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THE ANAEROBIC ATTACHED FILM EXPANDED BED REACTOR FOR THE TREATMENT OF DILUTE ORGANIC WASTES

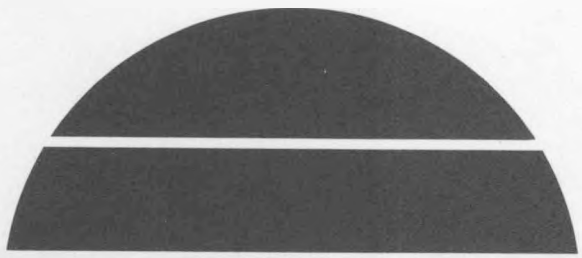
Technical Report

By
Michael S. Switzenbaum
William J. Jewell

August 1978

Work Performed Under Contract No. EY-76-S-02-2981

Cornell University
Department of Agricultural Engineering
Ithaca, New York



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Michael S. Switzenbaum
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William J. Jewell,
Project Director

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Department of Agricultural Engineering
New York State College of Agriculture and Life Sciences
A Statutory College of the State University at Cornell University
Ithaca, New York 14853

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THE ANAEROBIC ATTACHED FILM EXPANDED BED REACTOR FOR THE TREATMENT OF DILUTE ORGANIC WASTES

Michael S. Switzenbaum, Ph.D.

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A new process, the anaerobic attached film expanded bed reactor (AAFEB) has been found to be effective for the treatment of low strength soluble organic wastes anaerobically, at reduced temperatures, at short retention times, and at high organic loading rates. The process consists of inert particles, approximately 500 microns in apparent diameter, packed in a cylindrical column which expand slightly with the upward flow of liquid through the column. The AAFEB permits the maintenance of high solids retention times (SRT) values with low hydraulic retention times (HRT) values and it appears that most of the advantages of anaerobic treatment can be utilized with the disadvantages minimized with this process.

Three reactors fed a soluble synthetic waste consisting of glucose and nutrient salts at concentrations ranging from 50 to 600 mg/l COD were monitored over a period of nine months of start up and six months of operation. The effects of temperature, influent substrate concentration, and hydraulic flow rate on process efficiency were measured. Process efficiency was evaluated in terms of soluble COD removal and other parameters. In addition an

organic carbon mass balance was evaluated to verify the experimental results. Biofilm thickness and biomass concentration were also measured.

This study presents an analysis of the key process variables which affect AAFEB operation and presents two simplified first order equations relating the process efficiency to the net specific growth rate of the film and specific substrate utilization - two widely used operational parameters which are based on fundamentals of microbial growth and energetics.

The high effectiveness of this process is believed to be due to the large surface area to volume ratio created by the inert support media which enabled a large active mass of attached microorganisms to remain in the reactor at high liquid flow rates. Microbial mass concentrations exceeding 30 grams per liter were common in this reactor. The rate limiting step in the overall process was determined to be the biochemical reactions and not mass transfer. Finally, a preliminary energy consumption comparison was made between the AAFEB and conventional aerobic treatment processes for the treatment of low strength organic wastewater.

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CHAPTER 1. INTRODUCTION

A new anaerobic attached microbial film process has been found to be effective for the treatment of low strength soluble organic wastes at low temperatures and at short retention times. The process permits the maintenance of exceptionally high solids retention time (SRT) values, and with this process it appears that most of the advantages of anaerobic waste treatment can be utilized, with disadvantages minimized.

In the past, broad scale application of the anaerobic treatment process has been largely with the treatment of municipal sewage sludges to achieve waste stabilization and solids destruction. Over the more recent past new process configurations have evolved for application to both municipal and industrial waste waters. The anaerobic contact process (anaerobic activated sludge) has been applied to the treatment of wastes of several processes (McCarty (83), Schroepfer et al. (111)).

McCarty and Young (84, 131) developed the anaerobic filter for the treatment of concentrated soluble organic wastewaters. This submerged column filter, an attached microbial film process, was found to be able to treat dilute wastes (BOD greater than 750 mg/l) in a short period of time at 25°C. However, the process was found to be hampered by clogging and inefficient contact of the micro-organisms and the wastewater.

Recently research has been reported which appears to have the potential of developing an anaerobic system that would eliminate the problem of clogging and enable the use of short hydraulic retention times. Jewell and MacKenzie (63) showed that under aerobic conditions

static attached films had twice the organic removal capacity of a suspended microbial system under comparable conditions. In a subsequent study, Jewell (59) proposed the attached film expanded bed process (AFEB). This application of the attached film concept was based on the assumption that the concept could be most practically applied in an upflow filter bed composed of small light weight sand-sized particles.

While the two previous studies were limited to the aerobic stabilization process, Leuschner (75) and Jewell and Switzenbaum (64) demonstrated that it was possible to utilize this concept using an anaerobic film. Subsequent work has also shown that the concept can be applied to complex wastes such as cow manure (62). Jeris (56, 58), and Scott and Hancher (112) have used a similar process to remove nitrate from wastewater and found it to be highly efficient at short hydraulic retention times. More recently, Jeris (57) used a similar aerobic process for removal of BOD and nitrogen from wastewater, again with very promising results.

The purpose of this study was to examine the efficiency and mechanism of the anaerobic attached film expanded bed (AAFEB) as a process for the removal of low strength organics from a synthetic waste. The overall objectives of this study were:

1. To develop design relationships that can be used by designers for various applications of this study.

2. To determine the rate limiting steps and thereby understand the process in detail so that it can be improved even more and reach its potential.

3. To compare the AAFEB with conventional secondary wastewater treatment processes.

Specific objectives of this study were to:

1. Examine the efficiency of the AAFEB process in removing low strength organic over a range of temperatures.

2. Examine the effect of influent substrate concentration on process efficiency.

3. Determine the capacity of the process in terms of organic mass and volumetric loading rates.

4. Determine the maximum biomass concentration that can be maintained with the AAFEB process.

5. Examine the effect of film thickness on process efficiency.

6. Correlate the microbial yield with mass loading rate.

7. Correlate effluent suspended solids to process loading and flow rates.

CHAPTER 2. BACKGROUND

2.1 The anaerobic process for waste treatment

The anaerobic waste treatment process converts complex organic materials to carbon dioxide and methane in the absence of molecular oxygen. McCarty (84) states five important advantages of anaerobic processes. These are: 1) a higher degree of waste stabilization can be achieved, 2) a lower microbial yield, therefore, reducing the amount of sludge produced, 3) because of the lower yield, less nutrients are required, 4) this system requires no oxygen as do aerobic systems, and 5) methane gas is produced which is a useful end product.

McCarty (83), Kirsch and Sykes (66), Hobson et al. (52), Toerien and Hattingh (120), and Kotze et al. (68) have presented detailed review papers describing the microbiology, environmental factors, biochemistry, chemistry, and control of the anerobic digestion process. More recently, Mah (78) and Zeikus (132) have presented review papers discussing the physiology and biochemistry of the methane bacteria. Yet, despite its widespread use, the fundamental microbiology and chemistry of the anaerobic digestion process is poorly understood and at an elementary level (Lawrence and McCarty (74)).

Kirsch and Sykes (66) point out several disadvantages of the anaerobic waste treatment process: 1) Reasonably high rate digestion may require the elevation of the process temperature to one approaching 35°C, 2) the process has an apparent inability to

adjust quickly to sudden changes in nutrient concentration, nutrient composition, temperature and pH (this lack of accommodation is believed due to the slow growth of the methane forming organisms and the sensitivity of the organisms), 3) the inherent inability of anaerobic bacteria to degrade various species of organic compounds, and 4) a general feeling of unreliability of the process.

The methane forming bacteria are a fastidious group of organisms. They are obligate anaerobes and, as such, oxygen is extremely harmful. The optimum temperature for growth is in the range of 30 to 35°C for mesophiles and 50 to 60°C for thermophiles. pH is a very important factor. The general pH range is from 6.6 to 7.6 with the optimal range being from 7.0 to 7.2. Methane forming bacteria are very sensitive to changes in pH, as well as the pH itself. Consequently, anaerobic processes must be well buffered.

Kugelman and Chin (69), and Kugelman and McCarty (70) have studied the effects of potentially toxic materials on anaerobic fermentation. Substances such as sulfides, sodium, potassium, calcium, magnesium, ammonia, and soluble heavy metals all can have toxic effects on the anerobic fermentation process.

To summarize, the anaerobic fermentation has been used in the past for the stabilization of sludges and manures. Lack of fundamental knowledge of the process and its limitations have probably limited more widespread usage of the process.

With the increasing awareness of energy conservation, more attention is being forced on anaerobic fermentation because it produces methane as a byproduct. Jewell et al. (61) have pointed

out that aside from waste treatment, other benefits can be realized through the use of anaerobic fermentation of dairy manure such as energy production and nutrient conservation.

This study focusses on the potential of an anaerobic process for the treatment of low strength wastewaters. It will be shown that this new process has the potential of maximizing the benefits of anaerobic treatment and minimizing its disadvantages.

2.2 Conversion of complex organic wastes

The conversion of a complex organic waste to methane and carbon dioxide involves three stages, or three metabolic groups of bacteria. Figure 1 is a scheme showing the three general metabolic groups of bacteria or stages of the methane fermentation for a polysaccharide.

The fermentative bacteria are the first stage. They produce mainly short chain fatty acids - acetate, propionate, butyrate- and carbon dioxide and hydrogen from polysaccharides. The second group takes these volatile acids plus ethanol or lactate and produces acetate, carbon dioxide and hydrogen. This group has been called the hydrogen producing acetogenic bacteria (Bryant (16)). The third stage is the methanogenic bacteria. They are able to use few compounds and mainly get energy for growth by using electrons generated in their oxidation of H_2. Some methane bacteria can utilize acetate. This is of great importance since acetic acid is the immediate precursor of 72% of the methane formed from a complex waste (McCarty (83)). So acetate, hydrogen, and carbon dioxide, are produced by bacteria of groups 1 and 2 and the methane bacteria

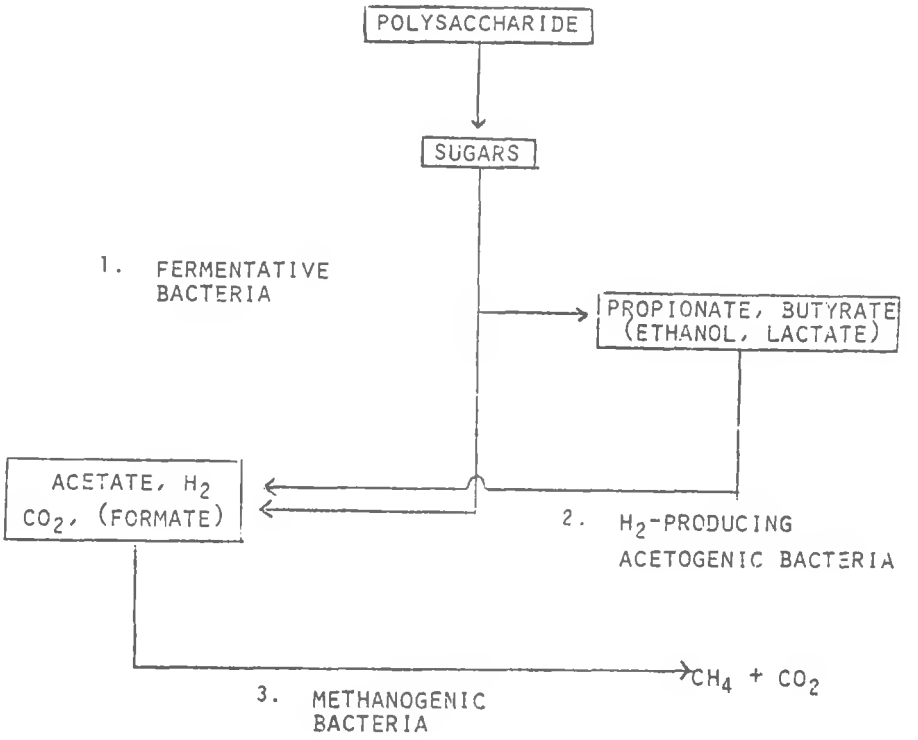

FIGURE 1. A SCHEME SHOWING THE THREE GENERAL METABOLIC GROUPS OF BACTERIA OR STAGES OF THE METHANE FERMENTATION (AFTER BRYANT (16))

catabolize these.

Bryant (16) has pointed out that hydrogen is a very important regulator in the methane fermentation. If the partial pressure of hydrogen builds up, it stops degradation of many organic compounds and instead forms reduced compounds which have as much energy as the original organic matter.

2.3 Anaerobic treatment of low strength wastes

In general, anaerobic fermentation has been used for the treatment of organic sludges and manures and high strength industrial wastes. The process is not considered to be practical for treating low strength wastes particularly substrates such as municipal wastewater with organics less than 500 mg/l (as volatile organics).

Winslow (128) used a biolytic tank to treat sewage anaerobically. This consisted of an inverted conical tank containing a blanket of digesting sludge. Sewage was fed into the bottom of the tank and flowed upward. Coulter (27, 29) made a modification of the biolytic tank on both a lab and pilot scale. At retention times down to 13 hours with sewage, good suspended solids removals were achieved (77%) but poor BOD removal (34 to 67%). This level of BOD removal is commonly reported for such diverse anaerobic processes as the septic tank and the static fixed film anaerobic filter.

Pretorius (106) developed a two stage anaerobic treatment process for raw sewage. The process consists of a sludge contact chamber and a biophysical filter. He was able to achieve 78 to 90% total COD reduction and 36 to 52% soluble COD reduction at retention

times on the order of one to two days at 20°C.

Young (130) in the first comprehensive study of the anaerobic filter concluded that the process was not able successfully to treat wastes with an influent COD of less than 750 mg/l which is greater than the normal range of values of domestic wastewater. VanDerMeer (122) using synthetic wastes with unconventional upflow filters found that the COD of these model dilute wastes with an influent COD of 500 to 1500 mg/l could be reduced to 10% at 30°C at retention times of 7 to 18 hours. This, however, is generally higher than normal settled domestic sewage and imposes a temperature constraint since sewage is generally colder than 30°C and is too dilute to produce enough gas by itself to maintain this reactor temperature.

In summary, the anaerobic fermentation process has not been considered practical for treating low strength wastes. Various studies reported in the literature report that wastes should be more concentrated than normal domestic sewage and warmer in order for the treatment process to be considered.

2.4 Types of anaerobic processes

Anaerobic waste treatment processes are generally considered to fall into three basic types based on reactor configuration. These include the conventional, contact, and anaerobic filter processes. These and a fourth type - the expanded bed - are discussed in this section.

2.4.1 Conventional processes

The typical conventional process is a large holding tank into

which wastes, usually high strength domestic or industrial, are fed continuously or intermittently. The wastes are usually maintained in the reactor for 30 to 60 days. The primary purpose of such a process is for waste stabilization and solids destruction. When first introduced several decades ago, these units were unmixed and unheated; hence the process was very slow and inefficient. More recent designs of the conventional process have evolved into heated, high rate, completely mixed systems. These innovations lowered the detention time to 15 days or less and increased the loading rate to 1.6 - 8 kg volatile solids/m^3/day from 0.48 to 1.6 kg solids/m^3/day (100-500 lbs/1000 ft^3/day from 30 to 100 lbs solids/1000 ft^3/day) as was practiced with the low rate units.

Modifications of the conventional process have been toward faster and more efficient treatment. The modifications have not been toward obtaining economical treatment of low strength wastes.

2.4.2 Contact processes

The anaerobic contact process developed from the concept of recycling biological solids to obtain a larger biomass for a longer retention time. It is an "anaerobic activated sludge process". The effluent from this system is pumped to a settling unit where a portion of settled sludge is returned to the reactor enabling the contact unit to maintain a high concentration of active mass. Thus solids concentrations can be made independent of waste flow using the method of biomass solids recycle.

Schroepfer *et al.* (111) initially developed this process for

meat packing wastes. This waste is primarily a dilute, soluble, highly biodegradable waste discharged in copious quantities. The BOD of this type of waste ranges from 800 to 1800 mg/l, 4 to 10 times more concentrated than domestic sewage, but 10 times less than the BOD of raw sludge. With good solids separation, the anaerobic contact process can be operated with an average hydraulic retention time as low as 12 hours. The anaerobic contact process has also been applied to the treatment of wastes of several industrial processes. Among them have been whiskey distilling; brewing; wine making; molasses, yeast, and starch production; cotton textile manufacture; and citrus fruit processing (83).

In order forthe anaerobic contact process to operate at the lower retention times, it must be operated at the optimum mesophilic digestion temperature of 35°C. Since the incoming waste is low strength, the amount of gas produced per volume of waste is not sufficient to keep the digester at a temperature of 35°C. Consequently, either the waste must be one which is inherently warm as it arrives to the unit or heat must be supplied at a significant cost. Schroepfer et al. (111) reported that with soluble wastes, the biological solids do not separate well and good settling can be accomplished only with degasification or the addition of some inert solid material to promote flocculation.

A different type of anaerobic contact process has also been reported in which wastes are passed through a blanket of concentrated anaerobic biological solids. Winslow and Phelps (128) first proposed this type of process and called it the biolytic tank. With the

biolytic tank sewage was fed into the bottom of an inverted conical tank and flowed upward through a blanket of digesting sludge. Coulter et al. (27, 29) later modified the "biolytic tank" type of process on a lab and pilot scale. Using sewage, at retention times down to 13 hours, good suspended solids removals could be achieved, 77%, but BOD removals were much poorer, 34 to 67%. Coulter and Ettinger (28), Pettet et al. (103), and Hemens et al. (49) used this later type of anaerobic contact process with food processing wastes with similar results.

In summary, McCarty (83) reported that anaerobic contact processes are better suited for the treatment of either concentrated or naturally warm wastes and, in general, they have not proved satisfactory for waste containing less than about 2000 mg/l BOD at temperatures below 30°C.

2.4.3 Anaerobic filters

The anaerobic upflow filter, or packed bed, is composed of one or more vertical filter beds containing some inert support such as rocks or plastic media which act as a stationary support surface for microbial film attachment. This type of system tends to permit an adequate solids retention time for the methane producing bacteria and still allows a short hydraulic retention time.

Young (130) described an anaerobic filter column constructed of plexiglass, filled with quartzite stones with a holding volume of 28.5 liters without stones, and 12 liters with stones. A synthetic soluble feed consisting of nutrient broth and glucose was used to simulate a low strength balanced waste (COD > 750 mg/l). Two of the

primary variables studied were the effect of hydraulic flow rate and organic loading.

Young (130) reported that reductions of 60 to 90% of the organic strength were achieved at organic loadings in excess of 750 mg/l as COD and with hydraulic retention times ranging from 72 to 18 hours at 25°C. When the flow rate was increased so that the detention time was 4 to 5 hours or less, the treatment efficiency of the system began to decrease and increases in effluent solids as well as undegraded organics were obtained. Treatment efficiency at high loading levels was high providing that the hydraulic retention time was also increased.

Advantages of the anaerobic filter as suggested by Young and McCarty (131) are high treatment efficiency of dilute soluble wastes, operation at 25°C, simplicity in design, no recycle of cells is required, slow accumulation of biomass, and good adaptation of filter to changes in feed strength and loading rates. The major disadvantages are the inability to handle wastes with high suspended solids, problems with clogging and short circuiting and low strength wastes having a BOD (ultimate biological oxygen demand) of less than 750 mg/l would not be acceptable.

El-Shafie and Bloodgood (36) investigated a multistage anaerobic filter system. This system consisted of a battery of six filters arranged in series so that the effluent from one filter was supplied to the next filter until the liquid waste finally received six separate treatment passes. Using diluted liquid diet food "Metrecal" as a source of balanced nutrient the waste was applied to the filters at

a concentration of 10,000 mg/l as COD at a hydraulic retention time of 18 hours. Removal efficiencies of 75 to 90% were achieved in the system overall.

DeWalle and Chian (33) evaluated a completely mixed anerobic filter in which the influent organic matter concentration is diluted with recirculated effluent. The laboratory scale anaerobic filter column was packed with plastic "Surfpac" slabs. The unit was found to effectively remove organic matter in high strength acidic wastewaters at a range of organic loadings and shock loads. A fixed film model was formulated and a comparison of the model to measured effluent concentrations tended to indicate that the substrate removal rate of the unit was primarily affected by substrate concentrations, specific surface area, flow rate and temperature.

Many other successful operations of the anaerobic filter on both bench and pilot plant scale were performed on various types of industrial wastes following Young and McCarty's initial study. Some of the various types of wastes included food processing wastes, Plummer and Malina (105); potato processing wastes, Pailthorp (100); wheat starch waste, Richter et al. (108), Taylor (119); brewery press liquor, Foree et al. (39, 40); pharmaceutical waste, Dennis and Jennett (32); and dilute waste sulfite liquor, Wilson and Timpany (127).

Mueller and Mancini (95) conducted a research effort to obtain a kinetic model of the anaerobic filter system based on laboratory studies. They concluded that the anaerobic filter appeared

to be a viable system of pretreatment of high temperature and high strength industrial wastes. Using easily degradable protein-carbohydrate waste at 35°C, the maximum attainable removal was found to be about 17.6 kg COD/m^3/day, (1100 lbs COD/1000 ft^3/day). At loading rates from approximately 3.2 to 27.2 kg COD/m^3/day (200 to 1700 lbs COD/1000 ft^3/day) and corresponding times of 24 to 3 hours, COD removals of about 90% and 50% were obtained respectively.

The anaerobic packed bed has potential application in industry in biochemical and chemical reactors utilizing immobilized enzymes (Weetall (125)) and for the conversion of waste materials to industrial intermediates including alcohols and organic acids (Compere and Griffith (23, 25, 44)).

2.4.4 Anaerboic expanded beds

The AAFEB process is similar to the chemical engineering processing filtration technology referred to as expanded or fluidized beds. The fundamentals and applications of fluidized beds to chemical engineering have been summarized (Zenz and Othmer (133)). Generally, the fluidized bed is used for gas-solids contacting, however, in some instances the presence of the gas or solid is used only to provide a fluidized bed to accomplish the end result (Perry and Clinton (102)). In most cases fluidization refers to more than a doubling in the reactor volume as caused by the high flow rate of gas through a filter composed of small particles. The term "expanded bed" has been used here to designate reactors that have an expansion of less than ten percent of the static volume.

More recently, the fluidized bed concept has been applied to biological systems. Enzymes can be immobilized on water insoluble materials retaining their catalytic activity (124) and these materials may then be used in a fluidized bed reactor to insure a large amount of contact between the enzymes and substrate. This can also be done with whole cells (Abbott (1)). These immobilization techniques and subsequent use in fluidized beds have been applied to the manufacturing of foods and beverages (124) and pharmaceutical compounds (1). Much effort during the last decades has been devoted to the possibility of converting batch microbial reaction processes onto a continuous flow basis (Lilly and Dunnill (77)). According to Denbigh and Turner (31) batch operation allows greater flexibility, particularly when small quantities of many different products are required. Continuous operation, though, has the following advantages: 1) diminished labor cost, 2) the facilitation of automatic control and 3) greater constancy in reaction conditions and hence greater constancy in the quality of the product.

Weetall (125) suggested that the fluidized bed is one basic type of reactor system applicable to immobilized enzyme technology. Several examples can be cited. Barker et al. (11) reported on the usage of a fluidized bed reactor for starch hydrolysis. Cheryan et al. (21) used the concept for skim milk coagulation. In both cases fluidized bed reactors successfully minimized column blockage. Fluidized bed reactors for the hydrolysis of particulate tributyrine in a fluidized lipase reactor was accomplished (Lieberman and Ollis (76)). O'Neill et al. (99) reported that fluidized bed reactor

efficiencies were greater than packed bed reactor during fluidization by viscous substrates.

Atkinson and Davies (6) proposed a completely mixed microbial film fermenter (CMMFF) based upon the fluidized bed principle for application to continuous operation when using growth associated systems. Unlike the conventional continuously stirred tank fermenter with suspended solids, the CMMFF allows high flow rates and according to the authors (6) there is no danger of the microbial mass being washed out. The fermenter contains microbial mass in the form of surface films and freely suspended flocs. In subsequent studies, the microbial film fermenter was described in a mathematical biological rate equation (7), and the CMMFF was evaluated in terms of any large scale operation involving any fermentation whether it be the production of biomass, primary or secondary metabolites, metal ions or BOD reduction (9).

The use of an expanded or fluidized bed prevents biomass buildup, accommodates particulates in the feed stream, is compatible with gas sparging, and allows easy removal or addition of the active material according to Scott and Hancher (112). They have developed laboratory scale tapered fluidized beds for aqueous bioprocesses in which adhering microorganisms or immobilized active biological fractions are used. According to the authors (112) the tapered reactor tends to stabilize the fluidized bed, thus allowing a much wider range of operating conditions. The tapered fluidized bed was evaluated for use in the enzymatic production of hydrogen, microbiological denitrification and microbial degradation and coal

conversion aqueous waste streams utilizing four tapered fluidized beds in series packed with coal particles with adhering microorganisms. Nitrate levels in a waste stream were reduced from 3500 to 13 ppm in residence times of less than 1 hour.

The fluidized bed concept has been used by Jeris et al. (56, 58) both in laboratory and pilot scale studies for denitrification. Jeris has stated that the following advantages are obtained in using the fluidized bed technique: 1) greater surface area available for growth per unit of reactor volume, 2) very small head loss, 3) no danger of clogging and 4) easier carrier removal procedure (56). On the pilot scale, the system consistently produced greater than 99% removal of the influent nitrogen in less than 6.5 minutes at a flux rate of 620 $l/min/m^2$ (58). More recently, Jeris (57) has used aerobic fluidized beds for BOD and nitrogen removal of domestic wastewater.

Jewell and MacKenzie (63) showed that under aerobic conditions the static attached film process had twice the organic removal capacity of a suspended microbial system under comparable conditions. In a subsequent study, Jewell (59) proposed the attached film expanded bed process. This application of the attached film concept was based on the assumption that the concept could be most practically applied in an upflow filter bed composed of small light weight 1 mm sized particles. This system was characterized by low hydraulic retention times in the active zone (average of 3 hours) and high rate biological treatment. Volumetric loading of 11.2 kg $COD/m^3/day$ (700 lbs of COD per 1000 ft^3) of reactor per day were employed with treatment efficiencies

of greater than 80%. At lower influent COD concentrations of 300 mg/l the efficiency of BOD removal was nearly 100% and the effluent contained only 5 mg/l and a turbidity of 2 units in a hydraulic retention time of 8 hours.

While the two previous studies were limited to the aerobic stabilization process, Leuschner (75) demonstrated that it was possible to utilize this concept using an anaerobic attached film. Effluent soluble COD concentrations of less than 20 mg/l were achieved at hydraulic retention times of only 6 hours. Subsequent work has also shown that the concept can be applied to complex wastes such as cow manure (62).

Jewell and Switzenbaum (64) demonstrated that the anaerobic attached film expanded bed reactor was capable of removing 80% of the total COD of domestic sewage at 20°C, anaerobically at loadings of 48 to 60 kg COD/m^3/day (400-500 lbs COD/1000 ft^3/day) at hydraulic retention times greater than one hour. This study is a continuation of this initial feasibility study.

2.4.5 Summary

The success of any anaerobic digestion process is dependent on bringing the waste into intimate contact with the biomass for a sufficient period of time to allow fermentation and gasification to occur (Kirshand Sykes (66)). The conventional process achieves this objective using a long HRT with a long biomass residence time. The contact process further improves upon this as it allows the cell residence time to be maintained independently of the hydraulic

retention time. The anaerobic filter offers an alternate means for maintaining a long SRT while permitting a relatively short HRT. But, none of these is able to treat low strength wastes such as municipal wastewater anaerobically.

The expanded bed technique appears to permit the maintenance of exceptionally high SRT's with concomitant short HRT's, on the order of several hours or less (75) (64). It also appears to be able to effectively treat low strength organic waste (COD less than 500 mg/l) under anaerobic conditions and reduce the problems of short circuiting and clogging common to anaerobic filters. This study sought to define the role of the anaerobic attached films in treating dilute wastewaters at lowered temperatures. Successful development of this technology would have widespread application.

CHAPTER 3. THEORETICAL CONSIDERATIONS

3.1 Fluidization - physical description

A fluid flowing at low velocities through a porous bed of small solid particles may not cause the particles to move. However, if the fluid velocity is steadily increased, a point is eventually reached at which the particles no longer remain stable, but "fluidize" under the action of the fluid. At that point the particles are supported by the fluid. In water treatment, the process has been used in back washing rapid sand filters to clean them by the resulting scouring action. The hydraulics of fluidized beds for filter back washing has been described (Fair, Geyer, and Okun (38)). The process is widely used in many catalytic processes because fluidization ensures contact of the fluid with all parts of the solid particles.

When the fluid velocity through a bed of solids causes all of the particles to be entrained in the fluid and to be carried along with it, this is called continuous fluidization. This is used primarily to transport solids either in a liquid stream or a gas stream. In most industries the suspending fluid is a gas.

The mechanism of fluidization and the design of fluidized bed systems have been thoroughly described (Perry and Clinton (102), Zenz and Othmer (133), McCabe and Smith (82), Bennett and Myer (12)). The following will serve as a brief description of the mechanism of fluidization with a liquid system.

A short vertical tube is partly filled with a fine sand. A fluid is admitted at the bottom at a low rate and flows upward

through the sand without causing the grains to move. If the superficial fluid velocity is increased, the pressure drop through the bed increases as shown by line OA in Figure 2 from McCabe and Smith (82). Eventually the pressure drop equals the force of gravity on the particles and the grains begin to move. This is point A on the graph. Then the bed begins to expand, but the particles are still in contact. During this period the porosity increases and the pressure drop rises more slowly than before due to net effect of increased porosity and velocity. When point B is reached, the bed is in the loosest possible condition with the grains still in contact. Between point A and B the bed is unstable, the particles begin to lose contact and then adjust their position to present as little resistance to flow as possible.

As the velocity is further increased, the grains separate and true fluidization begins. This is point F on the graph. By the time this point is reached all particles are in motion and beyond that point the bed continues to expand and the particles move in more rapid and more independent motion. The bed continues to expand as the velocity is increased and maintains a uniform character. The particles move in random directions through all parts of the liquid. Strong transient currents with many particles temporarily traveling in the same direction can be observed, but, in general, the particles move as individuals. This phenomenon is known as "particulate fluidization".

During this phase, the linear velocity of the fluid between the particles is much higher than the superficial velocity. Consequently

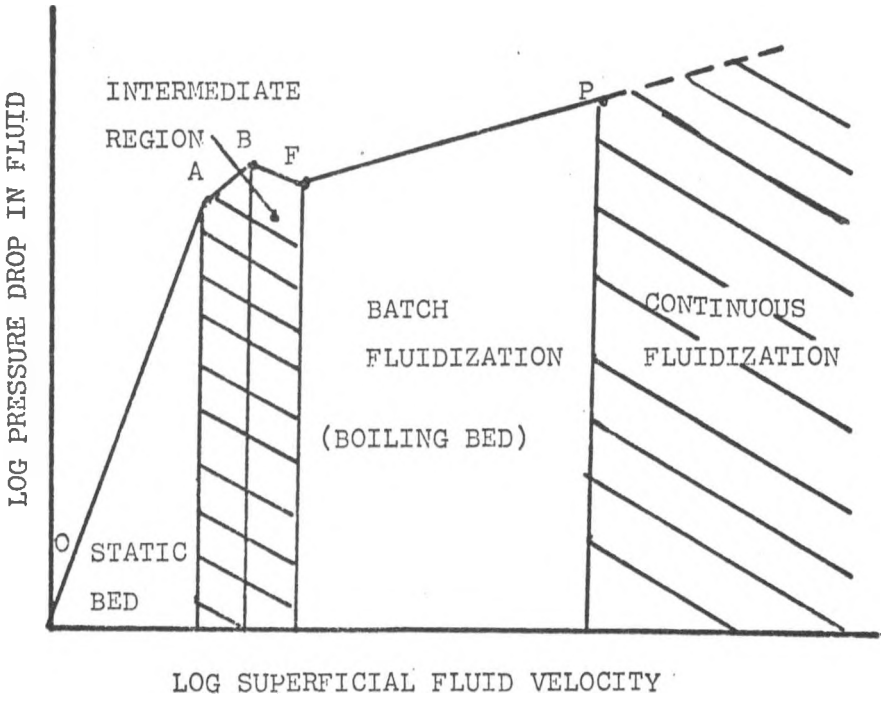

FIGURE 2. PRESSURE DROPS IN FLUIDIZED BED (McCABE AND SMITH(82))

any particles carried above the bed drop back into it. As the fluid velocity is increased further, the porosity increases, the bed of solids expands, and eventually at point P on the graph, all of the particles have been entrained in the fluid. The porosity approaches one and the bed ceases to exist. From this point on there exists the simultaneous flow of two phases. At point P, the superficial velocity is approximately equal to the terminal settling velocity of the particles.

As mentioned earlier, when point B is reached the bed is in its loosest possible condition with the grains still in contact. At that point the bed has expanded somewhat from its static position. The porosity of the bed when true fluidization begins is called the "minimum porosity for fluidization" and is designated by E_m. This corresponds to the porosity of a packed bed of hardly any weight. (Porosity is defined here as the volume fraction of voids in a bed of solids). Table 1 gives some representative values of E_m.

