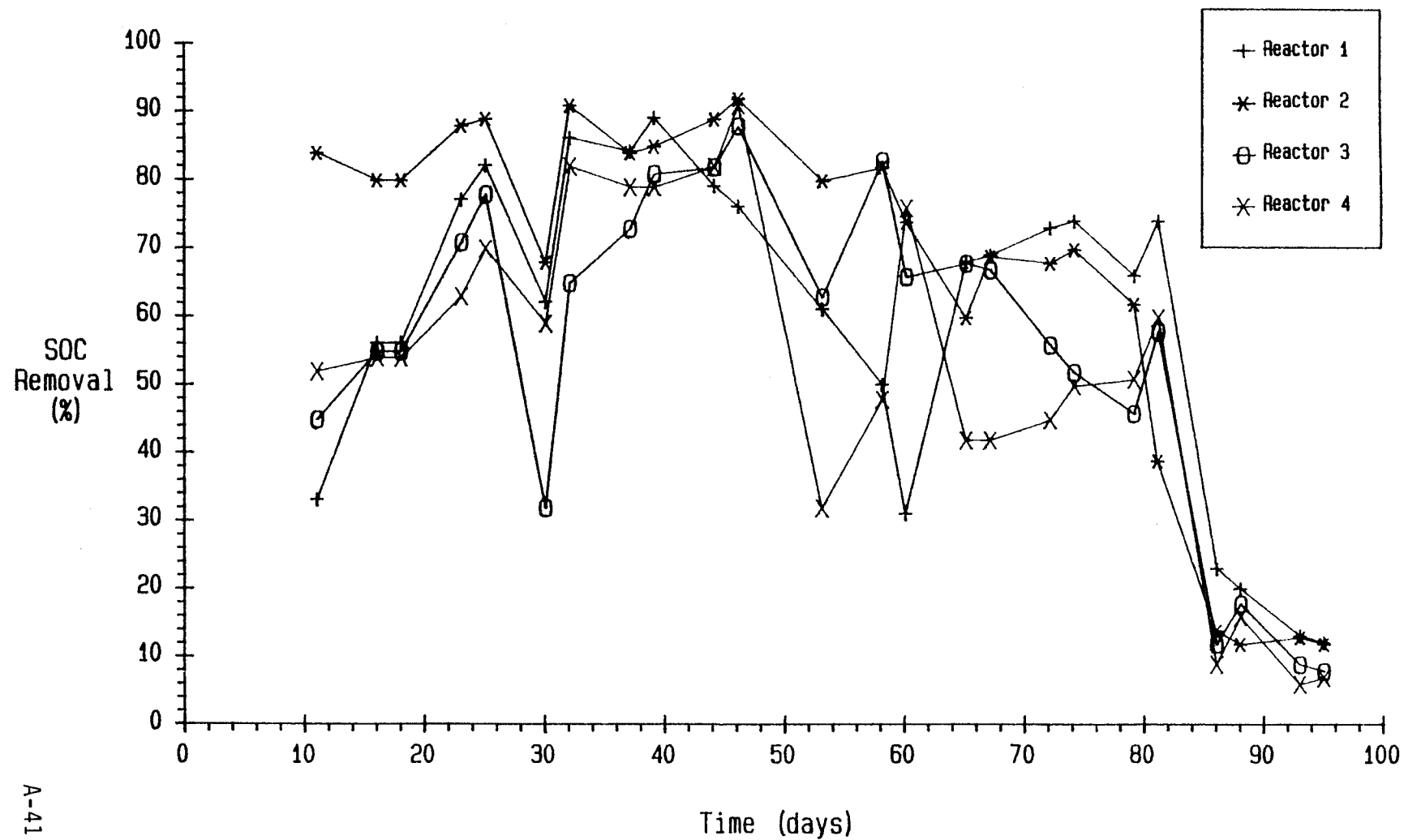
# Chemical Oxygen Demand Removal vs Time

## Reactor 4



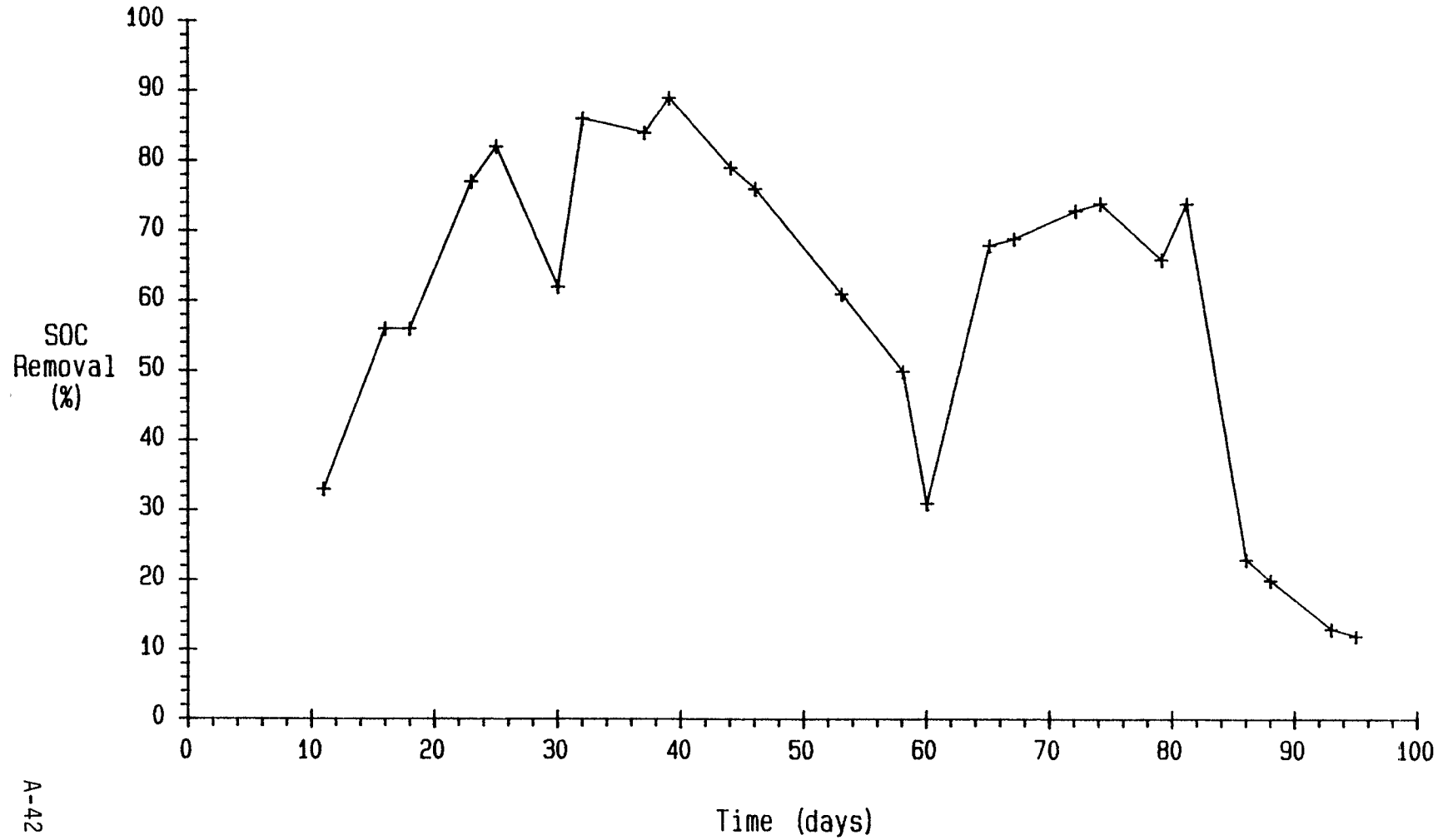
A-40

# Soluble Organic Carbon Removal vs Time



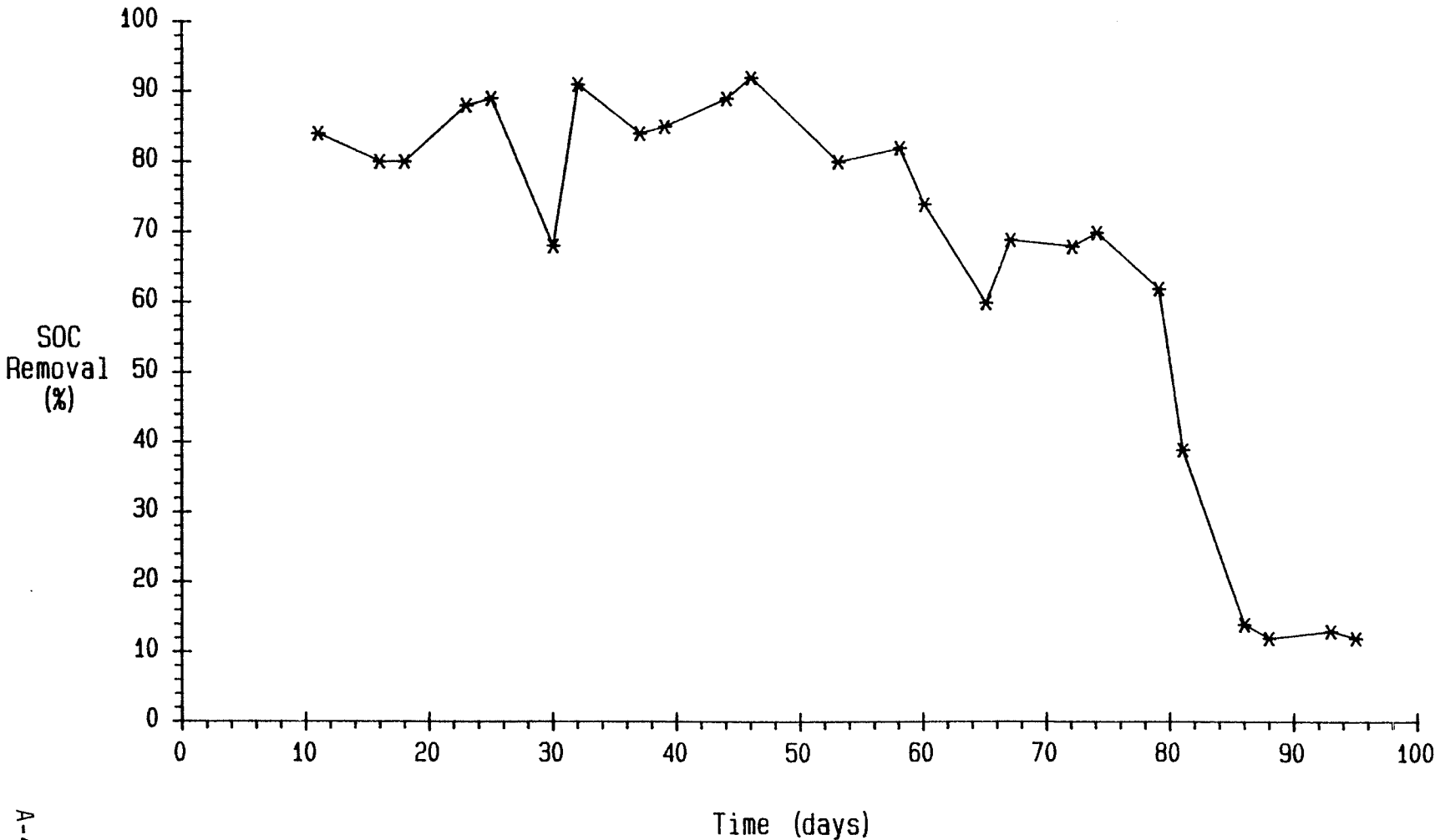
# Soluble Organic Carbon Removal vs Time

## Reactor 1



# Soluble Organic Carbon Removal vs Time

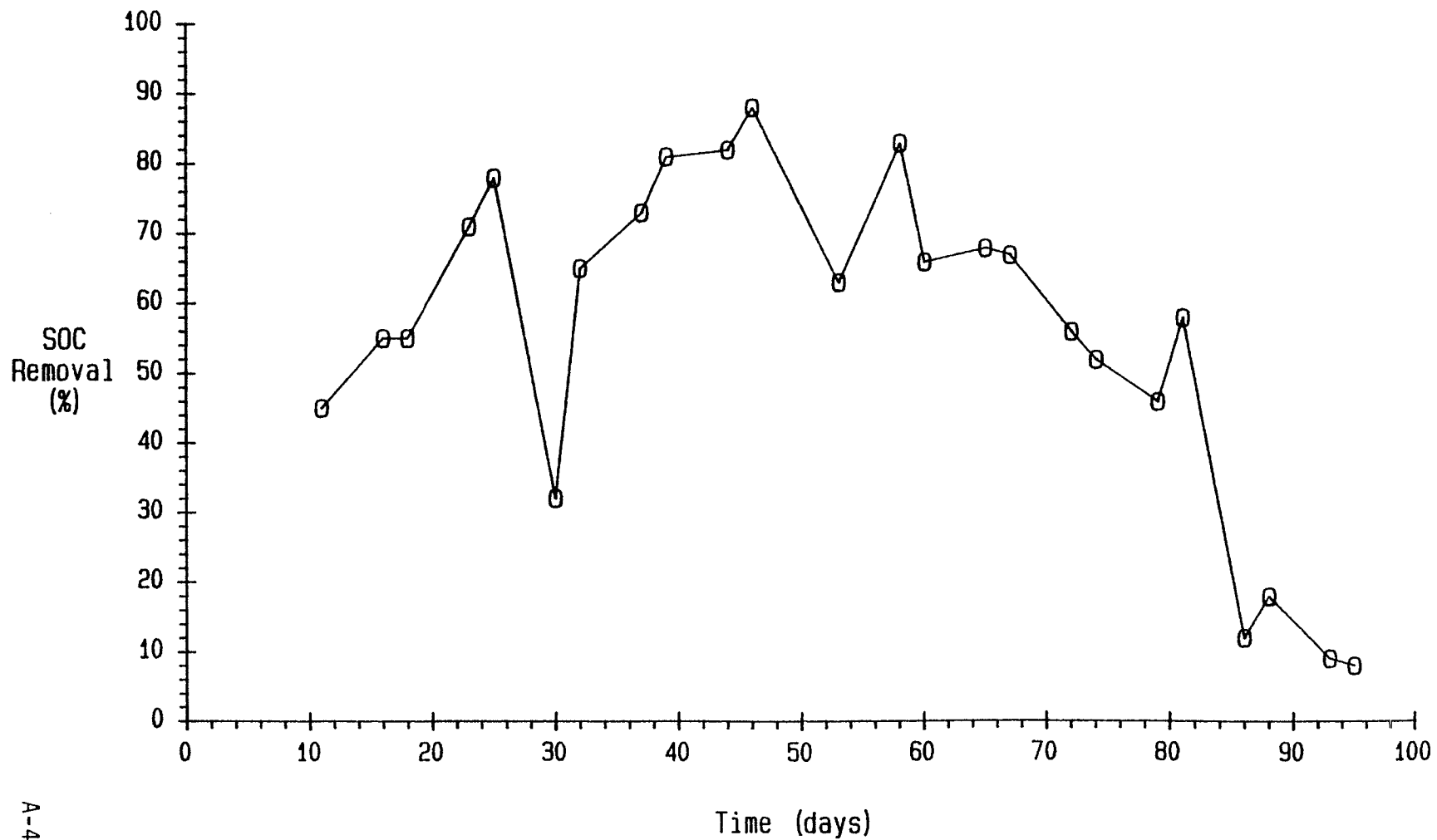
## Reactor 2





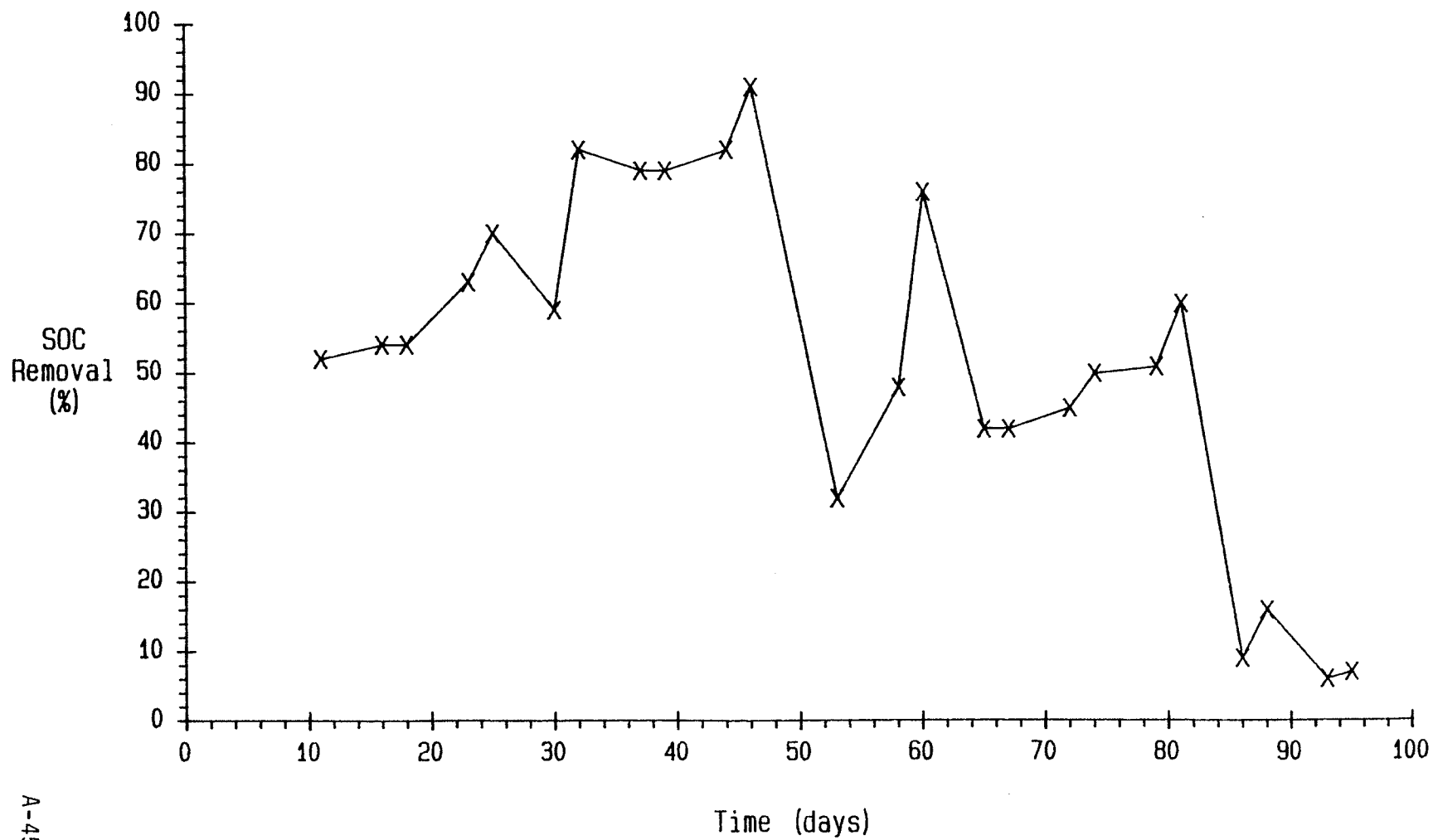
# Soluble Organic Carbon Removal vs Time

## Reactor 3



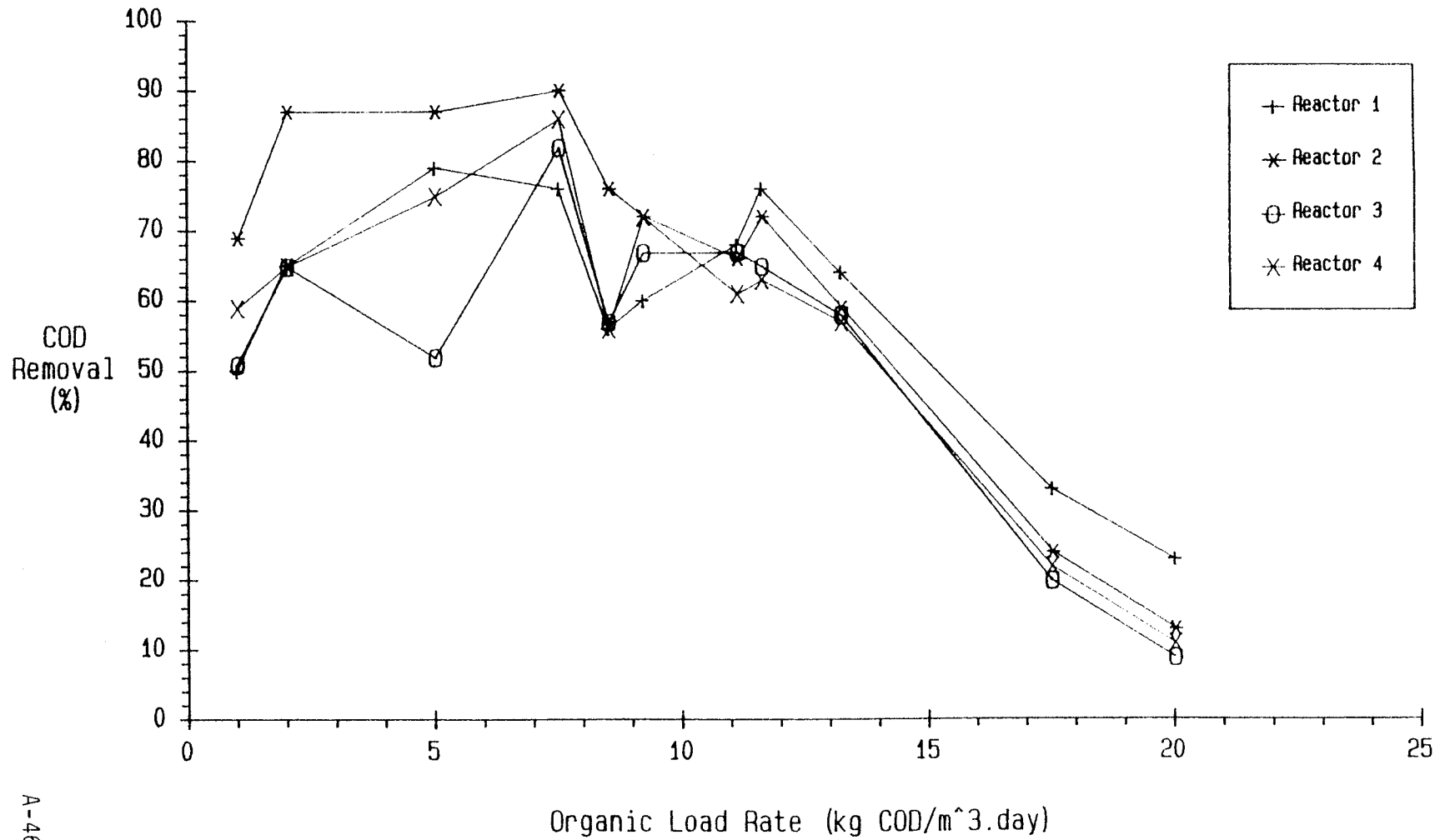
# Soluble Organic Carbon Removal vs Time

## Reactor 4



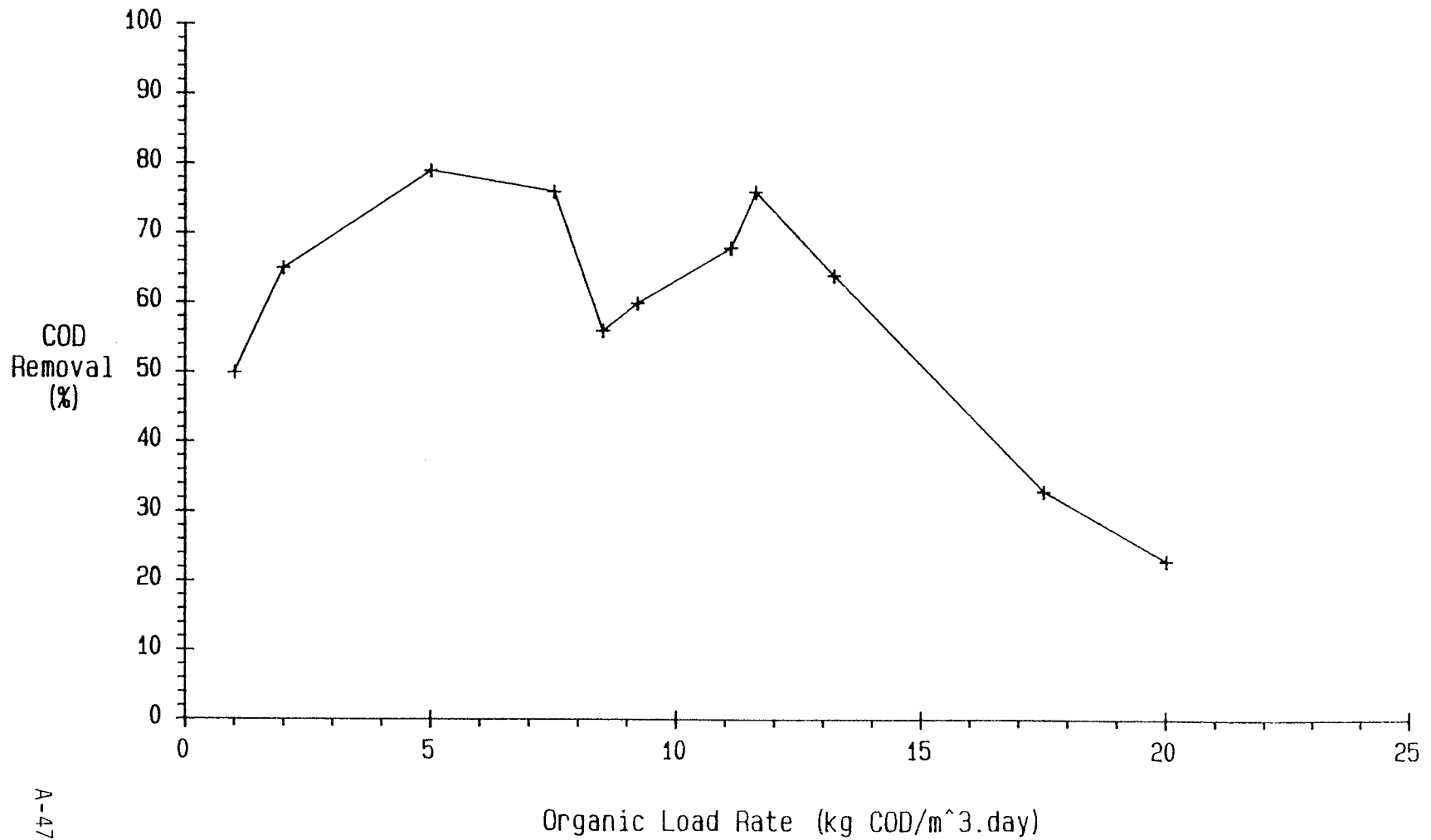
A-45

# Chemical Oxygen Demand Removal vs Organic Load Rate



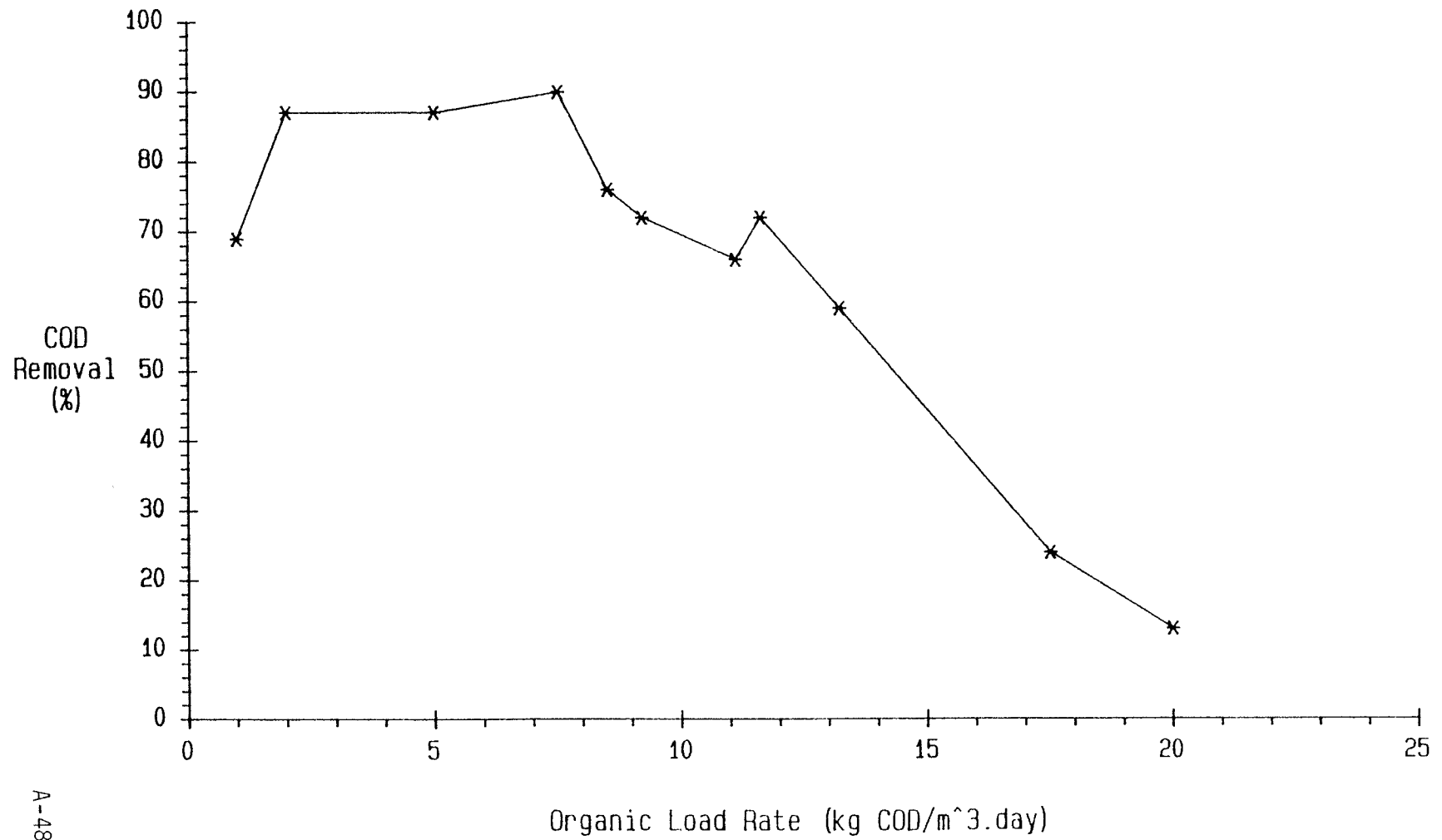
# Chemical Oxygen Demand Removal vs Organic Load Rate

## Reactor 1



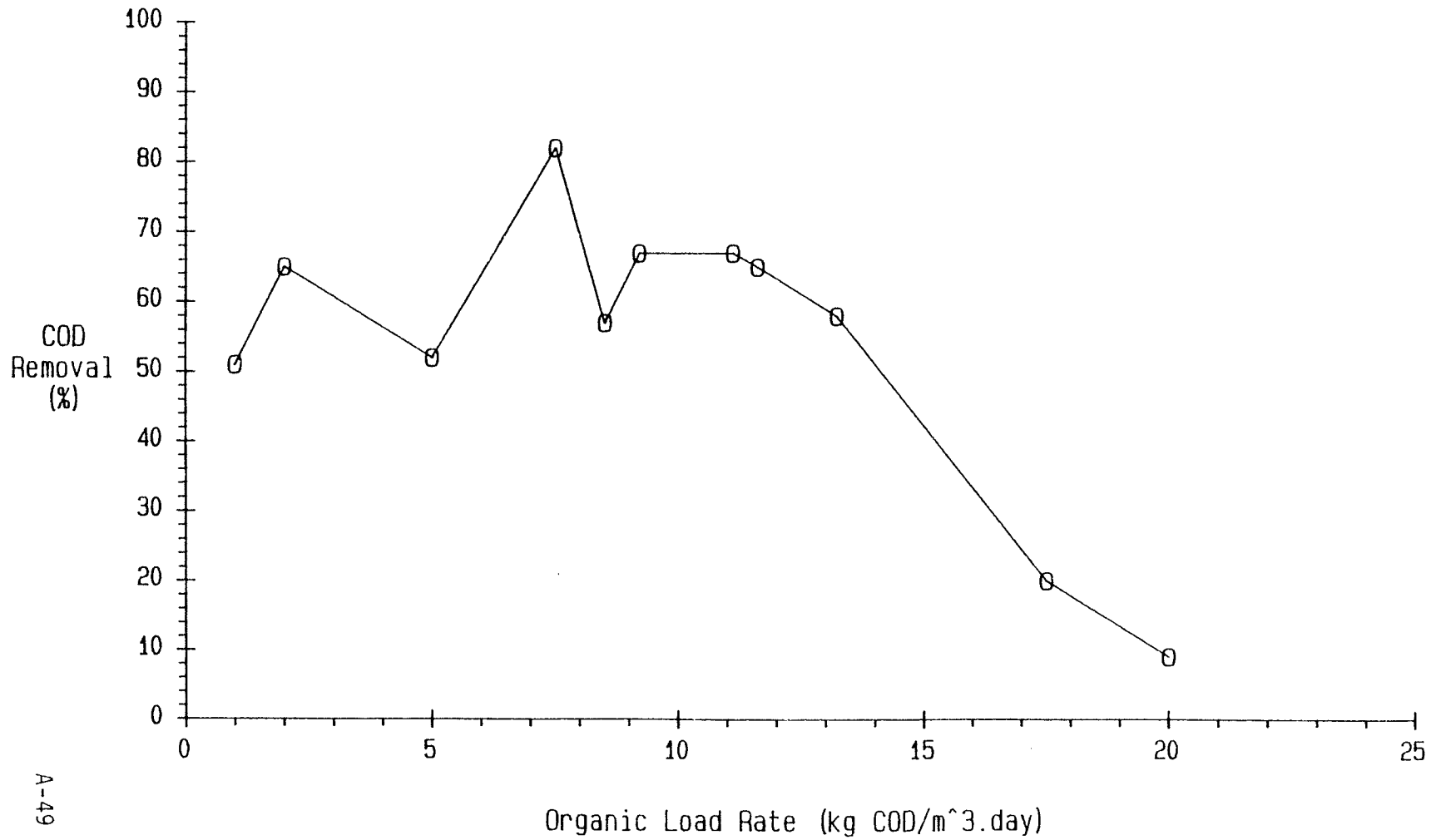
# Chemical Oxygen Demand Removal vs Organic Load Rate

## Reactor 2



# Chemical Oxygen Demand Removal vs Organic Load Rate

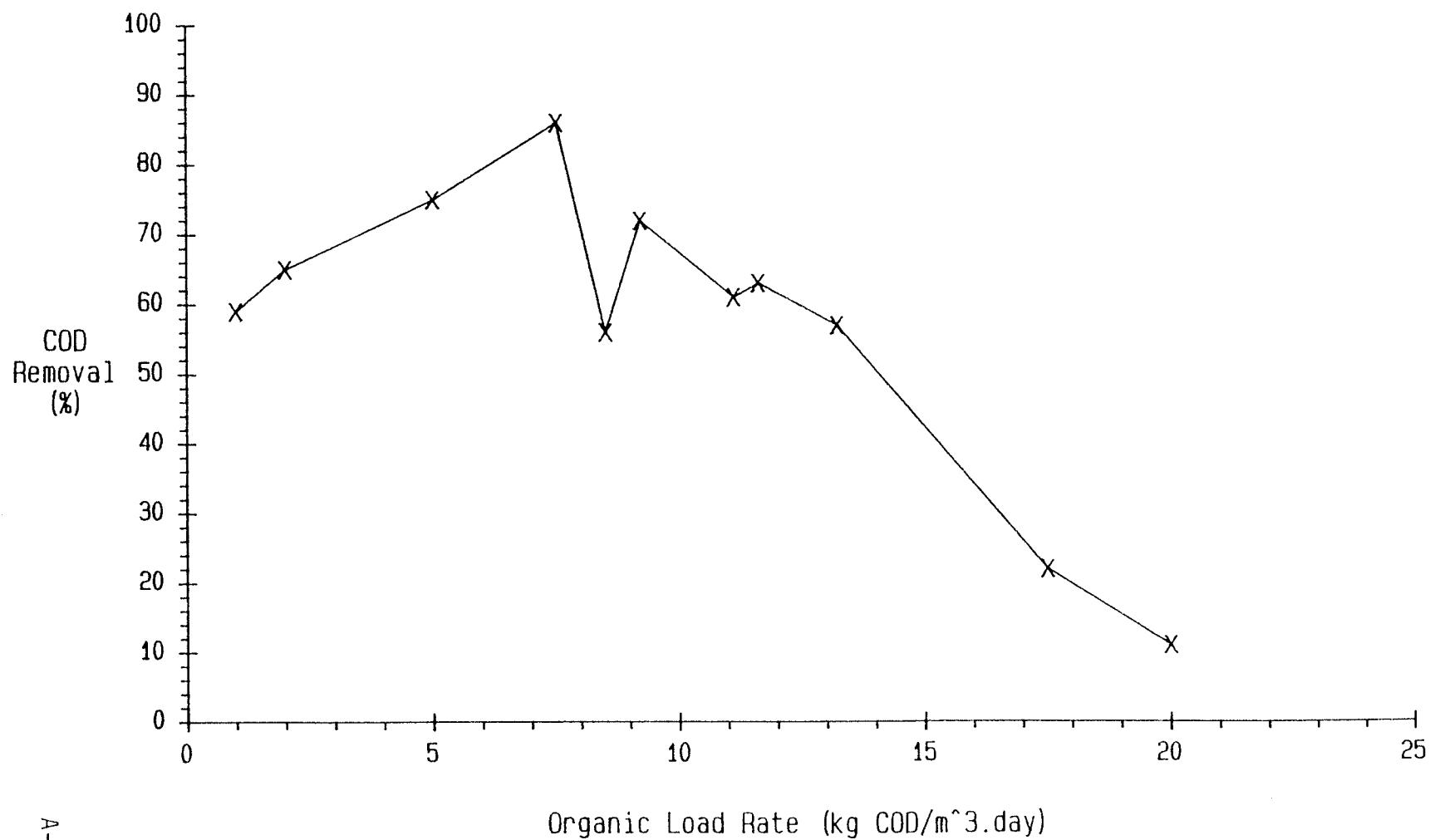
## Reactor 3



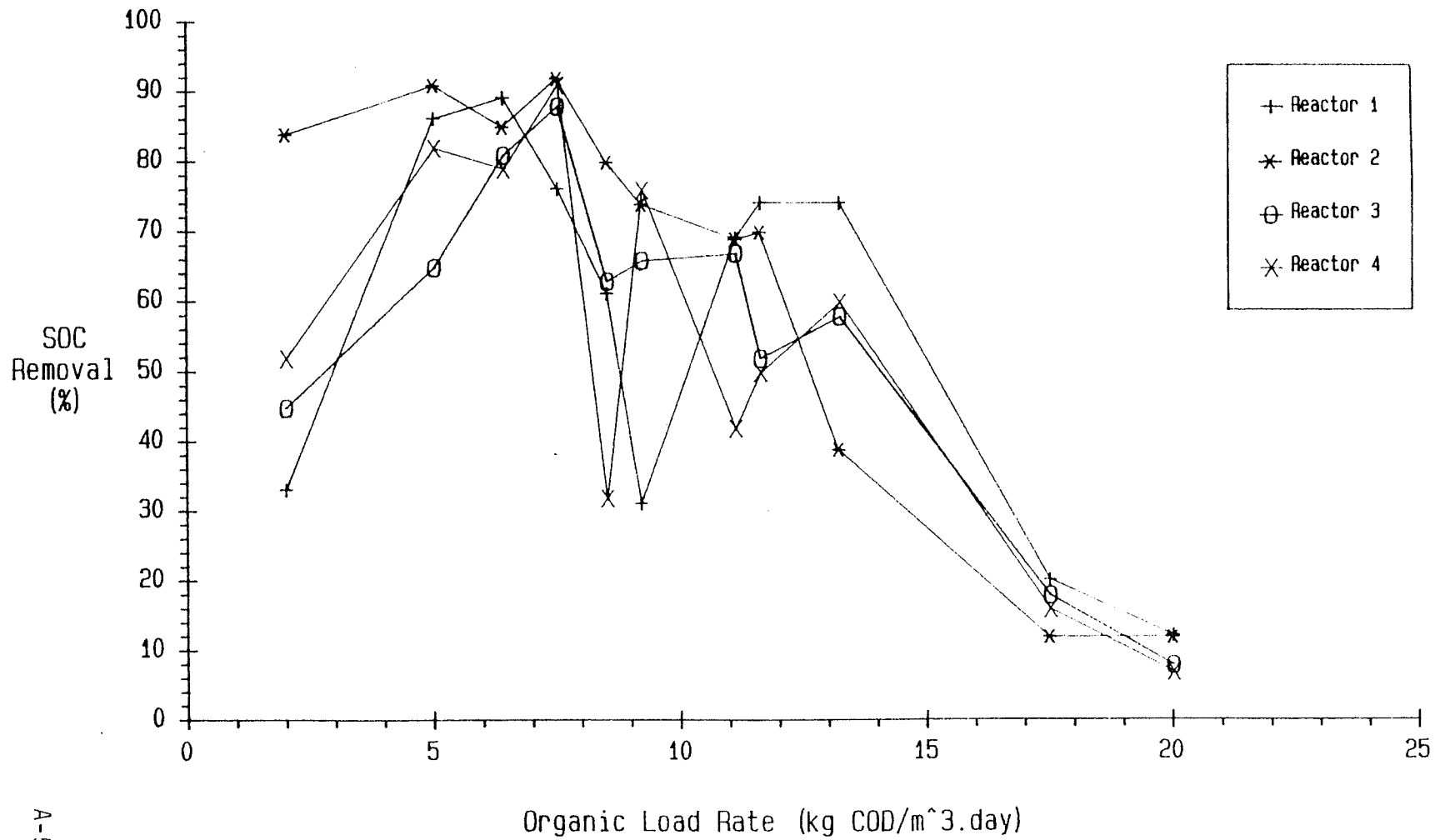
A-49

# Chemical Oxygen Demand Removal vs Organic Load Rate

## Reactor 4



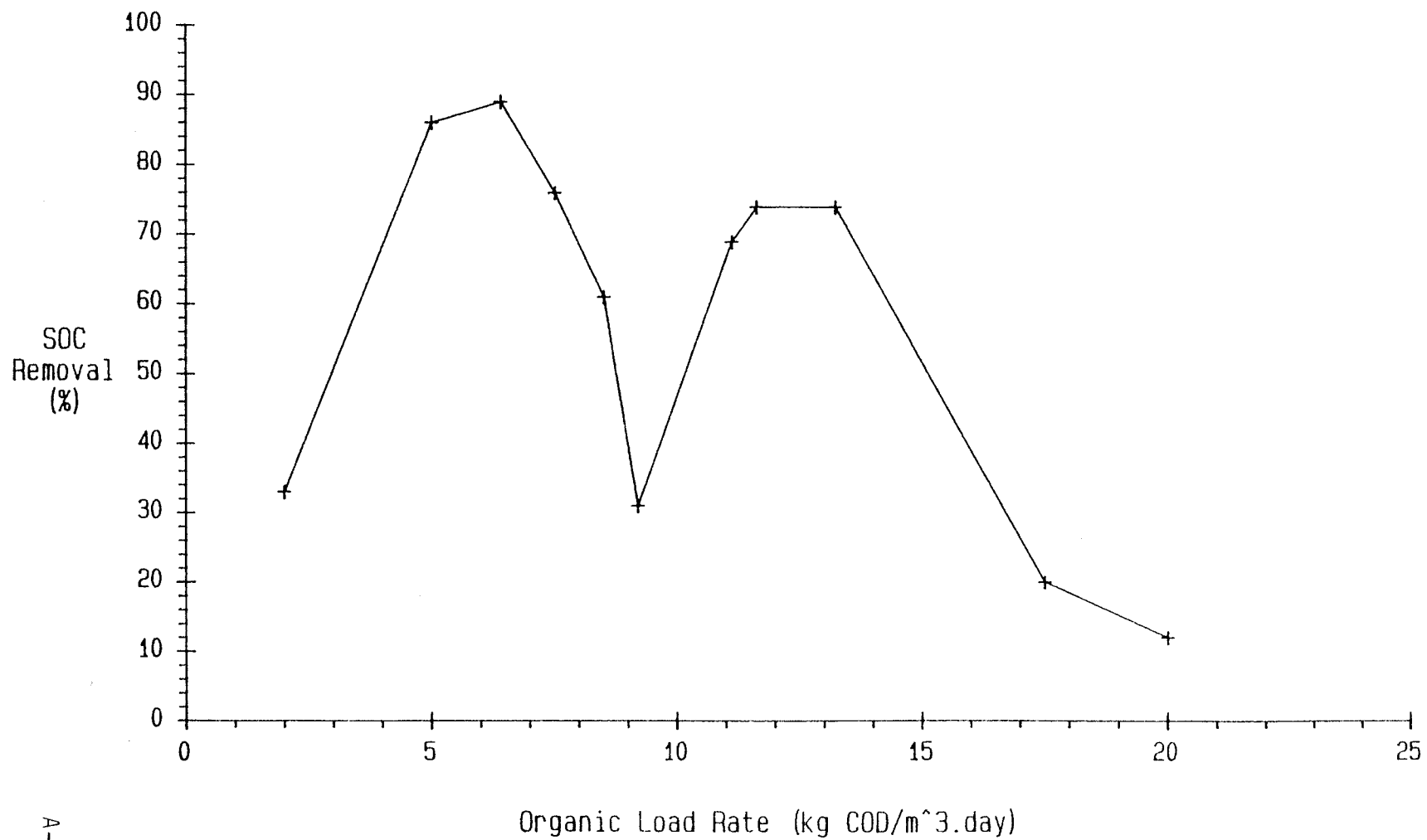
# Soluble Organic Carbon Removal vs Organic Load Rate