The following equations describe the relationship of flows and particle characteristics. The height of a fluidized bed can be predicted knowing the porosity of the unexpanded bed and expanded bed and the height of the unexpanded bed. When the fluid velocity is increased above the minimum required for fluidization, the bed expands and the porosity increases. Since the total mass of particles remains constant,

Table 1. The minimum porosity (E_m) of various media for fluidization

Size (mm)	0.02	0.05	0.07	0.10	0.20	0.30	0.40
Sharp sand ϕ = 0.67	-	0.60	0.59	0.58	0.54	0.50	0.49
Round sand ϕ = 0.86	-	0.56	0.52	0.48	0.44	0.42	-
Mixed round sand	-	-	0.42	0.42	0.41	-	-
Coal and glass powder	0.72	0.67	0.64	0.62	0.57	0.56	-
Anthracite coal ϕ = 0.63	-	0.62	0.61	0.60	0.56	0.53	0.51
Adsorption carbon	0.74	0.72	0.71	0.69	-	-	-
Carborundum	.-	0.61	0.59	0.56	0.48	-	-

ϕ = particle shape factor (dimensionless) for spheres ϕ = 1

E_M - (dimensionless)

$$L \cdot S\,(1-E)\,\rho p = L_E\,S\,(1-E_E)\,\rho p \qquad \text{Eq (1)}$$

where L = height of bed (unexpanded)

L_E = height of expanded bed

E_E = porosity of expanded bed

E = porosity of bed (unexpanded)

ρp = density of particle

S = cross sectional area

If S, the cross sectional area, of the bed is constant, then

$$L_E = L\,\frac{(1-E)}{(1-E_E)} \qquad \text{Eq (2)}$$

The pressure drop in a fluidized bed can also be calculated. As was shown in figure 2, the pressure drop prior to fluidization increases as the superficial velocity increases. During this period the pressure drop is given by either of the following equations:

For a uniform sand:

$$\Delta P_i = \frac{150}{N_{Rei}} \left[\frac{\Delta L \rho}{\phi_s D_{Pi}}\right] \left[\frac{(1-E)^2}{E^3}\right] \left[\frac{\bar{V}o^2}{g_c}\right] \qquad \text{Eq (3)}$$

For a stratified sand:

$$\Delta P = 150 \left[\frac{LP}{\phi_s}\right] \left[\frac{(1-E)^2}{E^3}\right] \left[\frac{\bar{V}o^2}{g_c}\right] \sum_i^n \frac{Xi}{N_{Rei}\,D_{Pi}} \qquad \text{Eq (4)}$$

where N_{Re} = Reynold's number = $\dfrac{\phi_s\, D_{Pi}\, \bar{V}o\, \rho}{\mu}$

where ΔP = head loss

ρ = density of liquid

D_p = diameter of particle

$\bar{V}o$ = superficial or empty tower velocity

g_c = acceleration due to gravity

ϕ_s = shape factor

X_i = volume fraction of particles of size 1 in a bed of mixed particles

μ^1 = absolute viscosity

After fluidization occurs, the pressure drop is essentially constant until point P is reached. For all practical purposes it can be considered as constant and equal to the pressure drop at the onset of fluidization.

When fluidization just begins, the pressure drop through the bed counter-balances the force of gravity of the particles. The pressure drop at incipient fluididation, that is as the bed just begins to expand may be found by equating the force it exerts on the solids to the force of gravity minus the buoyant force of the displaced liquid:

$$\Delta P \cdot A = \frac{g}{g_c} [\rho p \ (1-E_M) \ L_M \cdot A - \rho(1-E_M) \ L_M \cdot A] \qquad \text{Eq (5)}$$

$$-\Delta P = \frac{g L_M}{g_c} \ (1-E_M) \ (\rho p - \rho) \qquad \text{Eq (5a)}$$

where A = area

g = gravitational acceleration

E_M = minimum porosity for fluidization

L_M = bed height of incipient fluidization

L_m can be found with equation 6, the Burke Plummer equation since L_1E and E_m are known, or L and E can be substituted for L_m and E_m.

$$-\Delta P = 1.75 \left[\frac{L\rho}{\phi_s Dp}\right]\left[\frac{1-E}{E^2}\right]\left[\frac{\bar{V}o^2}{g_c}\right] \qquad \text{Eq (6)}$$

Finally the velocity ranges for fluidization can be calculated. For uniform particles, the critical superficial velocity required to fluidize a particle of diameter D_p can be calculated once E_m is known. Equation 5a gives the pressure drop at the onset of fluidization. Also, at the onset of fluidization the pressure drop can be estimated by the Carman-Kozeny equation provided the $N_{Re} < 10$. The Carman-Kozeny equation is as follows:

$$-\Delta P = 150 \left[\frac{L}{\phi_s{}^2 Dp^2}\right]\left[\frac{(1-E)^2}{E^3}\right]\left[\frac{\mu_1 \bar{V}o}{g_c}\right] \qquad \text{Eq (7)}$$

or in terms of the Reynolds number

$$-\Delta P = \frac{150}{N_{Re}}\left[\frac{L\rho}{\phi_s Dp}\right]\left[\frac{(1-E)^2}{E^3}\right]\left[\frac{\bar{V}o^2}{g_c}\right] \qquad \text{Eq (7a)}$$

Equating equations 6 and 7 results in

$$\bar{V}o_M = \frac{g\,(\rho p-\rho)\,\phi_s{}^2\,Dp^2}{150\mu_1}\left[\frac{E_M{}^3}{1-E_M}\right] \qquad \text{Eq (8)}$$

where $\bar{V}o_M$ is the minimum superficial velocity for fluidization

To calculate the limiting superficial velocity before entrainment, first as shown in Figure 3, the porosity increases as the superficial velocity is increased during fluidization. The porosity continues to increase until it approaches 1 and then the particles behave independently. Their motion is then governed by Stoke's law for most

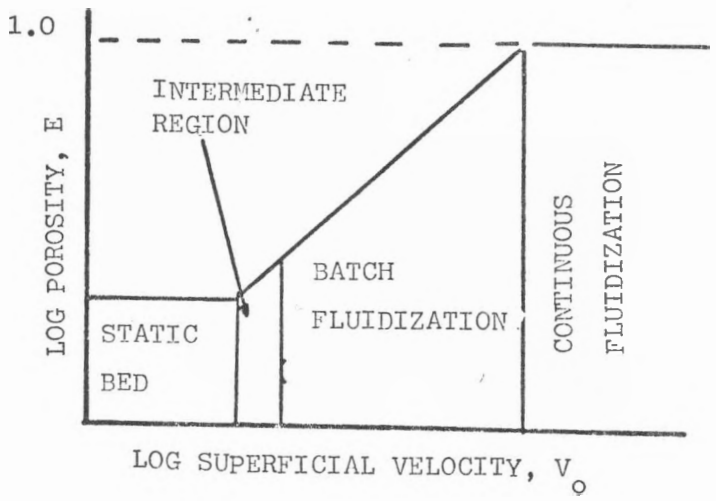

FIGURE 3. POROSITY OF FLUIDIZED BED (McCABE AND SMITH (82))

particles used and the limiting superficial velocity must be equal to the terminal settling velocity, as shown in equation 9.

$$Vo_L = Ut = \frac{g\ Dp^2\ (\rho p - \rho)}{18\ \mu} \qquad \text{Eq (9)}$$

where Vo_L = limiting superficial velocity

Ut = terminal settling velocity of particle

When large particles are fluidized by water it may be necessary to use the Burke-Plummer equation (6) to calculate the onset of fluidization and the Newton equation (9) to calculate the limiting superficial velocity.

For a range of particle sizes, rather than uniform particles, the critical superficial velocity for fluidization will be the velocity required to fluidize the largest particle size as calculated with the appropriate equation. The limiting superfical velocity will be that which would entrain the smallest size particle.

In this study, the expanded bed reactors were run just above incipient fluidization (area between points A and B, Figure 2). This allowed for energy conservation while ensuring that all of the film particles were in contact with the liquid flowing through the expanded bed.

The minimum porosity for fluidization E_M can be approximated for beds of particles smaller than 500 microns and larger than 50 microns in diameter from the empirical equation expressed by McCabe and Smith (82) as:

$$E_M = 1 - 0.356 \ (\log Dp^1 - 1) \qquad \text{Eq (10)}$$

where Dp^1 is the particle diameter in microns.

Since Dp^1 was measured to be 500 microns, $E_M = 0.38$

Now the minimum superficial velocity for fluidization, $\bar{V}o_M$ can be calculated from equation 8:

$$\bar{V}o_M = \frac{g \ (\rho p - \rho) \ \phi_s{}^2 \ Dp^2}{150 \ \mu^1} \left[\frac{E_M{}^3}{1-E_M}\right] \qquad \text{Eq (8)}$$

where ρp = density of particles = 2.79 gm/cm^3

ρ = density of liquid = 1.0 gm/cm^3

g = gravitational acceleration = 980 cm/sec

ϕ_s = shape factor, assume = 1

Dp = diameter of particle = 0.05 cm

E_M = minimum porosity to fluidization = 0.38

μ^1 = absolute viscosity = 0.01 = 0.01 gm/sec·cm

From equation 8, $\bar{V}o_M$ = 0.233 cm/sec. Multiplying this by the cross sectional area of the column, S, which equals 20.25 cm^2, gives a value of 4.71 ml/sec or 283 ml/min. This is in good agreement with the actual value of 300 ml/min at which the recycle pump was operated throughout this study.

3.2 Attachment of microorganisms to solid surfaces

A reasonable working hypothesis for the interactions between microorganisms and surfaces according to Atkinson (6) suggests that any surface in contact with a nutrient medium which contains suspended microorganisms will in time become biologically active due to the adhesion of microorganisms. Atkinson and Fowler (8),

Characklis (20), and Costerten et al. (26) have summarized the literature concerning the factors affecting the adhesion of microbial mass to surfaces. A general book on microbial attachment was recently published by Marshall (81).

Organisms are able to stick to surfaces by secreting a mass of tangled fibers of polysaccharides, or branching sugar molecules that extend from the bacterial surface and form a felt-like "glycocalyx" surrounding an individual cell or a colony of cells. Ironically, many microorganisms in pure culture do not form glycocalyx attachments. According to Costerton (26) cells that fabricate these elaborate coatings are usually eliminated from pure cultures by uncoated mutants that can devote more of their energy budget to proliferation. Since microbiologists work with pure cultures, most organisms studied in detail have been such naked mutants.

In most competitive natural environment which are populated by many species of microorganisms, natural selection favors those cells that can generate a glycocalyx and adhere to a desirable surface and be protected. Costerten (26) has said that the glycocalyx is essential to the biological success of most bacteria in most of the natural environment in which they are observed.

The glycocalyx may also be important in grouping bacteria in something approaching an organized community. This "consortium" effect has been pointed out by Costerten (26) with cells of a particular species adhering in a favorable niche close to the source of a necessary nutrient. This is important in carrying out

physiological processes. Costerton (26) has described consortia in which one species of hydrogen producing bacteria releases hydrogen from organic compounds and passes it on to another species of methane producing bacteria which uses the hydrogen as an electron donor to reduce carbon dioxide to methane.

Zobell (134) stated that the formation of microbial films, particularly of low substrate concentrations results from the fact that localized high substrate concentration existed due to the adsorption of substrate molecultes onto surfaces and that consequently microorganisms preferentially chose to reside in this area. Films, however, also may develop in conditions where there is no substrate limitation. Nordin et al. (98) showed the importance of ionic strength on the adhesion of microorganisms to solid surfaces.

Subsequent works on the mechanism of microbial adhesion and summarized by Atkinson and Fowler (8) revealed a very complex situation involving ionic strength, pH, liquid velocity, surface treatment and surface properties. They emphasized that pH appears to be particularly critical and can decide whether microbial films form at all. They concluded that in the formation of microbial films, complex biological and physio-chemical factors are involved, and there is no simple explanation for the phenomenon of adhesion of microorganims to surfaces.

3.3 Microbial films

Biological reactors can be broadly divided into two groups depending upon whether the microbial mass is in suspended flocs or

in films. Industrial fermentations using biological films include biological waste treatment, the "quick" vinegar process, animal tissue culture, and bacterial leaching. These are further described by Atkinson and Flowler (8).

In the laboratory, microbial film fermenters provide a useful tool for the study of microbial kinetics since certain operating characteristics are easier to control in a film fermenter than in a conventional stirred-tank fermenter (Atkinson and Fowler (8)). Examples of the uses of film fermenters in laboratory studies are also described further (8).

Most current methods for the production of industrial chemical products by fermentation are by batch rather than continuous procedures. Continuous systems would be more desirable than batch for several reasons. Generally, continous systems are smaller and less costly. Culture maintenance can be reduced since the system is seeded less frequently. Also for fermentation, less feed storage tankage is required (Compere and Griffith (24)). The major disadvantage in using a continuous system is wash-out of the culture. But this advantage can be overcome in a fixed film system (Atkinson and Davies (6)). A film system affords bound organisms some protection from toxic substances and sudden changes in the feed. Washout is frequently caused by toxic materials in the feed, phages, or toxic material produced by a contaminating culture.

Control of microbial film thickness is desirable. Films which are too thick cause problems in that organisms nearest the support surface are starved of nutrients. According to Atkinson and Knights

(9) this results in endogenous respiration when unwanted or even toxic products can be produced, followed by detachment from the support surface and collapse of the film. The fluidized bed concept allows the film thickness to be maintained at a constant level. Frequent particle-particle contacts occur causing the film to attain a dynamic steady state between the growth and attrition of the microbial mass (Atkinson and Davies (6)).

The productivity of growth associated fermentations is dependent upon the quantity of biomass contained in the fermenter. Atkinson (8) refers to this quantity as microbial hold up, and has summarized some of the measurements made of the thickness of the microbial films which occur in various reactor configurations. This is shown in Table 2 which divides films into four general groups, relating film thicknesses to general descriptions. These groups show a rough correlation between the weights and the general characters of the film.

The significance of thin films in regard to efficiency of waste conversion has long been recognized. McKinney (86) has stated that microbial mass should be kept as thin as possible in a trickling filter and that filter efficiency will be maximal when a thin layer of organisms persists. Hawkes (48) has pointed out that "a very thin film, indeed, perceptible only to the touch" was all that was necessary for efficient purification in trickling filters.

Kornegay and Andrews (67), Maier (79), Sanders (109), Tomlinson and Snaddon (121), and Hoehn and Ray (53) considered the effect of film thickness on the nutrient removal efficiencies of films.

Table 2. Groups of microbial films*

Group	Description	Film Thickness (mm)
I	Uncontrolled zoogloeal film	0.2 - 4.0
II	Zoogloeal film, subject to mechanical or hydrodynamic control	0.07-0.2
III	Pure cultures in CSTF	0.001-0.01
IV	Casual deposition	0.001

*Atkinson and Fowler (8).

Empirical evidence presented in these reports substantiates the qualitative statements of McKinney (86) and Hawkes (48) concerning the depth of effective film. Estimates of the effective film thickness from these reports ranged from 0.7 microns to 120 microns. Snaders (109) stated that a film reaches a sufficient thickness then uptake rate of nutrients is reduced with increasing film thickness. Maier (79), Kornegay and Andrews (67), and Tomlinson and Snaddon (121) postulated that a limiting thickness at which nutrient removal rates will be maximized will be obtained, but beyond that thickness, the rates will become constant and remain so until the film sloughs.

Hoehn (53) showed that the two theories advanced by these four investigations are not mutually exclusive. He stated that as films grow there will be a period of reduced nutrient capability once they obtain some limiting thickness. But given time, the films will adjust to the alteration of internal environmental conditions and will resume their activity at rates comparable to the former ones.

3.4 Film models

Toward the goal of obtaining a more rational design procedure for film processes, several attempts have been made based on fundamental theory, to mathematically describe the microbial process. Various attempts have described the systems as homogeneous (5) or heterogeneous (47, 126). In a homogeneous system a mass of microbes is completely dispersed throughout the reactor volume. No substrate

gradients exist between cells, and each microbe is randomly dispersed. A heterogeneous system is one in which the microbes are essentially separated from the fluid phase containing the substrate. In this system, an inert medium is used to support the growth of microorganisms. There are two phases of the system, one being the biofilm, the other being the liquid containing the substrate constituents. There exists a definite interface between the microbes and the liquid phase. The substrate must move across this interface in order to be utilized in the biofilm.

LaMotta (72) stated that when analyzing the different factors that affect the rate of substrate utilization by biological films, it is convenient to simplify all the steps occurring in the overall process into three major steps:

1. Diffusion of substrate from the bulk of the liquid to the interface between the liquid and the biological film.

2. Diffusion of substrate within the porous biological slime.

3. Biochemical reaction (substrate consumption) within the film.

In order to understand and adequately model the substrate uptake reaction by biological film, a proper identification of each one of these processes is required.

As summarized by Harris and Hansford (46), Ames et al. (2), Grieves (43), Gulevich (45), Kehrberger and Busch (65), Kornegay and Andrews (67), Maier et al. (80), Mehta et al. (88), Monadjemi (91), Pirt (104), Saunders and Bazin (110), Vaughn et al. (123), Jank and Drynan (55), LaMotta (71, 72), and Sylvester and Pitayagulsarm (118))

have done a significant amount of theoretical work, much of which was backed by experimental investigations. In all the above investigations, rate limitation was restricted to one reactant and assumed to be due either to mass transfer in the liquid, diffusion in the slime, or biochemical reaction.

If the AAFEB reactor is operated with high fluid velocities to bring the reactor operation to the reactor controlled regime and thus avoid external diffusional resistances, and if film thickness is small so that full substrate penetration within the entire film thickness can be obtained, then the process is limited by biochemical reactions within the film. The system may then be described as a homogeneous one. For simplicity, this will be assumed to be the case in this study.

The reaction of organics and microorganisms in a homogeneous system may be described in equation 11.

$$\frac{ds}{dt} = \frac{kSX}{Ks + S} \qquad \text{Eq (11)}$$

where S = substrate concentration (mass/volume)

k = maximum utilization coefficient (mass/time)

X = concentration of active microorganisms - (mass/volume)

Ks = the "half velocity coefficient", the substrate concentration at which the rate of reaction is one-half the maximum rate (mass/volume)

By dividing this equation by X, the microbial mass concentration, the result is:

$$U = \frac{kS}{Ks + S} \qquad \text{Eq (12)}$$

where the specific utilization, U equals ds/dt/X, which is the rate of removal of substrate per unit weight of microorganisms, $time^{-1}$.

Another equation is used to describe the relationship between net rate of growth of microorganisms and rate of substrate utilization as:

$$\frac{dX}{dt} = Y\left[\frac{ds}{dt}\right] - bx \qquad \text{Eq (13)}$$

where: dx/dt = net growth rate of microorganisms per unit of volume of reactor, (mass/volume-$time^{1}$

Y = growth yield coefficient, (mass/mass)

b = microorganisms decay coefficient, ($time^{-1}$).

Dividing equation 13 by X, yields the following equation:

$$\mu = YU - b \qquad \text{Eq (14)}$$

where μ = (dx/dt)/x- the net specific growth rate of microorganisms, ($time^{-1}$).

By combining equations 11 and 13 the relationship between growth rate and substrate concentration is as follows:

$$\mu = \frac{YkS}{Ks + S} - b \qquad \text{Eq (15)}$$

Equation 15 may also be expressed in the following form:

$$\mu \quad \frac{\mu Max\ S}{Ks + S} \qquad \text{Eq (16)}$$

where μ_{Max} = maximum net specific growth rate of microorganisms, ($time^{-1}$).

Equation 16 is widely accepted kinetic model proposed by Monod (92) for continuously fed, completely mixed, suspended pure cultures.

The solids rentention time, SRT, is directly related to μ. The SRT is defined as the amount of biomass in a given system divided by the amount of mass leaving the system over a given time period, or by definition.

SRT = X/(dX/dt) - the solids retention time, (time).

From this definition, the solids retention time is equal to the inverse of the net specific growth as defined previously. The solids retention time is a widely used operational control and design parameter in biological waste water treatment systems which is based on fundamental microbial concepts of growth and energetics.

Another approach can be used to describe the homogeneous system. Eckenfelder (34, 35) proposed an empirical mass balance equation relating process efficiency to the hydraulic retention time. This equation is expressed as follows:

$$\frac{Se}{So} = \frac{1}{1 + K^1 t} \qquad \text{Eq (17)}$$

where: S_o = influent substarte concentration, (mass/volume)

S_e = effluent substrate concentration, (mass/volume)

K = substrate removal coefficient, ($time^{-1}$)

t = hydraulic retention time, (time)

Both approaches, that of Monod and Eckenfelder, will be used in the analysis of data obtained during the laboratory investigation.

CHAPTER 4. EXPERIMENTAL PROCEDURE

4.1 Scope of Study

In accomplishing the objectives outlined in the first chapter, a study was conducted in the laboratory over a 15 month period varying temperature, organic volumetric loading rate, hydraulic loading rate, and influent substrate concentration and quantifying these parameters in terms of process efficiency. The variables tested and parameters measured are outlined in Table 3.

The study was done to begin to define the phenomena responsible for the efficiency of the process. This, in turn, would be useful in leading toward a rational design of the AAFEB process. The methods and materials employed toward reaching this objective are described in this chapter.

4.2 Experimental operating procedure

4.2.1 Apparatus

Three laboratory scale anaerobic attached film expanded bed reactors were constructed for use in the experimental study. Laboratory expanded beds were constructed of plexiglass columns, 6.35 cm in outside diameter and 49.36 cm tall. The lab scale AAFEB is shown in Figures 4 and 5, and repeated later in Figures 6 and 7. The inside of the column was 5.08 cm and the total volume was one liter. The base was constructed to disperse the waste uniformly across the bottom by using baffles and dispersion plates.

Table 3. Scope of study - variables tested and parameters measured

Variables tested	
Hydraulic retention time	6 to 0.33 hours
Influent substrate concentration	50 to 600 mg/l
Organic loading rate	0.8 to 43.2 kg COD/m^3/day
	0.05 to 2.7 lbs/ft^3/day
Temperature	10 to 30°C
Parameters measured	
Soluble COD removal	
Effluent suspended solids	
Effluent volatile suspended solids	
pH	
Volatile acids	
Gas production	
Composition of gas (% methane)	
Attached microbial mass	
Entrapped microbial mass	
Attached film thickness	

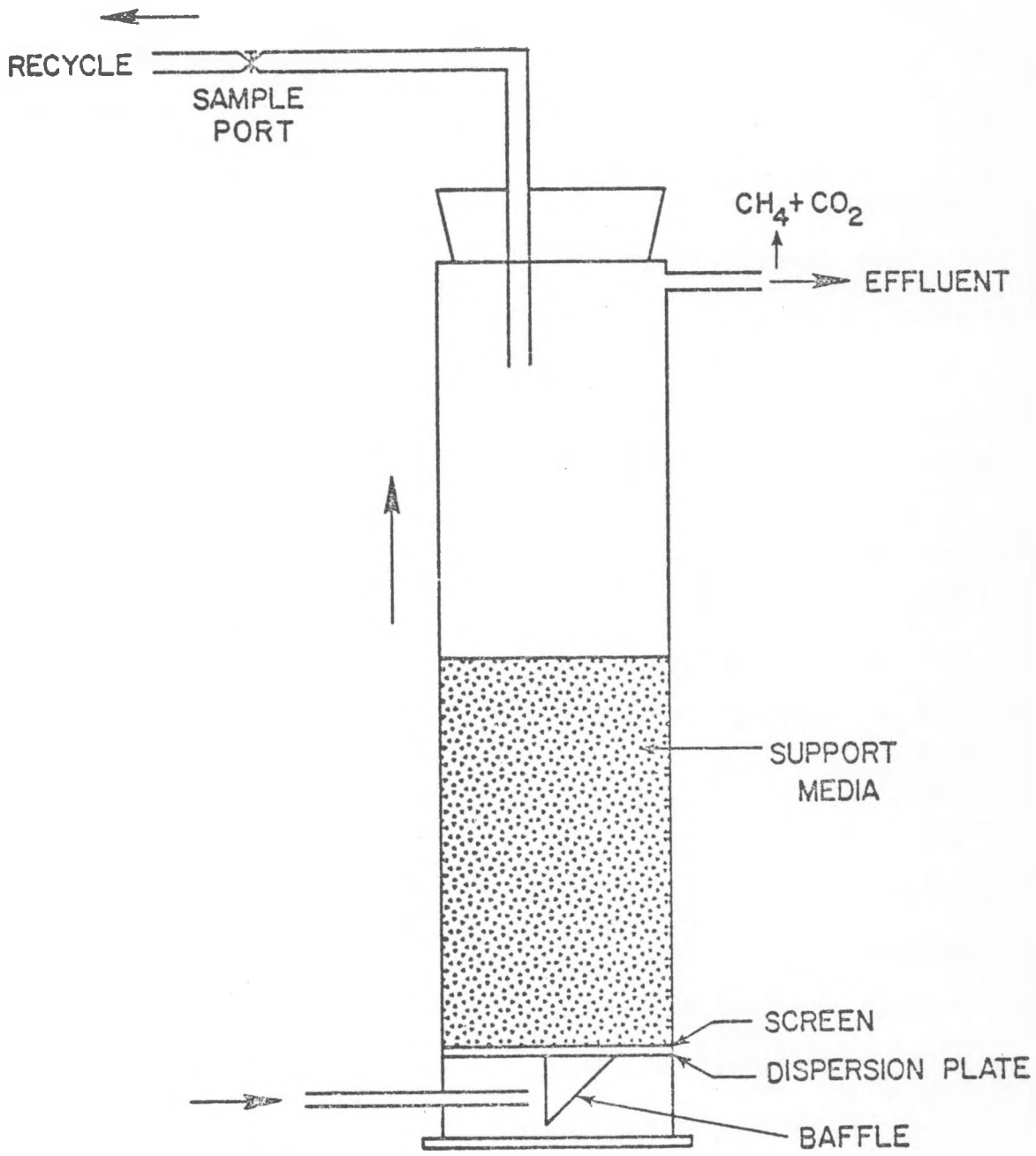

FIGURE 4. DIAGRAM OF THE ANAEROBIC ATTACHED FILM EXPANDED BED REACTOR

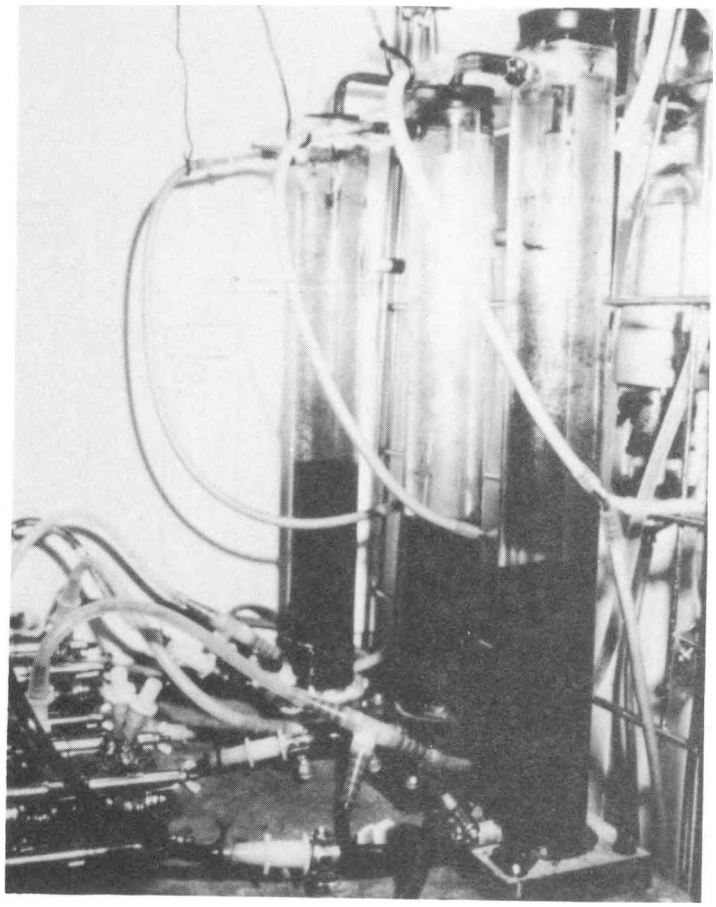

FIGURE 5. PHOTOGRAPH OF THE AAFEB REACTOR

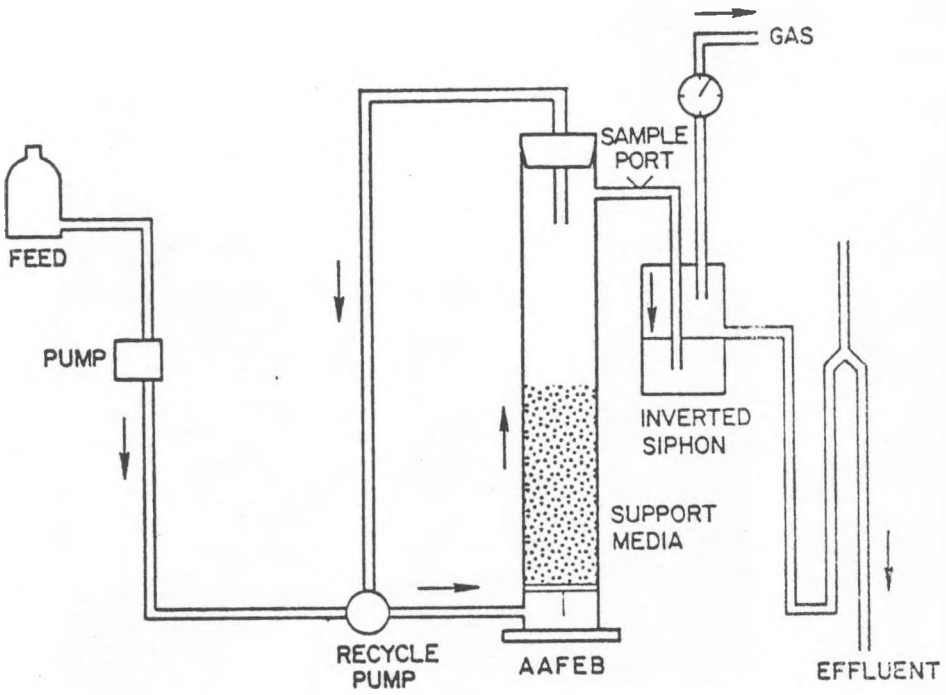

FIGURE 6. SCHEMATIC DIAGRAM OF THE AAFEB SYSTEM USED IN LAB STUDY

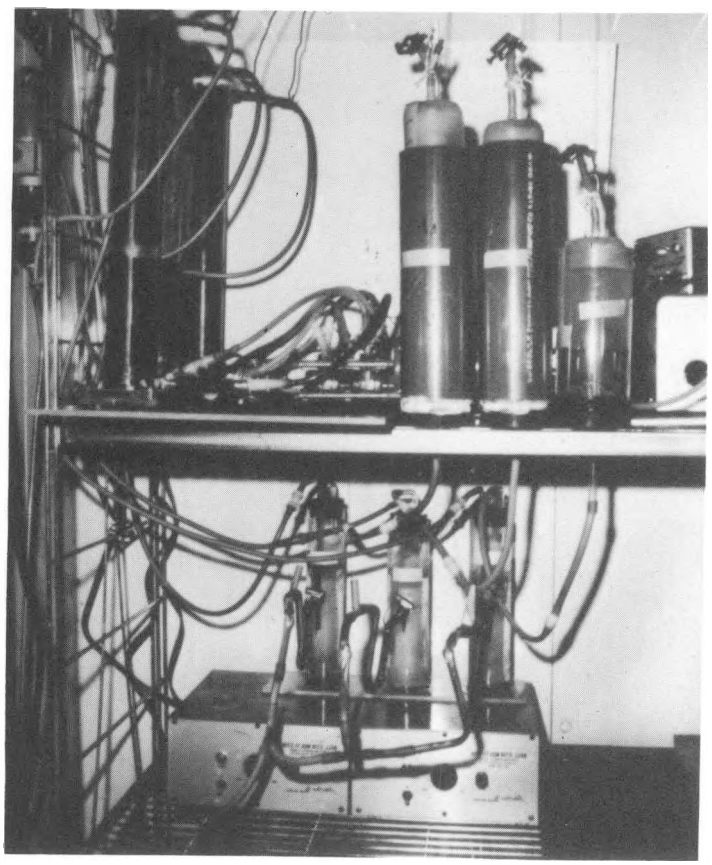

FIGURE 7. PHOTOGRAPH OF AAFEB SYSTEM USED IN LAB STUDY

After the substrate entered through the bottom of the column it passed over a baffle system. Above the baffles was a distribution plate consisting of 1/8 inch holes arranged in a concentric fashion. Above the distribution plate a 50 mesh nylon screen was secured to the plate to prevent the media from plugging up the distribution plate and feed lines.

Each column was filled with 160 g of support media which unexpanded occupied a total volume of 400 ml. The support medium was composed of aluminum oxide, a porous water insoluble inorganic biomaterial support manufactured commercially by Corning Glass Company. The particles were sieved for uniformity and the resulting particles had an apparent diameter of approximately 500 microns. The particle density of the support media was determined to be 2.79 gm/cm^3; the bulk density was 0.6 gm/cm^3. The support medium was estimated to displace a liquid volume of 90 ml. This material was chosen because of its uniformity and also because it could be fired to measure dry organic matter concentrations. It does not represent a material that could be considered in full scale application because of its high cost.