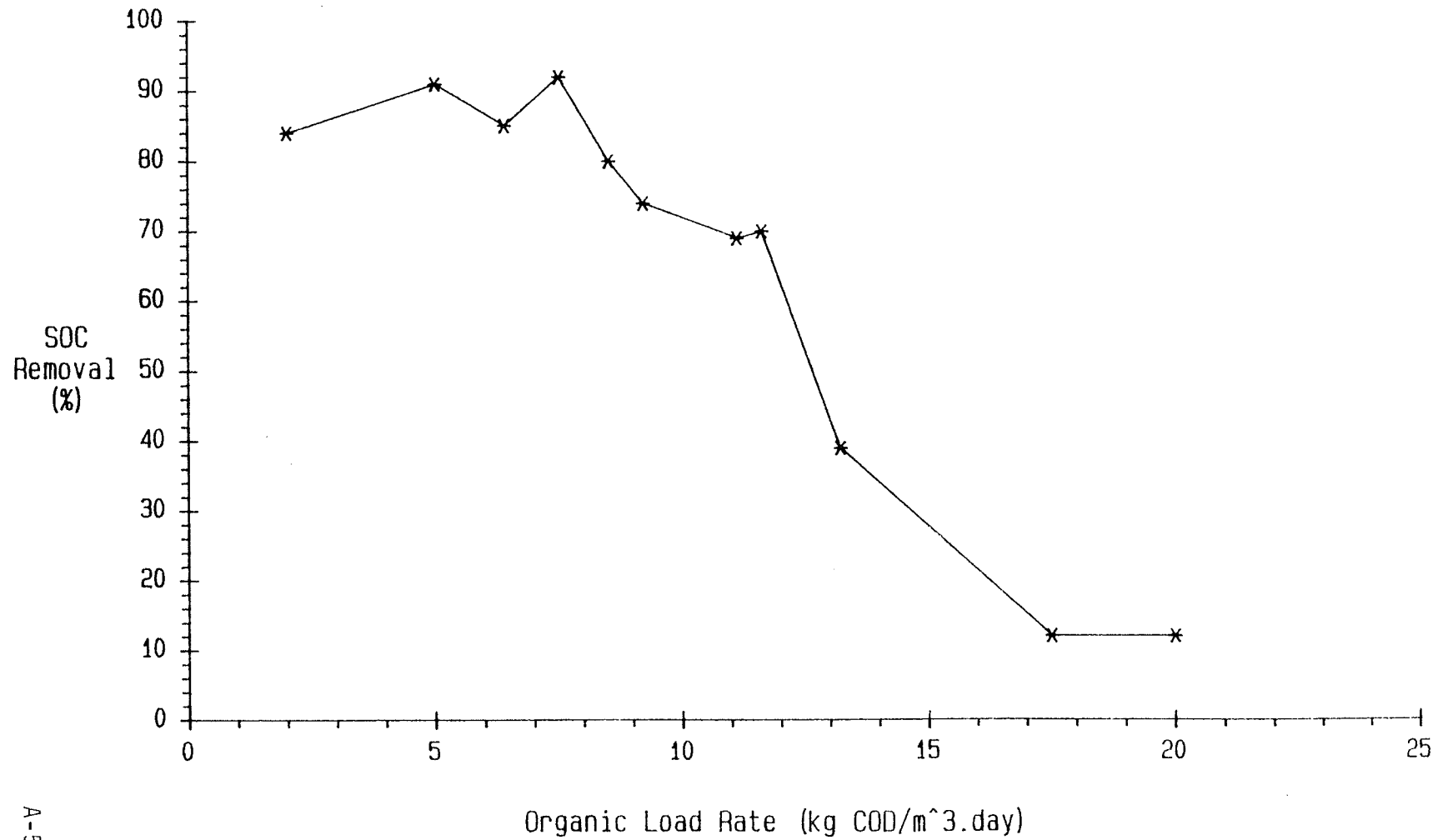
# Soluble Organic Carbon Removal vs Organic Load Rate

## Reactor 1



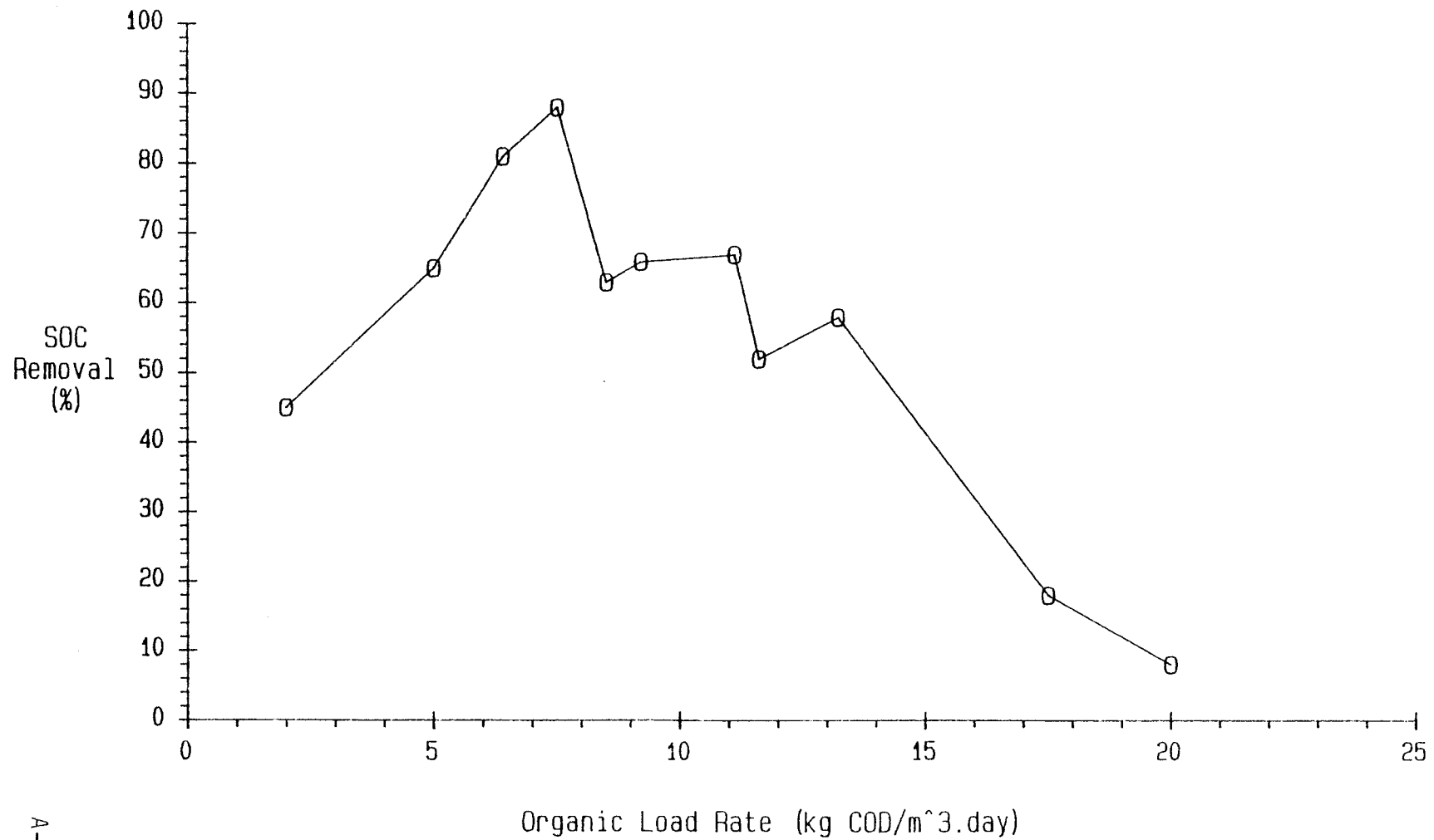
# Soluble Organic Carbon Removal vs Organic Load Rate

## Reactor 2



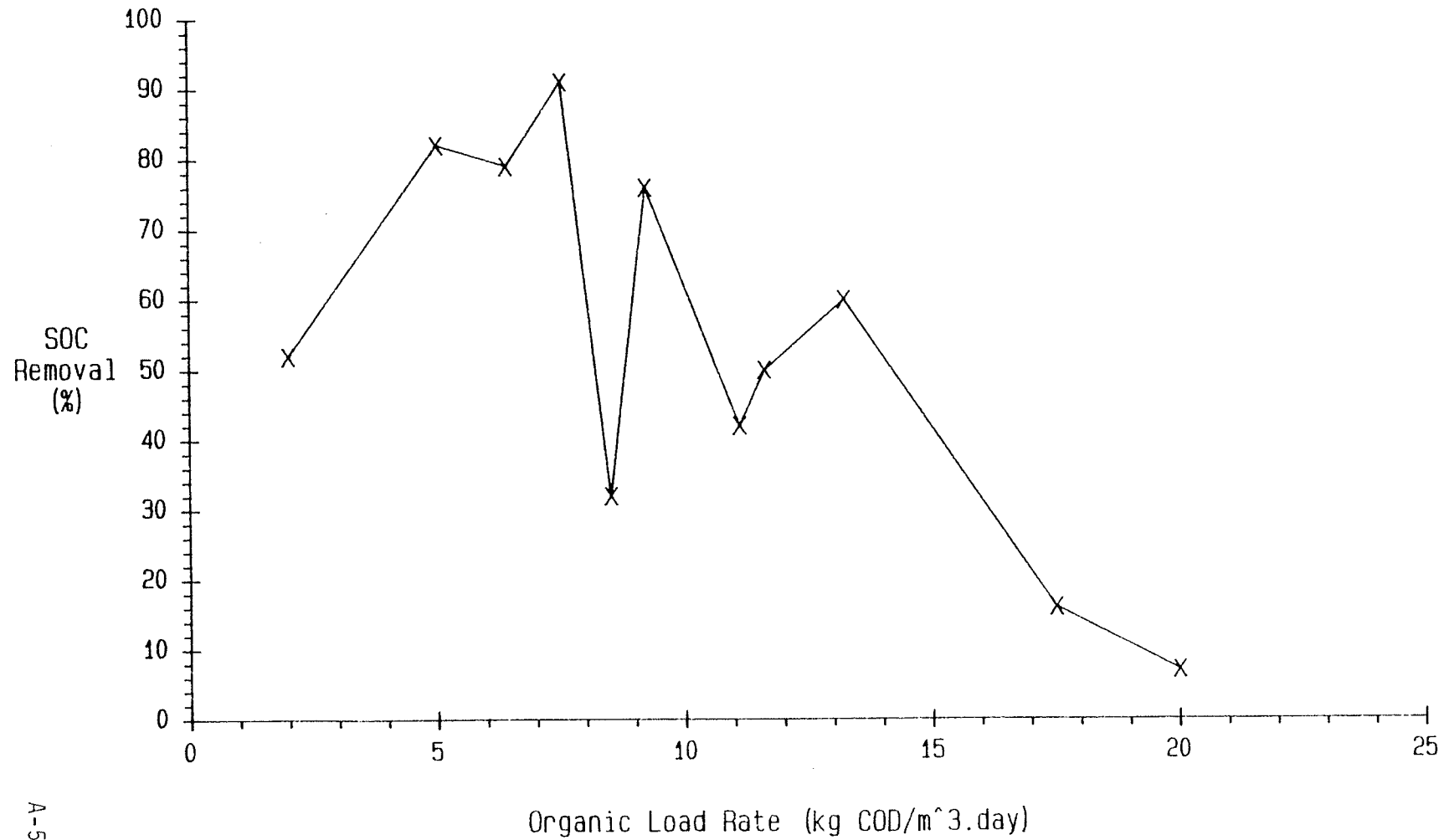
# Soluble Organic Carbon Removal vs Organic Load Rate

## Reactor 3

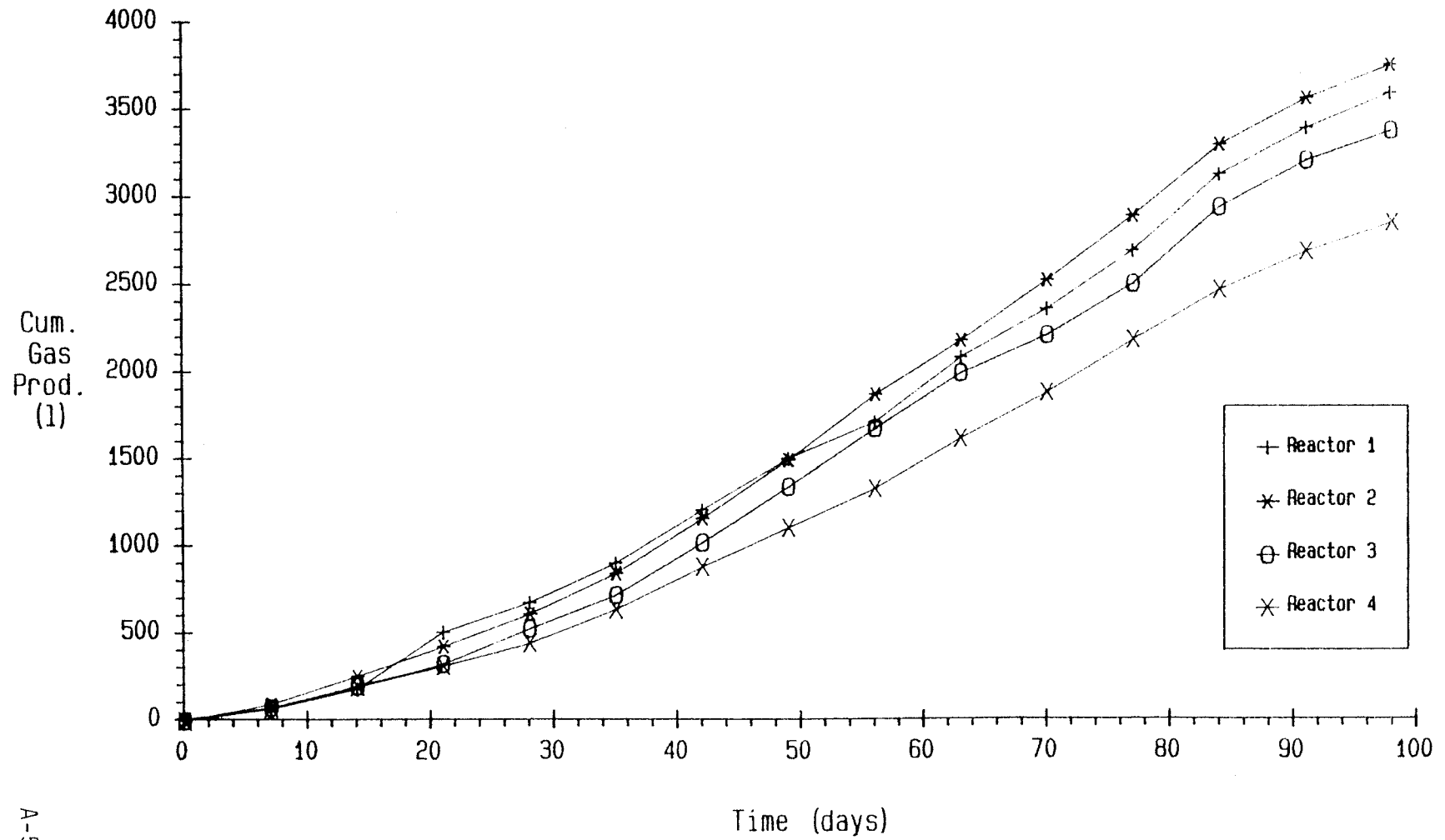


# Soluble Organic Carbon Removal vs Organic Load Rate

## Reactor 4

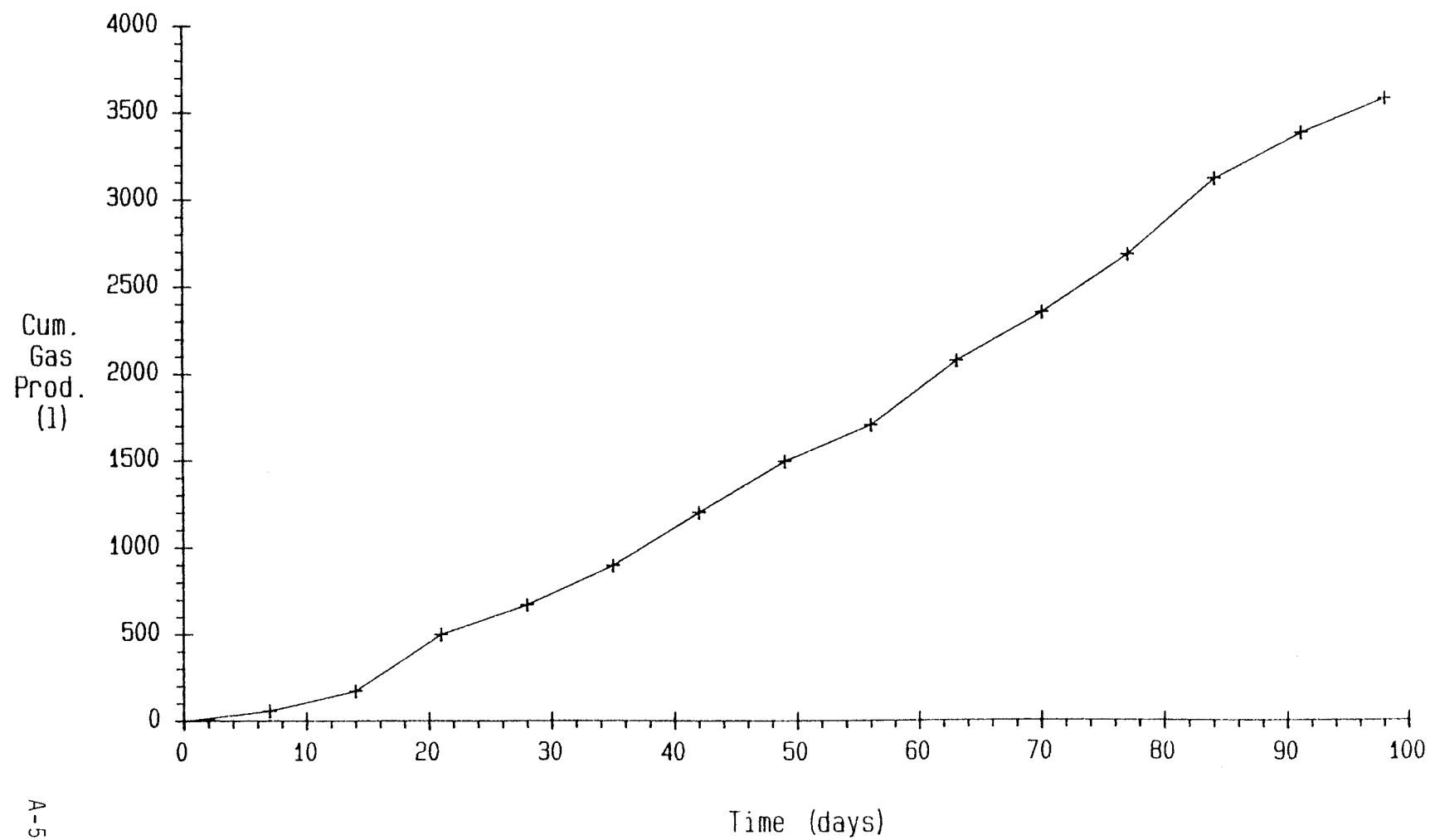


# Cumulative Gas Production vs Time



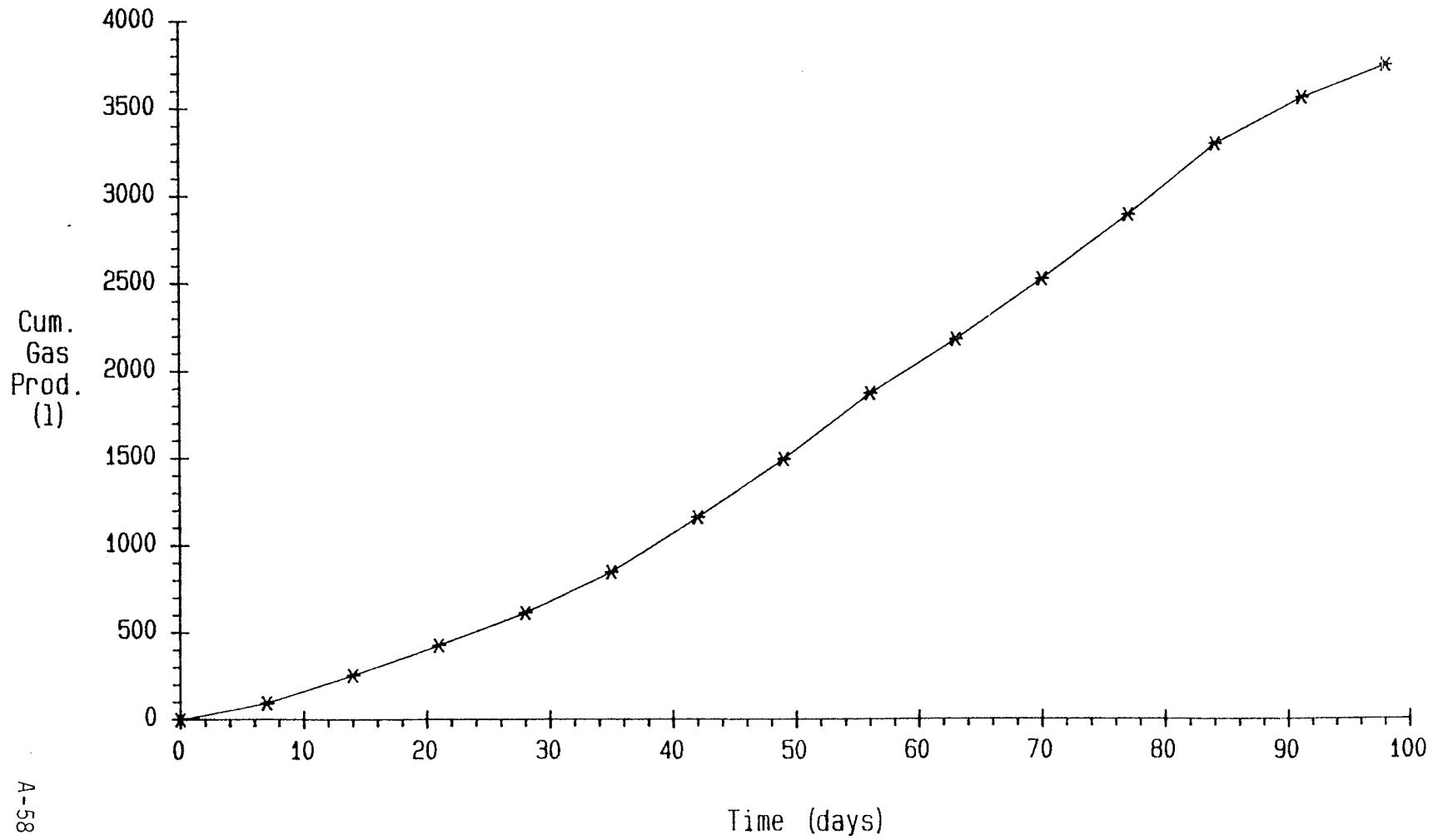
# Cumulative Gas Production vs Time

## Reactor 1



# Cumulative Gas Production vs Time

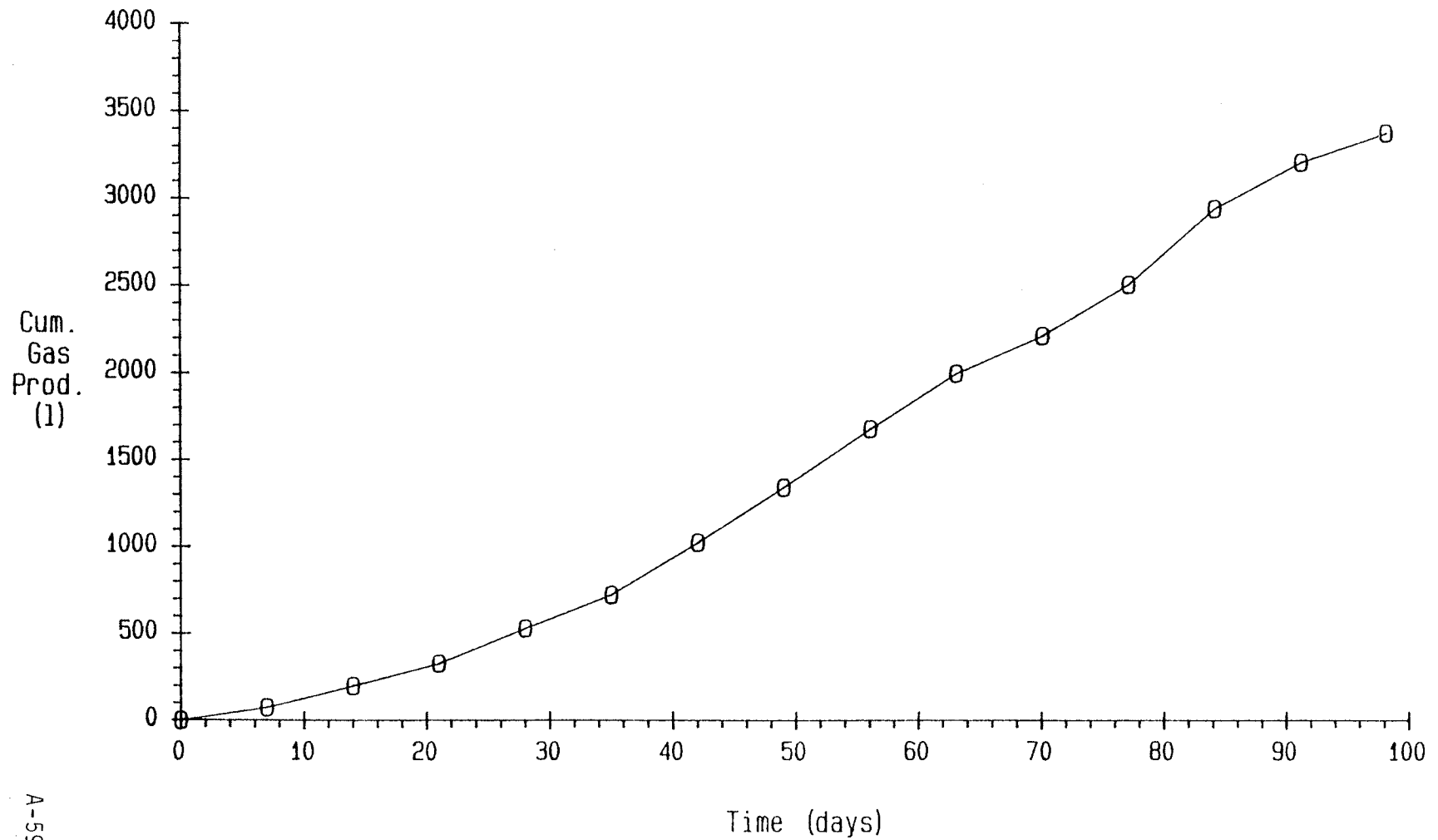
## Reactor 2



A-58

# Cumulative Gas Production vs Time

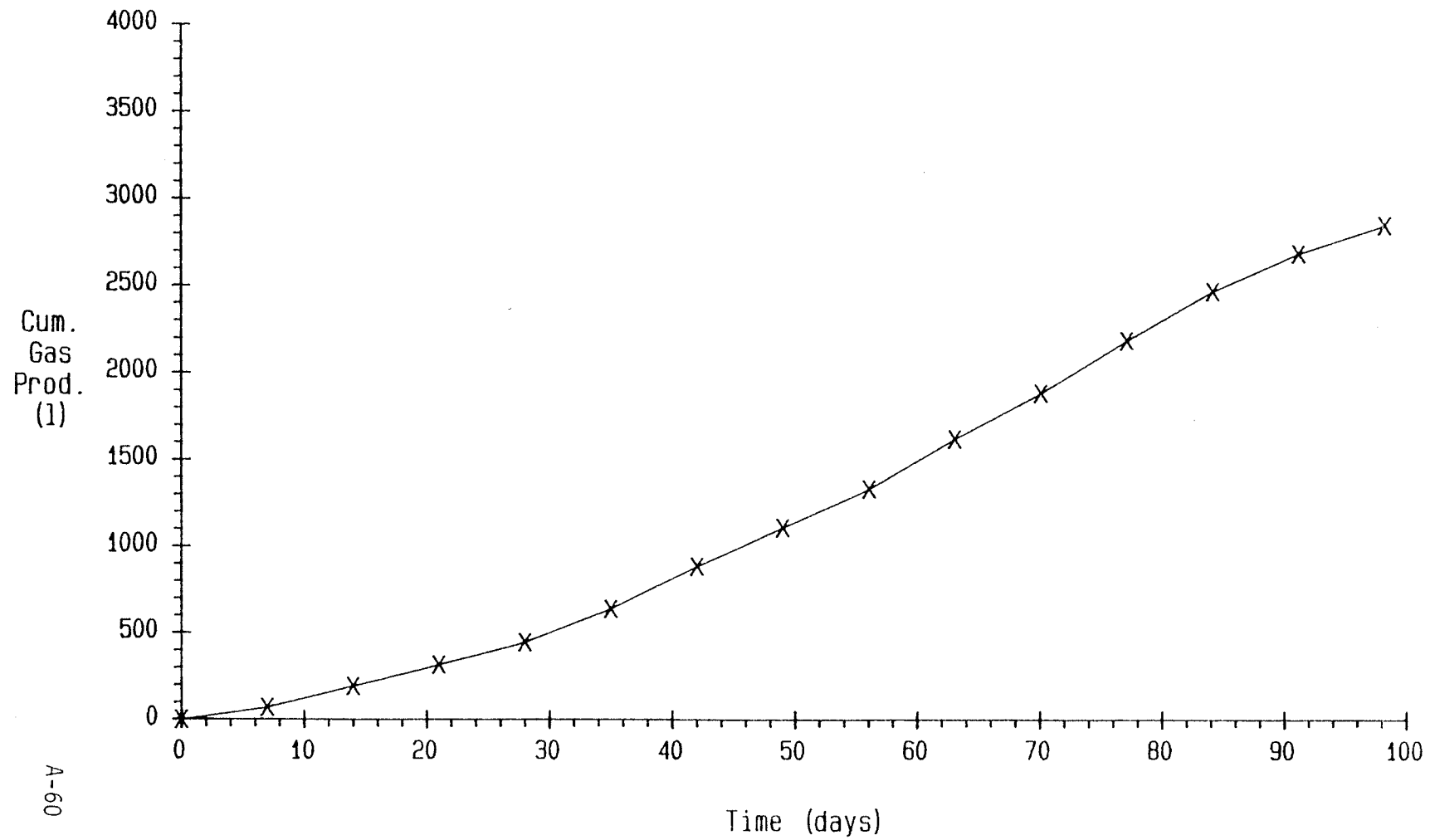
## Reactor 3





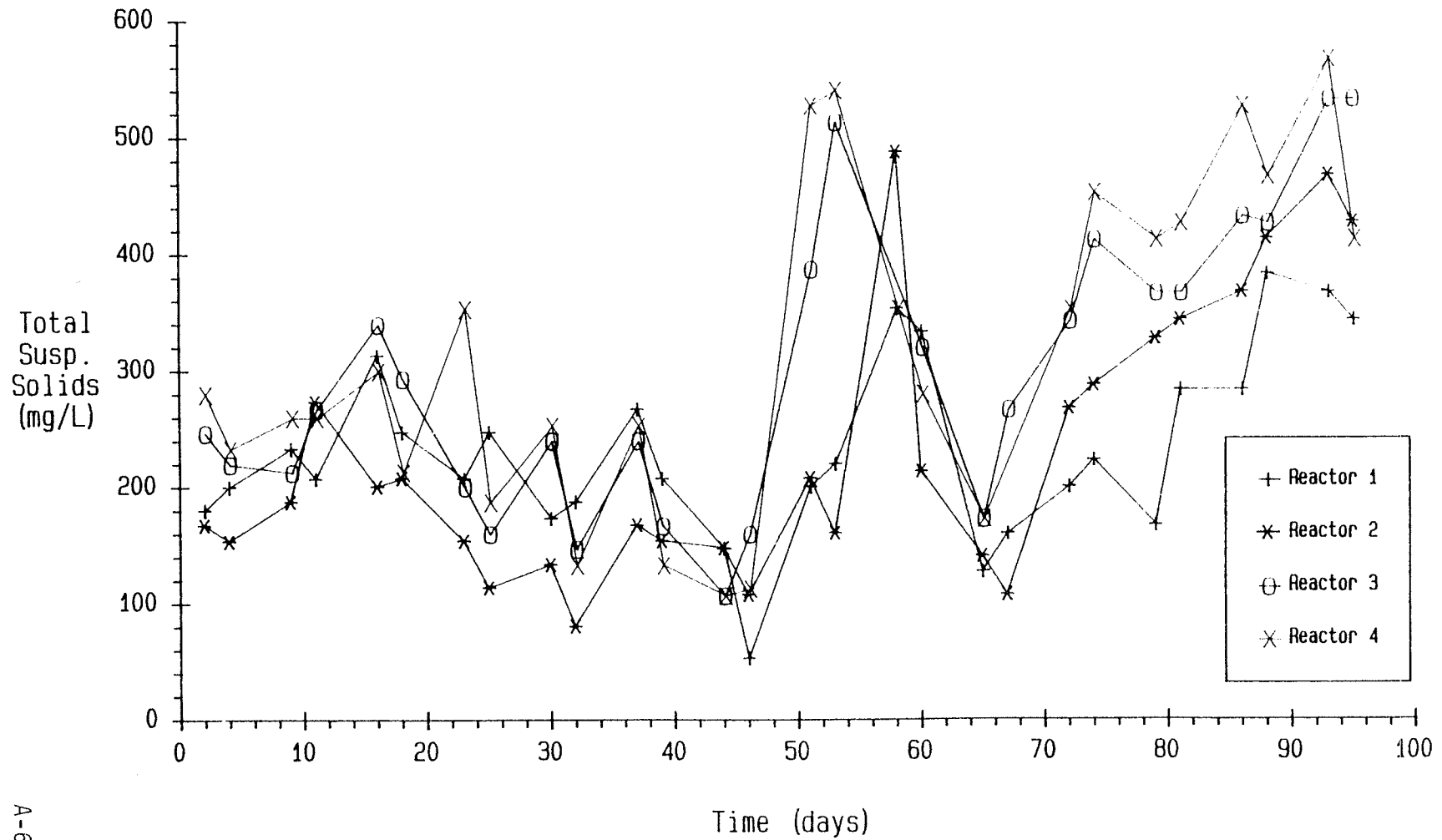
# Cumulative Gas Production vs Time

## Reactor 4

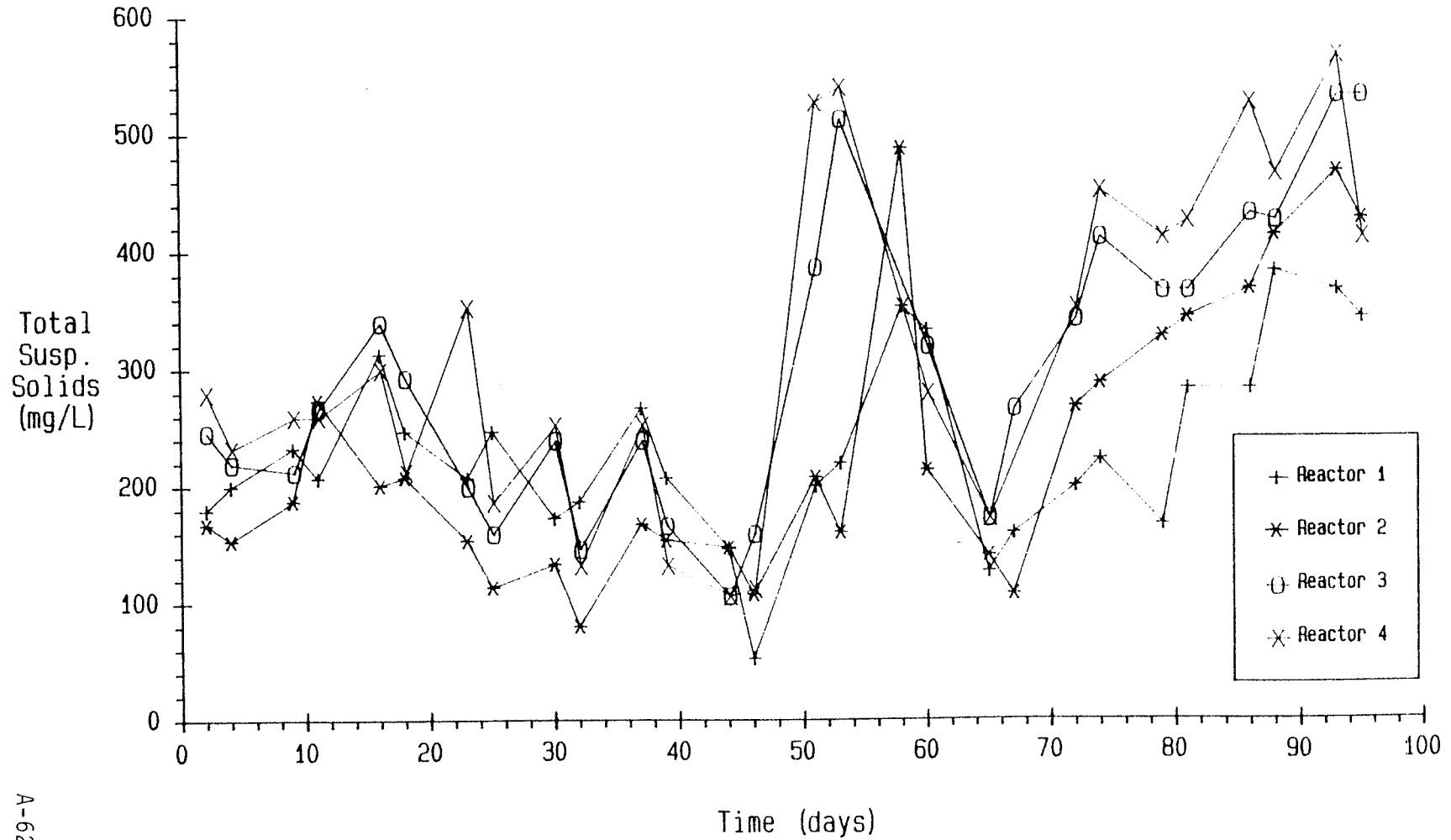


A-60

# Total Suspended Solids vs Time

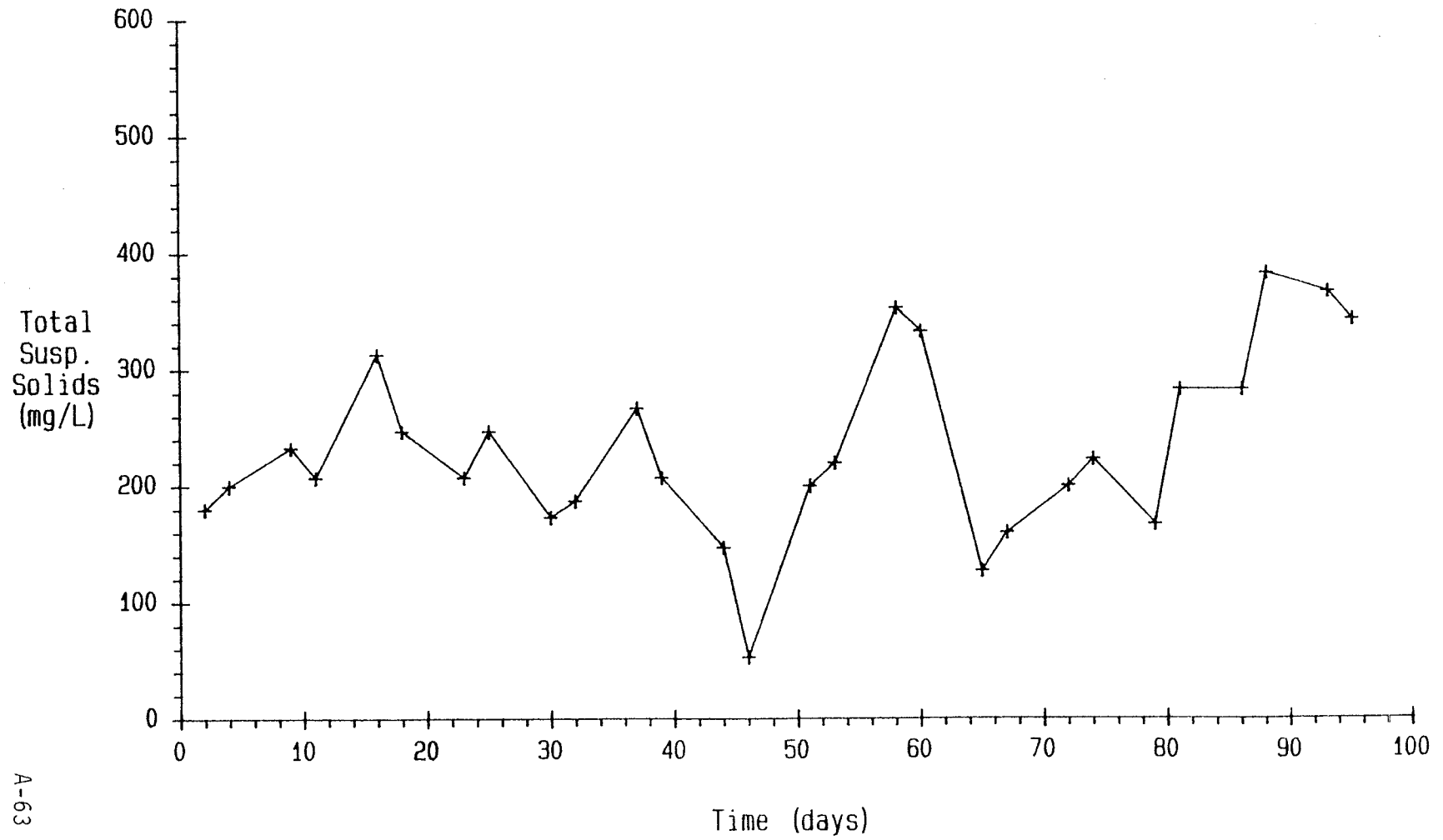


# Total Suspended Solids vs Time



# Total Suspended Solids vs Time

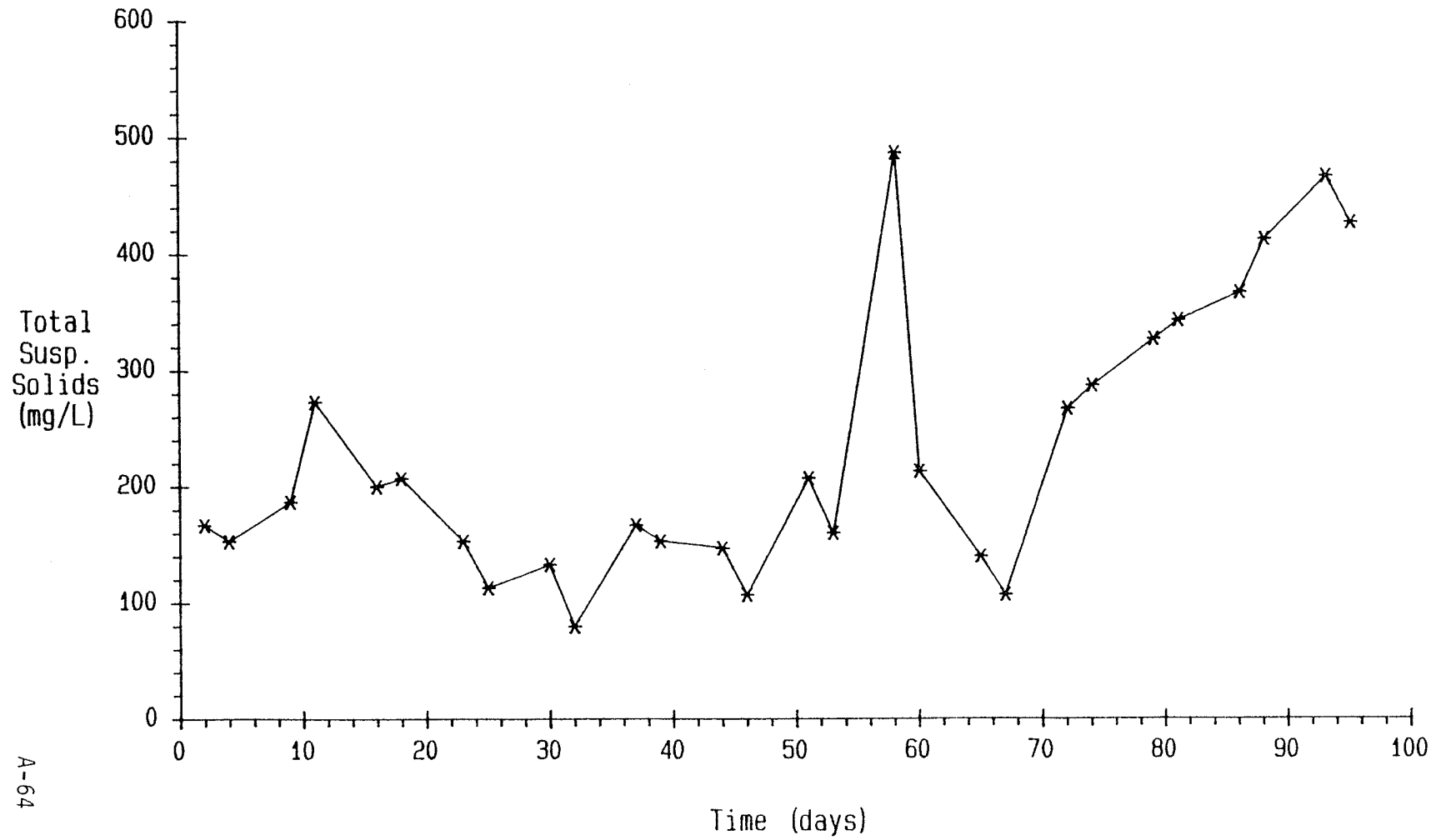
## Reactor 1



A-63

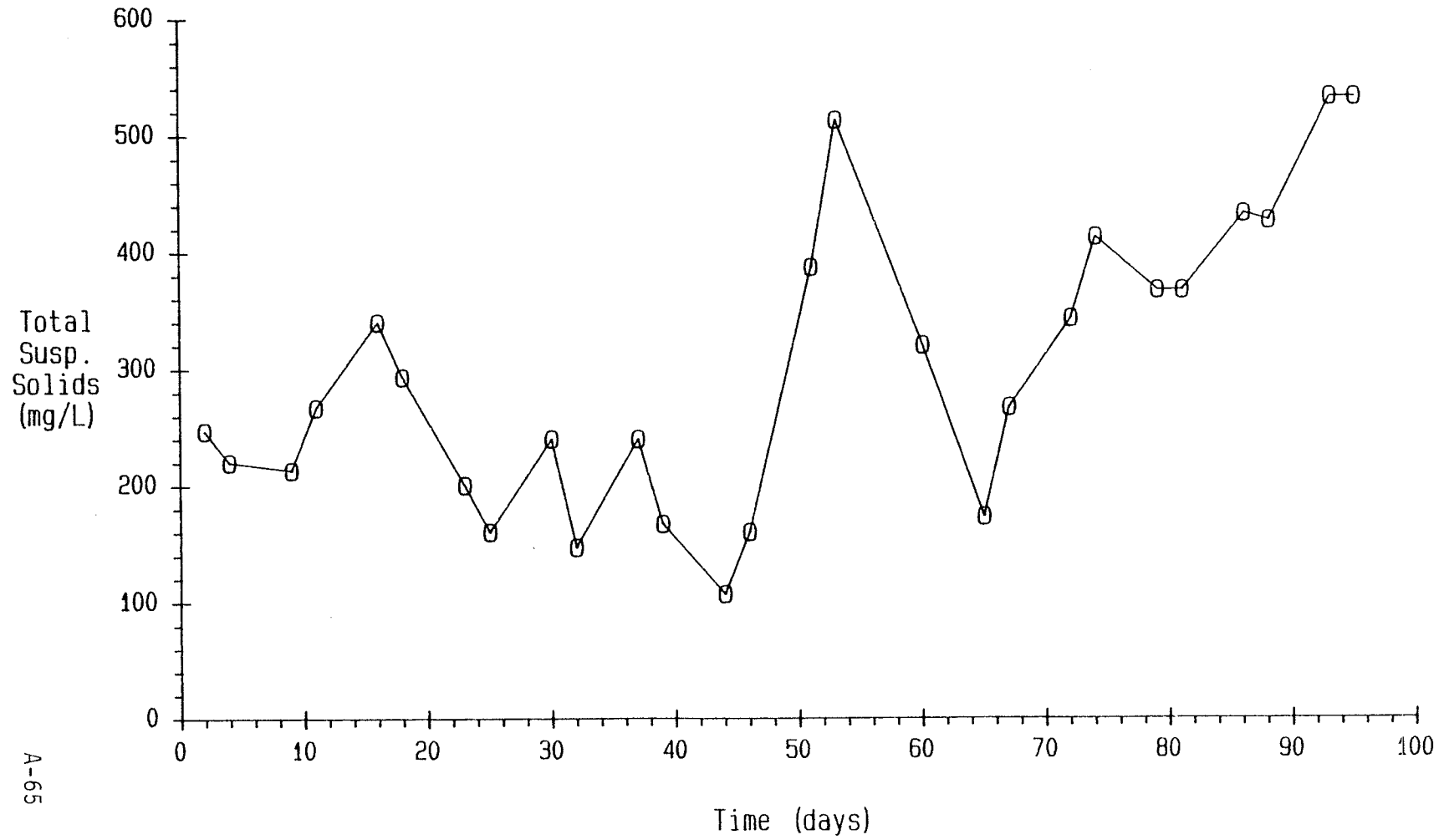
# Total Suspended Solids vs Time

## Reactor 2



# Total Suspended Solids vs Time

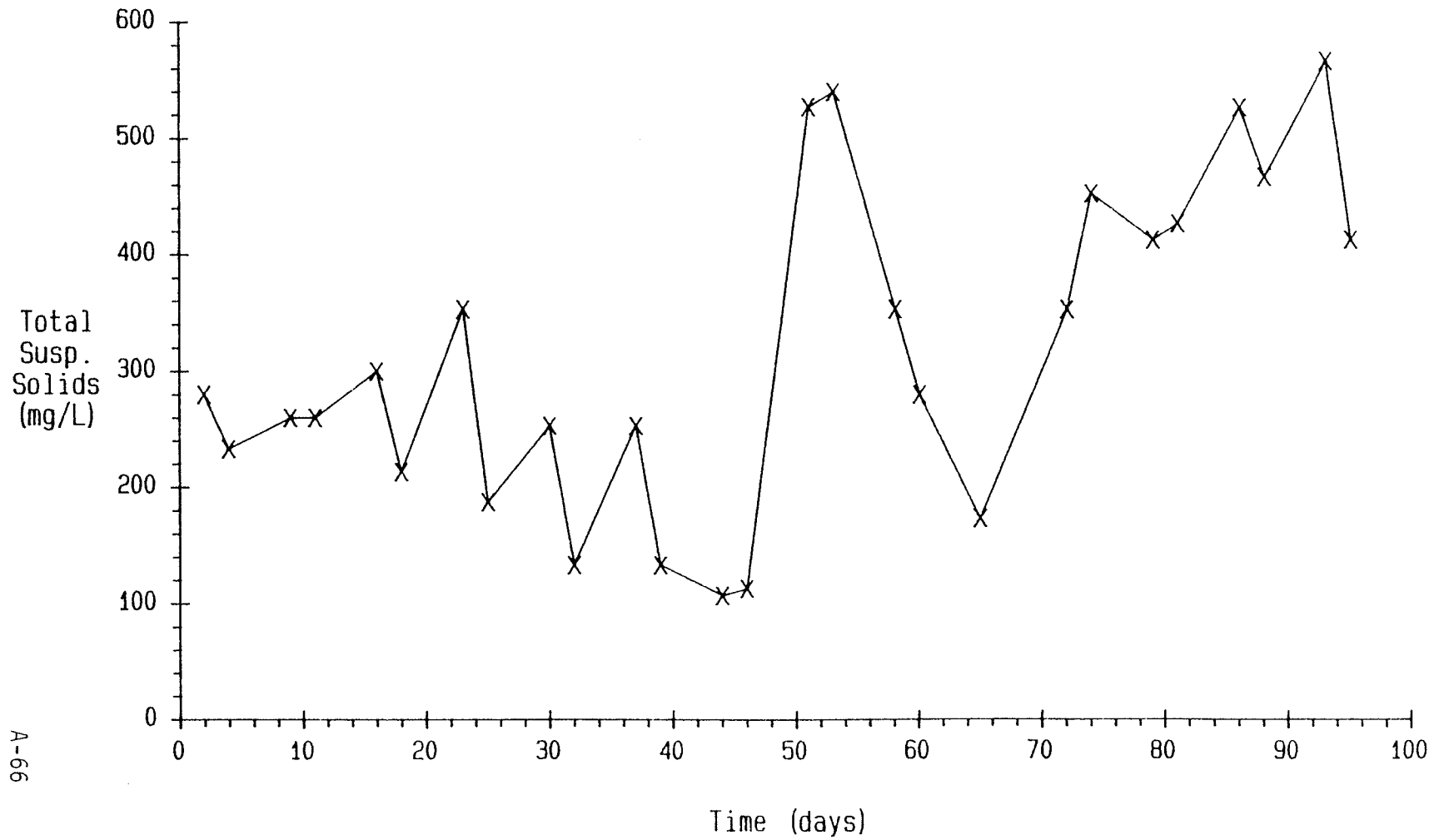
## Reactor 3



A-65