During operation the bed was only slightly expanded, to the 500 ml line by means of a recycle pump. The pumping rate was adjusted periodically to account for varying biomass in order to keep a constant expanded level.

The bottom of each reactor was scaled by means of a rubber "O" ring supported between two plates bolted together. These plates could

be separated by unscrewing the bolts when maintenance in the bottom plates was necessary. The top of each reactor was sealed by means of a rubber stopper.

A schematic diagram of the anaerobic attached film expanded bed system used in the lab study is shown in Figure 6, and a photograph is shown in Figure 7. Feed for each of the three expanded bed reactors was pumped directly from a refrigerator holding three feed reservoirs by an Ultraslow Speed Pump, Spectroderm Model MK-III, into the recycle lines.

The other pumps used in the system were a recycle pump, a Masterflex variable Speed Drive with a solid state controller (Model 7545, Cole Parmer Instrument Company). These pumps drew effluent from near the top of the reactor and pumped it into the bottom assembly. They were used to expand the beds as well as for recycle of the reactor effluents. These pumps were maintained at a rate of approximately 300 ml/min.

The reactor effluent first passed through a sealed contact chamber on a magnetic stirrer. The stirrer provided agitation to remove gas bubbles from the liquid portion. The effluent then passed through an inverted siphon to separate the gas produced from the liquid. The drain side of this inverted siphon was common to all three reactors. In this manner, sufficient flow could be maintained in the siphon to prevent its becoming plugged.

The separated gas was collected in an inverted calibrated cylinder in a column filled with water. At the top of each cylinder was a septum from which a gas sample could be obtained to determine

composition of the gas.

Tygon tubing was used to connect the feed reservoirs, pumps, and reactors in the AAFEB system. The tubing in the pump heads was changed each week to decrease the possibility of a variable feed rate resulting from worn tubing. The other tubing, outside of the pump heads was cleaned weekly to avoid microbial growth in the lines.

The entire study took place in a walk-in controlled temperature chamber. The chamber was manufactured by Scherr-Gillet and was able to maintain temperatures between 0° and 50°C, with temperature variation of $\pm$ 1°C. The synthetic feed used in the experimental program will be discussed in the next section.

4.2.2 Program of experimentation

A previous study (64) had shown that the anaerobic attached film expanded bed reactor was able to treat primary sewage anaerobically at 20°C at very low hydraulic detention times. Since the primary purpose of this study was process capability definition, it was decided to use a defined synthetic substrate. The stock feed solution was composed of various components to provide the system with a substrate of organic carbon, nitrogen, phosphorus, miscellaneous nutrients, and alkalinity which was used for pH control. The organic carbon and miscellaneous nutrients were provided in the form of glucose and yeast extract (Difco brand). Nitrogen was provided in the form of ammonium chloride, and phosphorus was in the form of mono- and dibasic potassium phosphate. Alkalinity was

added in the form of sodium bicarbonate. Table 4 shows the components of the synthetic feed and the amounts of each component which made up the concentrated feed solution. The average COD:N:P ratio of the feed solution was 100:5:1.

The stock feed solution was prepared in one liter quantities weekly or less and stored at 4°C. From this stock feed, dilute feed was prepared in concentrations ranging from 50 to 600 mg/l COD. The program of experimentation was designed to determine the performance of the expanded bed when treating a low strength substrate. Concentrations of less than 600 mg/l were selected since such wastes cannot normally be treated by any known anaerobic process. Both the stock and standard feeds were made with tap water. The variation in feed strengths is given in Table 5. The variation is seen to be small, $\pm$ 4.3% and deemed acceptable.

Table 6 shows the variation in feed rates for the synthetic waste pumped by the Ultraslow Spectroderm pump. These variations were calculated by making 15 measurements with a graduated cylinder over a one hour time period at three different flow rates. These variations are also small and decrease with increasing flow rates. Nonetheless, in order to minimize the effect of this variation on the analytical results, the feed rates were checked daily before sampling and adjusted when necessary.

Since three expanded beds were in preparation, three different substrate concentrations could be used for each hydraulic retention time. This corresponded to three organic loading rates for each HRT since only one pump was used for the three feed reservoirs.

Table 4. Concentrations of all components of stock feed solution

Component	Concentration g/l
Glucose	29.9
Yeast extract	0.1
Ammonium chloride	7.5
Monobasic potassium phosphate	2.5
Dibasic potassium phosphate	1.0
Sodium bicarbonate	55.0

Table 5. Variations in feed strength

For COD = 200 mg/l, made from concentrated feed; COD = 30,000 mg/l

	200 mg/l	30,000 mg/l
1	195.8	29,376
2	193.0	28,956
3	199.6	29,940
4	206.7	31,005
5	217.2	32,580
6	188.1	28,215
7	214.4	32,160
8	214.4	32,160
9	207.7	31,155
10	207.7	31,155
11	207.3	31,095
12	205.9	30,885
13	195.2	29,280
14	199.9	29,985
15	215.6	32,346
16	197.3	29,595
	Mean = 204.11 mg/l Std Dev = ± 8.84 mg/l, ± 4.3% N = 16	Mean = 30,600.93 mg/l Std Dev = 13.41 mg/l, ± 4.3% N = 16

Table 6. Variations in feed pumping rates, 1 hr measurements, Ultraslow Spectroderm MK-III Pump

at 1.0 ml/min	at 5.0 ml/min	at 10.0 ml/min
63	314	596
62	310	598
55	294	598
54	298	602
59	290	606
57	302	602
56	306	594
56	308	598
64	292	602
56	294	604
57	308	604
59	302	602
61	309	598
58	298	596
57	296	590
N = 15 Mean = 58.2 Std Dev = 3.01 ml/hr ± 5.1%	N = 15 Mean = 309.3 Std Dev = 7.39 ml/hr ± 2.4%	N = 15 Mean = 600 Std Dev = ± 5.01 ml/hr ± 0.8%

Table 7 shows the range of corresponding loadings for each hydraulic retention time and substrate concentration. Also given are the feed rates used with the expanded bed. In all six hydraulic retention times were used for each temperature studied ranging from 6 hours to 20 minutes. Using three substrate concentrations of 200, 400, and 600 mg/l COD gave a wide range of organic loadings from 6 to 324 $kg/m^3/day$ (0.050 to 2.70 $lbs/ft^3/day$). Hydraulic and organic loading are based on the volume of the fluidized portion of the column which was kept at 500 ml throughout the study.

The methane produced from the anaerobic treatment of dilute waste would likely not be sufficient to heat the incoming waste significantly. Therefore, it is important that a process for treating dilute waste operate at lower temperatures. Previous studies (64, 75) demonstrated that the AAFEB was able to operate at 20°C successfully. This study looked at three temperatures of operation - 10°, 20°, and 30°C, to observe the effect of this important variable on process efficiency.

Thus, for each temperature, six different hydraulic retention times were used, and at each HRT, three substrate concentrations with three organic loading rates. Twenty hydraulic retention times were allowed after each change before data were taken, and three data points, generally on three successive days, were taken for each HRT. In general, only small differences in removal efficiency were observed over the three measurements made at each condition and this was interpreted as an indication that the system was at steady state.

Table 7. Organic loadings corresponding to various combinations in hydraulic flow rate and waste strength used in the experimental study

Hydraulic flow rate:							
Detention time - hrs		6	4	2	1	0.66	0.33
Liters/day		2	3	6	12	18	36
Pump rate - ml/min		1.38	2.08	4.16	8.32	12.50	25.00
		Organic loading: Kg COD/m^3/day (lbs COD/ft^3/day)					
Waste							
COD - mg/l	200	0.8 (0.050)	1.2 (0.075)	2.4 (0.150)	4.8 (0.300	7.2 (0.450)	14.4 (0.906)
	400	1.6 (0.100)	2.4 (0.150)	4.8 (0.300)	9.6 (0.600)	14.4 (0.900)	28.8 (1.800)
	600	2.4 (0.150)	3.6 (0.225)	7.2 (0.450)	14.4 (0.900)	21.6 (1.350)	43.2 (2.700)

NOTE: HRT and organic loading based on reactor volume of 500 ml

(See steady state data time series plots in Appendix 4).

In addition to the temperature, organic loading, hydraulic loading and substrate concentration study, side studies were conducted at each temperature. One side study was the effect of very low substrate concentrations (50, 100, and 150 mg/l COD) on process efficiency. This is referred to as the dilute studies. The other side study concerns the effects of shock loadings on the system; this was accomplished by batch studies at each temperature. This is referred to as the acclimation studies.

4.3 Sampling and analysis program

The three expanded beds were set as described in the previous section and seeded with anaerobic sewage sludge and rumen fluid. The units were then fed the synthetic substrate continuously and operated at 30°C to develop film on the support media. After ten months of operation to allow biomass accumulation of the anaerobic film and monitoring effluent COD, data were taken for the three temperatures studied.

For each hydraulic retention time, pH, effluent soluble COD, effluent suspended and volatiles suspended solids, volatile acids, gas production, and gas composition were measured daily, usually for three successive days. Film thickness, attached mass and entrapped mass were measured once for each HRT. After each change in hydraulic detention time, twenty HRTs were allowed to pass before the next set of data was taken. Procedures and analytical techniques for each parameter are described in the next section.

Effluent soluble COD, suspended solids, volatile suspended solids, pH, and volatile fatty acids samples were taken from the top of the column before the effluent reached the contact chamber which separated liquid and gas. This minimized any effect of storage on the effluent quality. Gas composition samples were taken through the septa in the inverted cylinders. Film thickness and entrapped and attached mass samples were taken by removing the top of reactors and inserting a serological pipette fitted with a Fisher Pi-Pump to take a representative core from the middle of the expanded portion of media.

All samples except for volatile acids were analyzed immediately after withdrawal from units so that no sample storage technique was utilized. Volatile acid samples were frozen and run when each temperature group was finished. All glassware was washed with hot soapy water, rinsed with tap water, then washed with concentrated dichromate acid cleaning solution. The glassware was finally rinsed three times with distilled water and dried at 103° in a drying oven.

4.4 Analytical technique

4.4.1 pH

pH was determined by the glass electrode method using an Orion Model 701 pH meter. The sensitivity of the meter was 0.01 pH units.

4.4.2 Suspended and volatile suspended solids

The glass fiber filter method was used to measure the suspended

solids in this study. An appropriate sample volume was filtered through a fiber filter supported on a Millipore filtration apparatus (Millipore Corp., Bedford, Mass.). The filter was then washed with an equal amount of distilled water. After filtration, the filter pad and retained solids were dried at 103°C for one hour, cooled in a desiccator, and weighed. To determine the volatile portion of these solids, the dried pad and solids were placed in a muffle furnace at 560°C, ignited for 15 minutes, cooled in a desiccator, and weighed. A blank pad was carried through all steps in order to correct for moisture changes or volatilization of the filter. The glass fiber filters used were 4.25 cm Whatman Glass papers, grade GFC, manufactured by W. and R. Balston, Ltd. of England. Wyckoff (129) presented data which show a coefficient of variation of 11.2% and 13.8% for suspended and volatile suspended solids respectively using the above techniques.

4.4.3 Chemical oxygen demand

Chemical oxygen demand was determined by the dichromate reflux method as described on page 495 of Standard Methods for the Examination of Water and Wastewater, 13th edition, 1971 (115). For soluble COD's, the sample was first filtered as described in the previous section for suspended solids analysis, and the filtrate was used for COD analysis. Nitrogen gas was bubbled through the filtrate for 10 minutes at a rate of about 150 ml per minute to remove dissolved hydrogen sulfide gas. No significant loss of organic carbon resulted from this procedure, as was determined by comparison testing.

4.4.4 Volatile acids

Volatile acid analysis were performed according to the procedure on page 577 of Standard Methods, 1971 (115). The method is a column-partition chromatographic technique for measuring total organic acids. This method yields average efficiencies of about 95 percent for volatile acid concentration in excess of 200 mg/l as acetic acid.

4.4.5 Alkalinity

Alkalinity was determined by the methyl orange alkalinity method as described on page 54 of Standard Methods (115).

4.4.6 Gas analysis

Total gas production for each column was measured continuously by buoyancy displacement described in the previous part of the chapter. Determinations for methane and carbon dioxide content were made using a Gow Mac-550 thermal conductivity gas chromatograph.

The column used in the gas chromatograph was stainless steel, 6 feet long by 1/4 inch in diameter and packed with 60/80 mesh Porapak Q. The column was operated at 40°C, the detector and injector at 100°C. Helium was used as a carrier gas at 60 ml/min. Gas samples were introduced by a 1 ml gas-tight syringe.

4.4.7 Attached and entrapped mass

Measurement of biological mass in the expanded bed was performed in the following manner. Using a 10 ml serological pipet, a 5 ml (settled portion) sample of the bed was removed. This sample was

placed in a 100 mesh sieve and placed over an evaporating dish. The sample was then rinsed with 50 ml of distilled water. The water was sprayed in a stream onto the particles with sufficient velocity to agitate and mix the particles. The filtrate then contained the entrapped mass portion. Next the sieve was turned over and the particles with attached mass were rinsed into another evaporating dish. This portion of the sample was defined as the attached mass.

Both samples were then dried at 103°C for twenty-four hours, allowed to cool in a desiccator, and weighed. After this, the samples were ashed at 600°C for one hour, removed, allowed to cool, and weighed. This loss of material due to ignition is defined as the volatile mass and is measured as concentration of total volatile solids.

The entrapped and attached portions are added together and called total mass. Ten replicates for this test showed a standard deviation of $\pm$ 5.4%, or 1568.58 mg/l total volatile solids for an average mass of 28760 mg/l total volatile solids.

4.4.8 Film thickness

Film thickness was determined statistically by taking the difference between the diameter of a sample of uncoated support particles and coated (covered with a microbial film) particles.

The uncoated, clean, particles were particles taken from an AAFEB reactor after nine months of operations and then removed of attached film. By using particles which had been in an active unit, the effect of attrition was minimized. The measurement of the

apparent diameter of the uncoated particles was done only once using a sample size of 100.

The coated particles, that is with attached microbial films, were taken from each of the AAFEB reactors at each experimental condition. Here again a sample size of 100 was used. The apparent diameters of both the coated and uncoated particles were determined by viewing each one under a light microscope with a calibrated ocular. Attached film was removed from the control particles by ashing the particles at 600°C for one hour. Film thickness was calculated by taking one-half of the difference between the average diameter of the coated and uncoated particles. Photomicrographs of new particles (never in a reactor), uncoated particles (formerly coated) and coated particles are shown in Figure 8.

4.5 Analytical errors

A summary of the variations which could be expected for the analyses used in this study is given in Table 8. The systematic and analytical errors associated with the laboratory measurements were considered to be within the range of values published in Standard Methods (115). Other systematic and analytical errors were calculated as necessary for the study.

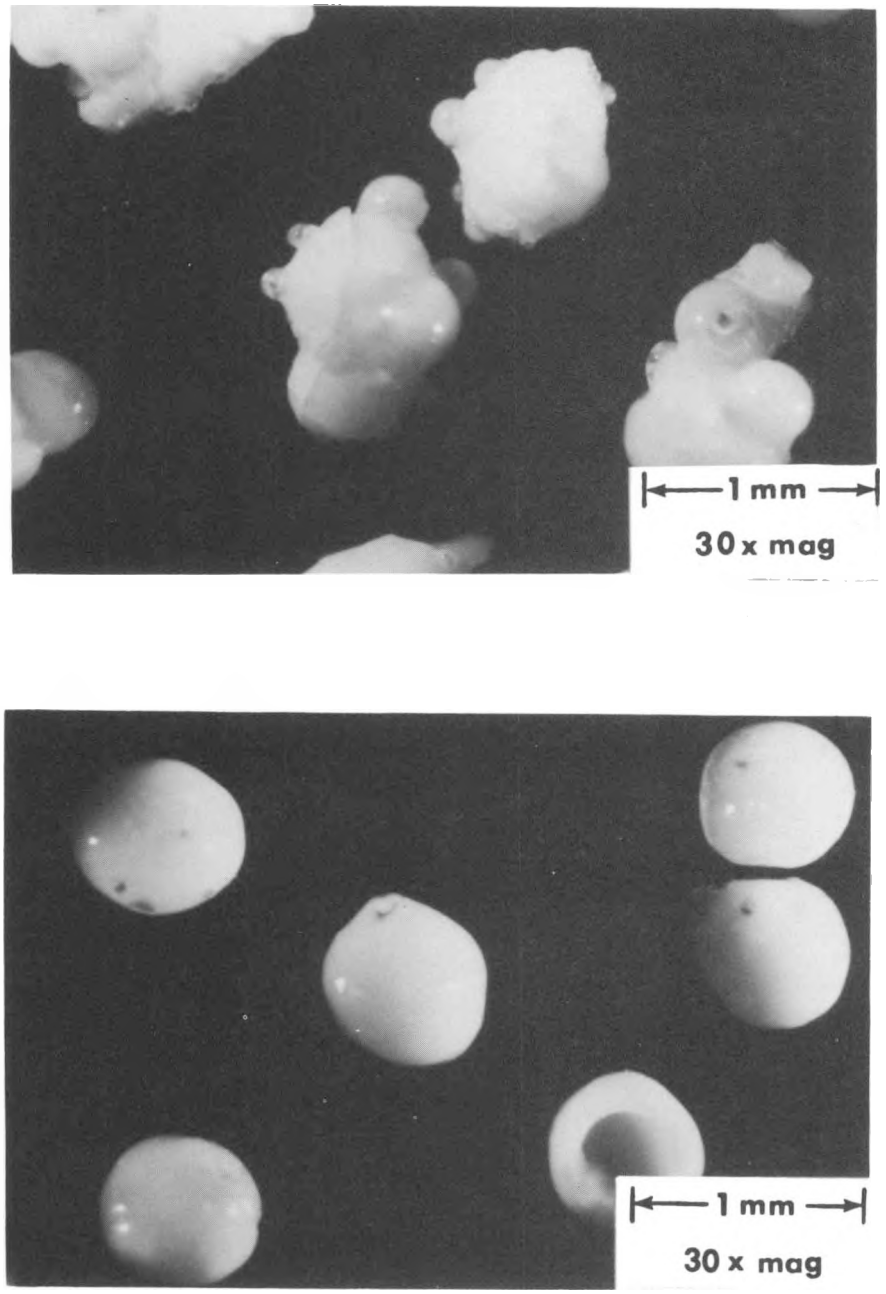

FIGURE 8. PHOTOGRAPHS OF SUPPORT PARTICLES.

TOP: PARTICLES UNCOATED WITH FILM, NEVER USED IN THE AAFEB

BOTTOM: PARTICLES USED IN THE AAFEB WHICH HAVE BEEN REMOVED OF ATTACHED FILM

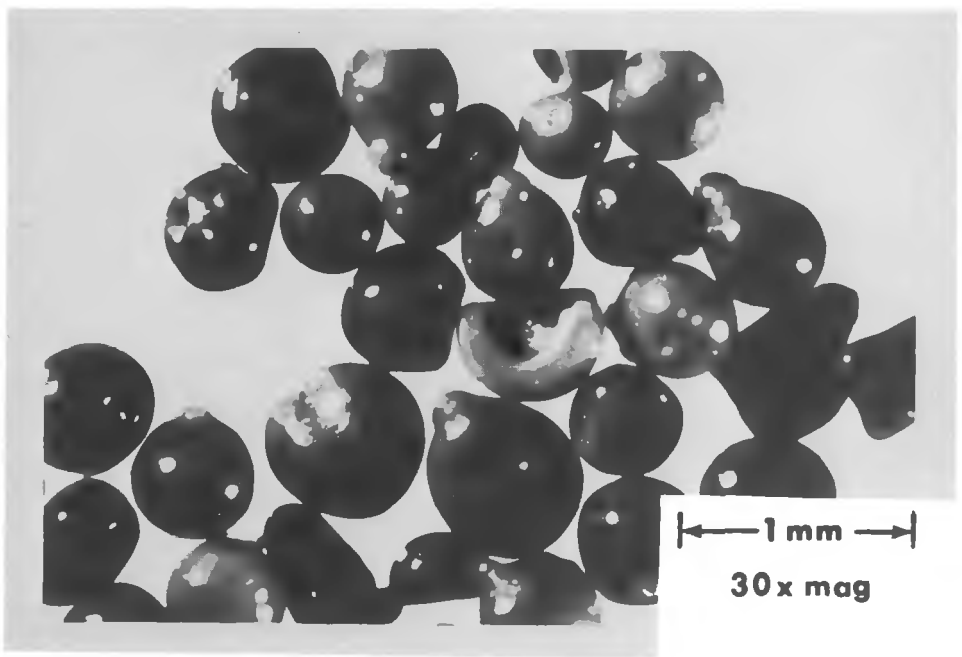

FIGURE 8. COATED PARTICLES FROM THE AAFEB

Table 8. Summary of analytical errors

Measurement	Test	Recovery	Standard Deviation	Reference
pH	Glass electrode	-	± 0.1 pH unit	Standard Methods (115)
SS	Glass pad	-	< 20%	Wyckoff (129)
VSS	Glass pad	-	± 13%	Wyckoff (129)
COD	Dichromate-$AgSO_4$ + H_2SO_4	97.99%	± 8%	Standard Methods (115)
VFA	Column partition chromatography	95% for VFA > 200 mg/l	± 12%	Standard Methods (115)
Alkalinity	Methyl orange			
Total Gas	Buoyancy displacement	100% subject to calibration	-	This study
Total Mass	Ashing method		± 5.4%	This study
Gas Composition	Gas chromatography	Relative to std.	± 1.0%	This study

CHAPTER 5. EXPERIMENTAL RESULTS

The AAFEB reactors were operated for a period of over 15 months. The first nine months were for start up and developing experimental techniques. The last six months were for obtaining the experimental data. The major parameters examined during the experimental data portion of this investigation were the influence of influent substrate concentration, hydraulic retention time, organic loading rate, and temperature on the AAFEB process efficiency. In addition side studies were conducted investigating very dilute wastes and shock loadings (acclimation studies). Results of the main experiments are presented in this chapter. The side studies are presented in Appendix 1 and 2.

5.1 Influence of influent substrate concentration, hydraulic retention time, organic loading rate, and temperature on process efficiency.

5.1.1 Data

The composition summary data for each of the three temperatures evaluated in the laboratory investigation are presented in Tables 9, 10, and 11. In each of these tables, results are presented for effluent soluble COD, suspended solids, volatile suspended solids, and volatile fatty acids concentrations; daily gas production composition (i.e., percent methane); attached film thickness; and attached and entrapped biomass concentrations. Each of the data points is the average of three measurements except for the biomass which is only one measurement and the film thickness measurements which are the result

Table 9. Data summary for all steady state test conditions at 30°C

Hydraulic retention time (hours)	Organic loading rate, Kg COD/m^3/day (lbs COD/1000 ft^3/day)	Calculated influent COD, mg/l	Soluble effluent COD, mg/l	% COD removal	Suspended solids, mg/l	Volatile suspended solids, mg/l	Gas production rate, ml/day	% CH_4	Attached microbial film thickness, microns	Attached mass, mg/l, total volatile solids	Entrapped mass (unattached), mg/l, total volatile solids	Total mass mg/l total volatile solids	Volatile fatty acids, mg/l as acetate
6	0.8 (50)	200	42.4	78.9	10.6	10.6	100	53	7.15	14,000	600	14,600	10.2
6	1.6 (100)	400	72.7	81.8	8.0	8.0	250	71	8.30	15,000	600	15,600	10.2
6	2.4 (150)	600	107.9	82.0	17.3	13.3	400	79	9.86	15,600	400	16,000	102.0
4	1.2 (75)	200	56.0	72.0	16.0	12.0	133	51.6	9.00	14,200	1200	15,400	20.4
4	2.4 (150)	400	69.3	82.6	22.6	16.0	366.6	61.3	7.21	15,800	400	16,200	25.5
4	3.6 (224)	600	88.0	85.3	37.3	24.0	566.6	74.0	8.25	18,400	1200	19,600	71.4
2	2.4 (150)	200	68.1	65.9	16.0	12.6	116.6	58.0	9.23	18,800	1200	20,000	5.1
2	4.8 (300)	400	91.2	77.2	26.0	20.0	466.6	71.6	10.50	21,600	1600	23,200	71.4
2	7.2 (450)	600	152.7	74.5	32.0	23.0	1000	66.0	10.09	23,400	2000	25,460	163.2
1	4.8 (300)	200	78.3	60.8	12	12	166.6	64.3	8.59	17,800	1000	18,800	30.6
1	9.6 (600)	400	118.4	70.4	36	32	933.3	69.6	10.03	22,600	2600	25,200	132.6
1	14.4 (900)	600	205.3	65.7	41.3	30.6	1666.6	63	11.25	26,000	2250	28,250	163.2
0.66	7.2 (450)	200	94.8	52.6	30	20	200						20.4
0.66	14.4 (900)	400	130.3	67.4	40	30	1200						224.4
0.66	21.6 (1350)	600	225.1	62.4	50	40	3000						346.8
0.33	14.4 (900)	200	148.9	25.5	10	10	0						25.5
0.33	28.8 (1800)	400	219.5	45.1	20	20	1500						275.4
0.33	43.2 (2700)	600	315.5	47.4	30	30	3200						489.4

Table 10. Data summary for all steady state test conditions at 20°C

Hydraulic retention time (hours)	Organic loading rate, Kg COD/m^3/day (lbs COD/1000 ft^3/day)	Calculated influent COD, mg/l	Soluble effluent COD, mg/l	% COD removal	Suspended solids, mg/l	Volatile suspended solids, mg/l	Gas production rate, ml/day	% CH_4	Attached microbial film thickness, microns	Attached mass, mg/l, total volatile solids	Entrapped mass (unattached), mg/l, total volatile solids	Total mass mg/l total volatile solids	Volatile fatty acids, mg/l as acetate
6	0.8 (50)	200	34.1	82.9	13.3	6.6	100	60.0	7.90	14,800	400	15,200	21.6
6	1.6 (100)	400	48.6	87.8	17.3	10.6	166.6	60.3	7.09	16,000	800	16,800	21.6
6	2.4 (150)	600	79.5	86.7	22.6	16.0	266.6	76.6	12.69	17,800	800	18,400	43.2
4	1.2 (75)	200	51.8	74.1	10.6	9.3	133.3	56.6	9.80	26,600	1200	27,800	32.4
4	2.4 (150)	400	55.1	86.2	17.3	13.3	333.3	81.3	10.44	21,400	2400	23,800	32.4
4	3.6 (225)	600	101.7	83.0	28.0	24.0	633.3	75.3	10.96	24,400	1600	26,000	43.2
2	2.4 (150)	200	55.4	72.3	10.6	8.0	83.3	73.3	9.11	17,600	400	18,000	32.4
2	4.8 (300)	400	76.4	80.9	16.0	10.6	500.0	83.6	9.00	25,600	800	26,400	32.4
2	7.2 (450)	600	124.0	79.3	28.0	22.6	833.3	76.3	13.27	30,600	1000	31,600	54.0
1	4.8 (300)	200	86.4	56.8	14	12	150	75.5	8.59	17,200	800	18,000	54.0
1	9.6 (600)	400	131.3	67.1	26	22	800	85.0	9.69	27,800	600	28,400	75.6
1	14.4 (900)	600	204.8	65.8	34	26	1400	80.0	12.98	33,600	1000	34,600	108.0
0.66	7.2 (450)	200	108	46.0	20	20	25	87					64.8
0.66	14.4 (900)	400	172	57.0	30	20	700	84					86.4
0.66	21.6 (1350)	600	268	55.3	40	30	1500	81					162.0
0.33	14.4 (900)	200	124	38.0	10	10	0	84					151.2
0.33	28.8 (1800)	400	232	42.0	20	20	800	84					118.8
0.33	43.2 (2700)	600	340	43.0	30	30	2000	80					151.2

Table 11. Data summary for all steady state test conditions at 10°C

Hydraulic retention time (hours)	Organic loading rate, Kg COD/m^3/day (lbs COD/1000 ft^3/day)	Calculated influent COD, mg/l	Soluble effluent COD, mg/l	% COD removal	Suspended solids, mg/l	Volatile suspended solids, mg/l	Gas production rate, ml/day	% CH_4	Attached microbial film thickness, microns	Attached mass, mg/l, total volatile solids	Entrapped mass (unattached), mg/l, total volatile solids	Total mass mg/l total volatile solids	Volatile fatty acids, mg/l as acetate
6	0.8 (50)	200	54.3	72.8	29.3	21.3	50	66.0	8.65	15,400	600	16,000	0
6	1.6 (100)	400	66.9	83.2	48.0	34.6	133.3	79.0	10.39	25,200	800	20,000	20.8
6	2.4 (150)	600	136.6	77.2	50.6	42.6	266.6	81.3	12.80	32,400	1400	33,800	52.2
4	1.2 (75)	200	60.2	69.9	18.6	16.0	100	66.0	10.96	20,000	800	20,800	0
4	2.4 (150)	400	77.2	80.7	29.3	21.3	200	75.3	11.07	28,200	800	20,000	20.8
4	3.6 (225)	600	142.7	76.2	34.6	26.6	400	82.0	14.19	32,000	1400	33,400	41.7
2	2.4 (150)	200	90.5	54.7	21.3	16.0	50	66.0	11.30	23,800	1000	24,800	20.8
2	4.8 (300)	400	139.1	65.2	32.0	24.0	300	81.0	13.27	33,000	1600	34,600	41.7
2	7.2 (450)	600	240.3	59.9	32.0	29.3	533.3	82.3	13.73	37,400	1800	39,200	104.4
1	4.8 (300)	200	99.1	50.4	29.3	21.3	33.3	63.0	12.50	22,800	1600	24,400	31.3
1	9.6 (600)	400	185.1	53.7	34.6	26.6	333.3	73.0	13.85	33,400	2000	35,400	73.0
1	14.4 (900)	600	321.0	46.5	37.3	29.3	733.3	81.0	14.37	38,000	2800	40,800	281.8
0.66	7.2 (450)	200	110.0	45.0	24.0	16.0	25	66					73.0
0.66	14.4 (900)	400	212.2	46.9	32.0	24.0	600	58					73.0
0.66	21.6 (1350)	600	345.8	42.3	48.0	40.0	800	79					177.4
0.33	14.4 (900)	200	129.6	35.2	32.0	24.0	0	66					41.7
0.33	28.8 (1800)	400	243.5	39.1	40.0	32.0	500	58					104.4
0.33	43.2 (2700)	600	385.1	35.8	40.0	32.0	1200	79					198.3

of one hundred measurements from one sample. Also each of the steady state measurements was done on three separate days except for the points at 0.66 and 0.33 hours hydraulic retention time which were done on the same day. Biomass values were not measured at the two highest hydraulic retention times (0.66 and 0.33 hours) and the values measured at one hour HRT were used in later computations for specific utilization and the net specific growth rate. This irregularity in data collection occurred because at the highest HRT's, relatively little time was needed to begin taking data after changing the volumetric loading rate (20 HRT's). In order to conserve the amount of media withdrawn from the expanded beds, the biomass concentrations at one hour were used for 0.66 and 0.33 hours since the later two periods were taken one day after the one hour data.

The hydraulic retention time and organic loading rates are based on a reactor volume of 500 ml which is the volume of the expanded bed portion of the reactor and not the clarification zone. This volume also does not take into account the volume occupied by the support particles. As an example, for a flow of 2000 ml per day (or 2 liters) the HRT is equal to the volume of the reactor divided by the flow rate or 500 ml divided by 2000 ml/day, which is equal to 0.25 days or 6 hours. In effect, the actual detention period was somewhat less than this since the particles and biomass occupied a significant volume.

The particles were not subtracted out to give the void volume, which would be less than 500 ml partly because their size is always

changing due to attrition and biofilm growth. Therefore, the organic and hydraulic loading rates as calculated are conservative.

5.1.2 Organic carbon sources

Organic carbon balances can be evaluated for the three AAFEB reactors over the periods of "steady state" operation. These balances are useful because they give additional support to the experimental results obtained by acting as a cross checking mechanism.

A mass balance on organic carbon may be evaluated in the following manner: Carbon In = Soluble carbon (COD) in effluent + Gaseous Methane production + Methane dissolved in effluent + Effluent volatile suspended solids + Accumulated organic carbon in the AAFEB (biomass). Organic carbon balances for each of the three temperatures evaluated are shown in Tables 12, 13, and 14.

The mass balances were evaluated for only three hydraulic retention times 4, 2, and 1 hours as they were the only conditions where accurate data could be obtained for the accumulated carbon number which is based on biomass measurements. The columns in Tables 12, 13, and 14 are explained as follows. The "Carbon In" column was calculated by multiplying the influent substrate concentration by the flow rate. "Carbon In" is expressed as mg COD/day as are all the other column entries. The "effluent column" is the product of the effluent soluble COD times the flow rate. The "CH_4 production" was calculated by multiplying the daily gas production by the percentage of CH_4 as was determined by gas chromatography, and then converting this quantity to its COD equivalent.

Table 12. Organic carbon balance at 30°C

Hydraulic retention time (hrs)	Carbon in mg COD/day	Effluent mg COD/day	CH_4 prod[1] mg COD/day	CH_4 in eff[2] mg COD/day	Eff. VSS[3] mg COD/day	Accumulated[3] mg COD/day	Total out	Rate in/out
4	600	168.0	174.3	249.9	51.1	64.9	708.2	0.0847
4	1200	207.9	570.8	249.9	68.1	48.9	1145.3	01.047
4	1800	264.0	1064.9	249.9	102.2	292.0	1973.0	0.912
2	1200	408.6	171.7	499.8	107.3	435.4	1622.8	0.739
2	2400	547.2	848.5	499.8	170.4	662.6	2728.5	0.879
2	3600	916.2	1676.4	499.8	195.9	549.0	3837.3	0.938
1	2400	939.6	272.0	999.7	204.5	0	2415.8	0.993
1	4800	1420.8	1649.9	999.7	545.2	162.2	4779.8	1.004
1	7200	2463.6	2666.8	999.7	521.4	231.2	6882.7	1.046
								Mean = 0.933

[1]Assume 1 gram COD = 393 ml CH_4 at 30°C

[2]Assume effluent is saturated with methane at 32.8 ml CH_4/liter

[3]Assume 1.0 mg solids = 1.42 mg COD

Table 13. Organic carbon balance at 20°C

Hydraulic retention time (hrs)	Carbon in mg COD/day	Effluent mg COD/day	CH_4 prod[1] mg COD/day	CH_4 in eff[2] mg COD/day	Eff. VSS[3] mg COD/day	Accumulated[3] mg COD/day	Total out	Ratio in/out
4	600	155.4	200.6	261.7	39.6	295.4	952.7	0.629
4	1200	165.3	720.7	261.7	56.6	795.2	1999.5	0.600
4	1800	305.1	1268.4	261.7	102.2	863.3	2800.7	0.642
2	1200	332.4	162.4	523.4	68.1	16.2	1102.5	1.088
2	2400	458.4	111.8	523.4	96.3	210.9	2400.8	0.999
2	3600	744.0	1691.2	523.4	189.1	454.4	3602.1	0.999
1	2400	1036.8	301.2	1046.3	204.5	0	2588.8	0.927
1	4800	1575.6	1808.8	1046.3	374.9	284.0	4919.2	0.975
1	7200	2457.6	2979.2	1046.3	443.0	426.0	7352.1	0.979
								Mean = 0.870

[1]Assume 1 gram COD = 376 ml CH_4 at 20°C

[2]Assume effluent is saturated with methane at 32.8 ml CH_4/liter

[3]Assume 1.0 mg solids = 1.42 mg COD

Table 14. Organic carbon balance at 10°C

Hydraulic retention time (hrs)	Carbon in mg COD/day	Effluent mg COD/day	CH_4 prod[1] mg COD/day	CH_4 in eff[2] mg COD/day	Eff. VSS[3] mg COD/day	Accumulated mg/COD/day	Total out	Ratio in/out
4	600	180.6	182.1	271.5	68.1	389.4	1091.7	0.549
4	1200	231.6	415.6	271.5	90.7	243.4	1252.8	0.957
4	1800	428.0	905.2	271.5	111.7	0	1716.5	1.048
2	1200	543.0	91.0	543.1	134.4	568.0	1879.5	0.638
2	2400	834.6	670.6	543.1	204.5	795.2	3048.0	0.787
2	3600	1441.8	1211.3	543.1	249.6	823.6	4269.4	0.843
1	2400	1189.2	57.9	1086.3	362.9	0	2696.3	0.890
1	4800	2221.2	671.5	1086.3	453.2	90.8	4523.0	1.061
1	7200	3852.0	1639.3	1086.3	499.2	181.7	7258.5	0.991

[1]Assume 1 gram COD = 362 ml CH_4 at 10°C

[2]Assume effluent is saturated with methane at 32.8 ml CH_4/liter

[3]Assume 1.0 mg solids = 1.42 mg COD

Methane can be converted to a COD equivalent (and vice versa) by the relationship that at 0°C and one atmosphere, one gram of COD is equivalent to 350 ml methane (93). This relationship can be further modified for temperature effects by the following relationship between temperature and gas volume (93).

$$V_{STP} = V_{Ta} \cdot \frac{T_{STP}}{Ta} \qquad \text{Eq (18)}$$

where V_{STP} = volume at standard temperature and pressure (0°C,

V_{Ta} = volume at temperature "a"

T_{STP} = temperature at STP in °R

Ta = temperature "a" in °R

Using these relationships, it was calculated that one gram of soluble COD is equal to 362 ml at 10°C, 376 ml methane at 20°C, and 393 ml methane at 30°C assuming a standard pressure of one atmosphere.

The "CH_4 in effluent" column expresses the rate of loss of dissolved methane in the effluent stream. The rate of loss of dissolved methane was assumed to be the product of the solubility of methane at 25°C and one atmosphere pressure (0.0328 liters CH_4/ liter of solution (19)) and the waste flow in liters per day. This assumes that the effluent is saturated with methane in solution. This assumption has previously been used with an anaerobic filter study by Young (130). The volume of dissolved gas is converted to its COD equivalent.

The "Effluent VSS" accounts for organic carbon lost as solids in the effluent. This is equal to the concentration of effluent volatile

suspended solids times the flow rate. The effluent solids quantity is then converted to a COD equivalent by assuming that 1.0 mg VSS is equal to 1.42 mg COD (93). The "accumulated" entry accounts for solids accumulating in the reactor. This was determined by finding the differences between successive biomass measurements and then dividing the difference by the number of days in between measurements. This solids measurement is also converted to COD by multiplying by 1.42.

The sum of the five previously mentioned columns is equal to the "Total out" column. This is compared to the "Total in" and the resulting ratio is reported in the last column in Tables 12, 13, and 14. The mean ratios of the in to out quantities were found to be 0.933, 0.870, and 0.862 for 30, 20 and 10°C, respectively. These high values indicate a good account of organic carbon coming in and out of the expanded beds and verify the reliability of the experimental results obtained in this study. A sample organic carbon balance calculation is shown in detail in Appendix 3.

In a similar type of calculation, Tables 15, 16, and 17 show comparisons of COD removed to methane production at 30, 20 and 10°C, respectively. In these tables, the total methane production (including methane captured as gas and methane dissolved in the effluent) is compared to the methane equivalent of the COD removed by the AAFEB process as determined by the difference between influent and effluent soluble COD multiplied by the daily flow rate and then converted to its CH_4 equivalent. The same conversion factors were used (CH_4 to COD equivalent) as in the previous organic carbon balances.

Table 15. Comparison of COD removed to methane production at 30°C

Hydraulic retention time (hrs)	Influent waste mg COD/day	Avg COD removed mg COD/day	Total CH_4 prod ml CH_4/day	CH_4 equivalent of COD removed	Conversion of COD to CH_4
6	400	315.2	118.6	124.0	95.6
6	800	654.6	243.1	257.7	94.3
6	1200	984.2	381.6	387.4	98.5
4	600	432.0	167.0	170.0	98.2
4	1200	992.1	323.1	390.5	82.7
4	1800	1536.0	517.6	604.7	85.5
2	1200	791.4	264.4	311.5	84.8
2	2400	1852.8	530.8	729.4	72.7
2	3600	2683.8	856.8	1056.6	81.0
1	2400	1460.4	500.7	574.9	87.0
1	4800	3379.2	1043.1	1330.3	78.4
1	7200	4736.4	1443.5	1864.7	77.4
		Totals:	6390.3	7802.4	81.9

Table 16. Comparison of COD removed to methane production at 20°C

Hydraulic retention time (hrs)	Influent waste mg COD/day	Avg COD removed mg COD/day	Total CH_4 prod ml CH_4/day	CH_4 equivalent of COD removed	Conversion COD to CH_4
6	400	331.8	125.6	124.7	100.7
6	800	702.8	166.0	264.2	62.8
6	1200	1041.0	269.8	391.3	68.9
4	600	444.6	173.8	167.1	104.0
4	1200	1034.7	369.3	388.9	94.9
4	1800	1494.9	575.2	561.9	102.3
2	1200	867.6	257.8	326.1	79.0
2	2400	1941.6	614.8	729.9	84.2
2	3600	2856.0	832.6	1073.6	77.5
1	2400	1363.2	506.8	512.4	98.9
1	4800	3224.4	1073.6	1212.1	88.5
1	7200	4742.4	1513.6	1782.8	84.9
0.66	3600	1656.0	612.1	622.5	98.3
0.66	7200	4104.0	1178.4	1542.8	76.3
0.66	10800	5976.0	1805.8	2246.6	80.3
0.33	7200	2736.0	1180.8	1028.5	114.8
0.33	14400	6048.0	1852.8	2273.6	81.4
0.33	21600	9360.0	2780.8	3518.7	79.0
		Totals:	15389.2	18767.7	84.6

Table 17. Comparison of COD removed to methane production at 10°C

Hydraulic retention time (hrs)	Influent waste mg COD/day	Avg COD removed mg COD/day	Total CH_4 prod ml CH_4/day	CH_4 equivalent of COD removed	Conversion of COD to CH_4
6	400	291.4	98.6	105.5	93.4
6	800	666.2	170.9	241.3	70.8
6	1200	966.8	282.3	350.2	80.6
4	600	419.4	164.4	151.9	108.2
4	1200	968.4	249.0	350.8	70.9
4	1800	1371.9	426.4	497.0	85.7
2	1200	657.0	229.8	238.0	96.5
2	2400	1565.4	439.8	567.1	77.5
2	3600	2158.2	635.7	781.9	81.3
1	2400	1210.8	414.5	438.6	94.5
1	4800	2578.8	636.9	934.3	68.1
1	7200	3348.0	987.5	1213.0	81.4
0.66	3600	1620.0	606.9	586.9	103.4
0.66	7200	3380.4	938.4	1224.7	76.6
0.66	10800	4575.6	1222.4	1658.1	73.7
0.33	7200	2634.4	1180.8	918.2	128.5
0.33	14400	5630.4	1470.8	2040.0	72.0
0.33	21600	7736.4	2128.8	2803.0	75.9
			Totals: 12283.4	15100.5	81.3

The results indicate that greater than 80 percent of the soluble COD was recovered as methane. These results suggest that the net biological solids synthesis for the AAFEB using the glucose waste was low. This is further verified with observed yield measurements shown later in this chapter.

5.1.3 Organic removal efficiency vs. hydraulic and organic loading rates

The organic removal efficiency of the AAFEB (as percentage of initial COD removal) compared to the hydraulic retention time and organic volumetric loading rate are shown in Figures 9 to 12. The influence of temperature for each influent substrate concentration is shown in Figures 9 and 11 while Figures 10 and 12 show the influence of influent substrate concentration for each temperature evaluated. Clearly, the efficiency of substrate removal is a function of the hydraulic retention time and concomitant organic loading rate, but is also influenced by the influent substrate concentration and temperature. However, it is important to note that 80 percent COD removal or better was achieved at low HRT's and a high organic loading rates regardless of the influent substrate concentration or temperature. Also the influence of temperature on process is not as pronounced as one would think - that is, the reaction rate does not double for each 10°C incremental rise in temperature. The statistical significance between the curves showing temperature and influent substrate concentration effects is confirmed later in the chapter with an analysis of variance of temperature and So effects.

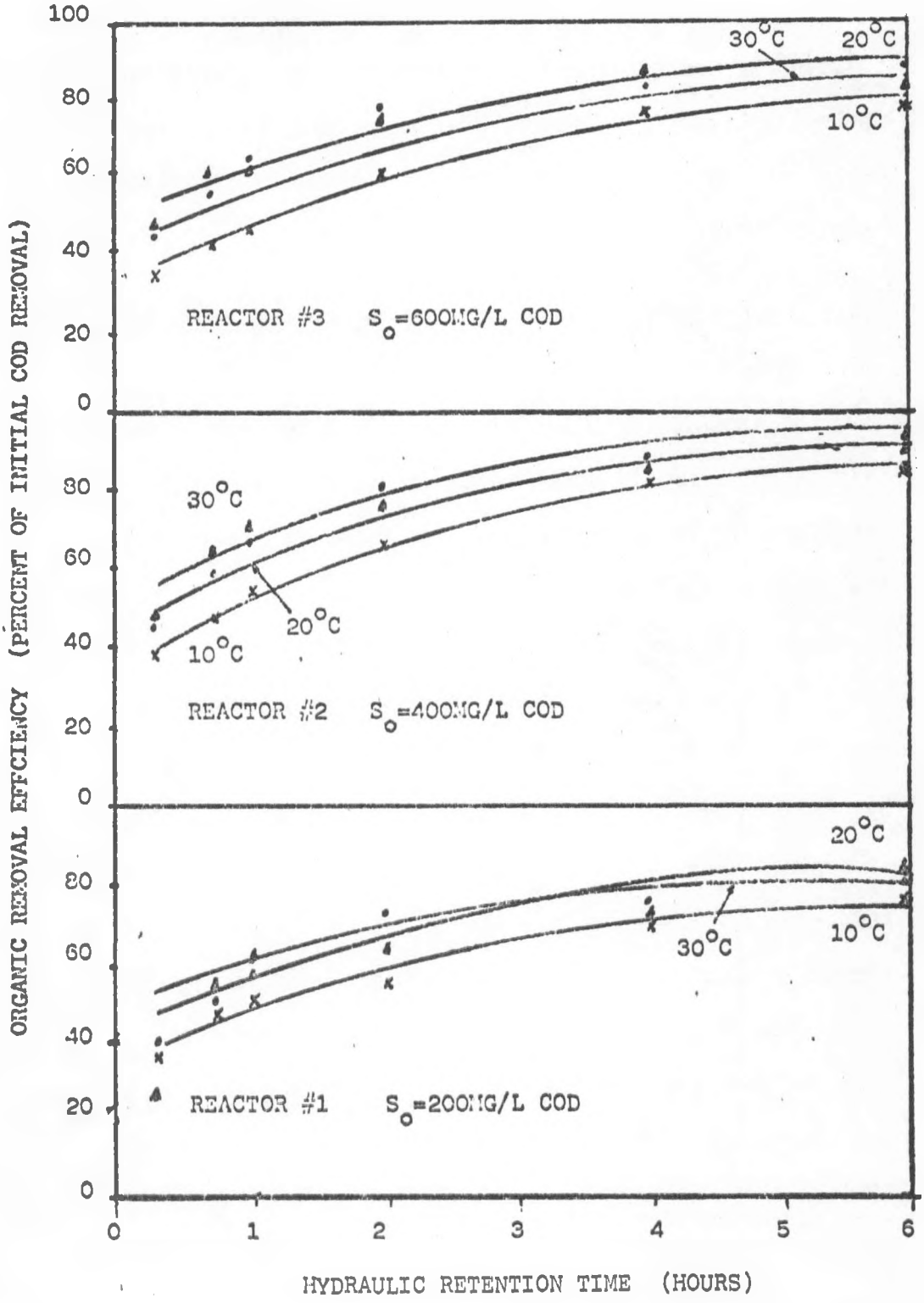

FIGURE 9. ORGANIC REMOVAL EFFICIENCY VS. HYDRAULIC RETENTION TIME-INFLUENCE OF TEMPERATURE

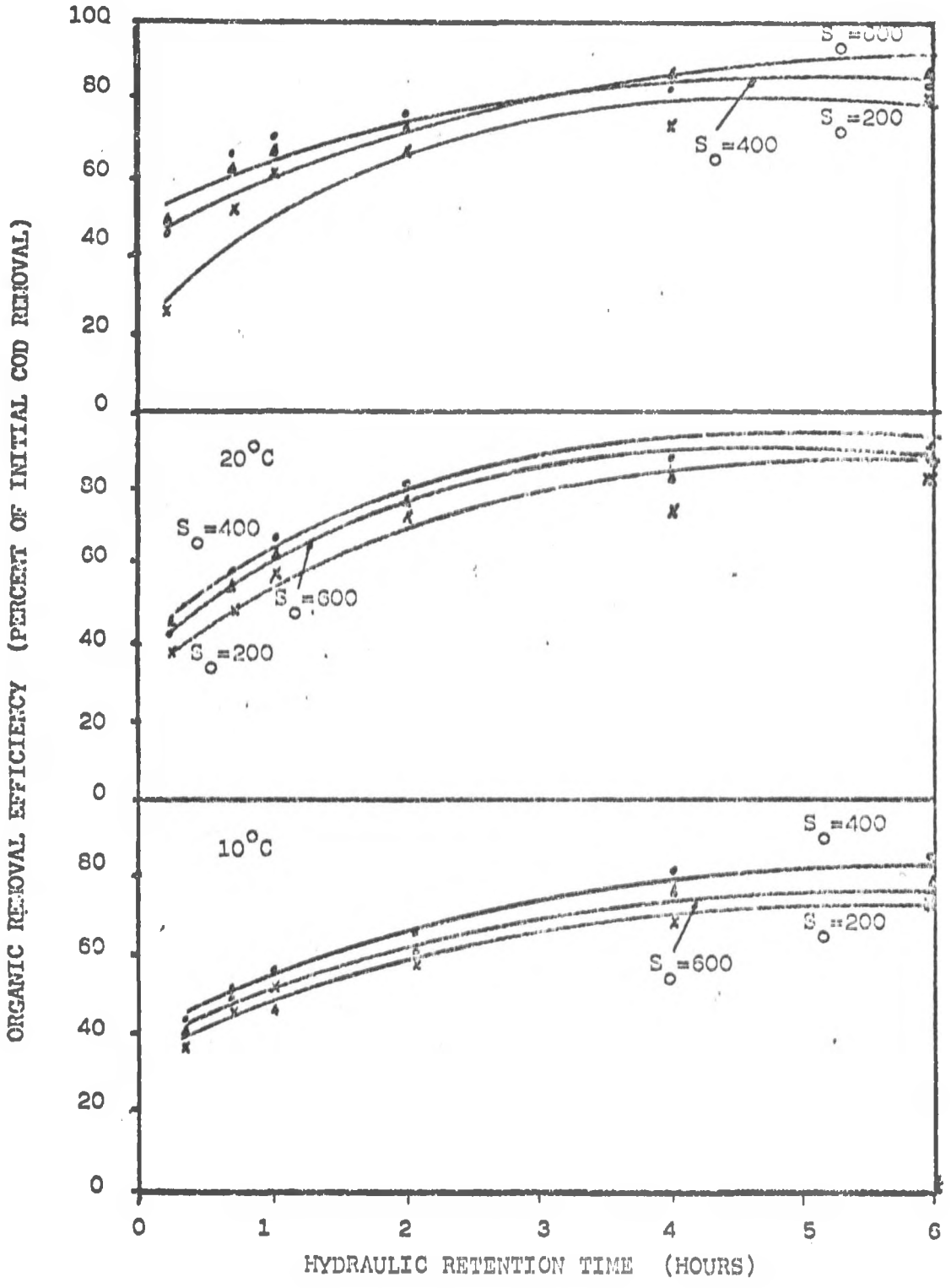

FIGURE 10. ORGANIC REMOVAL EFFICIENCY VS. HYDRAULIC RETENTION TIME-INFLUENCE OF INFLUENT SUBSTRATE CONCENTRATION

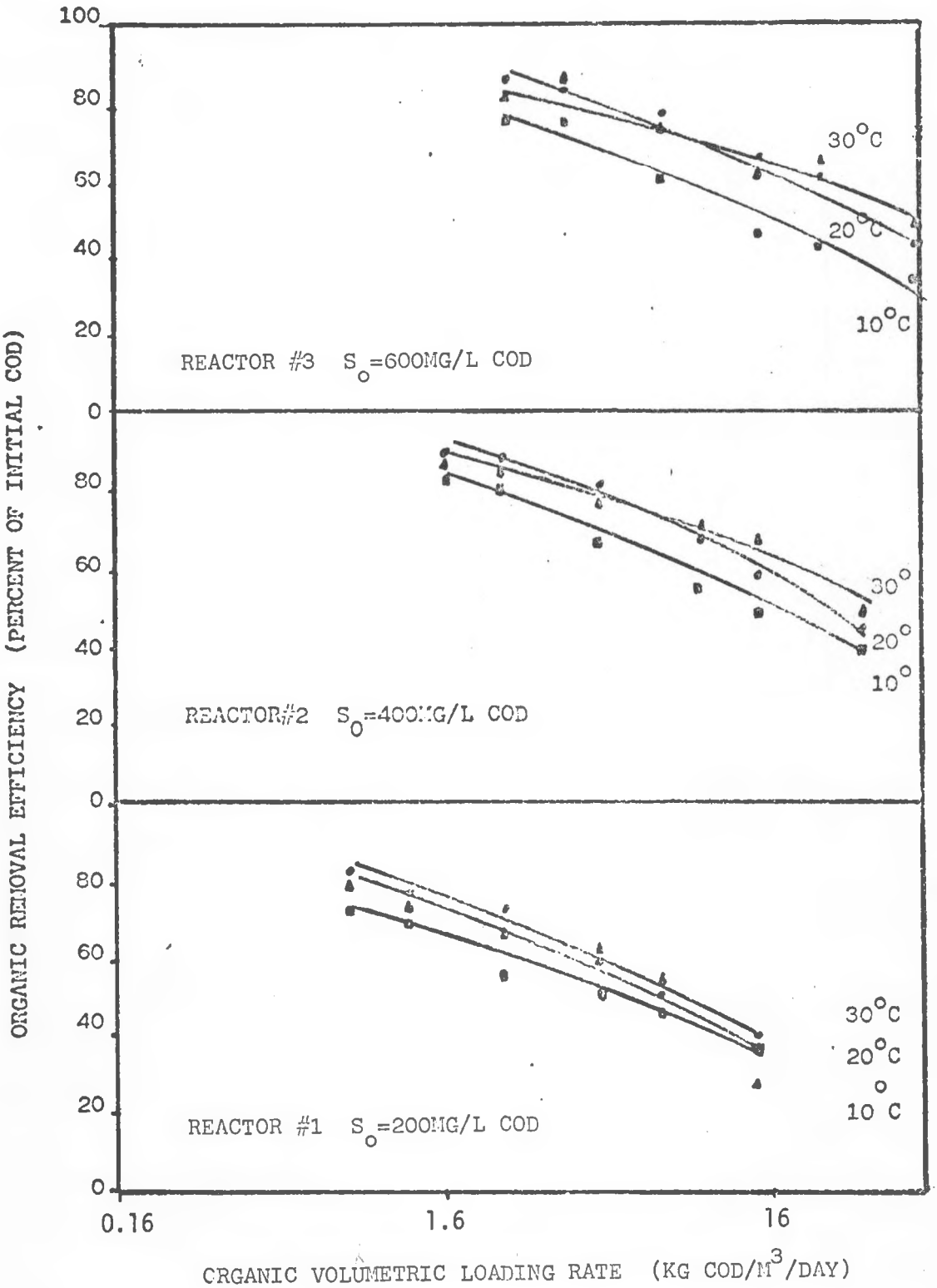

FIGURE 11. ORGANIC REMOVAL EFFICIENCY VS. ORGANIC LOADING RATE-INFLUENCE OF TEMPERATURE

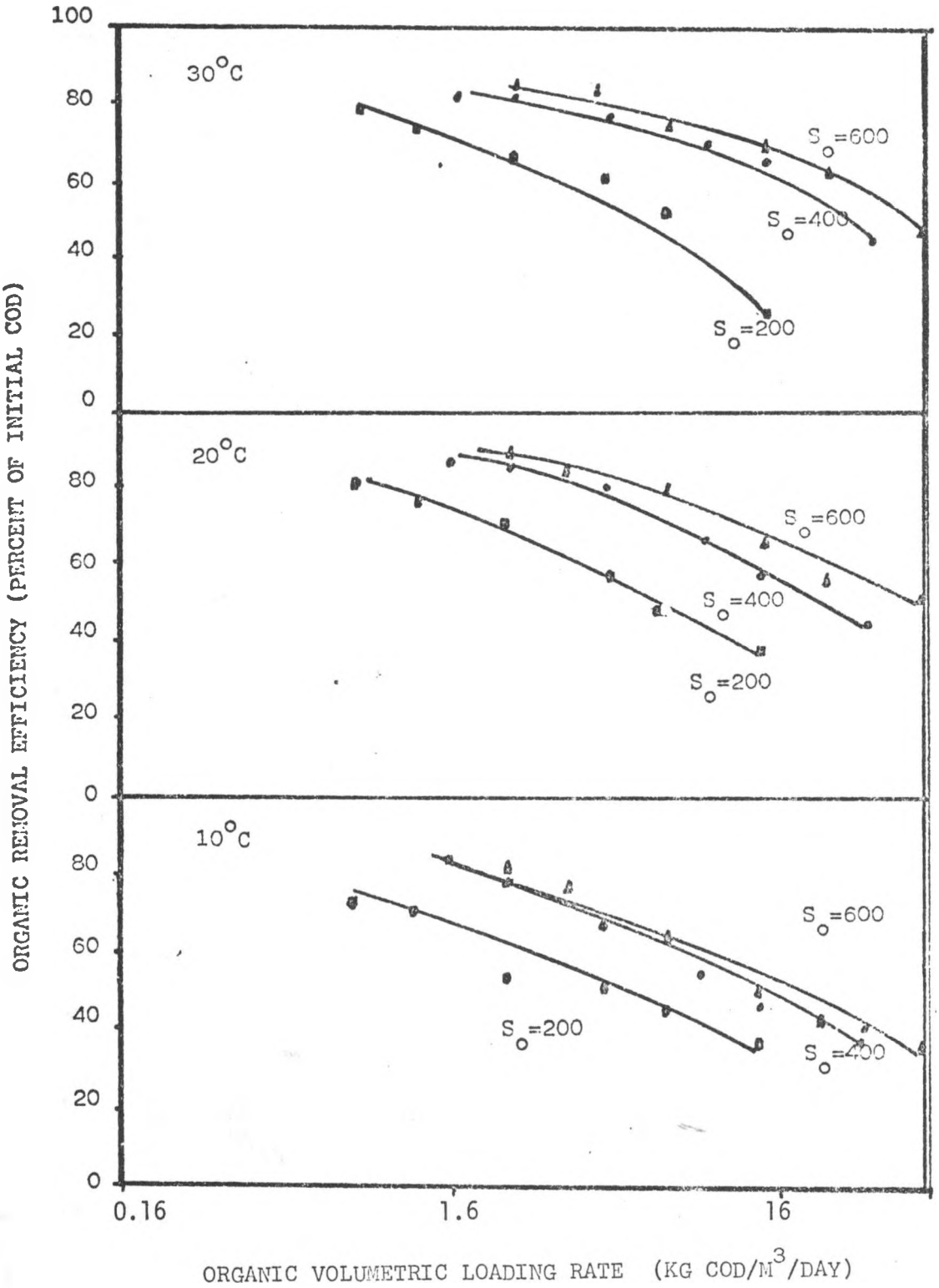

FIGURE 12. ORGANIC REMOVAL EFFICIENCY VS. ORGANIC LOADING RATE-INFLUENCE OF INFLUENT SUBSTRATE CONCENTRATION

5.1.4 Film thickness and biomass concentration

Figures 13 and 14 are plots of film thickness of the microbial film coated particles formed at varying organic loading rates, temperatures, and influent substrate concentrations. Thicker films were observed at lower temperature and higher organic loading rates. At 20°C thicker films were observed at So = 200 mg/l than at So = 400 mg/l. Also at 30°C, some observations showed thicker films in Reactor #1 than at 20°C. But, in general, thicker films were observed at lower temperatures and at higher loading rates. The two exceptions cited above are probably due to experimental error.

It should also be pointed out that the films observed were relatively thin, never exceeding 15 microns in thickness. This compares to 140-1700 microns thickness commonly found in trickling filters (15) and 40 microns, the diameter of a typical floc of suspended solids found in an activated sludge aeration basin (10).

Figures 15 and 16 are plots of the total organic mass concentration in the expanded beds (measured in unexpanded volume) measured as total volatile solids as influenced by the organic loading rate, temperature, and influent substrate concentration. As explained previously in Chapter 4, the total microbial mass is composed of attached mass (that portion attached to the support media) and entrapped mass (that portion entrapped between the support particles). Again, higher mass concentrations were observed at lower temperatures and higher loading rates. High mass concentrations were observed exceeding 40,000 mg/l. These exceptionally high concentrations of microbial mass appear to be responsible for efficient organic