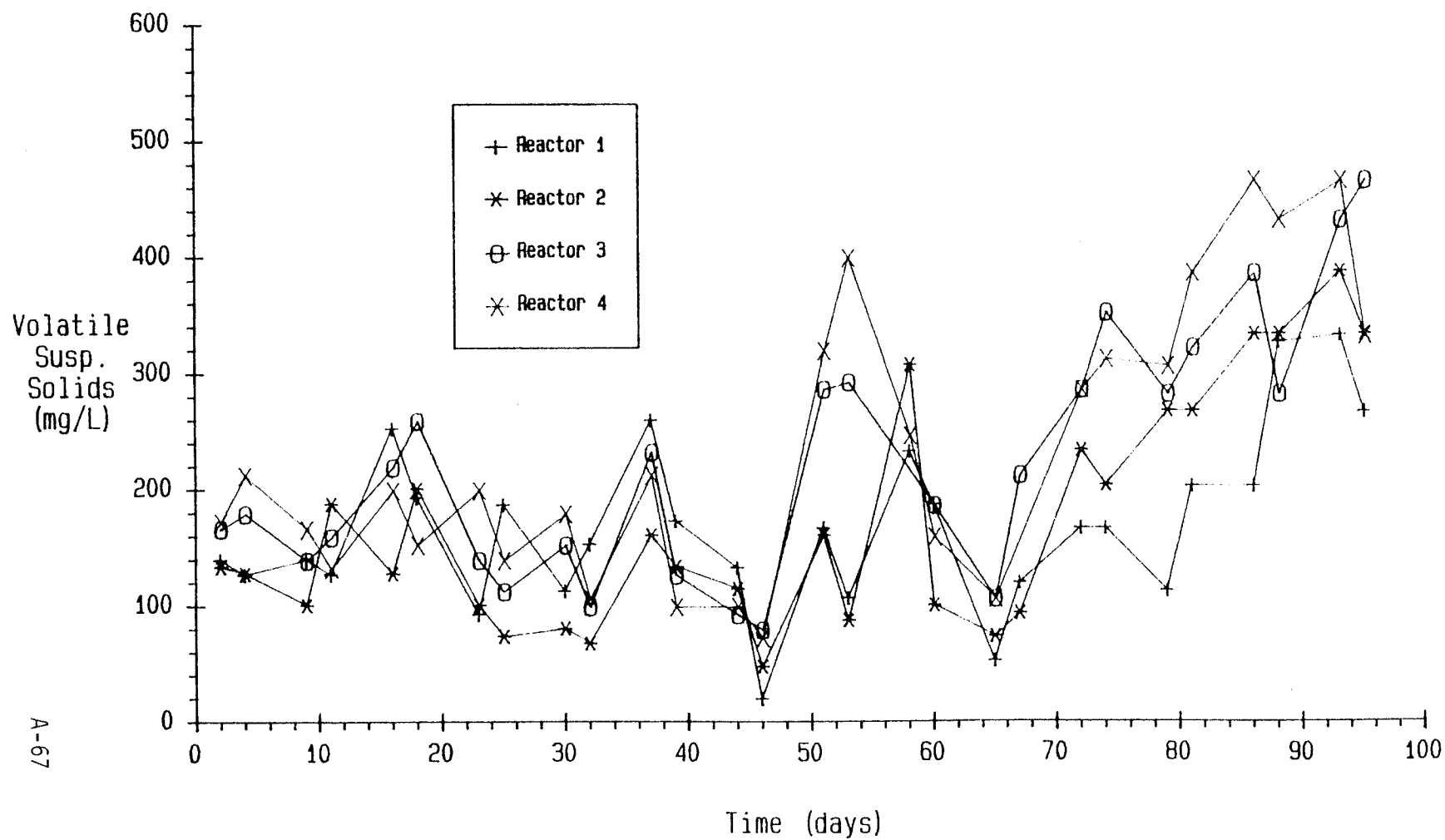
# Total Suspended Solids vs Time

## Reactor 4



A-66

# Volatile Suspended Solids vs Time

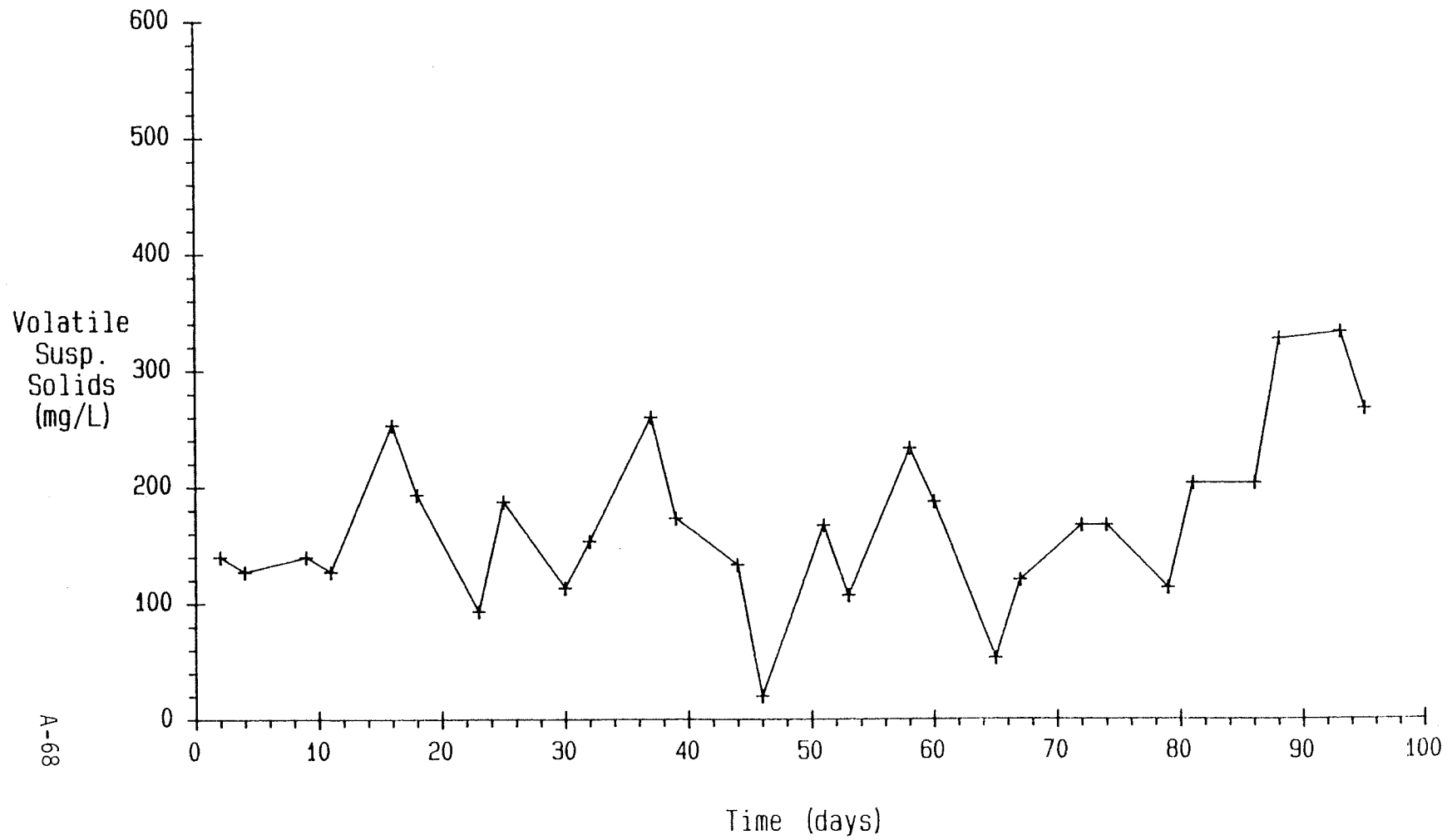


A-67



# Volatile Suspended Solids vs Time

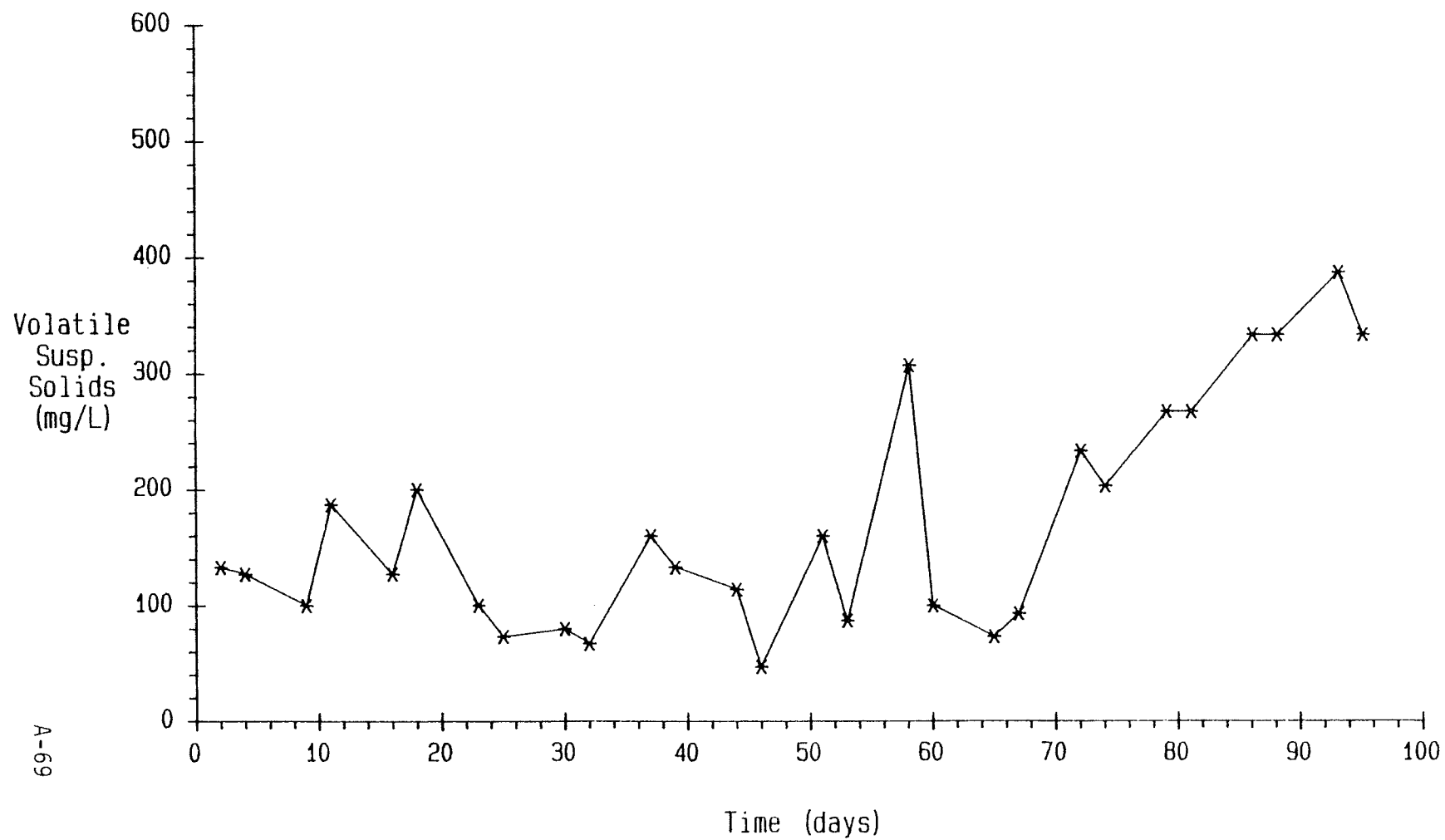
## Reactor 1



A-68

# Volatile Suspended Solids vs Time

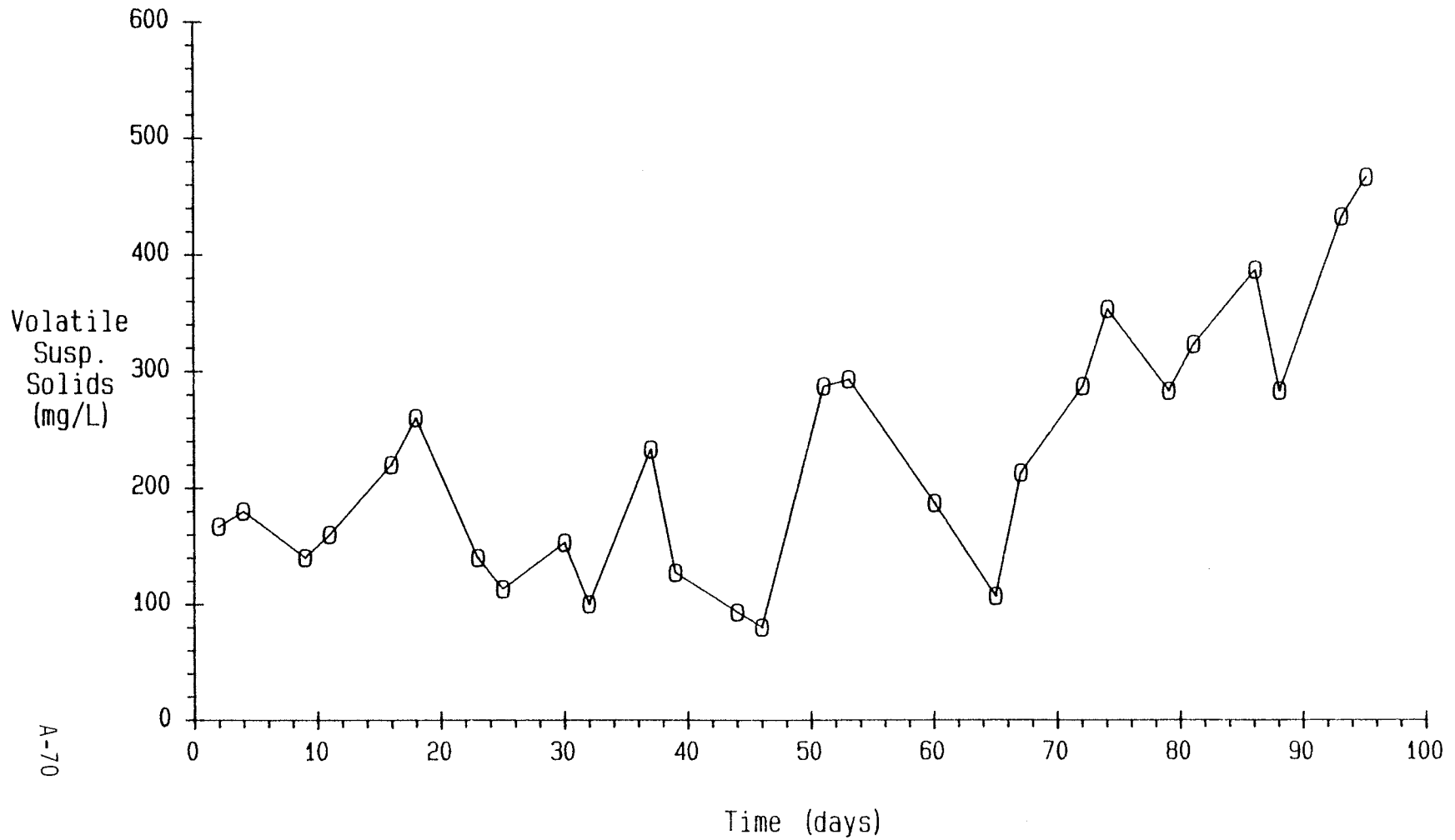
## Reactor 2



A-69

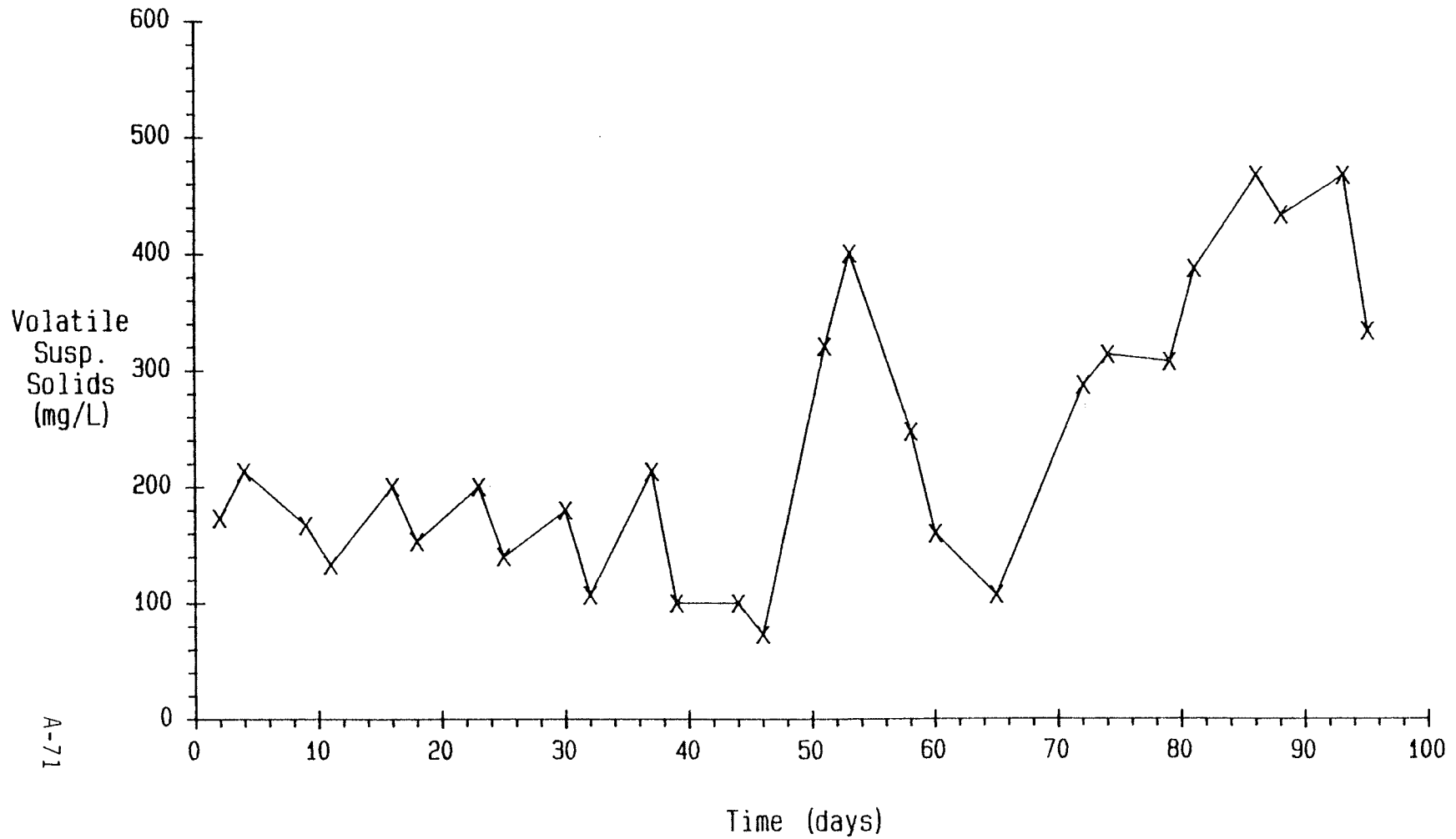
# Volatile Suspended Solids vs Time

## Reactor 3



# Volatile Suspended Solids vs Time

## Reactor 4



A-71

APPENDIX B

TRACER TEST DATA









## Concentration vs. Time Data

Reactor # 1 Date: 30/7/85  
 Flowrate: .22 l/m HRT: 1L  
 Gas Production: - Slug Concentration:  $\frac{100 \text{ mg/L}}{(30 \text{ ml})}$

Time (minutes)	Concentration ( $\mu\text{g/l}$ )	Time (minutes)	Concentration ( $\mu\text{g/l}$ )
0	23	120	39
5	21	150	37
10	32	180	33
15	36	210	29
20	47	240	28
25	53	270	30
30	51	300	28
35	89	330	29
40	115	360	29
45	152		
50	141		
55	120		
60	112		
65	101		
70	82		
75	64		