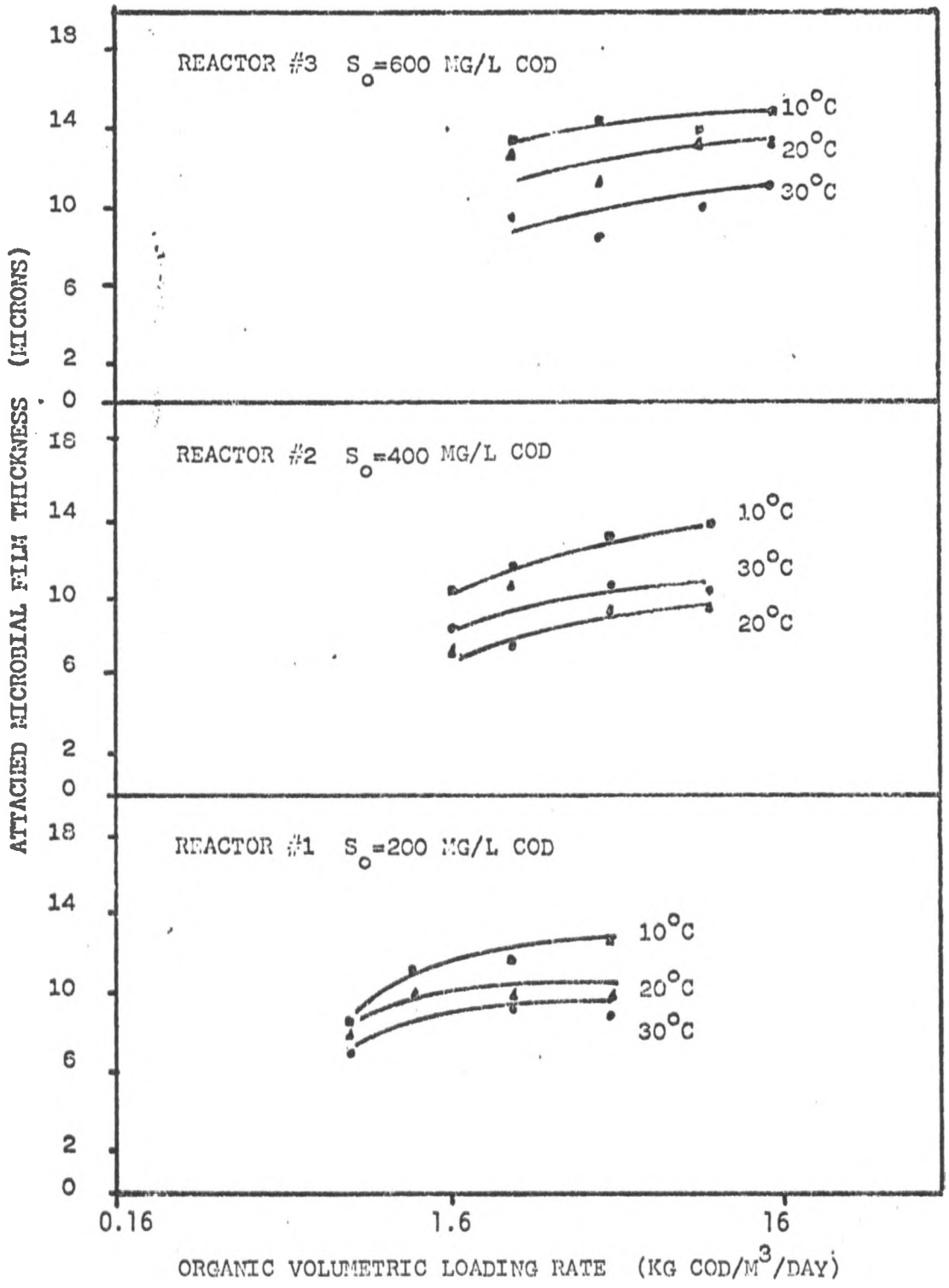

FIGURE 13. ATTACHED FILM THICKNESS VS. ORGANIC LOADING RATE-INFLUENCE OF TEMPERATURE

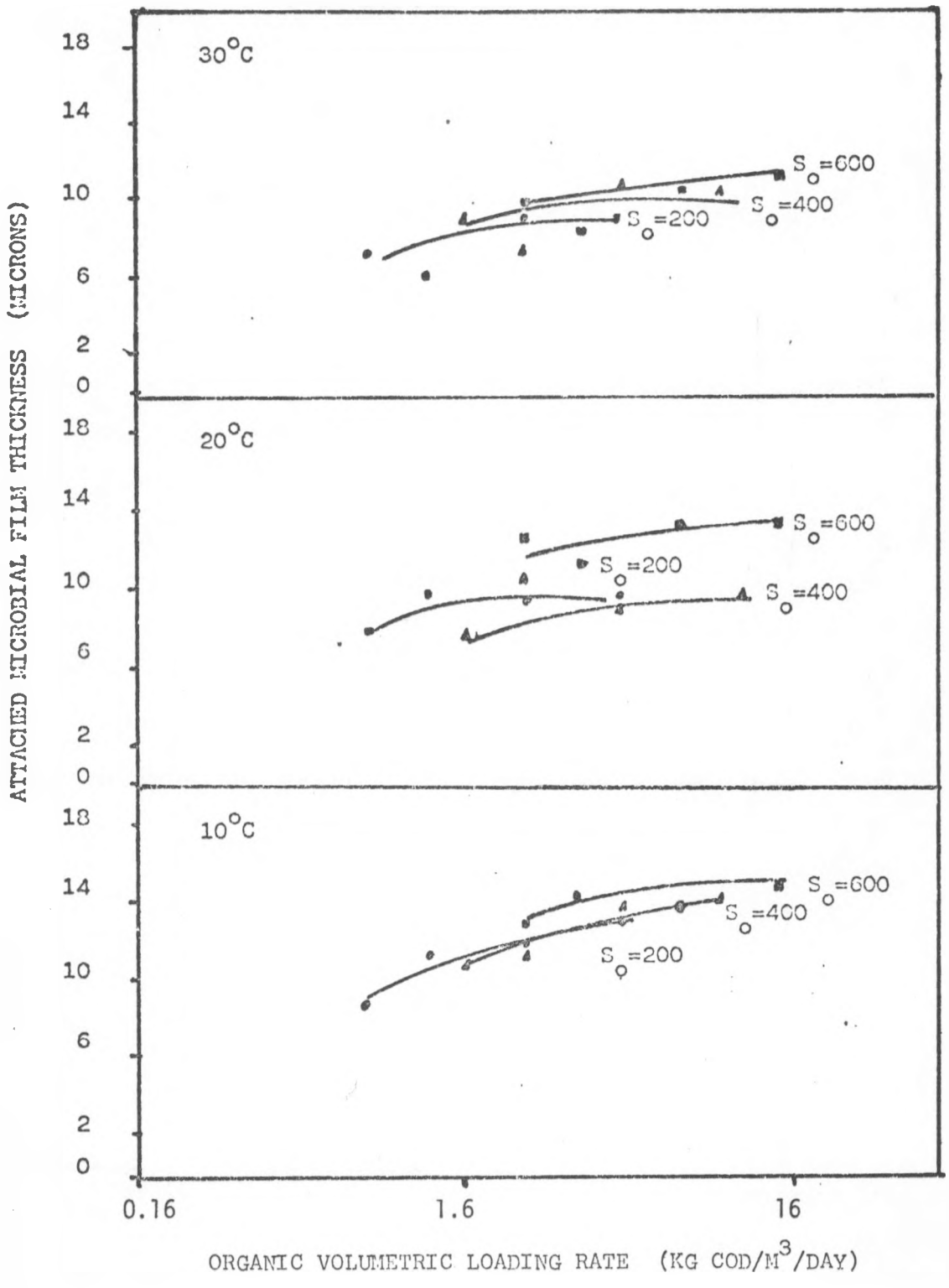

FIGURE 14. ATTACHED FILM THICKNESS VS. ORGANIC LOADING RATE-INFLUENCE OF INFLUENT SUBSTRATE CONCENTRATION

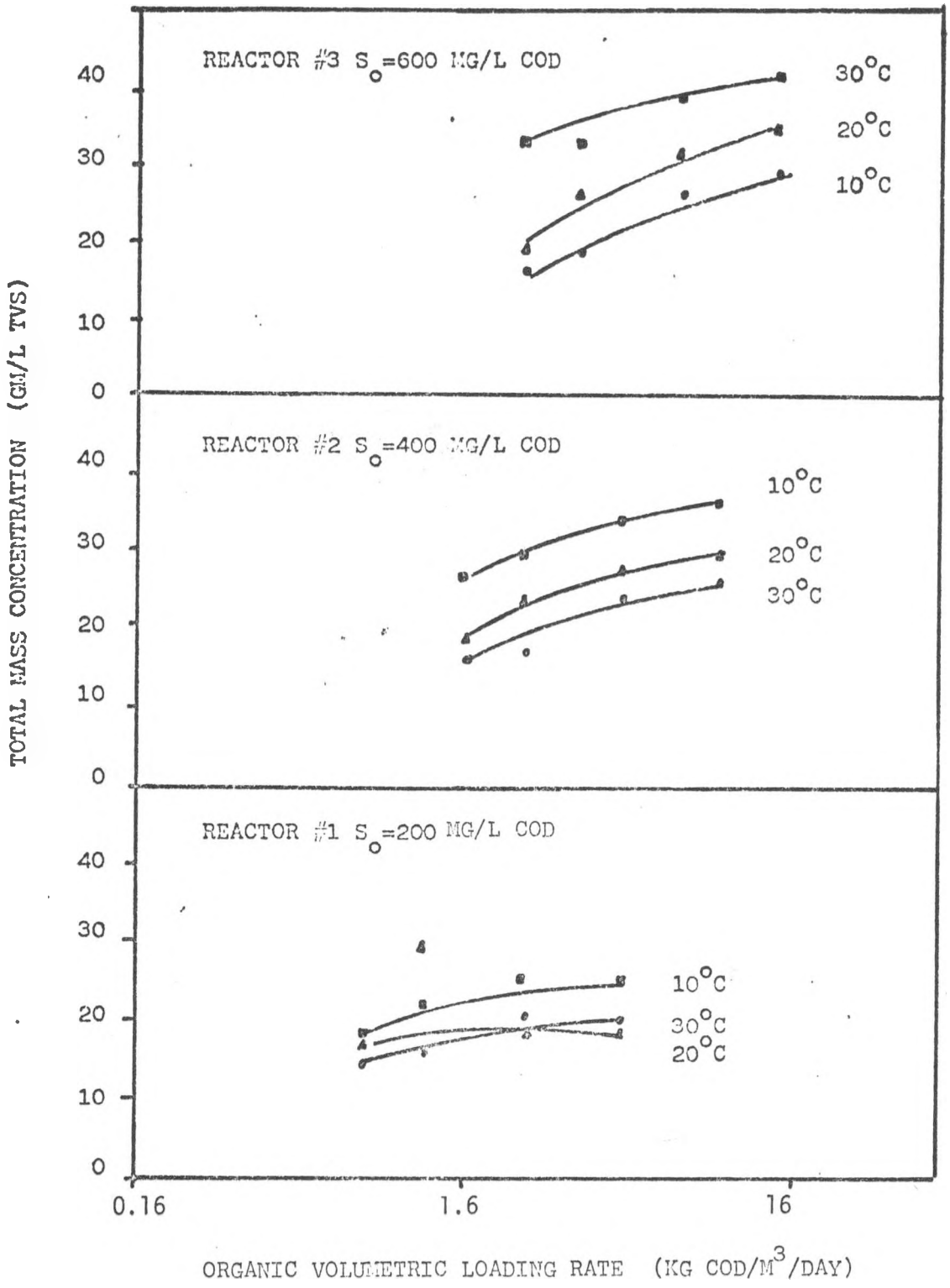

FIGURE 15. TOTAL MASS CONCENTRATION VS. ORGANIC LOADING RATE-INFLUENCE OF TEMPERATURE

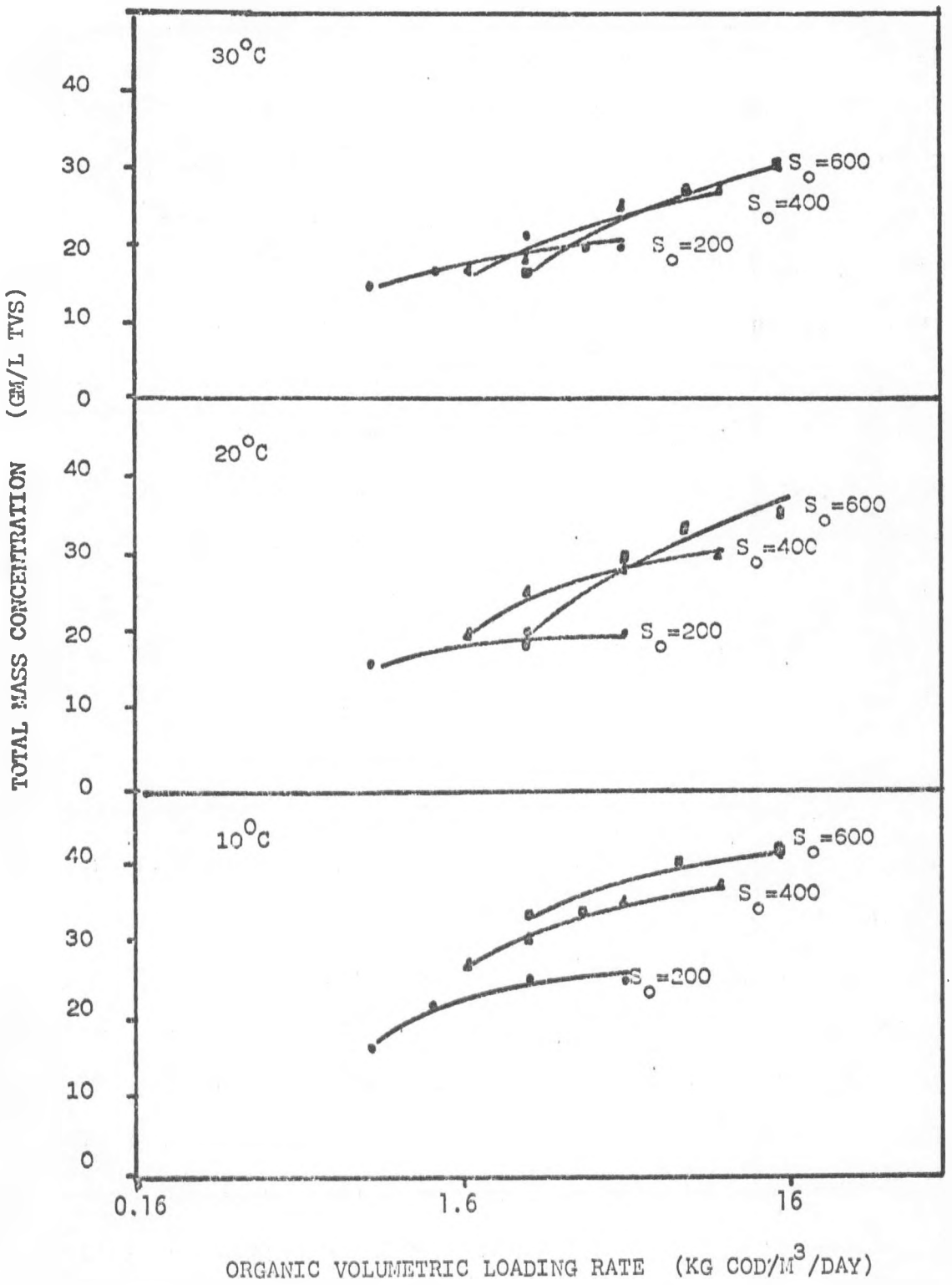

FIGURE 16. TOTAL MASS CONCENTRATION VS. ORGANIC LOADING RATE-INFLUENCE OF INFLUENT SUBSTRATE CONCENTRATION

carbon removal even at low temperatures.

The amount of entrapped microbial mass relative to the total mass concentration is shown in Table 18. All biomass measurements are listed in Table 18; the first column is the total mass concentration, the second attached mass, and the third entrapped mass. All are expressed as mg/l total volatile solids. The last column expresses the ratio of entrapped mass to total mass as a percentage. It is seen that the percentages are low. An average value for each temperature was computed and is shown in Table 18. Mean values of 5.9%, 4.1% and 4.5% entrapped to total mass were calculated for 30, 20, and 10°C, respectively. This indicated that the large majority of biomass was retained as attached film rather than entrapped flocs in the AAFEB under the conditions maintained during the laboratory investigation.

As was mentioned previously, the entrapped portion of the biomass measurement was determined by vigorously washing the mass sample. The fact that such a small percentage of the mass was entrapped indicates that the attached film was very tightly attached to the support particles. Also, since such large masses were observed with concomitant thin films, the density of the attached films must have been large. The significance of thin films with a large biomass concentration will be further discussed in the next chapter.

One other point to be noted is that the sequence of testing in the laboratory investigation was to complete all 30°C testing first, then the 20°C data, and the 10°C datawere taken last. It could be argued that the increasing biomass with lower temperatures is not

Table 18. Ratio of entrapped to total mass - relative amount of entrapped mass making up the total mass concentration

Total Mass (mg/l TVS)	Attached Mass (mg/l TVS)	Entrapped Mass (mg/l TVS)	% Entrapped/Total
30°C			
14,600	14,000	600	4.2
15,600	15,000	600	3.8
16,000	15,600	400	2.5
15,400	14,200	1200	7.8
16,200	15,800	400	2.5
19,600	18,400	1200	6.1
20,000	18,800	1200	6.0
23,200	21,600	1600	6.9
25,400	23,400	2000	7.9
18,800	17,800	1000	5.3
25,200	22,600	2600	10.3
28,250	26,000	2250	7.9
			5.9% - Mean 30°C

Table 18. (Continued)

20°C			
15,200	14,800	400	2.6
16,800	16,000	800	4.7
18,400	17,800	600	3.3
27,800	26,600	1200	4.3
23,800	21,400	2400	10.0
26,000	24,400	1600	6.2
18,000	17,600	400	2.2
26,400	25,600	800	3.0
31,600	30,600	1000	3.2
18,000	17,200	800	4.4
28,400	27,800	600	2.1
34,600	33,600	1000	2.9
			4.1% - Mean 20°C

Table 18. (Continued)

10°C			
16,000	15,400	600	3.7
26,000	25,200	800	3.0
33,800	32,400	1400	4.1
20,800	20.000	800	3.8
29,000	28,200	800	2.8
33,400	32,000	1400	4.2
24,800	23,800	1000	4.0
34,600	33,000	1600	4.6
39,200	37,400	1800	4.6
24,400	22,800	1600	6.6
35,400	33,400	2000	5.6
40,800	38,000	2800	6.9
			4.5% Mean 10°C

related to a compensation factor, but is only due to accumulation with time. However, it should be noted that after each change in temperature, biomass was lost from each reactor and then gradually developed to a new equilibrium thickness and concentration for each condition investigated. For example, at the highest hydraulic loading rates at 30°C, total mass concentrations of 18800, 25200 , and 28250 mg/l TVS were measured for reactors 1, 2, and 3 respectively (see Table 9). Then after changing the temperature to 20°C, at the lowest hydraulic loading rates after the acclimation period of 20 HRT's, mass measurements were 15200, 16800, and 18400 mg/l TVS indicating a net loss of biomass after the temperature change. Thus it is felt that the time factor of accumulation is not as important a factor as the temperature compensation factor. Additional studies need to be conducted to clarify this phenomenon.

5.1.5 Effluent suspended solids

An important variable for effluent quality is the suspended solids concentration. Values for this parameter are listed in Tables 9, 10, and 11 for the three temperatures evaluated. The effluent suspended solids concentration as a function of the hydraulic retention time is shown in Figure 17.

The major point to be recognized from Figure 17 is that the effluent suspended solids concentration increased only slightly for large increases in the hydraulic loading rate. Most of the measurements for suspended solids were less than 30 mg/l even at HRT's down to 0.33 hours. This is one of the major advantages of the AAFEB system over

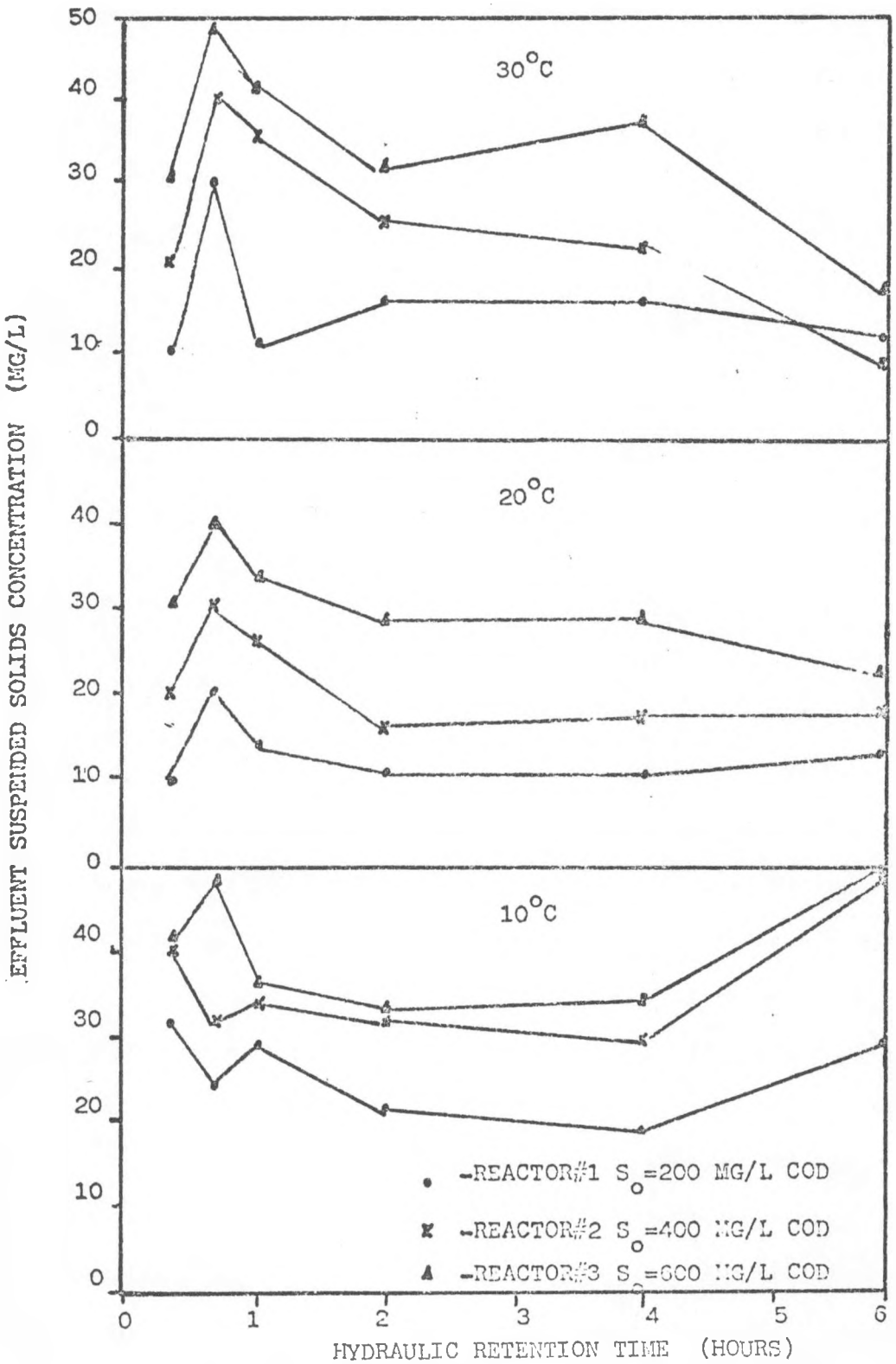

FIGURE 17. EFFLUENT SUSPENDED SOLIDS AS AFFECTED BY FLOW RATE (HYDRAULIC RETENTION TIME)

a suspended microbial system in that even at very high loading conditions, washout is not a problem. This will be further discussed in the following chapters.

5.1.6 Refractory organics

In examining Figures 9 and 10, it is noted that the removal efficiency of the AAFEB reactors was never much greater than 85 percent (of the initial influent COD concentration) even at the lowest loading rates. All of the curves level off at approximately 80 percent removal. Since the influent substrate was composed of glucose and mineral salts, which can be assumed to be close to 100 percent biodegradable, it can be assumed that a large fraction of the soluble COD measured in the effluent is not biodegradable or is, in other words, refractory organic material of microbial origin.

Table 19 provides further evidence for this assumption. Volatile fatty acids, especially acetic acid, are the immediate precursors for most of the methane generated in an anaerobic digestor. When the volatile acids start to accumulate in the anaerobic reactor, it is a signal that the reactor is being overloaded (83) since the rate limiting step in anaerobic fermentation according to Lawrence and McCarty (74) is the conversion of acetate to methane. In Table 19, the relative amount of the soluble effluent COD which is volatile fatty acids is shown. The data are scattered, but it can be seen that even at the lowest hydraulic loading rates, the volatile fatty acids only make up a fraction of the effluent soluble COD. Since it was assumed that the influent is close to 100 percent biodegradable, the rest of the effluent COD is refractory organics,

Table 19. Evaluation of effluent soluble organic matter - percent of volatile fatty acids (as COD) to effluent soluble COD

Hydraulic retention time (hrs)	Organic loading rate (kg COD/m^3/day)	10°C	20°C	30°C
6	0.8	0	67.7	25.7
6	1.6	33.2	47.5	15.0
6	2.4	40.8	58.1	101.1
4	1.2	0	66.9	38.9
4	2.4	28.8	62.9	39.3
4	3.6	31.2	45.4	86.8
2	2.4	24.5	62.5	8.0
2	4.8	32.0	45.3	83.7
2	7.2	46.4	46.5	114.3
1	4.8	33.7	66.8	41.8
1	9.6	42.1	61.6	119.8
1	14.4	93.9	56.4	85.0
0.66	7.2	71.0	64.2	23.0
0.66	14.4	36.8	53.7	184.2
0.66	21.6	54.8	64.6	164.8
0.33	14.4	34.4	130.4	18.3
0.33	28.8	45.8	54.7	134.2
0.33	43.2	55.0	47.5	165.9

most likely of microbial origin.

The significance of this fact is that the reactor is a generator of organic matter as well as a destroyer of organic matter. It also implies that the reactor efficiency is higher based on percent of initial removal of biodegradable COD. The refractory phenomenon is further discussed in the next chapter.

5.1.7 Summary of data

The data indicate that the AAFEB process is able to treat low strength organic wastes at low temperatures, short dentention times, and at high organic loading rates. An organic carbon balance revealed that the net biosynthesis of solids in the reactor is low and verified the reliability of the experimental results. The reactors were able to achieve very high biomass concentrations with concomitant thin films which increased with lower temperatures. Also effluent suspended solids concentrations did not greatly increase with high flow rates (i.e., washout did not occur).

5.2 Linear approximation of rational design relationship equations

This section of the results chapter will present simplified design equations. These equations relate process efficiency to two fundamental microbial parameters. These equations will be used later in the discussion section to demonstrate temperature and influent substrate concentration effects on process efficiency.

5.2.1 Computation of specific film utilization rate and net specific film growth rate

In describing the growth rate of the film and the uptake of

substrate by the film equations 12 and 16 may be modified to:

$$A = \frac{ks}{Ks + s} \qquad \text{Eq (19)}$$

where A = specific film utilization which is the rate of substrate removal per unit weight of film, $time^{-1}$

and

$$B = \frac{Bmax\ s}{Ks + s} \qquad \text{Eq (20)}$$

where B = the net specific film growth rate, $time^{-1}$

and B max is the minimum net specific film growth rate, $time^{-1}$.

The values of A and B for each of the three temperatures evaluated during the laboratory investigation are presented in Table 20. The net specific film growth rates were obtained by dividing the mass leaving the system (mg volatile suspended solids per day) by the mass in the system (mg total volatile solids). Since the feed to the units was totally soluble, that is with no particulates, it can be assumed that no biomass is being introduced with the feed. Also implicit in the calculation of the kinetic parameters is the assumption that the system is at steady state conditions and the biomass concentration remains constant during the length of the evaluation testing. Because of the large biomass concentrations which exist in the AAFEB and because of the low growth rate of the methane bacteria, it can be assumed that the biomass concentration remains constant for any given day. For example, assuming a biomass concentration of 30,000 mg/l TVS, at HRT = 2 hours, and So = 400 mg/l, and assuming that the yield is equal to 0.1, the biomass concentration would increase by only 500 mg/l (less than 2%) during the length of one day assuming 80 percent of the influent substrate

Table 20. Calculated organic removal rates and microbial growth rates at varying hydraulic retention times, influent conditions and temperature

HRT hrs	50 mg/l	10°C		20°C		30°C	
		B $days^{-1}$	A $days^{-1}$	B $days^{-1}$	A $days^{-1}$	B $days^{-1}$	A $days^{-1}$
6	200	.0066	.0434	.0021	.0544	.0036	.0540
	400	.0066	.0638	.0031	.1044	.0025	.1046
	600	.0063	.0684	.0043	.1411	.0041	.1536
4	200	.0057	.0499	.0025	.0398	.0058	.0700
	400	.0055	.0832	.0041	.1084	.0074	.1528
	600	.0059	.1024	.0069	.1435	.0091	.1956
2	200	.0096	.0660	.0066	.1204	.0094	.0986
	400	.0104	.1130	.0060	.1838	.0129	.1994
	600	.0108	.1375	.0107	.2258	.0135	.2637
1	200	.0261	.1238	.0200	.1891	.0191	.1939
	400	.0225	.1819	.0232	.2832	.0380	.3350
	600	.0215	.2049	.0225	.3422	.0324	.4185
0.66	200	.0295	.1658	.0500	.2299	.0478	.2517
	400	.0304	.2383	.0316	.3612	.0535	.4812
	600	.0441	.2803	.0390	.4315	.0714	.5968
0.33	200	.0885	.2594	.0500	.3799	.0478	.2440
	400	.0813	.3979	.0633	.5323	.0714	.6441
	600	.0705	.4740	.0780	.6710	.0955	.9060

was removed.

As mentioned previously, the net specific growth rate is equal to the inverse of the solids retention time which is a widely used design and operational parameter for biological waste treatment systems. By taking the inverse of the B values in Table 20, it is observed that high SRT values were maintained in the AAFEB exceeding 400 days at some instances. More important, even at the highest loadings, the SRT values were greater than 10 days and thus while the system efficiency was lowered, washout was avoided as organisms were still growing at a rate greater than they were being removed. This is a very important factor in maintaining a stable and reliable system.

The specific utilization rate was obtained by dividing the amount of substrate removed per day by the amount of mass in the system. Values shown in Table 20 are used for evaluation of system performance and kinetic analysis in the following sections.

5.2.2 Comparison of observed yield to measured yield

Equation 14 showed the relationship between net specific growth rate of microorganisms and specific utilization. Substituting the film growth and uptake parameters (B and A) gives an equation similar to 14.

$$B = Y - b \qquad \text{Eq (21)}$$

Equation 21 is in linear form and can be used to calulate Y the yield coefficient and b the decay coefficient. These constants are valuable for kinetic analysis and give valuable information concerning the amount of solids which accumulate in the expanded bed

reactor.

The constants Y and b can be easily determined by least squares regression analysis for the experimental data plotted in the above form. Growth constants were determined for the glucose synthetic substrate at 30, 20, and 10°C. These analyses are shown in Figure 18.

Observed growth yield coefficients, which are equal to the slope of the least squares regression line, were determined to be 0.1156, 0.1251, and 0.1945 mg TVS/mg COD for the 30, 20 and 10°C data, respectively. The values for "b" the decay coefficient, are obtained from the intersection of the least squares regression with the ordinate. For all three temperatures, however, the value of the ordinate intercept is negative albeit very small in absolute value. This is most likely due to the insensitivity of the analysis, which could easily account for a small difference between an expected small positive number and the actual very small negative one. Therefore, the value of b is assumed to be negligible in all further considerations.

Using the comparison of COD removed to CH_4 production tables (Tables 15, 16, and 17), a measured yield can be formulated by the difference between the COD removed and the CH_4 produced. This difference is assumed to be converted to biomass and hence is a measured yield. The values for the observed yields correspond well with the measured yields as is shown in Table 21, with the exception of the 30°C measured yield which appears to be too high. The otherwise good correlation between the two yield measurements

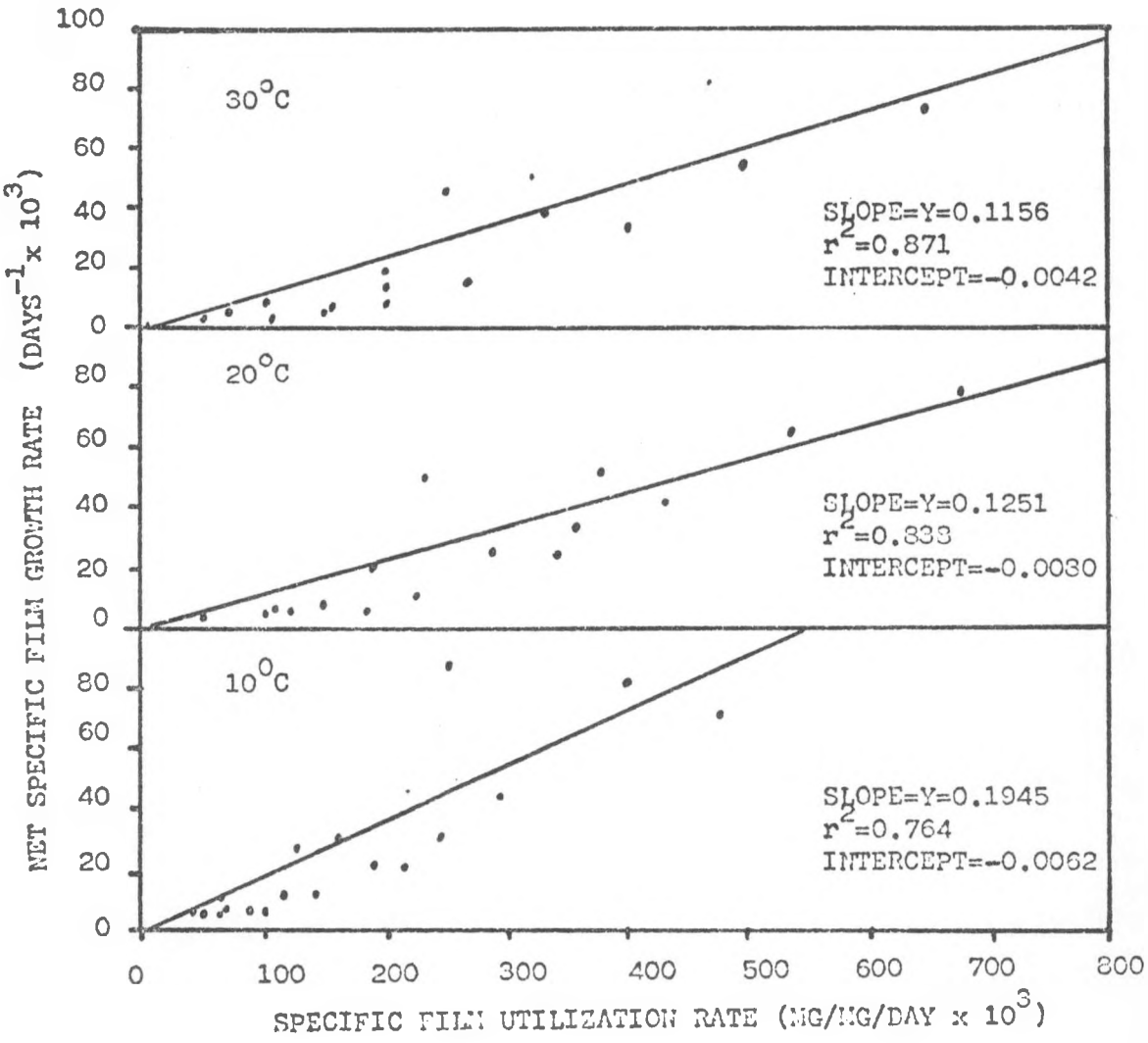

FIGURE 18. LEAST SQUARES REGRESSION ANALYSIS FOR THE DETERMINATION OF THE KINETIC GROWTH CONSTANTS Y (THE OBSERVED YIELD) AND b (THE DECAY COEFFICIENT)

Table 21. Comparison of observed yields and measured yields

Temperature	Observed yield	Measured yield
30°C	0.12	0.18
20°C	0.13	0.15
10°C	0.19	0.19

provides additional reliability to the calculated A and B values, which were used for the calculation of Y and b, and further demonstrates the low net synthesis of solids in the AAFEB, which has practical engineering significance. This will be further discussed in Chapter 7.

5.2.3 Specific utilization vs. hydraulic and organic loading rates

Figures 19-22 examine the specific utilization rate of the AAFEB (as mg of COD removed per mg of mass per day) against the hydraulic (Figures 19 and 20) and organic loading rates (Figures 21 and 22). The influence of temperature for each influent substrate concentration is shown in Figures 19 and 21 while Figures 20 and 22 show the influence of influent substrate concentration for each temperature evaluated.

The specific utilization by the films increases with increasing volumetric and hydraulic loading rates and is also higher with higher temperatures. Again, it should be noted that the influence of temperature on process efficiency is not as pronounced as might be expected. Also, the specific utilization rate has a greater rate of increase as the organic volumetric loading rate increases. This is in accord with the prediction of Atkinson and Davies (6) that uptake rates will be highest at high loading rates with a completely mixed microbial film fermenter which prevents microbial washout. Maximum organic utilization rates varied from 40 gm COD/l/day at 30°C (2.5 lbs/ft^3/day) down to about 10 gm COD/l/day (0.6 lbs/ft^3/day) at 10°C.

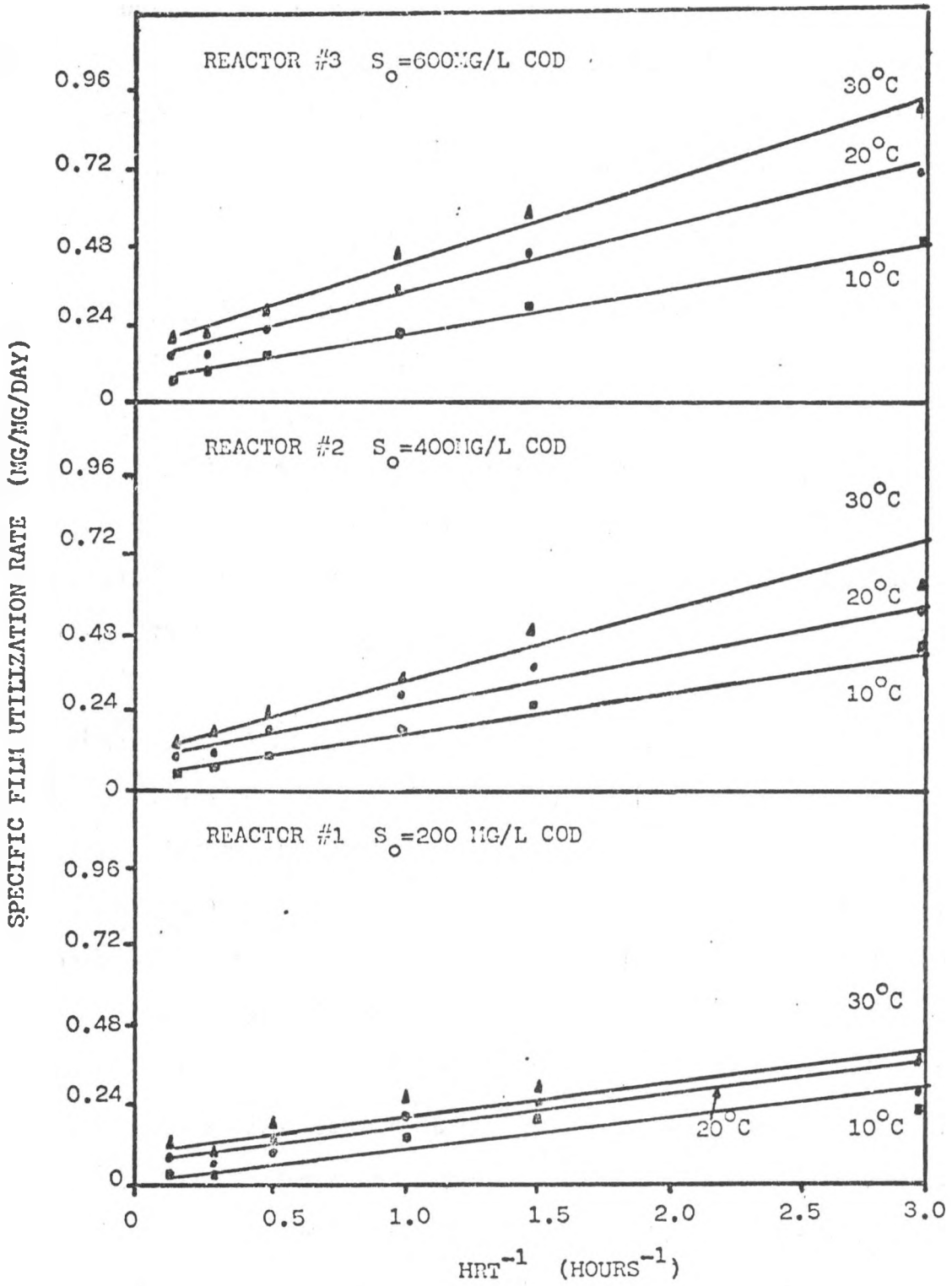

FIGURE 19. SPECIFIC FILM UTILIZATION RATE VS. HRT^{-1}- INFLUENCE OF TEMPERATURE

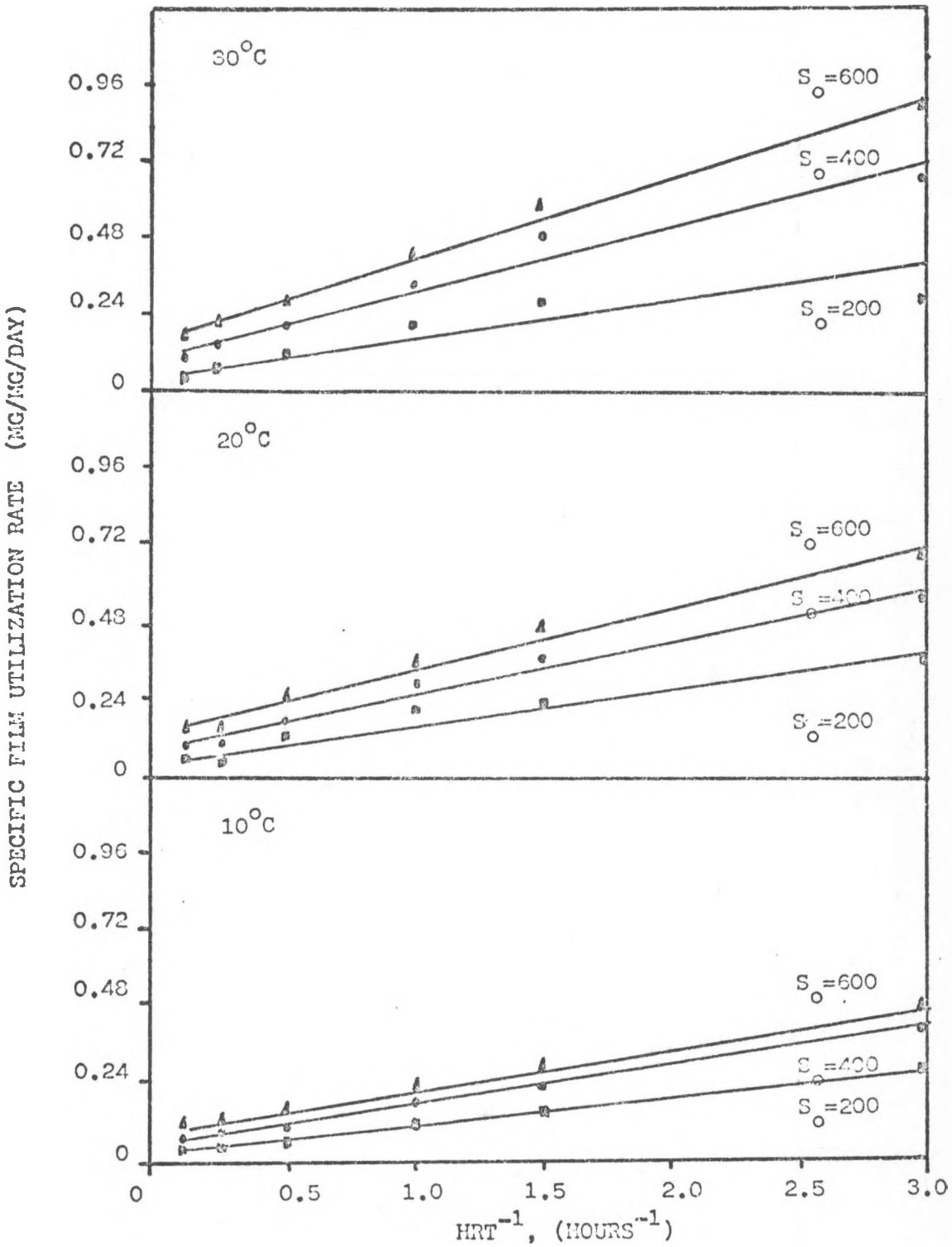

FIGURE 20. SPECIFIC FILM UTILIZATION RATE VS. HRT^{-1}_ INFLUENCE OF INFLUENT SUBSTRATE CONCENTRATION

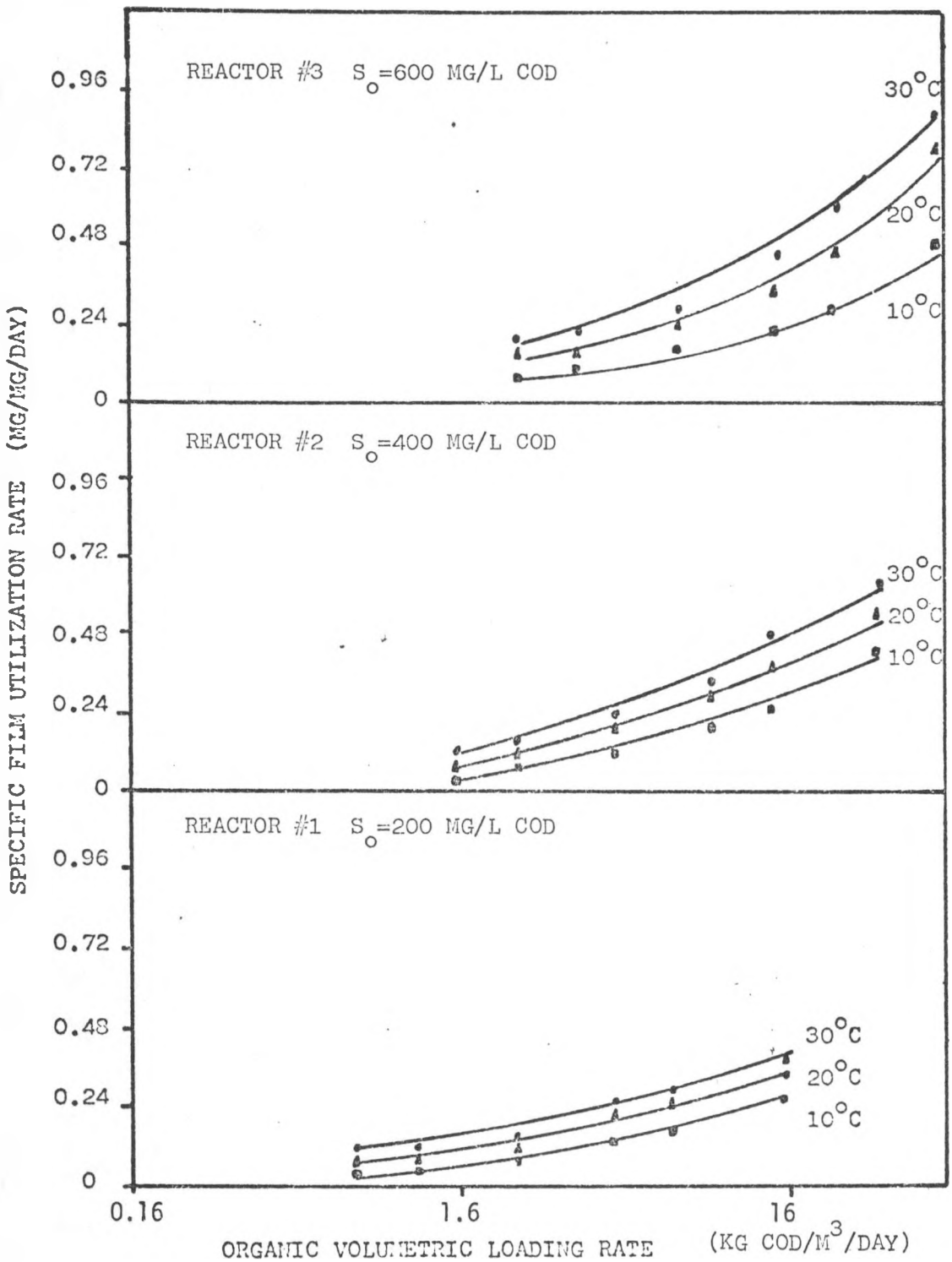

FIGURE 21. SPECIFIC FILM UTILIZATION RATE VS. ORGANIC LOADING RATE-INFLUENCE OF TEMPERATURE

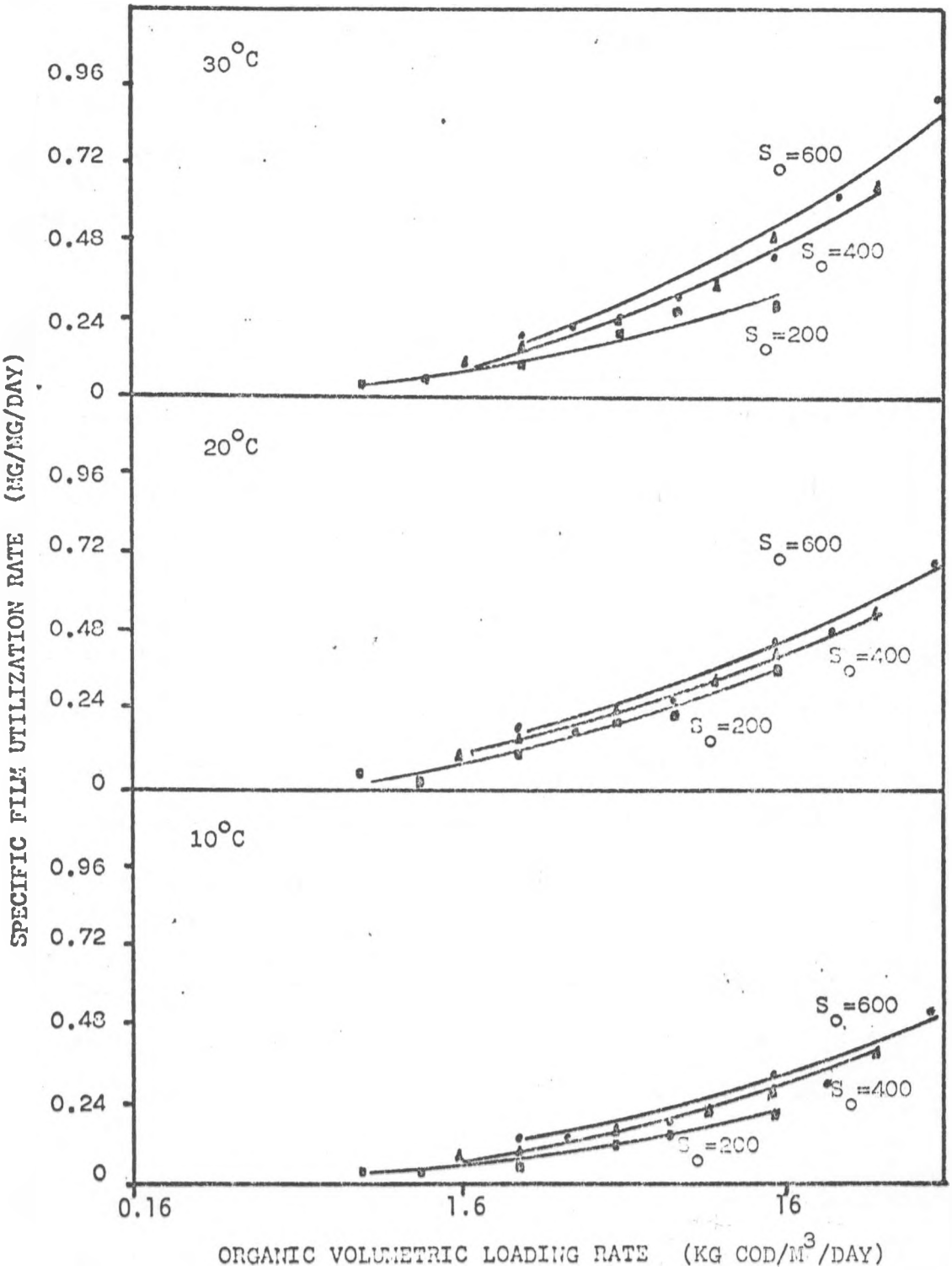

FIGURE 22. SPECIFIC FILM UTILIZATION RATE VS. ORGANIC LOADING RATE-INFLUENCE OF INFLUENT SUBSTRATE CONCENTRATION

5.2.4 First order approximation of the Monod equation and specific utilization equation

Equation 20, the Monod equation, can be further simplified into a first order equation in the manner of Garret and Sawyer (41) when the substrate concentration is low compared to the value of K_s. This leads to equation 22:

$$B = \frac{B_{MAX} \times S}{K_s} = K_1 S \qquad \text{Eq (22)}$$

where $K_1 = \frac{B_{MAX}}{Ks}$

The relatively high Ks values for anaerobic treatment and low substrate concentrations used in this study, provide the rationale for this equation. This equation will be used to demonstrate the effect of influent substrate concentration on process efficiency as expressed by the effluent substrate concentration, S_e. It will subsequently be used to demonstrate that the process efficiency is dependent on the net specific film growth rate, a parameter based on fundamental concepts. In the following paragraphs this analysis will be developed.

Table 22 presents the data for values of B at each value of S_e where S_e is the effluent COD concentration (equal to mixed liquor substrate concentration for a completely mixed reactor) at each influent substrate concentration (S_o) for each temperature. Equation 22 predicts that when S is equal to zero, B will also be equal to zero. Thus, the intercept of this equation should be zero and the slope will be equal to K_1. The data were thus analyzed by least squares linear regression forced through the origin. The results

Table 22. Data for developing design relationship between the net specific film growth rate and effluent substrate concentration

30°C					
So = 200 mg/l		So = 400 mg/l		So = 600 mg/l	
B days-1	Se mg/l	B	Se	B	Se
.0036	42.4	.0025	72.7	.0041	107.9
.0058	56.0	.0074	69.3	.0041	88.0
.0094	68.1	.0129	91.2	.0135	152.7
.0191	78.3	.0380	113.4	.0324	205.3
.0478	94.8	.0535	130.3	.0714	225.1
.0478	148.9	.0714	219.5	.0956	315.5
Slope = K_1 = .00304 r^2 = .772		Slope = K_1 = .00297 r^2 = .899		Slope = K_1 = .000238 r^2 = .911	
20°C					
B	Se	B	Se	B	Se
.0021	34.1	.0031	48.6	.0043	79.5
.0028	51.8	.0041	55.1	.0069	101.7
.0066	55.4	.0060	76.4	.0107	124.0
.0200	86.4	.0232	131.3	.0225	264.0
.0500	108.6	.0316	172.0	.0390	268.0
.0500	124.6	.0633	232.0	.0780	340.0
Slope = K_1 = .000335 r^2 = .908		Slope = K_1 = .00214 r^2 = .969		Slope = K_1 = .000168 r^2 = .930	

Table 22. (Continued)

10°C

B	Se	B	Se	B	Se
.0066	54.3	.0066	66.9	.0063	136.6
.0057	60.2	.0055	77.2	.0059	142.7
.0096	90.5	.0104	139.1	.0108	240.3
.0261	99.1	.0225	185.1	.0215	321.0
.0295	110.0	.0305	212.2	.0441	345.8
.0885	129.6	.0813	243.5	.0705	385.1
Slope = K_1 = .000201 r^2 = .820		Slope = K_1 = .000118 r^2 = .934		Slope = K_1 = .000174 r^2 = .766	

are calculated in Table 22 and shown graphically in Figure 23.

Figure 23 shows that the influent substrate concentration plays a significant role in determining the relationship between the effluent substrate concentration and the net specific film growth rate. In a manner similar to that employed by Grady and Williams (42), a simple model for the effects of influent substrate concentration and net specific film growth rate can be formulated for the AAFEB process under operating conditions that allow use of first order approximation of the Monod equation. This equation can be written as follows:

$$Se = K_1 \, So \, B \qquad \text{Eq (23)}$$

Using this equation, an analysis can be made for each temperature including all three influent substrate concentrations utilized in the investigation using the least squares recognition analysis forcing the equation through the origin. The results of this analysis is shown in Figure 24. Values of K_1 were found to be 6.44, 9.45 and 12.72 for 30°, 20° and 10° C with corresponding correlation coefficients (r^2) of 0.882, 0.914, and 0.765, respectively.

Equation 19 can similarly be further simplified into a first order equation when the substrate concentration is low compared to the value of K_s. This leads to equation 24:

$$A = \frac{k}{ks} \; S = K_2 S \qquad \text{Eq (24)}$$

where $K_2 = \frac{k}{ks}$

Table 23 presents the data for values of A at each value of S_e

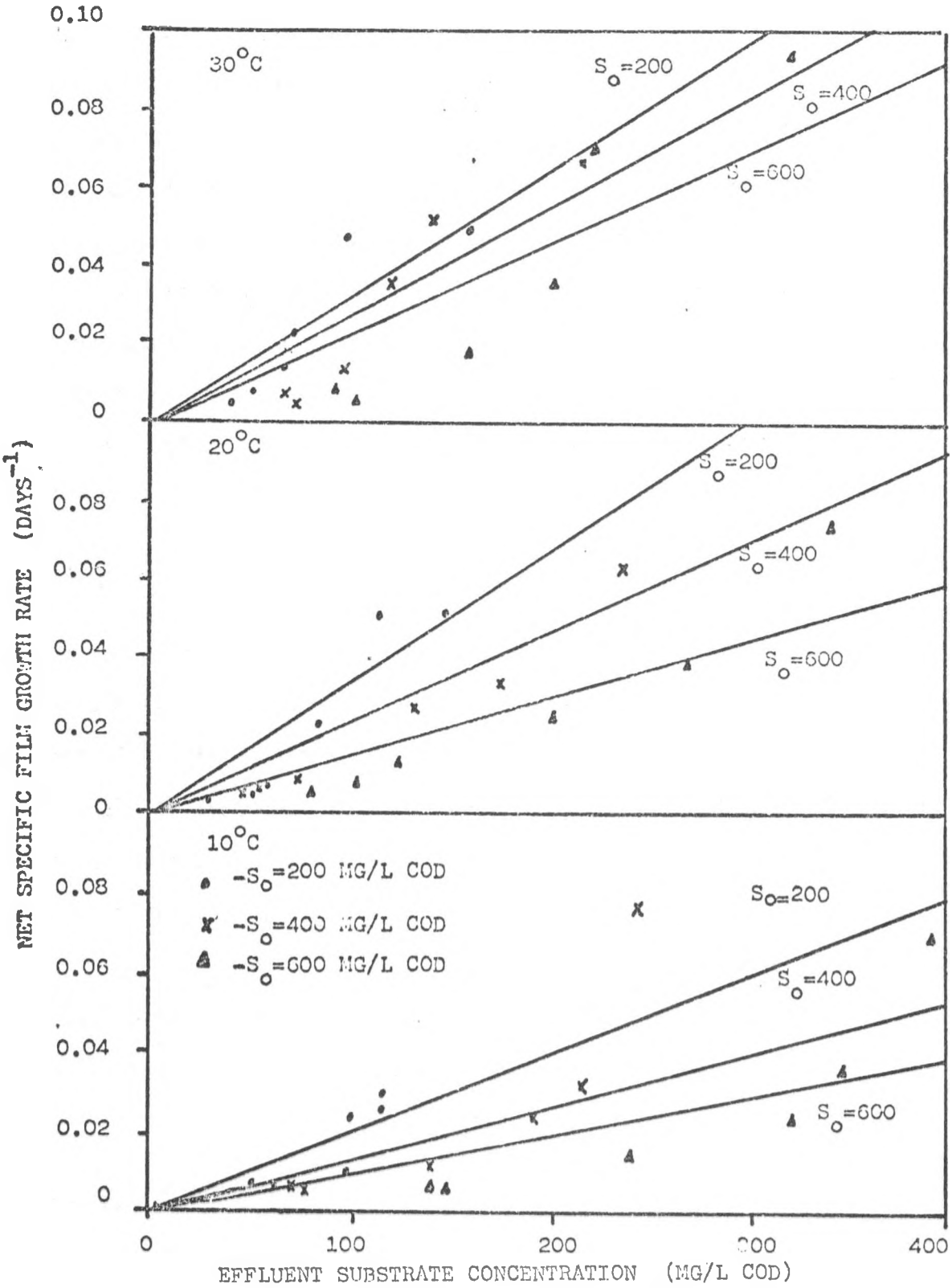

FIGURE 23. PLOTS OF NET SPECIFIC FILM GROWTH RATE VS. EFFLUENT SUBSTRATE CONCENTRATION DEMONSTRATING THE EFFECT OF INFLUENT SUBSTRATE CONCENTRATION

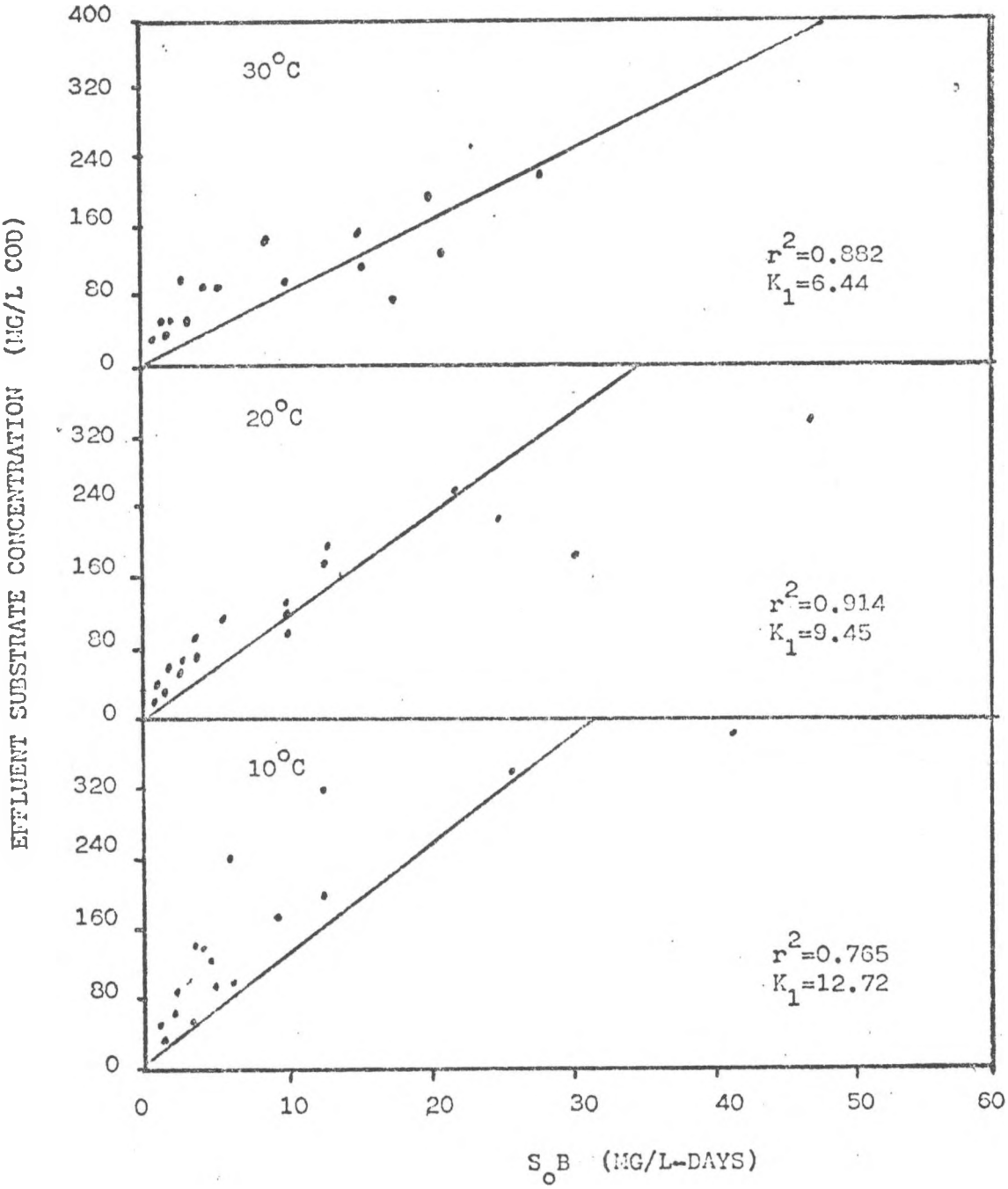

FIGURE 24. FIRST ORDER LINEAR APPROXIMATION OF THE MODIFIED MONOD EQUATION, $S_e=S_oBK_1$

Table 23. Data for developing design relationship between the specific film substrate utilization rate and effluent substrate concentration

30°C

So = 200 mg/l		So = 400 mg/l		So = 600 mg/l	
A $days^{-1}$	Se mg/l	A	Se	A	Se
.0540	42.4	.1046	72.7	.1536	107.9
.0700	56.0	.1528	69.3	.1956	88.0
.0986	68.1	.1994	91.2	.2637	152.7
.1939	78.3	.3350	118.4	.4185	205.3
.2517	94.8	.4812	130.3	.5961	225.1
.2440	148.9	.6441	219.5	.9060	315.5
Slope = K_2 = .0021 r^2 = .910		Slope = K_2 = .0029 r^2 = .922		Slope = K_2 = .0025 r^2 = .953	

20°C

A	Se	A	Se	A	Se
.0519	34.1	.1044	48.6	.1411	79.5
.0398	51.8	.1084	55.1	.1438	101.7
.1204	55.4	.1838	76.4	.2258	124.0
.1891	86.4	.2832	131.3	.3422	204.0
.2299	108.0	.3612	172.0	.4315	268.0
.3799	124.0	.5322	232.0	.6710	340.6
Slope = K_2 = .0023 r^2 = .907		Slope = K_2 = .0022 r^2 = .993		Slope = K_2 = .0013 r^2 = .970	

Table 23. (Continued)

10°C

So = 200 mg/l		So = 400 mg/l		So = 600 mg/l	
A	Se	A	Se	A	Se
.0434	54.3	.0638	66.9	.0684	136.6
.0499	60.2	.0832	77.2	.1024	142.7
.0660	90.5	.1130	139.1	.1375	240.3
.1238	99.1	.1819	185.1	.2049	321.0
.1658	110.0	.2383	212.2	.2803	345.8
.2594	129.6	.3979	243.6	.4740	385.1
Slope = K_2 = .0014 r^2 = .860^2		Slope = K_2 = .0012 r^2 = .845^2		Slope = K_2 = .0008 r^2 = .790^2	

at each influent substrate concentration (S_o) for each temperature. Again each set of data was analyzed by a least squares linear regression line forced through the origin.

Figure 25 shows that the influent substrate concentration also plays a significant role in determining the relationship between the effluent substrate concentration and the specific utilization by the film. Again, a simple model for the effect of influent substrate concentration and specific film utilization rate can be formulated for the AAFEB process under operating conditions that allow use of the first order approximation of the Monod equation.

This equation can be written as follows:

$$Se = K_2 \, So \, A \qquad \text{Eq (25)}$$

Using this equation an analysis can be made for each temperature covering all three influent substrate concentrations utilized in the laboratory investigation using a least squares regression analysis forcing the equation through the origin. The results of this analysis are shown in Figure 26. Values of K_2 were found to be 0.71, 0.99, and 1.84 for 30°, 20°, and 10° C with corresponding correlation coefficients (r^2) of 0.848, 0.805, and 0.861, respectively.