80	53		
85	48		
90	42		









## Concentration vs. Time Data

Reactor # 4 Date: 12/8/85  
 Flowrate: .22 l/m HRT: 1L  
 Gas Production: 5 l/d Slug Concentration:  $\frac{100 \text{ mg/L}}{(30 \text{ ml})}$

Time (minutes)	Concentration ( $\mu\text{g/l}$ )	Time (minutes)	Concentration ( $\mu\text{g/l}$ )
0	18	120	69
5	33	150	43
10	26	180	41
15	35	210	32
20	57	240	28
25	82	270	26
30	101	300	24
35	112	330	25
40	113	360	26
45	109		
50	115		
55	117		
60	89		
65	95		
70	98		
75	96		
80	101		
85	84		
90	78		



## Concentration vs. Time Data

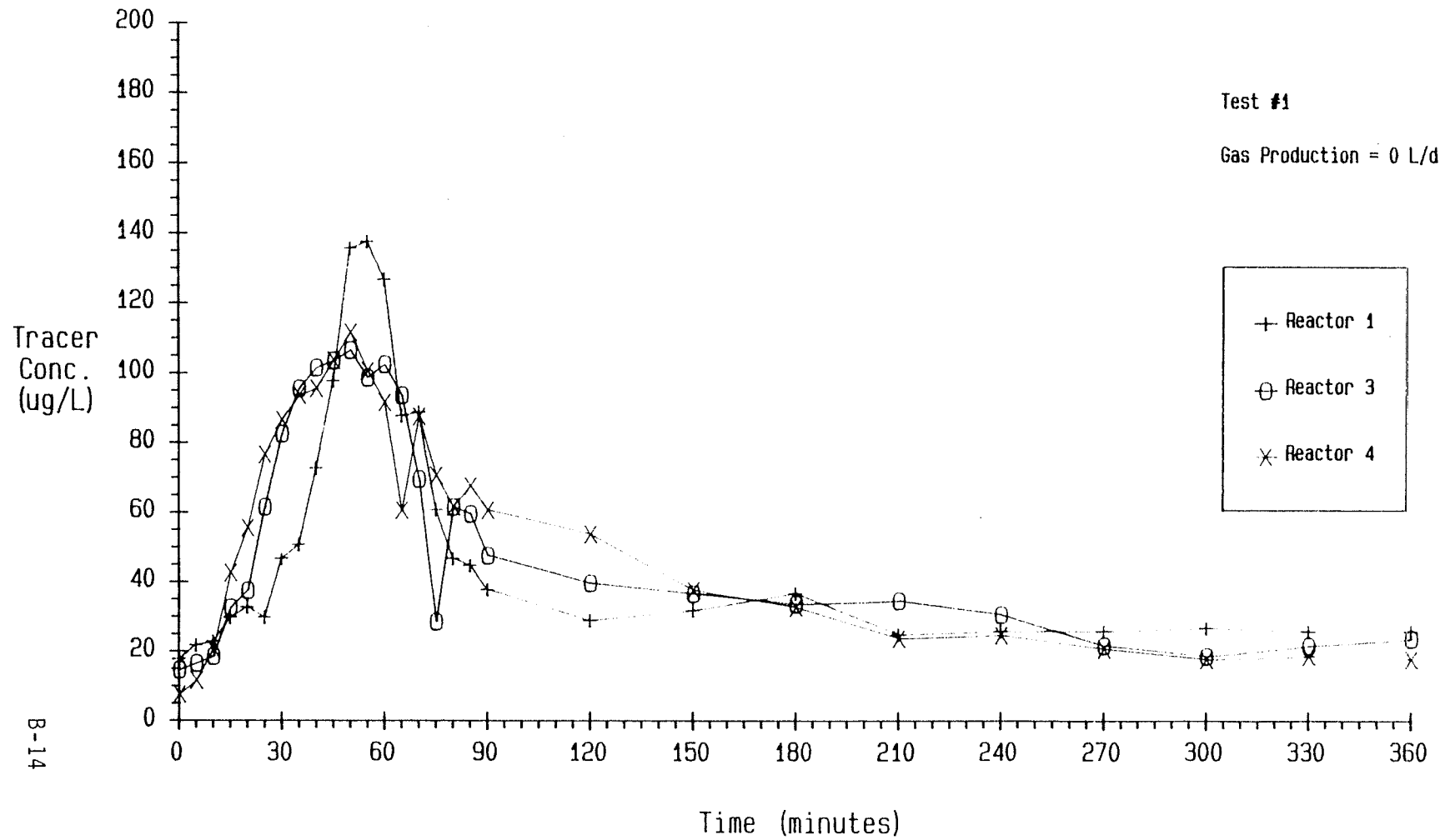
Reactor # 3 Date: 9/8/85  
 Flowrate: .22 l/m HRT: 1L  
 Gas Production: 5 l/d Slug Concentration:  $\frac{100 \text{ mg/L}}{(30 \text{ ml})}$

Time (minutes)	Concentration ( $\mu\text{g/l}$ )	Time (minutes)	Concentration ( $\mu\text{g/l}$ )
0	17	120	60
5	21	150	48
10	38	180	29
15	58	210	33
20	69	240	28
25	68	270	21
30	89	300	23
35	91	330	23
40	112	360	22
45	110		
50	114		
55	126		
60	102		
65	103		
70	99		
75	101		
80	87		
85	92		
90	65		