5.3 Mass balance model

As was previously mentioned, Eckenfelder (34, 35) developed a simplified mass balance predictive equation relating organic removal efficiency to hydraulic retention time. This equation is repeated below:

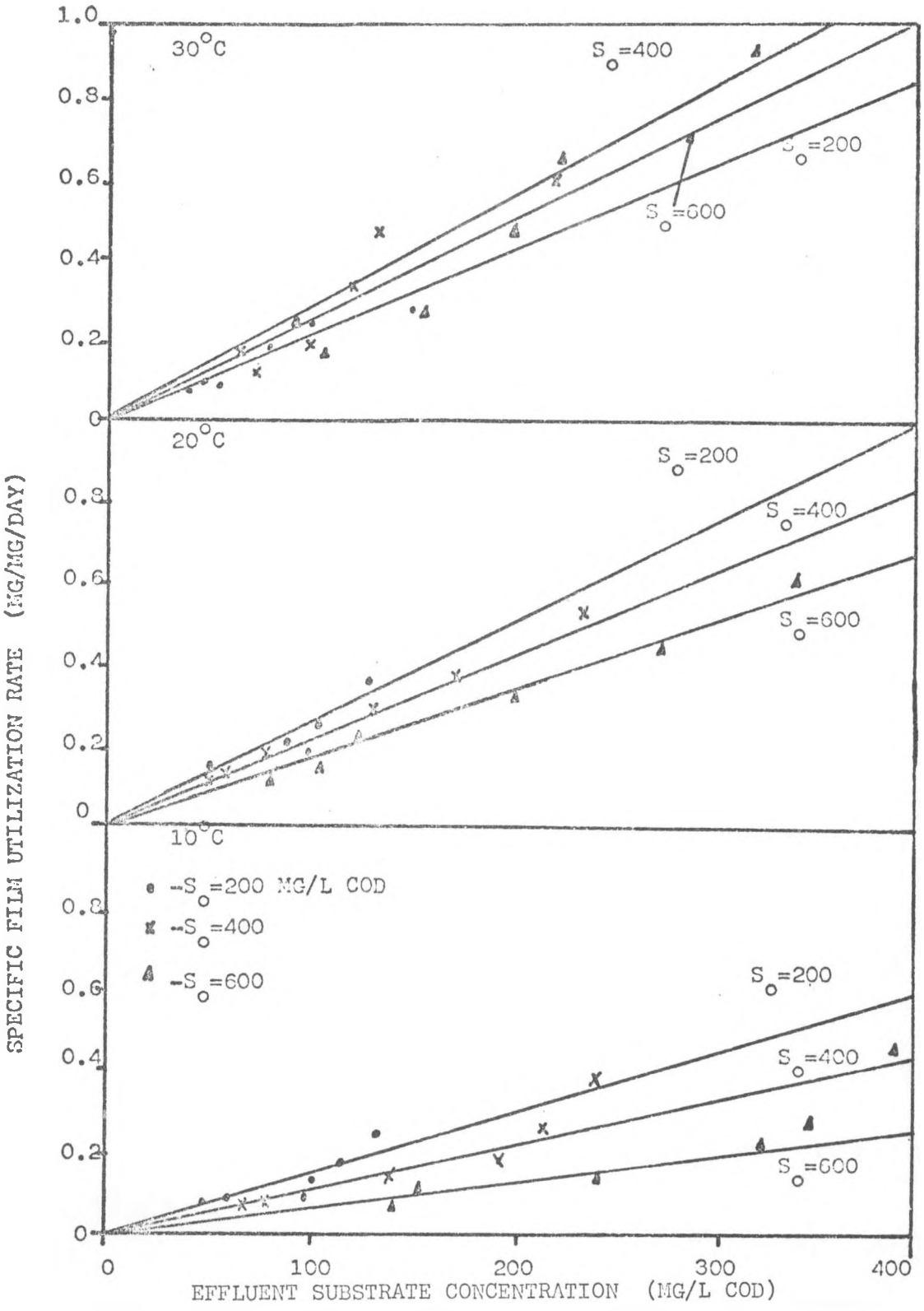

FIGURE 25. PLOTS OF SPECIFIC FILM SUBSTRATE UTILIZATION RATE VS. EFFLUENT SUBSTRATE CONCENTRATION DEMONSTRATING THE EFFECT OF INFLUENT SUBSTRATE CONCENTRATION

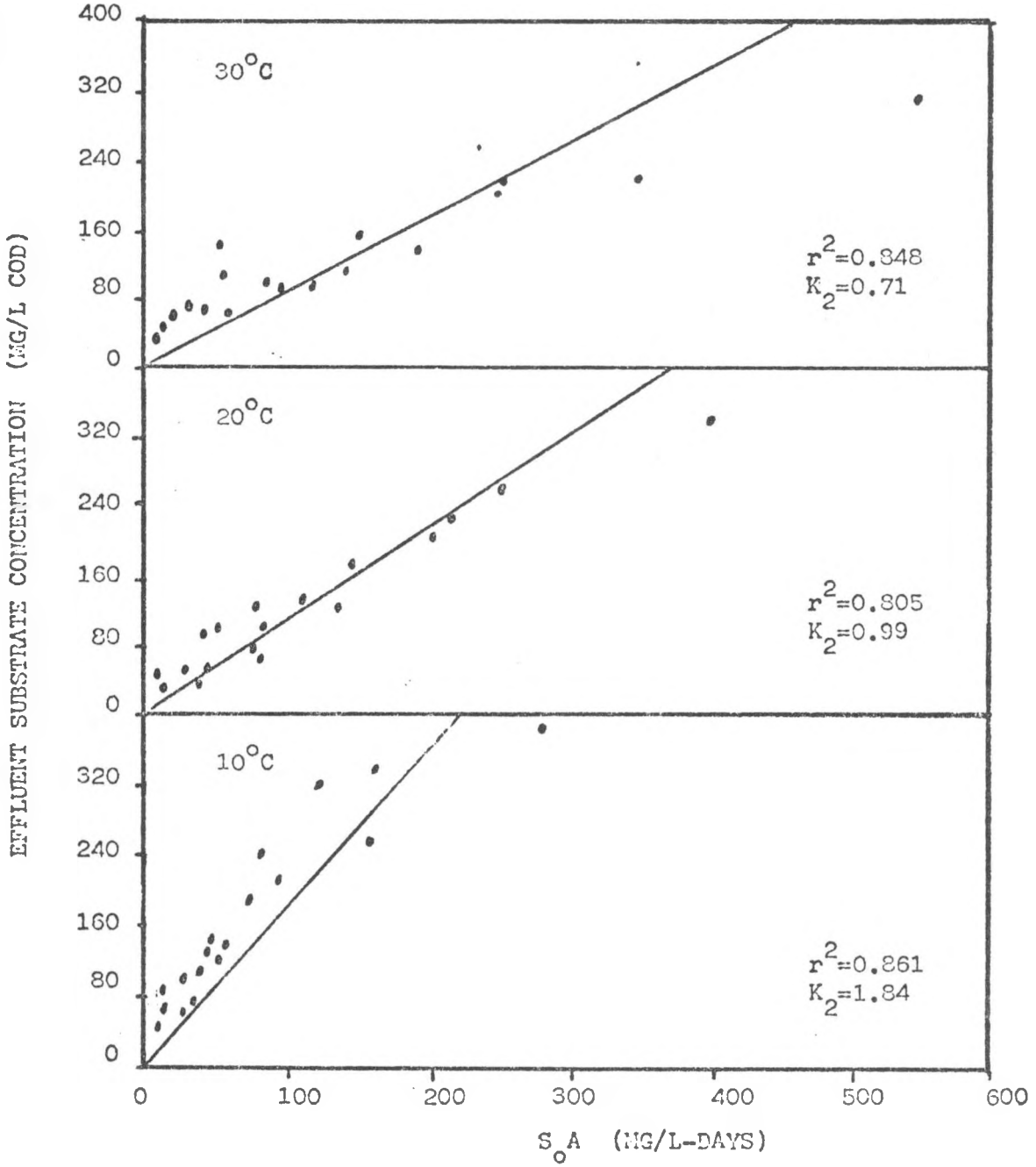

FIGURE 26. FIRST ORDER APPROXIMATION OF THE MODIFIED SPECIFIC UTILIZATION EQUATION, $S_e=S_oAK_2$

$$\frac{Se}{So} = \frac{1}{1 + K^1 t} \qquad \text{Eq (17)}$$

where: S_e = effluent substrate concentration, mass/volume

S_o = influent substrate concentration, mass/volume

K^1 = substrate removal coefficient, $time^{-1}$

t = hydraulic retention time, time

Inverting equation 17 and then subtracting the quantity one from each side results in:

$$\frac{So}{Se} - 1 = K^1 t \qquad \text{Eq (26)}$$

By equating the left side of equation 26 as the dependent variable (Y) and letting t be the independent variable (X), we have an equation of the form $Y = K^1X$, which is a linear curve which passes through the origin. Table 24 presents the data for each influent substrate concentration at each temperature evaluation in this form. Also in Table 24 are the results of the least squares regression analysis of the best fit curve passing through the origin. The slope (which equals K^1) and r^2 values are listed for each set of data.

Figures 27 and 28 are plots of the data with best fit lines drawn in. Figure 27 shows the influence of temperature at each influent substrate concentration at each temperature. These figures suggest that both temperature and influent substrate concentration exert significant influence on the slopes of these lines and consequently on the substrate removal coefficient, K^1. This can be further quantified by a two-way analysis of variance of the K' values testing row effects (temperature) and column effects (S_o) for significance. This is shown in Table 25. Analysis of variance and

Table 24. Mass balance model data

So = 200		So = 400		So = 600	
Y	X (hrs)	Y	X	Y	X
30°C					
3.71	6	4.56	6	4.46	6
2.75	4	4.77	4	5.81	4
1.93	2	3.38	2	2.92	2
1.55	1	2.37	1	1.92	1
1.10	0.66	2.06	0.66	1.66	0.66
0.34	0.33	0.82	0.33	0.90	0.33
$K^1 = 0.69\ hrs^{-1}$		$K^1 = 1.39\ hrs^{-1}$		$K^1 = 1.50\ hrs^{-1}$	
$r^2 = .936$		$r^2 = .920$		$r^2 = .992$	
20°C					
4.86	6	7.23	6	6.54	6
2.86	4	6.25	4	4.89	4
2.61	2	4.23	2	3.83	2
1.31	1	2.04	1	1.92	1
0.85	0.66	1.32	0.66	1.52	0.66
0.61	0.33	0.72	0.33	0.76	0.33
$K^1 = 0.83\ hrs^{-1}$		$K^1 = 1.39\ hrs^{-1}$		$K^1 = 1.21\ hrs^{-1}$	
$r^2 = .998$		$r^2 = .936$		$r^2 = .951$	
10°C					
2.68	6	4.97	6	3.39	6
2.32	4	4.18	4	3.20	4
1.20	2	1.87	2	1.49	2
1.01	1	1.96	1	0.86	1
0.81	0.66	0.88	0.66	0.73	0.66
0.54	0.33	0.64	0.33	0.55	0.33
$K^1 = 0.51\ hrs^{-1}$		$K^1 = 0.91\ hrs^{-1}$		$K^1 = 0.65\ hrs^{-1}$	
$r^2 = .970$		$r^2 = .976$		$r^2 = .941$	

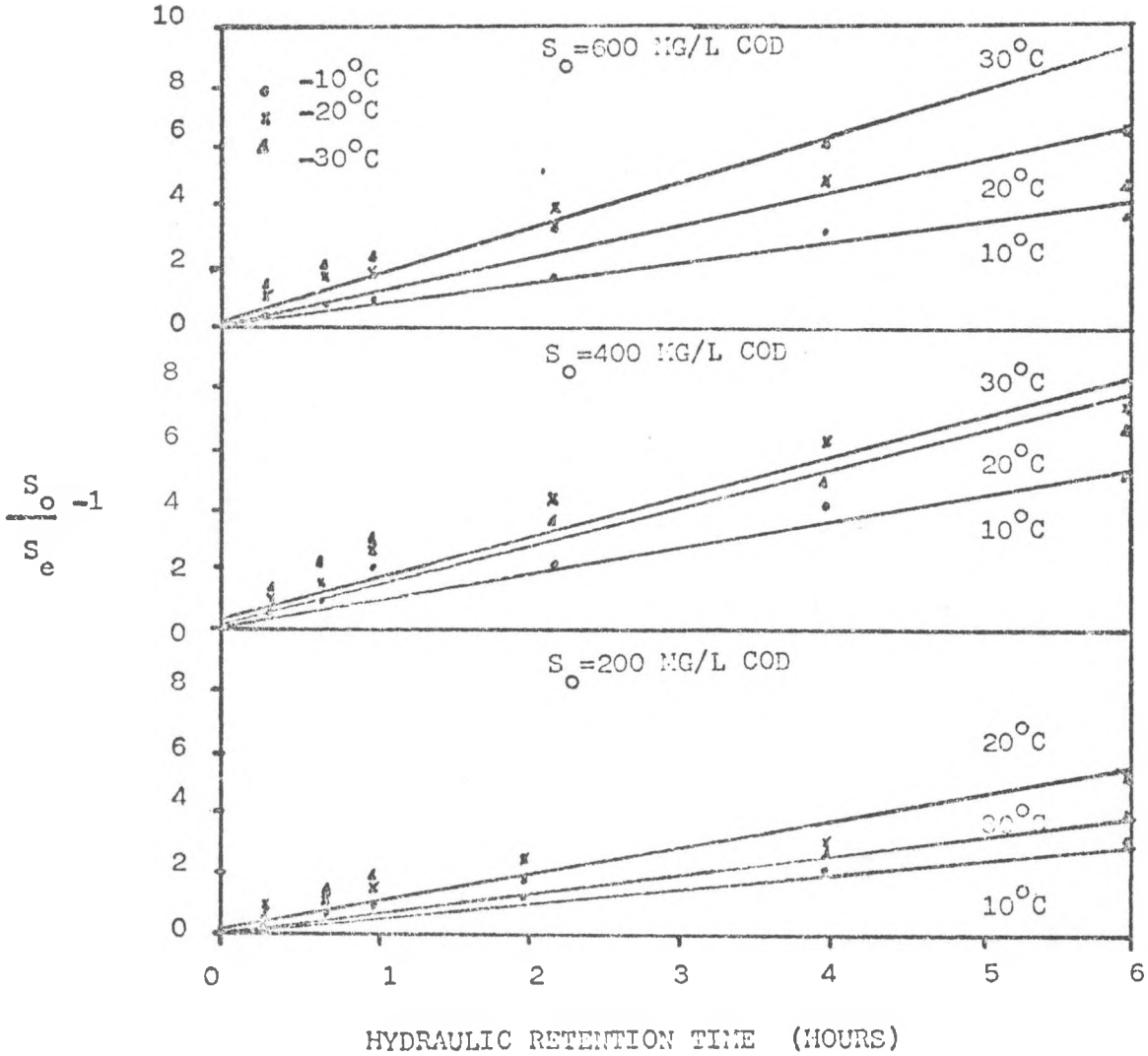

FIGURE 27. PLOT OF MASS BALANCE EQUATION-EFFECT OF TEMPERATURE

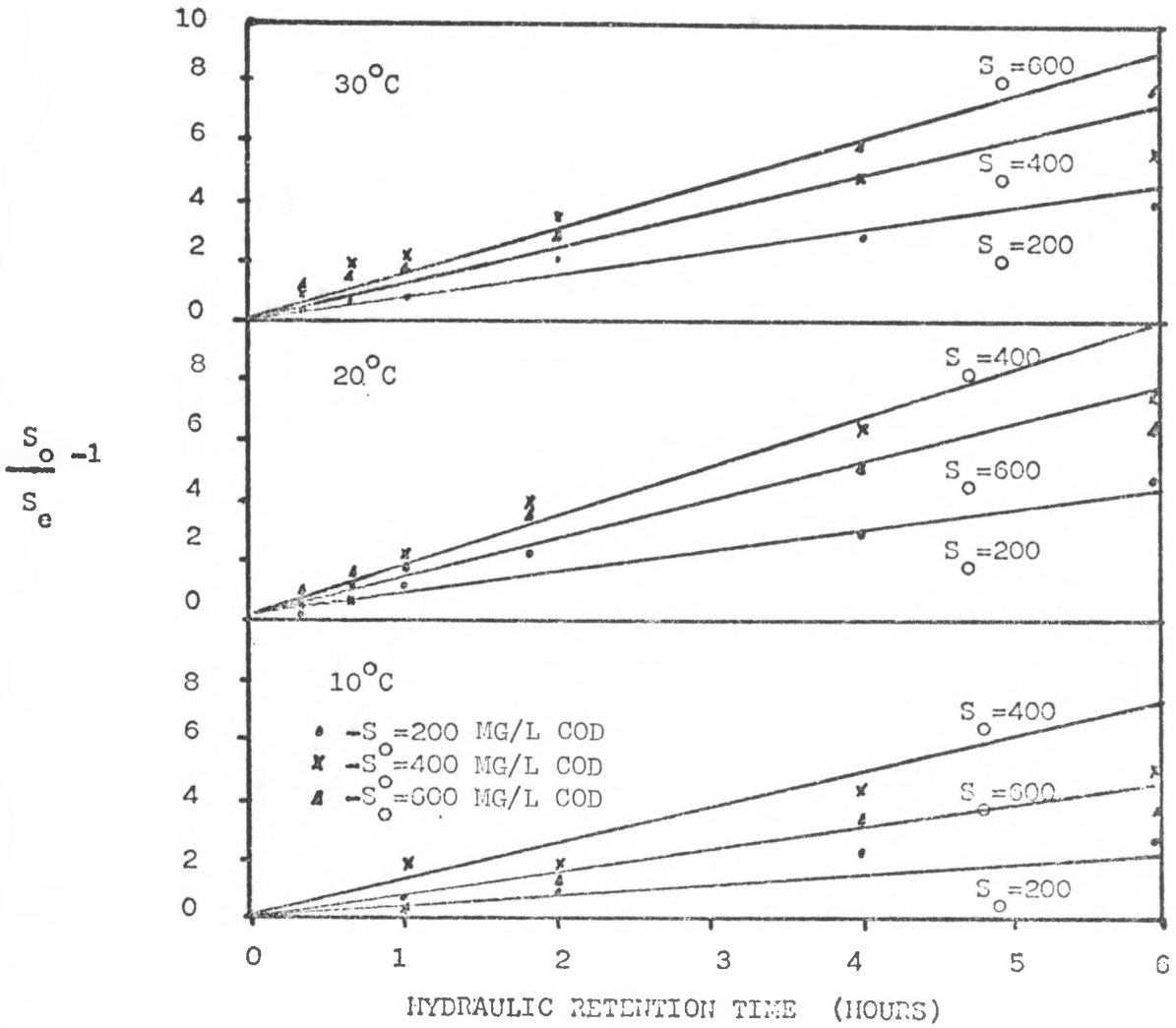

FIGURE 28. PLOT OF MASS BALANCE EQUATION-EFFECT OF INFLUENT SUBSTRATE CONCENTRATION

Table 25. Analysis of variance of K^1 values of mass balance model data for temperature and influent substrate concentration effects

50	200	400	600	
Temperature				
30	0.69	1.39	1.50	$RS_1 = 3.58$
20	0.83	1.39	1.21	$RS_2 = 3.43$
10	0.51	0.91	0.65	$RS_3 = 2.07$
	$CS_1 = 2.03$	$CS_2 = 3.69$	$CS_3 = 3.36$	

	SS	df	F ratio	
Row	0.46	2	7.88	- temperature
Column	0.51	2	8.79	- So
Error	1.09	4		
Total	2.06			

$F_{.05}(2.4) = 6.94$

the methodology for its computation is explained in Sokal and Rohlt (114).

From Table 25 it is seen that both the row and column effects are significant at the 5% testing level. In fact, the influent substrate concentration effect is slightly more significant than the temperature effects.

5.4 Summary of design relationship equations

Two simplified approaches have been developed for predicting the efficiency of the AAFEB process for the treatment of low strength waste organics. These equations can be used to demonstrate the effects of temperature, influent substrate concentration, and organic and hydraulic loading on process efficiency.

Modeling is important toward the goal of obtaining a more rational design procedure for the AAFEB process. These equations can be used as a simplified basis for modeling. The evaluation of these equations is presented in the next section.

CHAPTER 6. DISCUSSION OF RESULTS

The results presented verify that the anaerobic attached film expanded bed reactor is a feasible process for the treatment of low strength soluble organics and demonstrates that the process is able to effectively operate at low temperatures and high loading rates. High biomass concentrations were observed with concomitant thin films. The high efficiency of the AAFEB and evaluation of both the mass balance and modified Monod equations will be discussed in this chapter.

6.1 Comparison of AAFEB to other anaerobic processes

Table 26 presents a summary of literature showing operational performances of conventional, contact and anaerobic filter processes as compared to the AAFEB performance observed in both this study and a previous study done with sewage. The literature studies cover a range of both synthetic and naturally occurring industrial wastes. All studies incorporated an HRT greater than 12 hours and resulted in significant efficient concentrations of biodegradable organics. In contrast, the AAFEB was able to operate at ranges of lower temperatures, higher organic and hydraulic loading rates and was able to achieve much higher organic removal efficiencies.

Mueller and Mancini (95) summarized the literature concerning organic removal efficiencies of anaerobic filters as a function of HRT. This is shown graphically in Figure 29. In general, maximum efficiency for anaerobic filters is achieved at HRT's on the order

Table 26. Anaerobic processes - comparison of biological waste treatment criteria

Waste	Hydraulic retention time (hrs)	Temperature °C	Raw waste mg/l	Loading lbs/1000 ft^3/day	Percent removed	Reference
Conventional Process:						
Butanol	240	-	17,000 BOD	114 BOD	70	Buswell (17)
Acetic acid	720	35	620,000 BOD	975 BOD	99	McCarty et al. (85)
Butyric acid	720	35	400,000 BOD	1000 BOD	98	McCarty et al. (85)
Contact Processes:						
Maize starch	79.2	23	6,280 BOD	110 BOD	88	Hemans et al. (49)
Whiskey distillery	148.8	33	25,000 BOD	250 BOD	95	Painter et al. (101)
Cotton kiering	31.2	30	1,600 BOD	74 BOD	67	Pettet et al. (103)
Citrus	31.2	33	4,600 BOD	214 BOD	87	McNary et al. (87)
Brewery	55.2	-	3,900 BOD	127 BOD	96	Newton et al. (96)
Meat packing	31.2	33	2,000 BOD	110 BOD	95	Pettet et al. (103)
Meat packing	12	33	1,380 BOD	156 BOD	91	Steffen et al. (116)
Meat packing	12	35	1,430 BOD	164 BOD	95	Schroepfer et al. (111)
Meat packing	12	29	1,310 BOD	152 BOD	94	Schroepfer et al. (111)
Meat packing	12	24	1,110 BOD	131 BOD	91	Schroepfer et al. (111)

Table 26. (Continued)

Anaerobic Filter:						
Volatile acid and protein carbohydrate	4.5-72	25	375-12000 COD	265-212 COD	56-98 COD	Young and McCarty (131)
Food Processing (carbohydrate)	13-83	35	8500	100-640 COD	30-86 COD	Plummer and Malina (105)
Acetic acid	12	33	6400 COD	370 COD	30-80 COD	Clark and Speece (22)
Potato processing waste	13-59	19-22	3000 COD	33-145 COD	41-79 COD	Pailthorp (102)
Wheat starch waste	22	-	5930-13100 COD	237 COD	65 COD	Richter et al. (108) Taylor (119)
Synthetic	17-46	32-35	2000 COD	35-130 COD	64-76 COD	Havlow et al. (54)
Petrochemical	72	34	2000-8000 COD	40-145 COD	10-13 COD	
Breney press liquor	15-330	35	6000-27000 COD	50-400 COD	30-97 COD	Force et al. (39)
'Metrecal'	18	30	11000 COD	427 COD	70-95 COD	El-Shafie & Bloodgood (36)
Pharmacetutical work	12-48	37	1250-16000 COD	14-220 COD	94-98 COD	Dennis and Jennett (32)
Sulfite liquor	89-95	35	1300-5300 BOD	125-375 BOD	27-58 BOD	Wilson and Timpany (127)

Table 26. (Continued)

AAFEB:						
Sewage	0.05-4.8	20	200 TCOD	66-3714-TCOD	0-85 TCOD	Jewell and Switzenbaum (164)
Glucose	0.33-6	30	200-600 SCOD	50-2700 SCOD	25-85 SCOD	This study
	0.33-6	20	200-600 SCOD	50-2700 SCOD	38-86 SCOD	This study
	0.33-6	10	200-600 SCOD	50-2700 SCOD	35-83 SCOD	This study

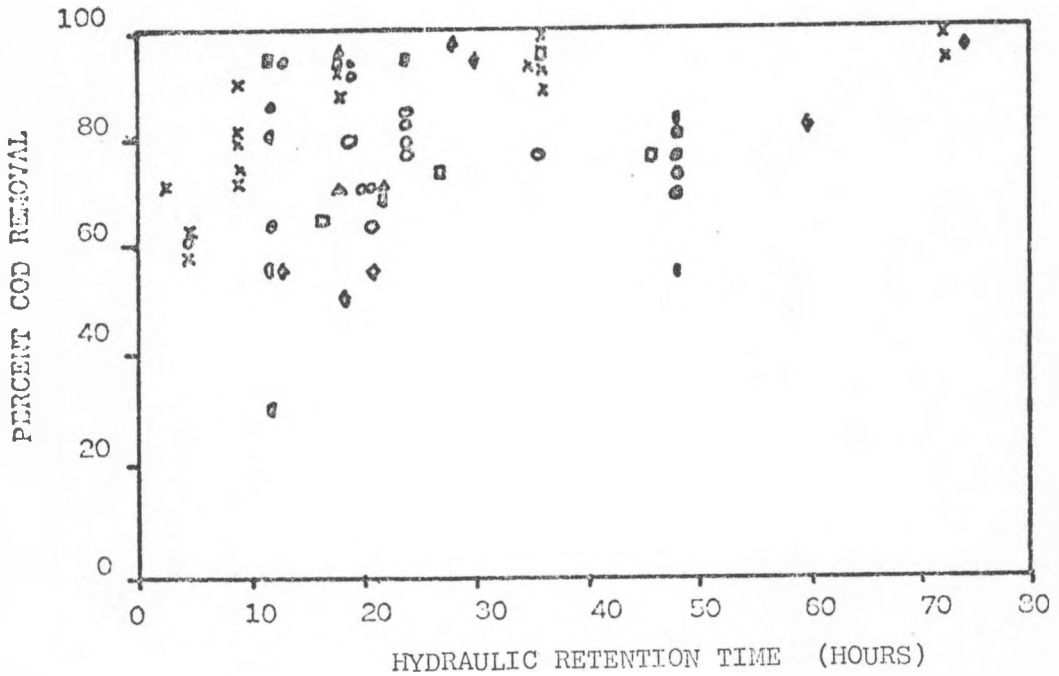

FIGURE 29. EFFECT OF HYDRAULIC RETENTION TIME ON COD REMOVAL EFFICIENCY OF ANAEROBIC FILTERS. (AFTER MUELLER AND MANCINI (95))

24 hours or more. The AAFEB performance as summarized in Figures 9 and 10 show maximum removal efficiencies occurring at only several hours. In addition, the AAFEB is not hampered by clogging and short-circuiting which are limitations with the anaerobic filter according to Young (130).

A very significant point is that the AAFEB is able to treat low strength organic wastes. In general little research has been done concerning anaerobic treatment of dilute wastes. This is because the process had been thought to be inefficient at low temperatures and organic concentrations. The anaerobic fermentation has more commonly been used for solids reduction and waste stabilization of concentrated sludges and manures. Table 27 presents a comparison of treatment of dilute wastes by the AAFEB and other studies concerning the treatment of dilute wastes with the anaerobic processes from the literature. Again, the AAFEB (numbers 5 and 6) is able to treat low strength substrates at lower temperature ranges, higher organic and hydraulic loading ranges and, in general, with higher organic removal efficiencies.

In summary, the AAFEB was demonstrated to treat low strength organic wastes (COD 200 to 400 mg/l) at low temperatures (10°C, 20°C), at short hydraulic retention times (several hours) and at high organic loading rates (up to 8.0 kg COD/m^3/day (0.5 lbs COD/ft^3/day)) anaerobically. A comparison made to other studies reported in the literature shows the AAFEB to be more effective in terms of higher loadings and lower temperatures and better able to

Table 27. Anaerobic processes - treatment of low strength wastes

Waste	Hydraulic retention time (hrs)	Temperature °C	Raw waste mg/l	Loading lbs/1000 ft^3/day	Percent removed	Reference
1. Acetate, sugar	7-18	30	1200-1800 COD	125-314 COD	40 COD	Van der Meer (122)
2. Sewage	35	4-25	181 BOD	8.2 BOD	53-78 BOD	Coulter et al. (27)
3. Sewage	24	20	500 TCOD 148 SCOD	0.03 TCOD	49 SCOD 90 TCOD	Pretorius (106)
4. Sewage	12	22	400-1700 COD	0.04 BOD	80 BOD	Simpson (113)
5. Sewage	0.05-4.8	20	200 TCOD	66-3714 TCOD	85 TCOD	Jewell and Switzenbaum (164)
6. Glucose	0.33-6	30	200-600 SCOD	50-2700 SCOD	25-85 SCOD	This study
	0.33-6	20	200-600 SCOD	50-2700 SCOD	38-86 SCOD	This study
	0.33-6	10	200-600 SCOD	50-2700 SCOD	35-83 SCOD	This study

treat dilute influent substrate concentrations.

6.2 Comparison of AAFEB to aerobic processes

Table 28 presents a comparison of treatment of dilute wastes by the AAFEB and biological waste treatment criteria of conventional activate sludge and modifications of activated sludge processes. The data used for the AAFEB entry (line 12)are that for the 4 hour hydraulic retention time over all temperatures and influent substrate concentrations.

The general, the ranges for percent removal, organic loading and the volumetric flow ratio (V/Q) are in the same range for both the AAFEB and the activated sludge processes. The AAFEB was observed to have higher biomass concentrations and therefore, similar V/Q ratios, higher SRT values, and lower F/M ratios. The AAFEB also produces less sludge than aerobic processes. Economic and energy comparisons between the AAFEB and conventional activated sludge processes will be made in the next chapter.

6.3 Effect of temperature on process efficiency

The effect of temperature on the process efficiency of the AAFEB can be evaluated by analyzing the kinetic parameters K_1, K_2, and Y by an Arrhenius temperature dependence plot. The relationship between temperature and reaction velocity as stated by Arrhenius (3) can be expressed as:

$$\bar{V} = A^1 \, e^{-\Delta H/RT} \qquad \text{Eq (27)}$$

Table 28. Aerobic processes - comparison of biological waste treatment criteria (Metcalf and Eddy, 1972) (89)

Modification	% BODs removal	F/M	SRT (days)	lbs BOD_5/1000 ft^3/day removed	MLSS (mg/l)	Volume to flow (V/Q) (hr)
1. Conventional	85-95	0.2-0.4	5-15	18-36	1500-300	4-8
2. Complete mix	85-95	0.2-0.6	5-15	45-108	3000-6000	3-5
3. Step aeration	85-95	0.2-0.4	5-15	36.54	2000-3500	3-5
4. Kraus	85-95	0.3-0.8	5-15	36-90	2000-3000	4-8
5. Contact stabilization	90-90	0.2-0.6	5-15	50-64	(1000-3000)[a] (4000-10000)[b]	(0.5-1.0)[a] (3-6)[b]
6. Modified	65-75	1.5-5.0	0.2-0.5	68-135	200-500	1.5-3
7. High rate	75-90	0.4-1.5	5-10	82-825	4000-10000	0.5-2
8. Extended	75-95	0.05-0.15	20-30	8.0-21	3000-6000	18-36
9. Pure oxygen	85-95	0.25-1.0	8-20	90-225	6000-8000	1-3
10. Aerobic lagoon	50-60	2.0	-	-	100	24-72
11. Aerobic facula-tive lagoon	80-90	2.0	0	0	50-100	84-720
12. AAFEB[c]	70-85	.02-.09	110-400	50-200[d]	15400-33400	4.0

[a]Contact unit; [b]Solid stabilization unit; [c]This study at HRT - 4 hrs (overall temperatures and Sos); [d]Expressed as soluble COD.

where: $\bar{V}$ = the velocity of the reaction

A^1 = a constant

ΔH = the activation energy

R = the gas constant

T = the absolute temperature

This relationship has found diverse use in describing the effect of temperature on many biological and microbiological parameters. In comparison, the activation energies are discussed more frequently than the constant A by most researchers. The Arrhenius equation is based on thermodynamic principles and finds wide usage in the scientific literature.

Figure 28 shows a semilog plot of Y vs. 1/T and the least squares line of best fit for this data. This is the Arrhenius plot of the Y results. It can be seen that the Y values fit the exponential relationship, but decrease with increasing temperature. Muck (94), Ng (97), and Brown and Rose (14) have found Y values to increase with temperatures up to some optimum point and then decrease as the temperature is increased. These studies were performed with both pure and mixed aerobic cultures in slurry. For an anaerobic slurry system, Lawrence (74) found that the yield constant showed no temperature trends.

Figures 31 and 32 show a semilog plot of $1/K_1$ and $1/K_2$ vs. 1/T and the least squares line of best fit for this data. These data fit very well to the Arrhenius plot and increase with increasing temperature. The activation energies were found to be 5986 and 8385 calories/mole for the K_1 and K_2 plots, respectively.