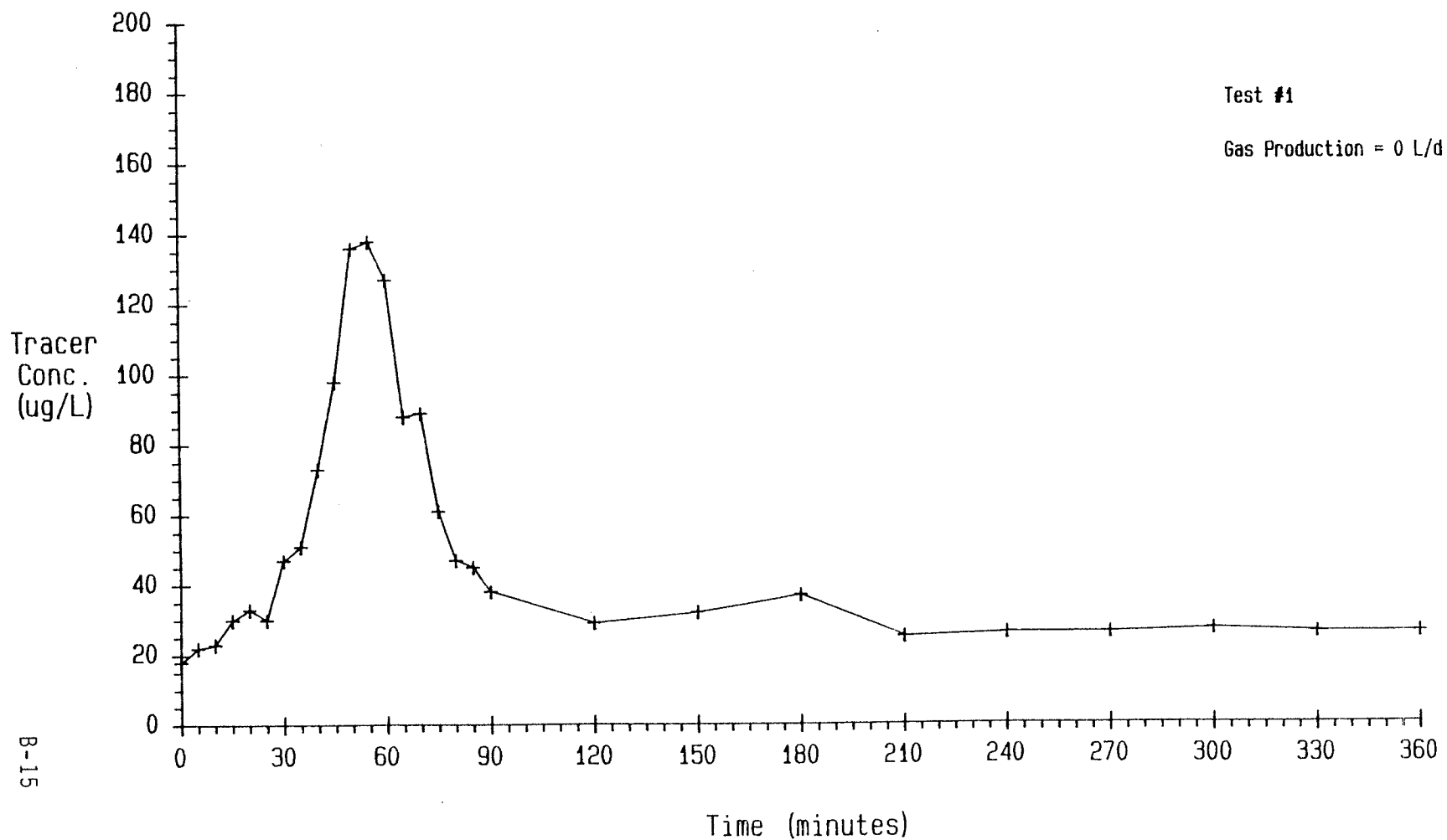
# Tracer Concentration vs Time



B-14

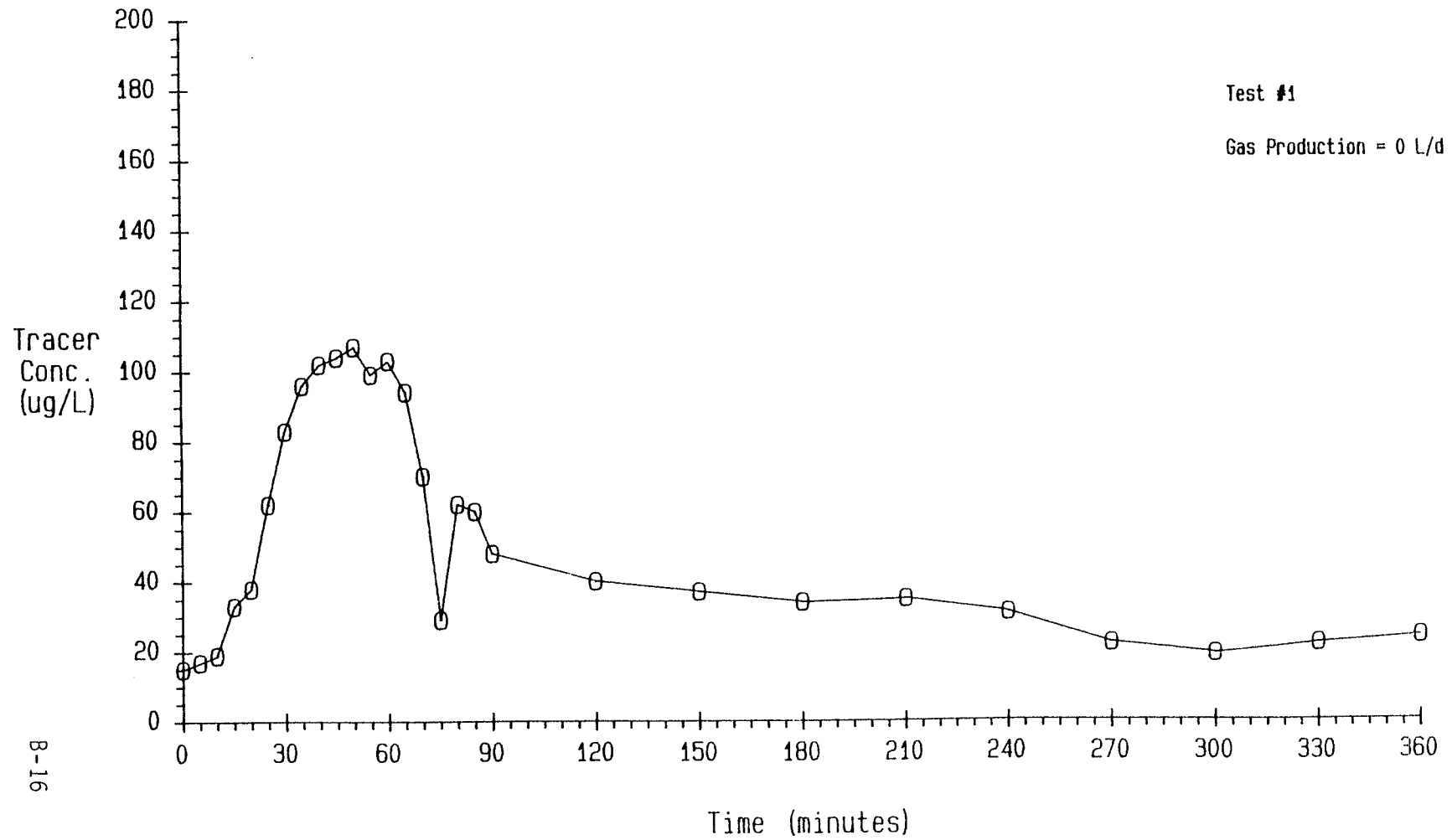
# Tracer Concentration vs Time

## Reactor 1



# Tracer Concentration vs Time

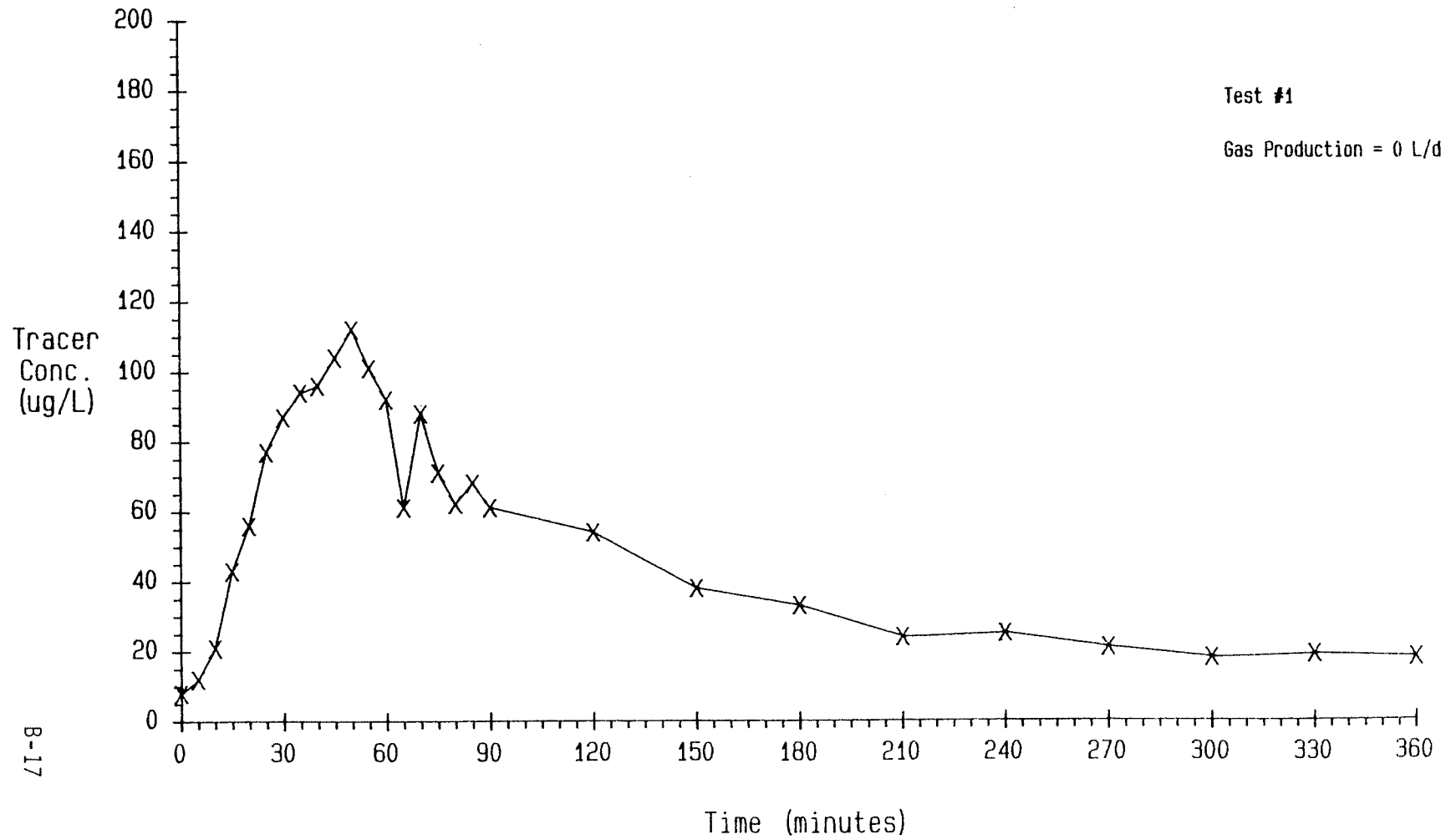
## Reactor 3



B-16

# Tracer Concentration vs Time

## Reactor 4

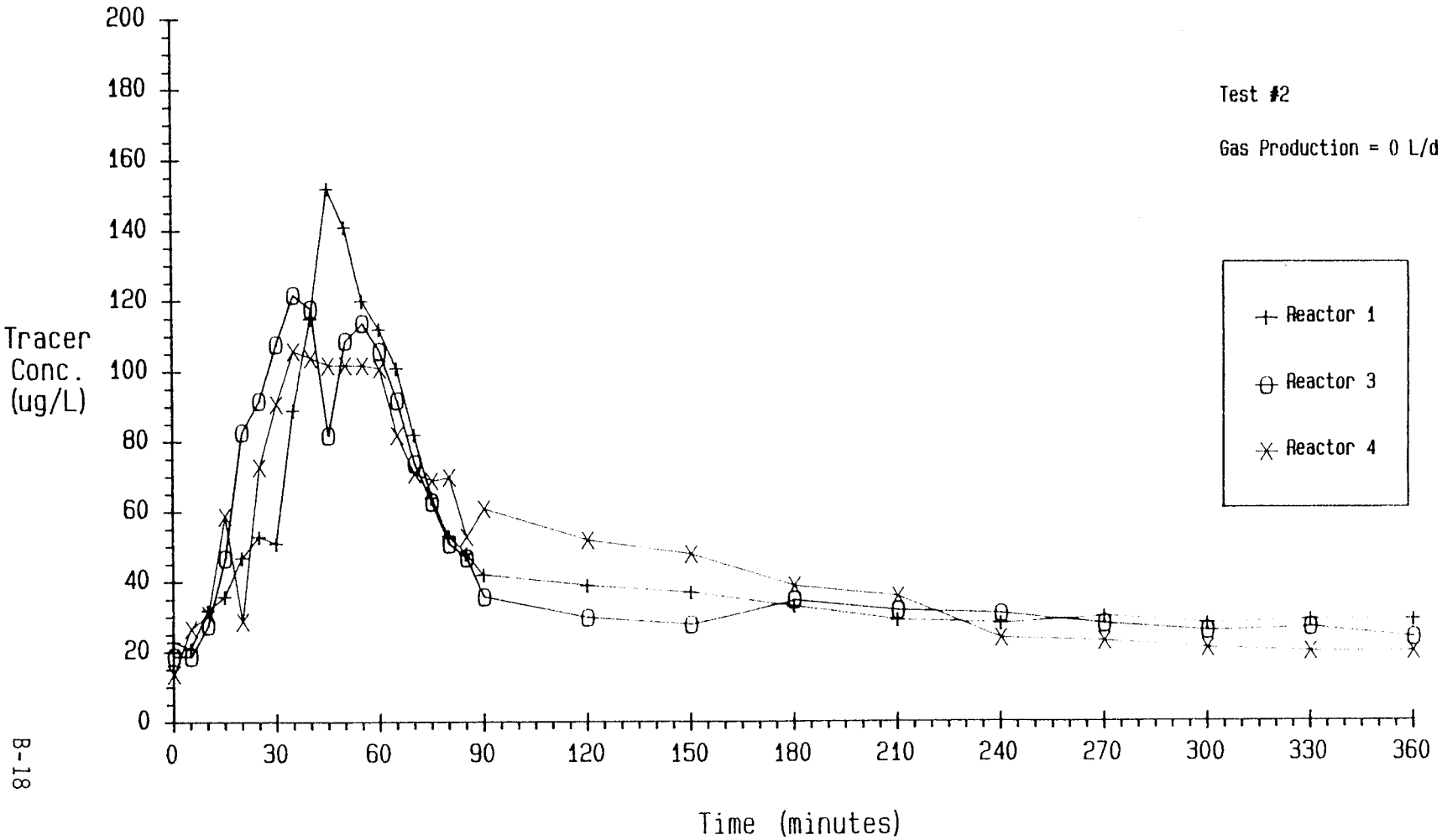


B-17

# Tracer Concentration vs Time

Test #2

Gas Production = 0 L/d



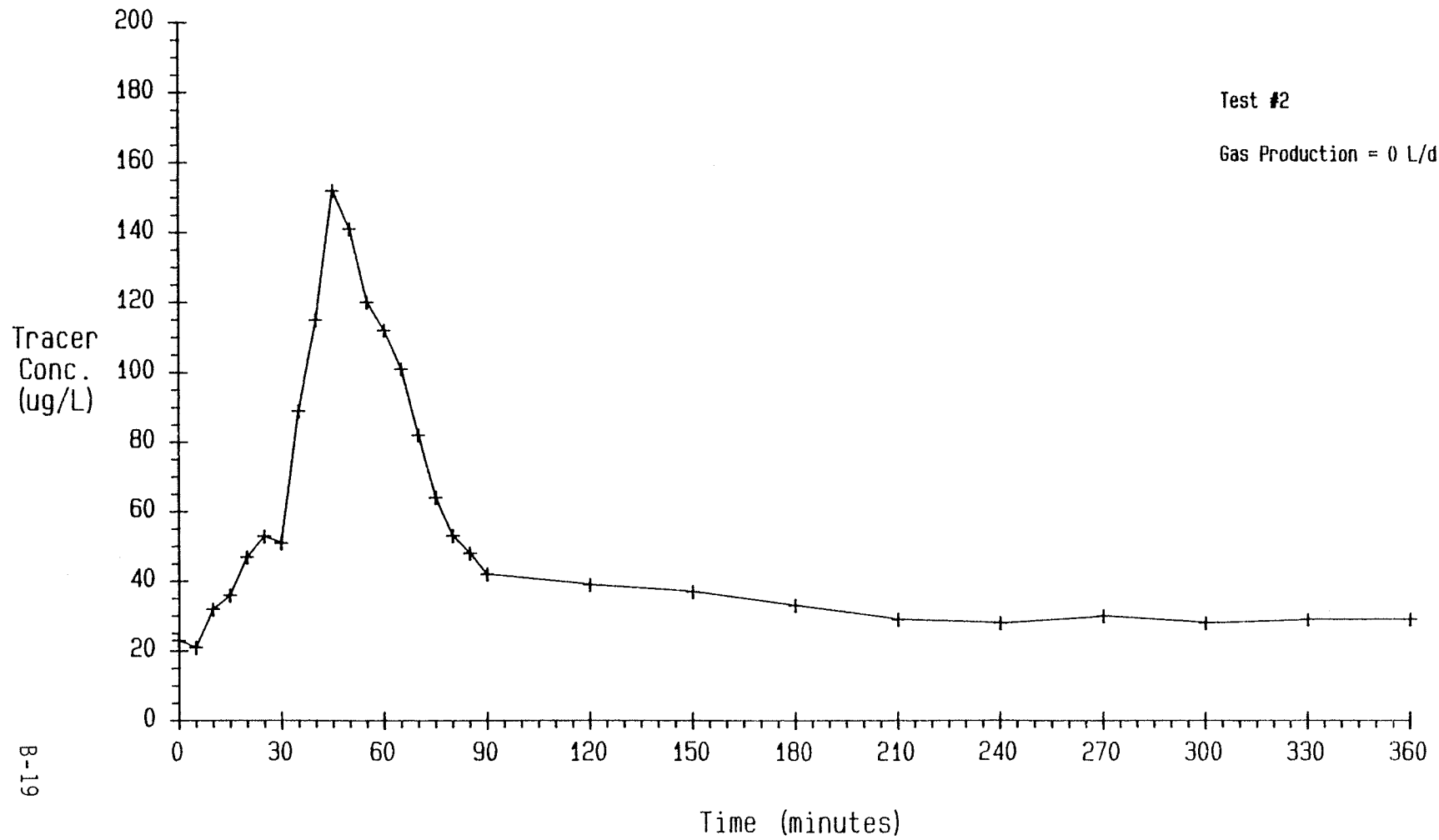
B-18

# Tracer Concentration vs Time

## Reactor 1

Test #2

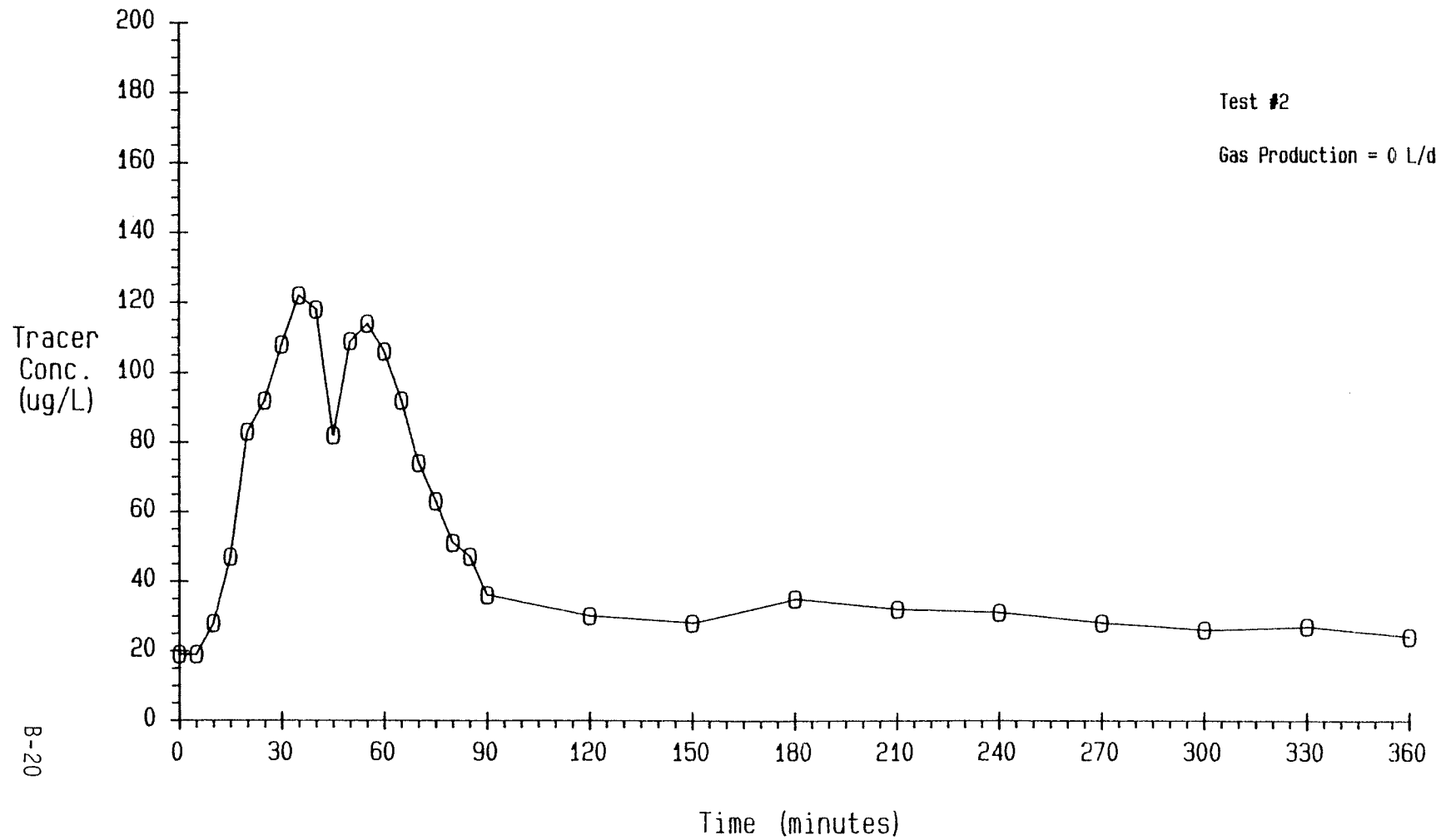
Gas Production = 0 L/d



B-19

# Tracer Concentration vs Time

## Reactor 3



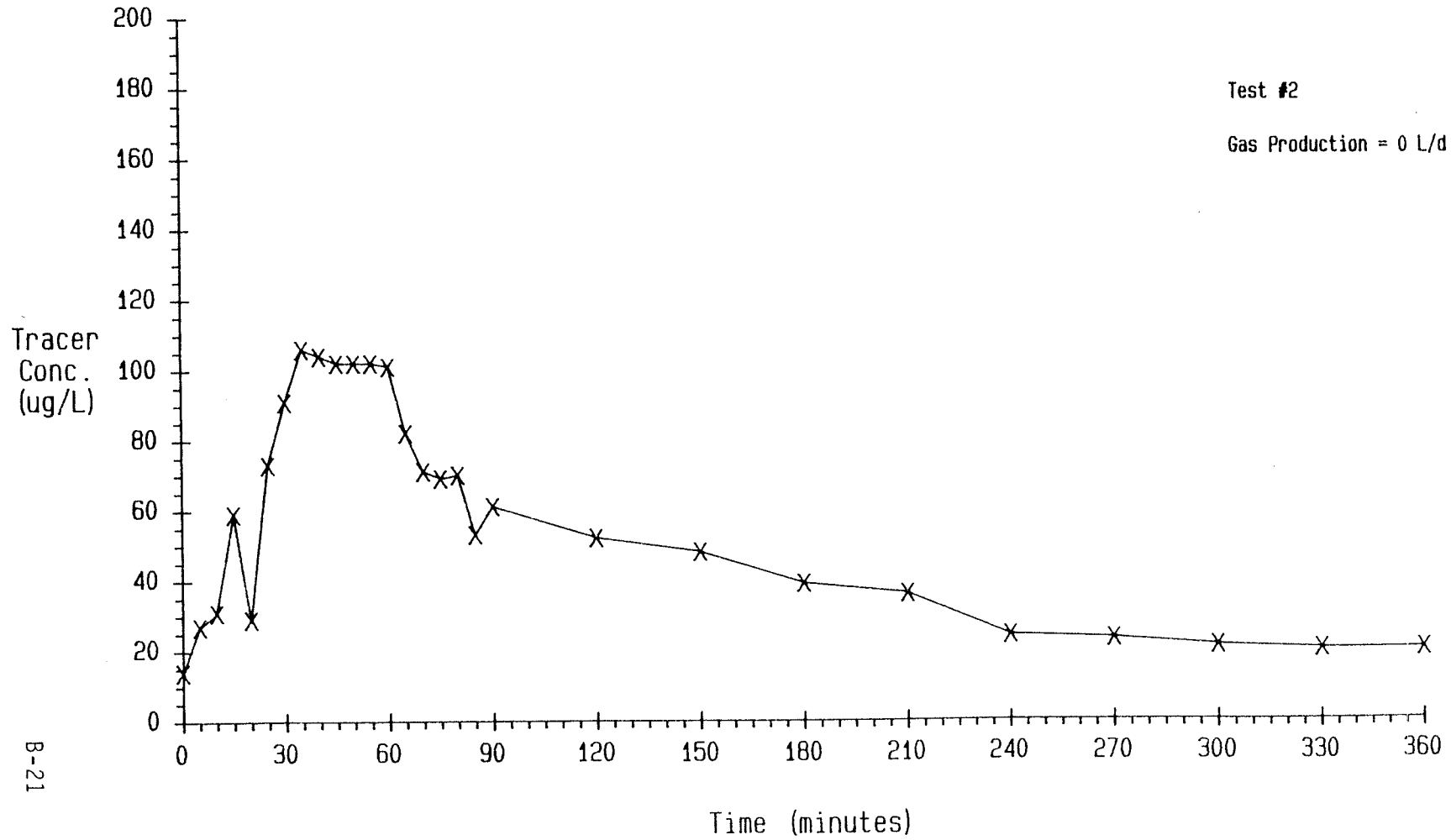


# Tracer Concentration vs Time

## Reactor 4

Test #2

Gas Production = 0 L/d

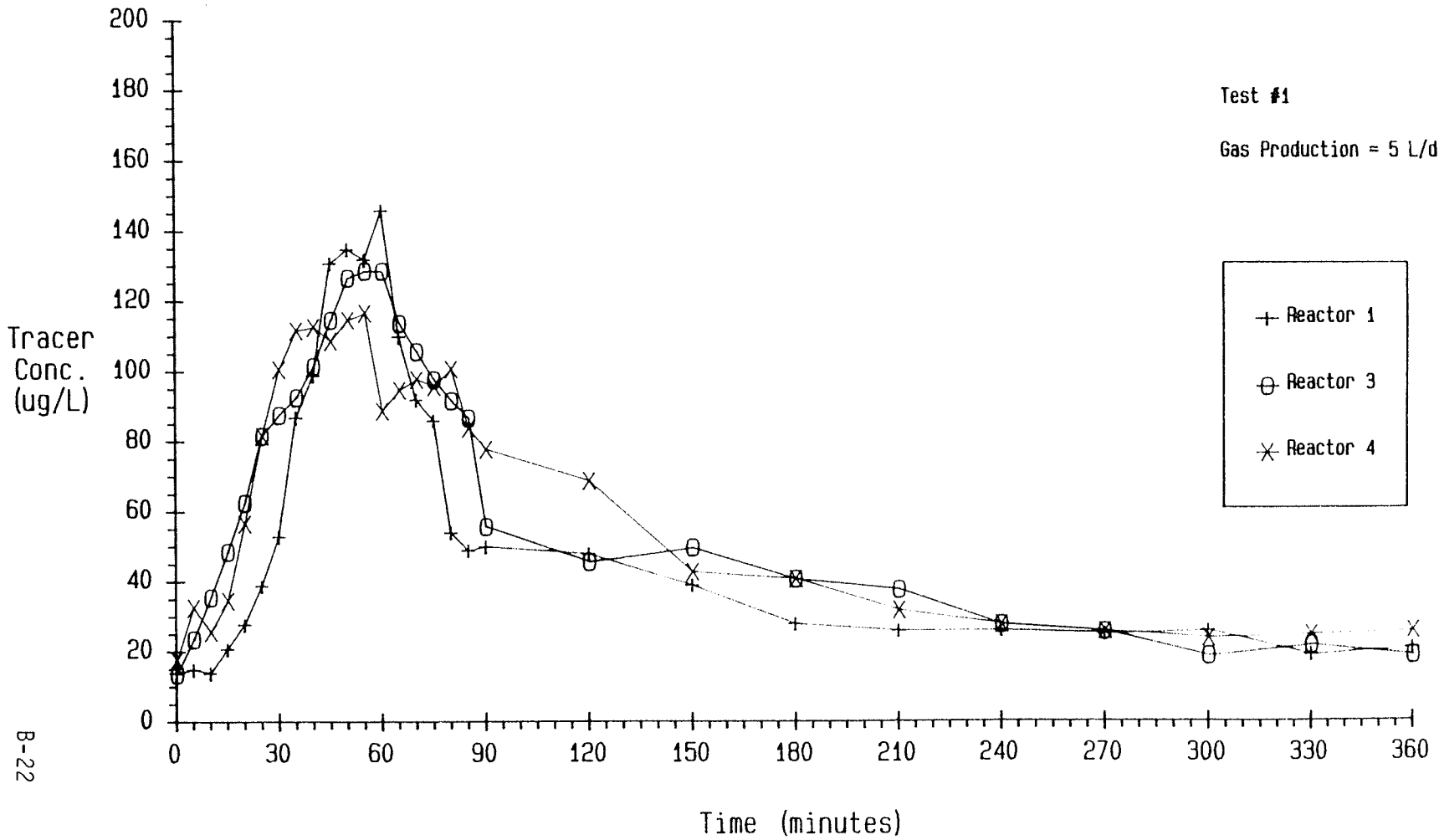


B-21

# Tracer Concentration vs Time

Test #1

Gas Production = 5 L/d

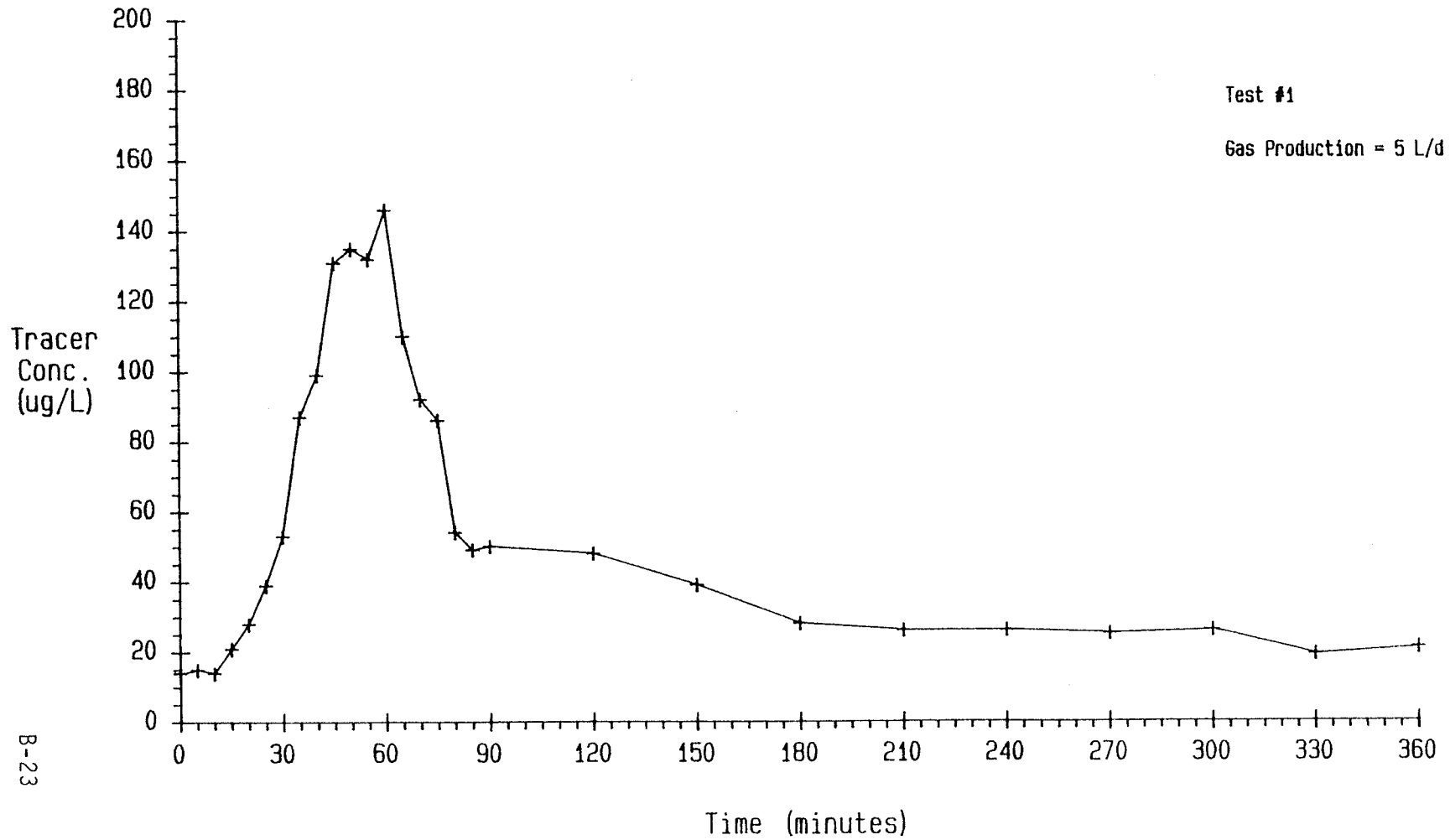


# Tracer Concentration vs Time

## Reactor 1

Test #1

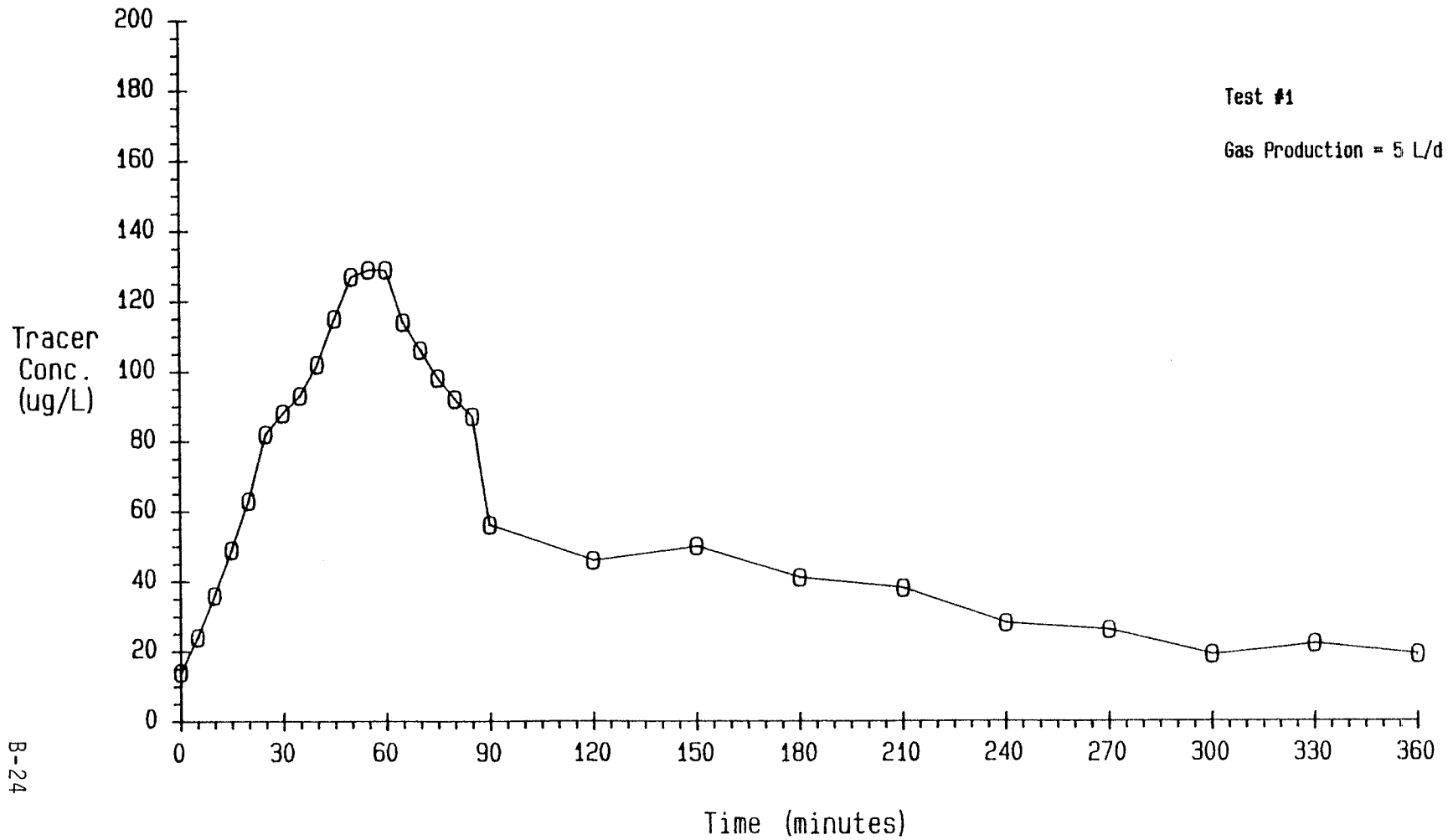
Gas Production = 5 L/d



B-23

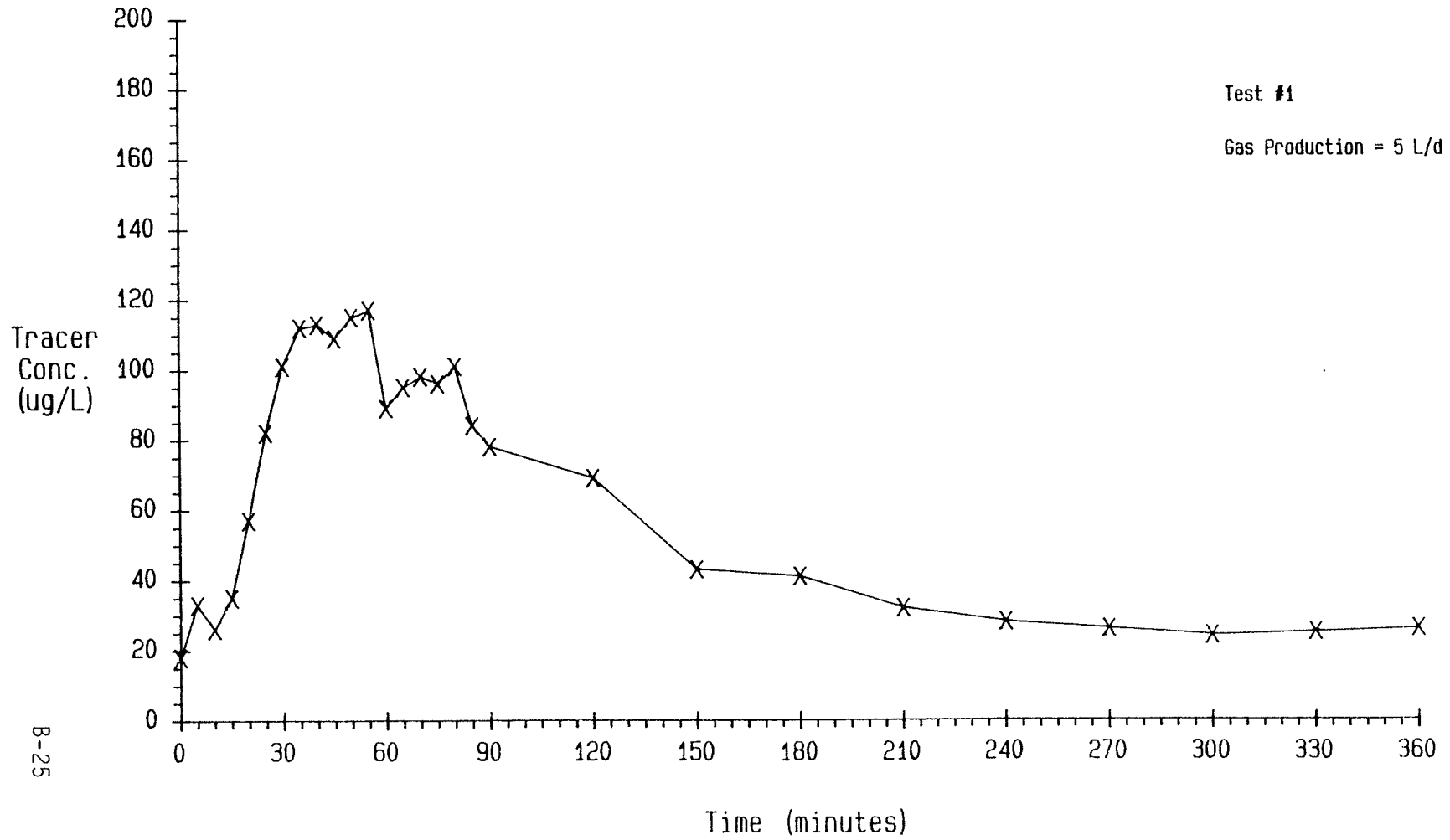
# Tracer Concentration vs Time

## Reactor 3



# Tracer Concentration vs Time

## Reactor 4

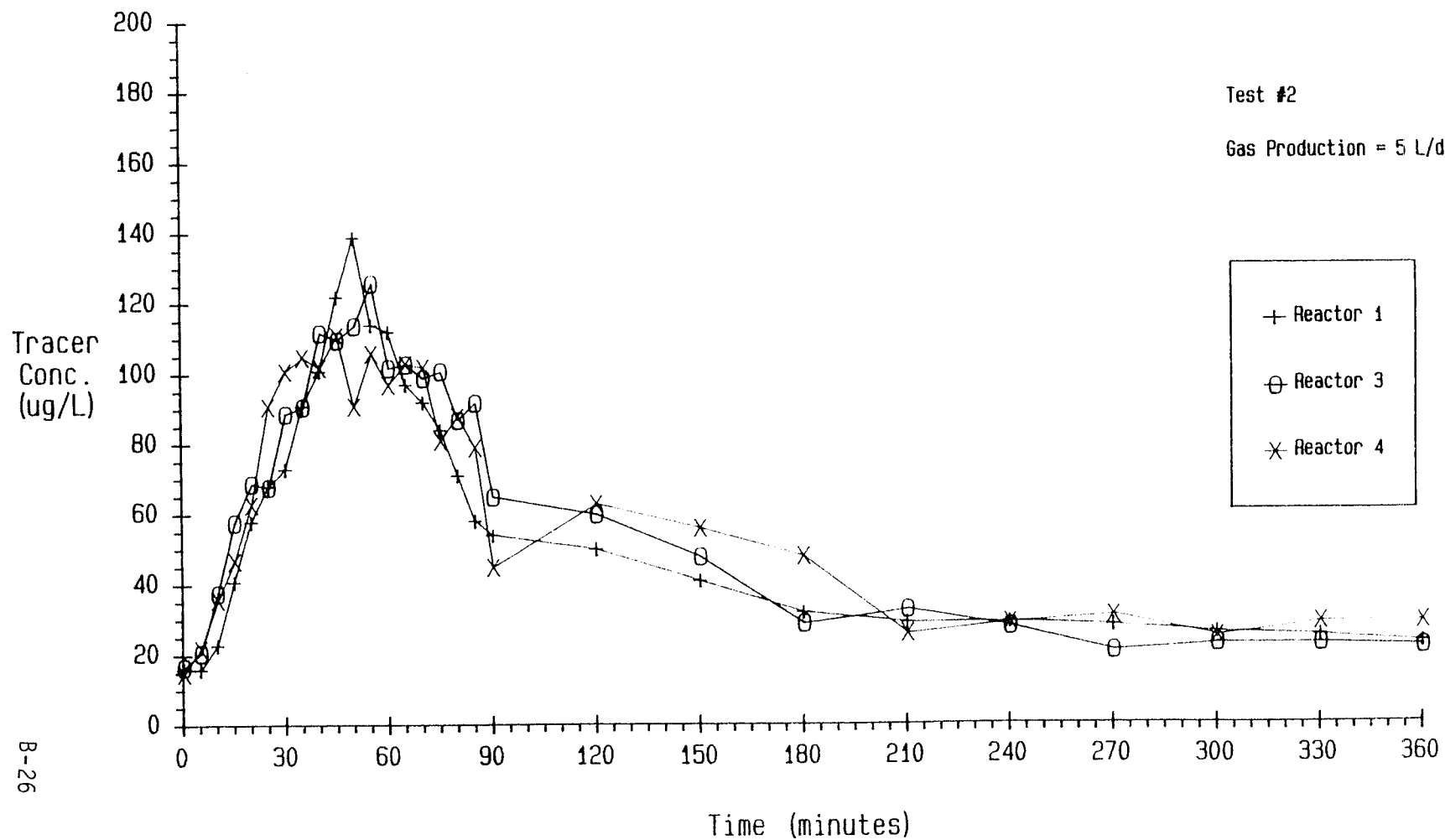


B-25

# Tracer Concentration vs Time

Test #2

Gas Production = 5 L/d

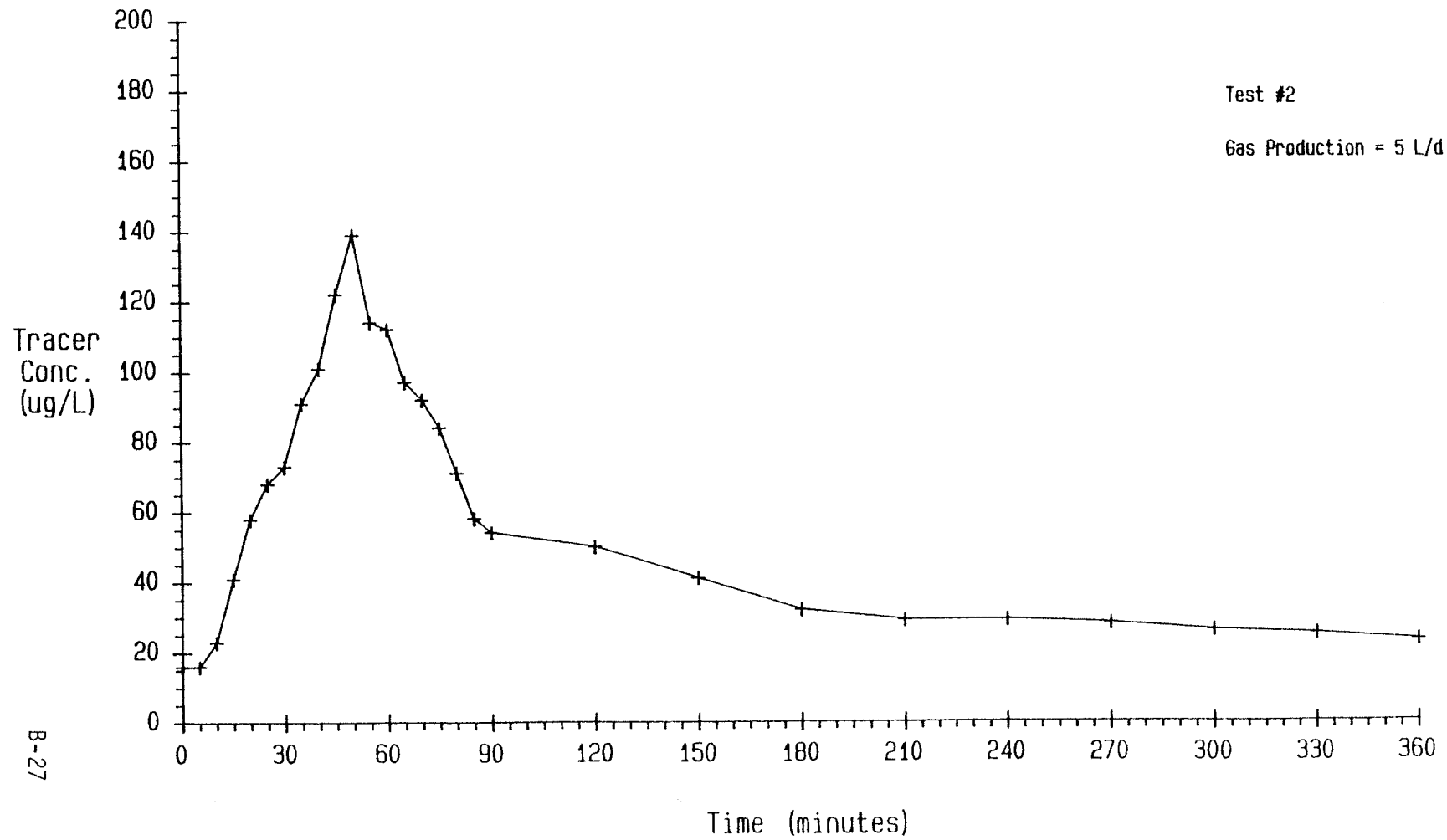


# Tracer Concentration vs Time

## Reactor 1

Test #2

Gas Production = 5 L/d



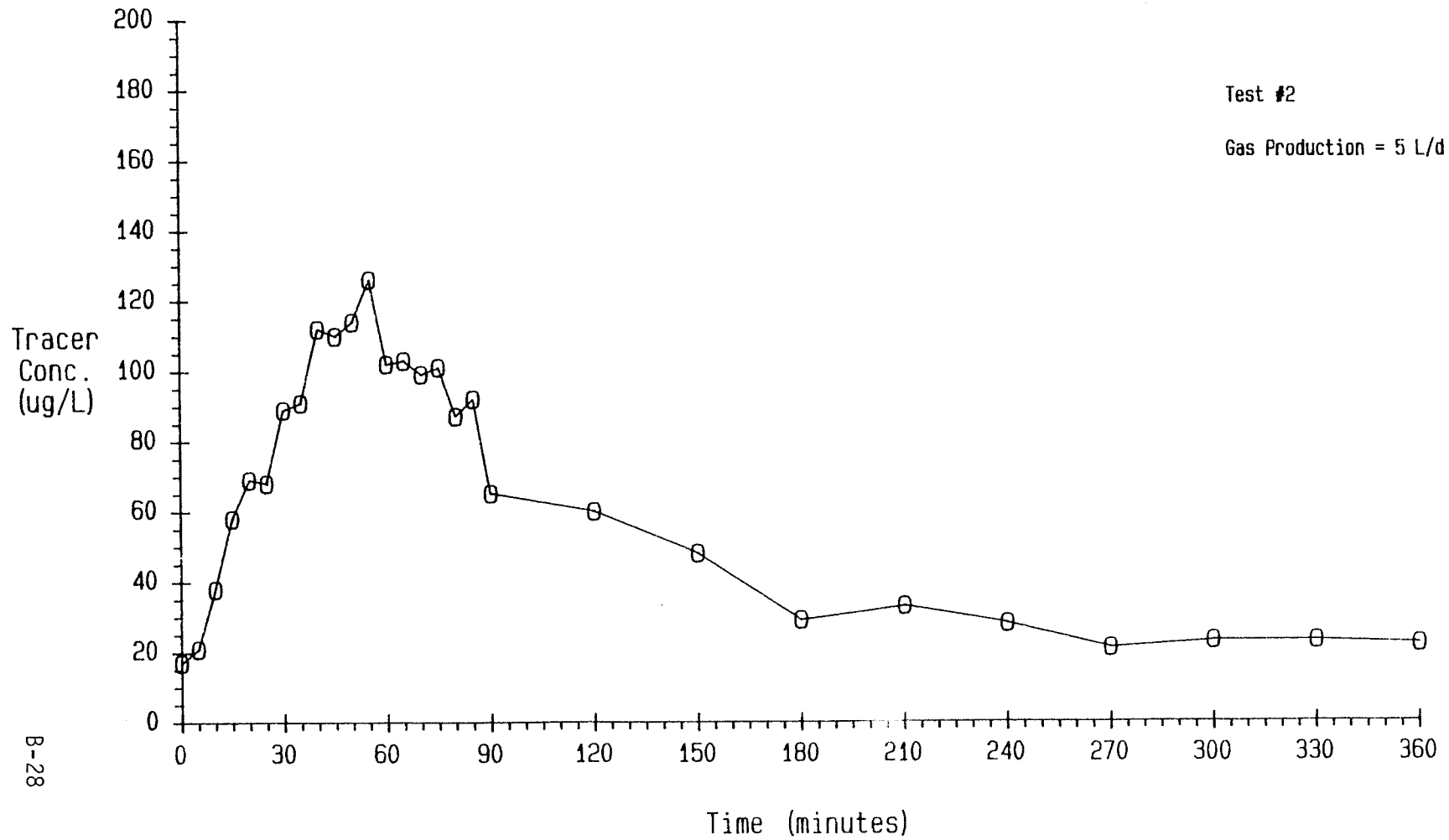
B-27

# Tracer Concentration vs Time

## Reactor 3

Test #2

Gas Production = 5 L/d



B-28



# Tracer Concentration vs Time

## Reactor 4

Test #2

Gas Production = 5 L/d