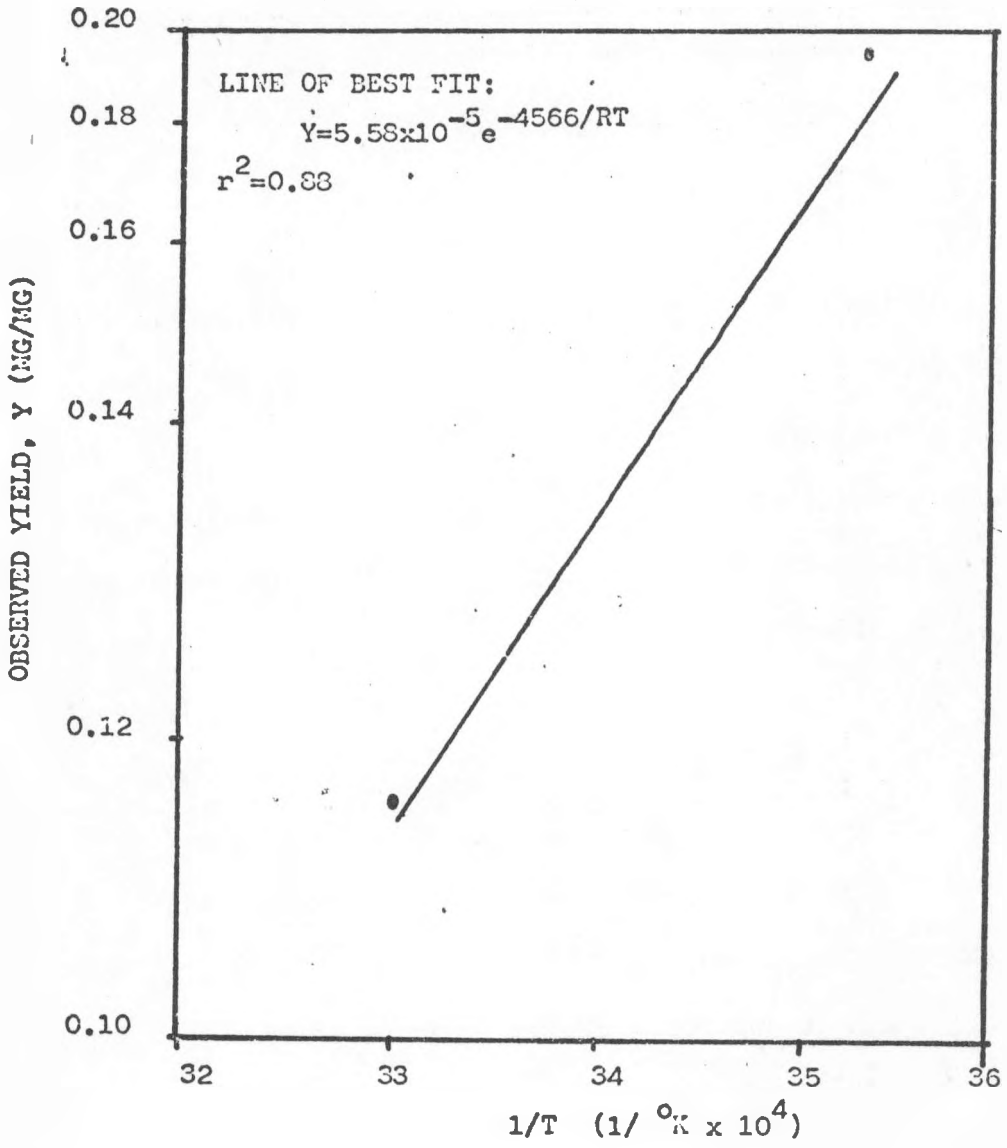

FIGURE 30. ARRHENIUS PLOT-EFFECT OF TEMPERATURE ON OBSERVED YIELD (Y)

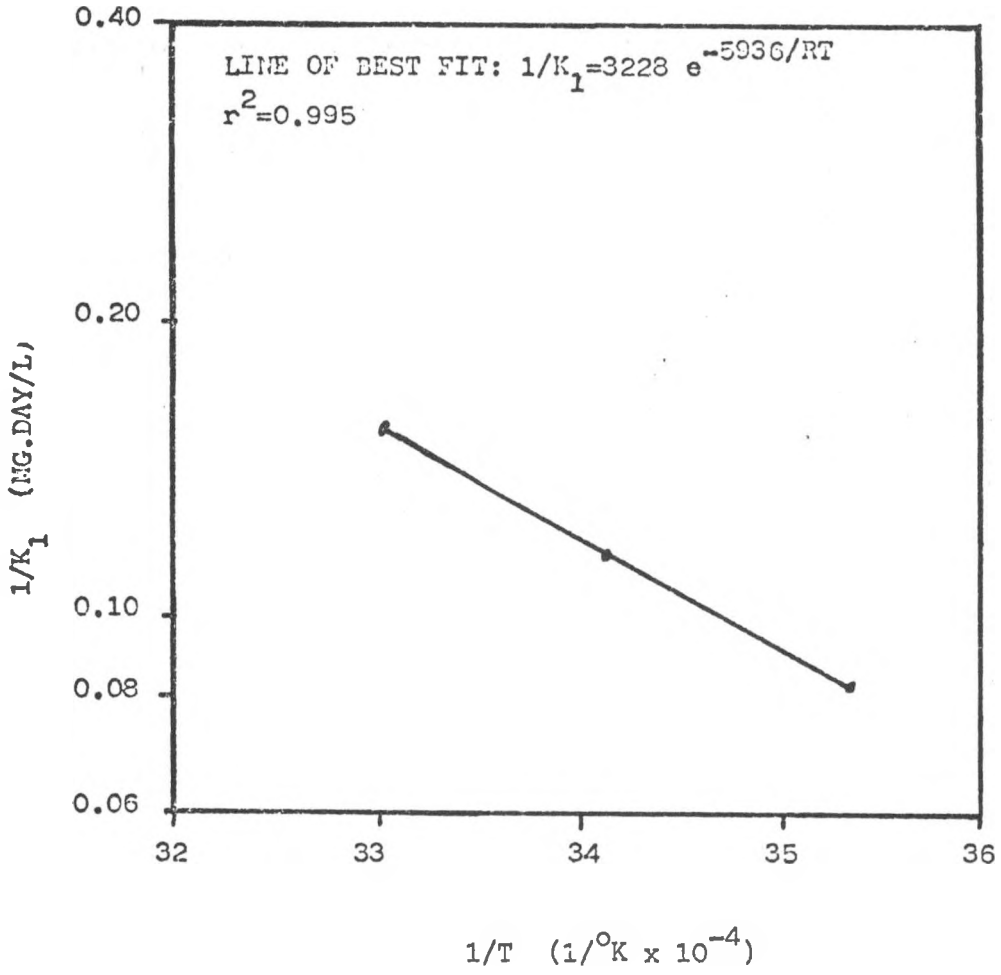

FIGURE 31. ARRHENIUS PLOT-EFFECT OF TEMPERATURE ON KINETIC COEFFICIENT K_1

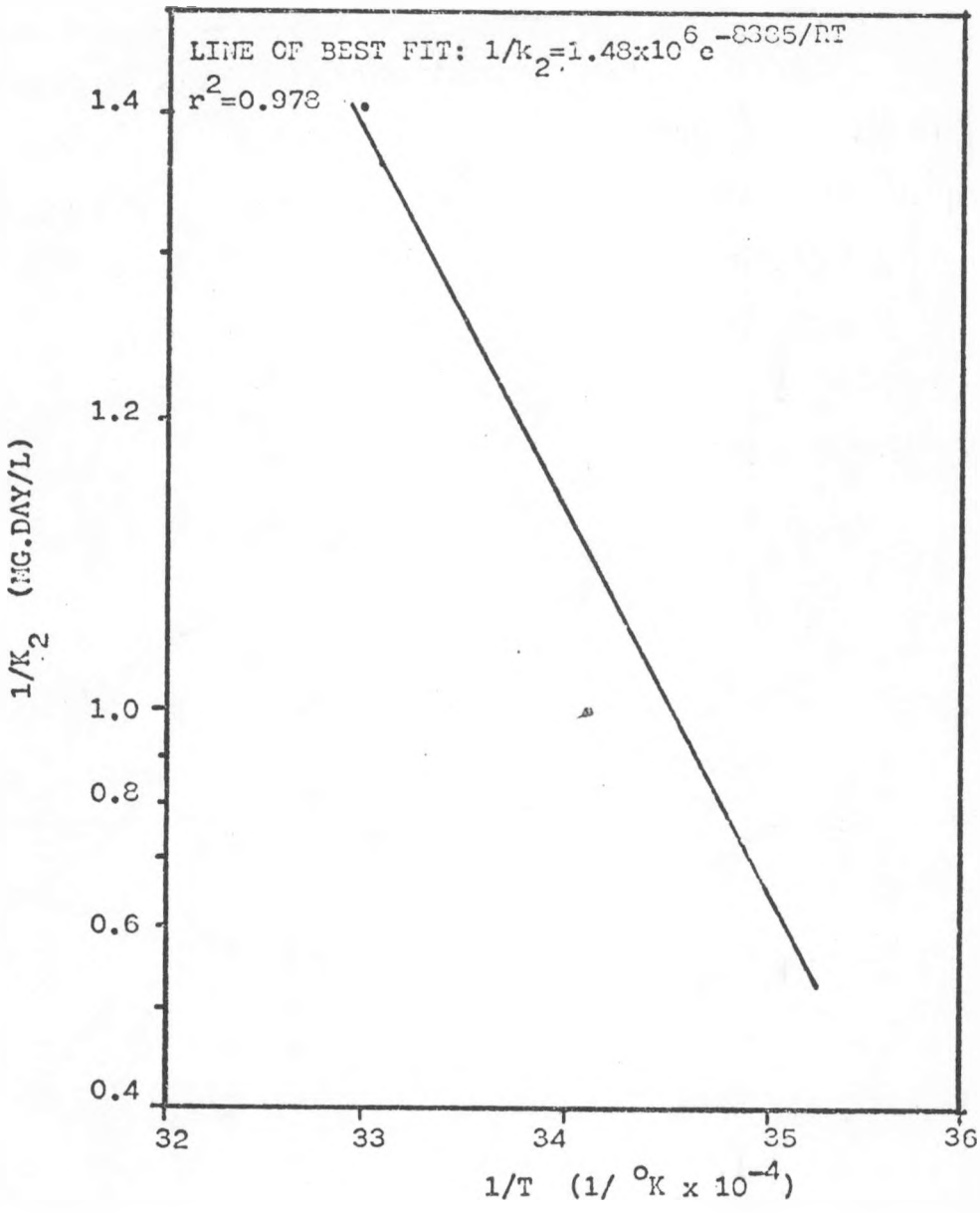

FIGURE 32. ARRHENIUS PLOT-EFFECT OF TEMPERATURE ON KINETIC COEFFICIENT K_2

These activation energies are relatively low compared to literature values. The K_1 and K_2 values do not follow the approximate rule of Van Hoff which states that the rate of chemical reactions doubles for each 10°C rise in temperature (37). Ashare et al. (4) have compiled a summary of studies concerning anaerobic slurry fermentations and have an Arrhenius temperature dependence for a rate constant, K, to fit the following least squares line:

$$K = 3.3 \times 10^9 \, e^{-15{,}000/RT} \qquad \text{Eq (28)}$$

Here, the activation energy was found to be 15,000 Cal/mole. For this activation energy, an increase of 10°C will result in an approximate doubling of the rate constant, K. It must be stressed again that the Ashare summary was concerning suspended growth systems only. The AAFEB is an adhered growth system.

The large difference between the activation energy for an anaerobic slurry system and the AAFEB points out a distinct advantage of the AAFEB process. The film system compensates for temperature much easier than the suspended growth systems. This is most likely due to the large mass of organisms present in the AAFEB and the fact that at lower temperatures, film thicknesses and concomitant biomass concentrations increase. It may also be due to electrochemical interphase process phenomena, to a lesser degree (81). This will be further discussed in the next section.

6.4 Film thickness and biomass concentration

The success of any biological waste treatment process is

dependent on bringing the waste into intimate contact with the biomass for a sufficient period of time to allow the desired biochemical reaction to occur. The advantages of the expanded bed for anaerobic fermentation over conventional, contact and anaerobic filter processes, was discussed previously. In short, the expanded bed process is able to achieve very high biomass build-ups with small film thicknesses because of the large surface area to volume ratios.

The high removal rates observed with the AAFEB can be attributed to this high surface area to volume ratio of the biologically coated support particles. Also because of the large biomass concentrations, smaller reactor volumes can be used and hence less land would be required. The thin films are important in that thick films may limit the reaction rate by mass transfer limitations. As was shown in Figures 11 and 12, films observed were relatively thin, never exceeding 15 microns in thickness.

The mass concentrations exceeded 40,000 mg/l with the vast majority (over 95%) being attached mass (Figures 15, 16, and Table 18). This is in general agreement with the figure of 32,800 mg/l volatile solids observed by Butts with an aerobic fluidized bed reactor (18). Young (130) measured a maximum mass concentration of 24,850 mg/l VSS in the bottom portion of the anaerobic filter. Theoretically, it can be calculated that the maximum attached mass would not exceed the 35,000-40,000 mg/l range (60). This figure is based on physical restrictions regarding maximum void space and maximum solids concentrations of films.

Figures 14 and 16 show the effect of varying influent substrate concentration on attached film thicknesses and biomass concentrations respectively. In general, thicker films and higher mass concentrations were observed with higher substrate concentrations. This is in agreement with several other studies (Reid (107) and Heukelekian (51)) which found that the amount of aerobic slime formed varied directly with the concentration of nutrient material. Heukelekian (51) also stated that abundant slimes which readily slough are characteristic of all growth of liquids high in oxidizable material while dense readily adhering slimes are most characteristic of low concentration nutrient solutions, such as found in this study.

Figures 13 and 15 illustrate the influence of temperature on attached film thickness and biomass concentrations. Thicker films and higher mass concentrations were observed at lower temperatures. Heukelekian (50) and Sullins and Galler (117) have observed the same phenomenon. Figure 33 shows the relationship of relative sludge amounts versus temperature for a full scale trickling filter used in the Sullins and Galler study.

The fact that higher mass concentrations were observed at lower temperatures is significant toward explaining the lower activation energies associated with the temperature dependence of the kinetic parameters. The system is able to compensate well because of the fact that the mass is able to increase at lower temperatures. This is a distinct advantage over suspended microbial systems.

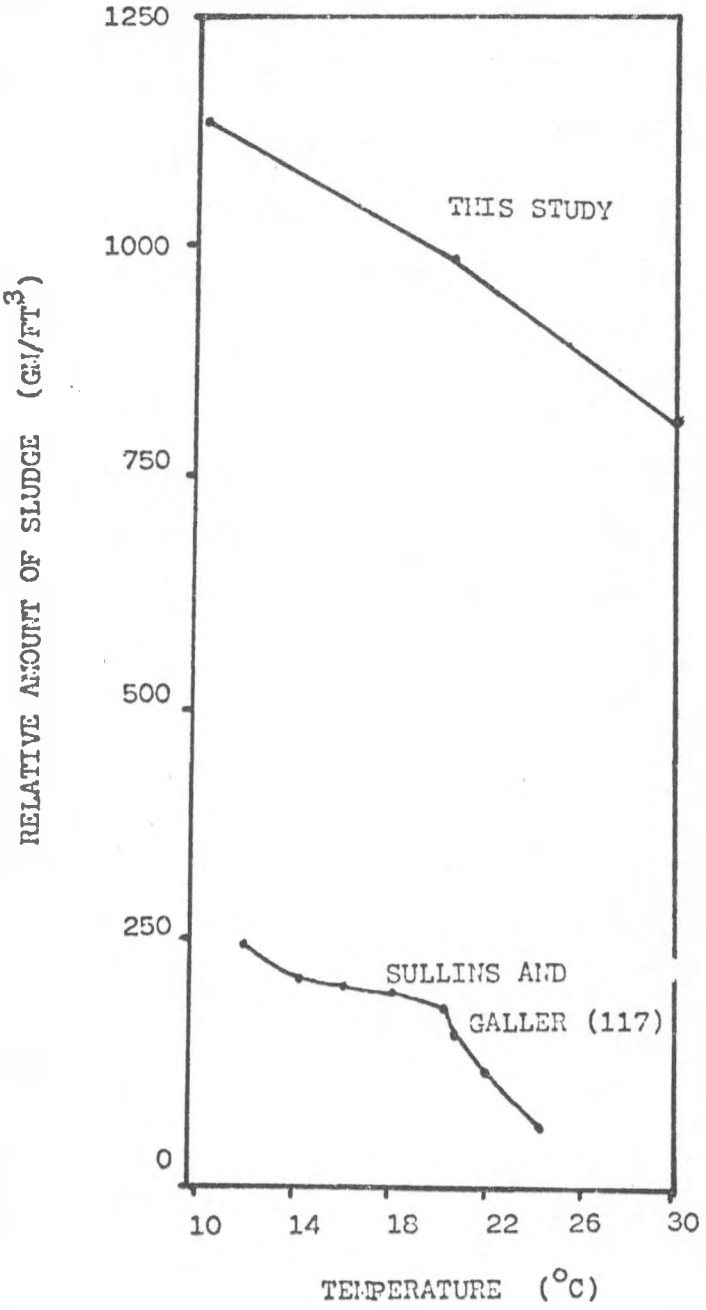

FIGURE 33. VARIATION OF FILM DISTRIBUTION WITH TEMPERATURE

6.5 Evaluation of first order approximation of Monod equation and specific utilization equation

Equations 23 and 25 repeated below were used to show that the overall process efficiency in terms of effluent substrate concentration is primarily dependent upon the influent substrate concentration, S_o, at a constant temperature

$$Se = K_1 \, So \, B \qquad \text{Eq (23)}$$

$$Se = K_2 \, So \, A \qquad \text{Eq (25)}$$

Figures 24 and 26 show the results of the data points against plots of equations 23 and 25, respectively.

Figures 34 and 35 are the same as Figures 24 and 26 except that the 95% confidence intervals are shown. They are represented by the dashed lines above and below the predictive equations. These confidence intervals are for mean values of S_e for any given So B value or So A value, respectively.

Most of the data points in Figure 34 fall within the confidence intervals and we could conclude that equation 23 is of value for predicting the effluent concentration or process organic removal efficiency. It must be noted, however, that most points fall above the predicted equation line. A better fit could be obtained by having the equation intersect the ordinate axis at some positive value rather than going through the origin. While this would have no actual meaning since S_e must be equal to zero when B equals zero, it points out that the actual observed effluent concentration is a result of the removal efficiency of the process and the fact

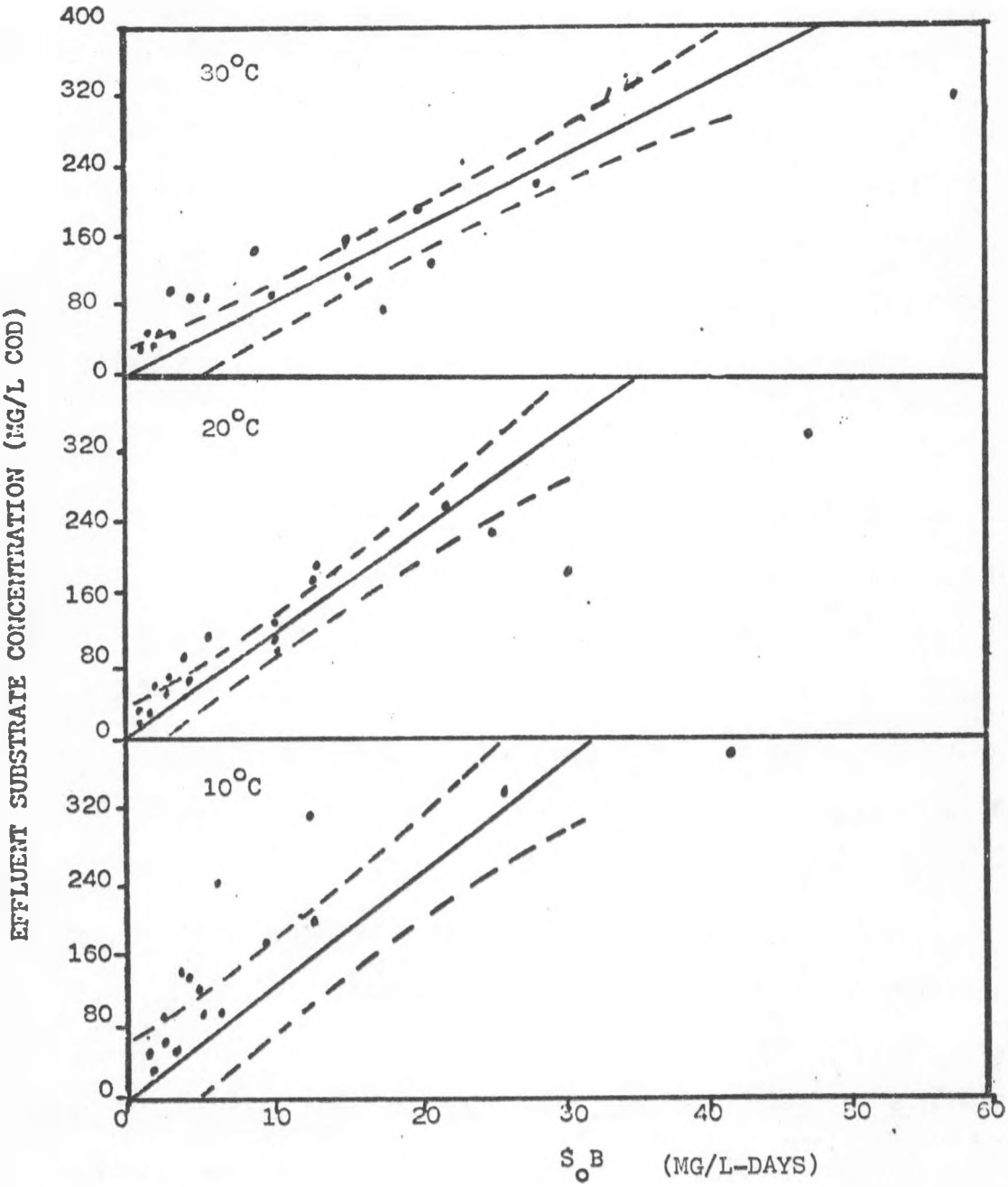

FIGURE 34. 95 PERCENT CONFIDENCE INTERVALS FOR MEAN VALUES OF S_e FOR EACH S_oB

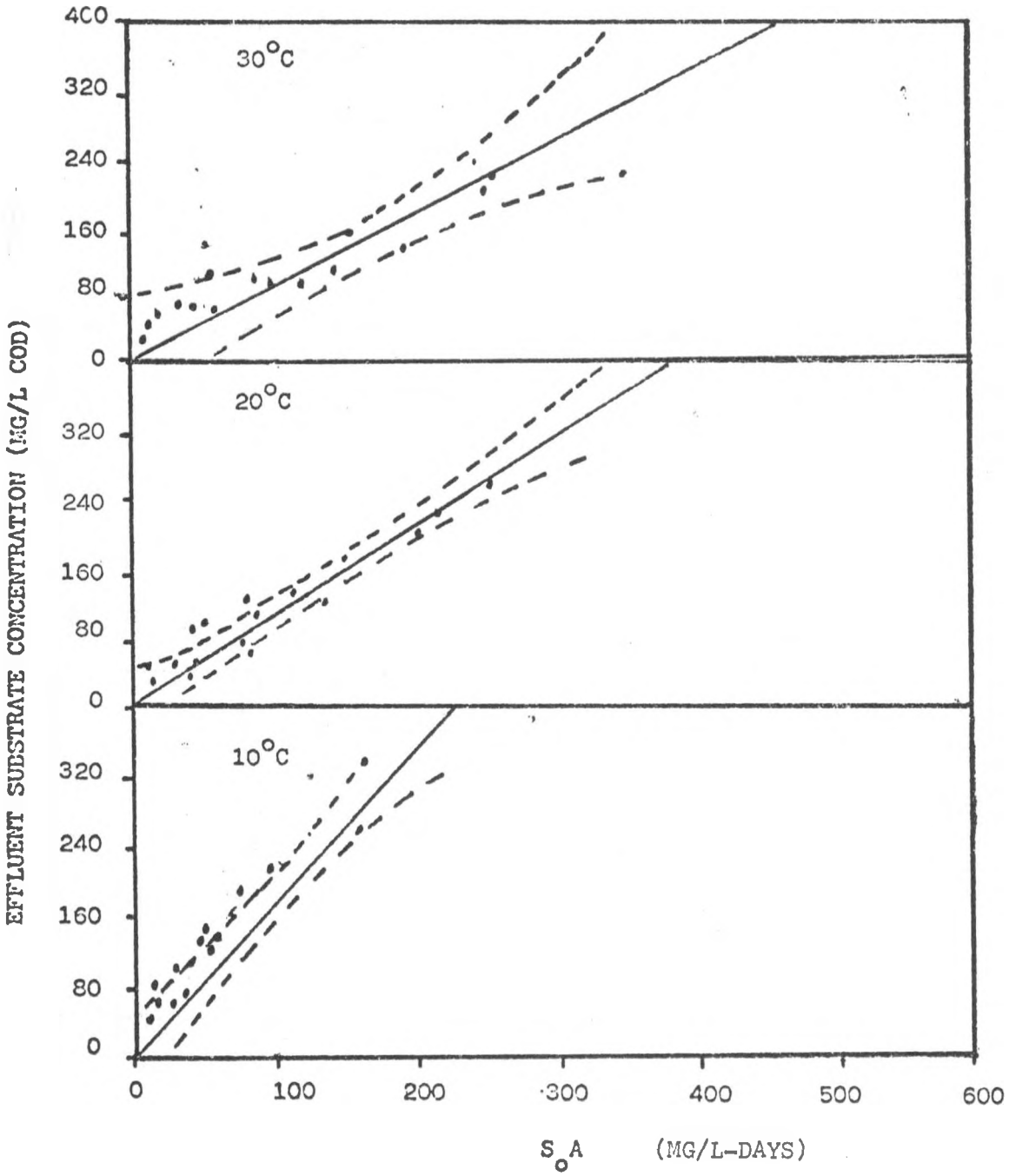

FIGURE 35. 95 PERCENT CONFIDENCE INTERVALS FOR MEAN VALUES OF S_e FOR EACH S_oA

that the biochemical reaction also produces organic matter. That is to say, some of the organic matter in the effluent is not from the original substrate and is probably of microbial origin. This phenomenon has been observed before (Daigger and Grady (30)). Also, while the correlation coefficients (r^2) show that it is reasonable to assume a linear fit of the data points, a better fit might be achieved by fitting the data to some order of reaction less than one. While the actual order of the reactor may be better described as first order, the reaction in the expanded bed reactor may be better described by an order less than one because of physical limitations imposed by the reactor vessel itself, such as short circuiting.

Figure 35 shows a somewhat better fit than Figure 34 by having more data points within the 95% confidence intervals. Again, however, most of the points fall above the prediction line, and visual inspection shows that a better fit may be obtained by a lower order equation.

In conclusion, the modified Monod and specific utilization equation are satisfactory for predicting the effluent concentration as a function of influent substrate concentration and net specific film growth rate or film substrate utilization rate at a given temperature. The equations were developed for low substrate concentrations (COD = 200-600 mg/l COD) for a temperature range of 10° to 30°C. Both equations are first order and pass through the origin. The deviations from fit of the actual to predicted values

suggest that the anaerobic expanded bed process is a producer of soluble organic matter as well as a consumer as are all biological systems. It is important to note that B (the net specific film growth rate) which is the inverse of the solids retention time and A (the specific film substrate utilization ratio) are widely used operational parameters which are based on fundamental concepts of microbiology derived from theories of the continuous culture of organisms. Also, they are controllable parameters as they are a function of the organic loading rate and mainly the microbial hold up time or biomass concentration. Atkinson and Knights (9) have pointed out that when other factors are held constant, the equilibrium biofilm thickness can be maintained in a biological expanded bed by controlling bed expansion. This, in turn, controls the biomass concentrations. In order to optimize the biomass concentration for reaction control, other physical parameters such as height of expanded bed, cross sectional area of the reactor, and support media characteristics must be considered.

6.6 Evaluation of mass balance equation

It was shown in section 5.3 that a mass balance equation can be used to show temperature and influent substrate concentration effects on process efficiency. This was shown by equation 26 which is repeated below:

$$\frac{So}{Se} - 1 = K^1 t \qquad \text{Eq (26)}$$

The substrate removal coefficient K^1, was demonstrated to be a

function of temperature and influent substrate concentration.

Figure 36 can be used to evaluate differences between hydraulic and organic loading rate on specific substrate utilization. In Figure 36 the specific film utilization rate A is plotted against the hydraulic retention time. Also, because of the redundancy built into the experimental program, (the same organic loading rate achieved by different combinations of So and HRT) the effect of the organic loading rate is shown. Each line in Figure 36 represents a group of points with the same organic loading rate achieved by different HRT's. Each line is labelled by its organic loading rate in Kg COD/m^3/day.

In general, higher specific utilization rates were observed at high organic loading rates. However, as the line of equal organic loading rates shows, there is little effect on the utilization rate exerted by the hydraulic retention time at equal organic loading rates. This is because of the fact that the bed is expanded by the recycle pump and in general the rate of recycle flow is much higher than the influent flow. Thus, the effect of the HRT is greatly minimized. Thus, in terms of process evaluation the organic volumetric loading rate is of more significance than the hydraulic retention time. This emphasizes the major advantages that attached microbial films have over systems that can be washed out at high flow rates.

Besides showing the effects of temperature and S_o on the process efficiency, the substrate removal coefficient, K', can be used in the analysis of the possible effect of mass transfer limitations on process efficiency. With most biofilm reactors,

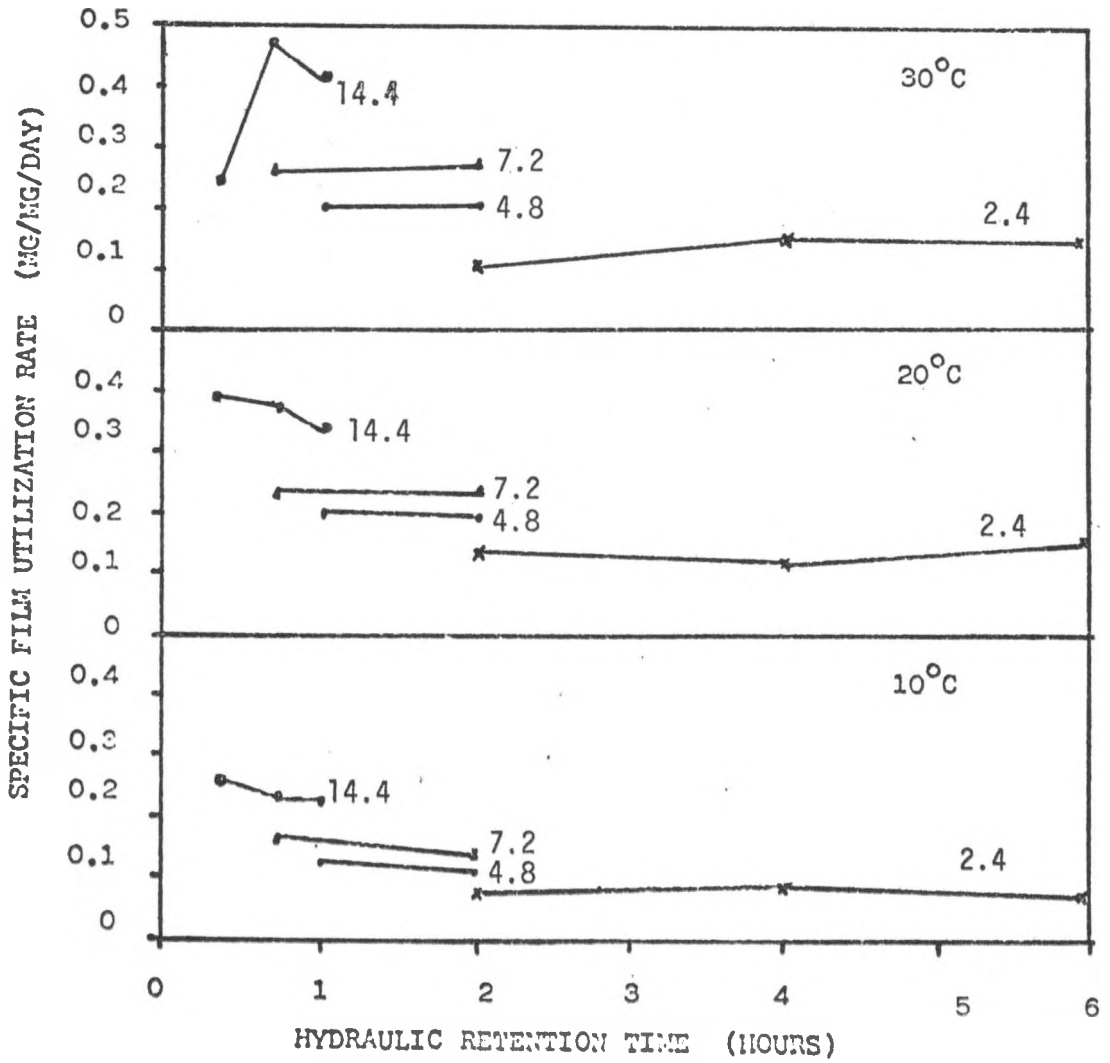

FIGURE 36. COMPARISON OF HYDRAULIC AND ORGANIC LOADING RATES

there is, in general, an associated mass transfer resistance. The mass transfer resistance is intrinsically incorporated in the kinetic equations. This can be misleading in the design of a system as the proper controlling mechanism is not identified in the data analysis. Thus, the effect of mass transfer as well as biological kinetics should be evaluated in data analysis.

Biofilm measurements were shown in Figures 13 and 14 and were observed to be very thin on the order of 15 microns and less. Because of previous studies showing that film thicknesses did not become limiting until reaching much larger thickness, it was assumed that any diffusional resistance afforded by the biofilm was insignificant. This leaves liquid film resistance to interphase transport as a possible means of diffusional resistance.

The magnitude of the liquid film resistance to interphase transport can be estimated in the manner of Baillod and Boyle (10), with the aid of a mass transfer correlation for flow past spheres, assuming each biosupport particle to be spherical. This is shown with equation (29) from Bird, Stewart, and Lightfoot (13).

$$Nu = 2.0 + 0.6\ (N_{RE})^{0.50}\ (Sc)^{0.33} \qquad \text{Eq (29)}$$

where: Nu = mass transfer Nusselt number = $\bar{k}d/Df$ (dimensionless)

N_{RE} = Reynold's number = ud/v (dimensionless)

S_c = Schmidt number = v/Df (dimensionless)

$\bar{k}$ = mass transfer coefficient (defined as mass flux) - concentration driving force (length/time)

v = kinematic viscosity of bulk fluid (length^2/time)

d = spherical particle diameter (length)

u = relative velocity between particle and fluid (l/t) (length/time)

D_f = diffusivity of substrate in bulk fluid (l^2/t) (length^2/time)

U, v, and d can be measured for the AAFEB process used in this study. The diffusivity of glucose in bulk fluid has been reported by LaMotta (73) for low concentrations at 22°C as 1.27×10^{-9} cm^2/sec. The Reynold's number was calculated to be 7.50; the Schmidt number 7.87×10^{-3}, and the mass transfer Nusselt number 33.63. This Nu was then used to calculate the mass transfer coefficient, $\bar{k}$, and this was found to be 8.54×10^{-4} cm/sec. This mass transfer coefficient must be multiplied by the ratio of surface area to volume of an individual particle ($4\ r^2/4/3\ r^3$) to get kLA, the mass transfer coefficient for any individual particle. For the AAFEB kLA was found to be 0.1025 sec^{-1}.

The overall observed kinetic coefficient, K^1, is actually composed of the chemical reaction coefficient K_c and the mass transfer coefficient kLA. This can be mathematically expressed as:

$$1/k^1 = 1/kLA + 1/kc \qquad \text{Eq (30)}$$

For an estimate of the effect of mass transfer, we can compare the kLA measured at low concentrations at 20°C to the K^1, substrate removal coefficient evaluated at 20°C and the lowest influent substrate concentration, 200 mg/l COD. From Table 23, K^1 at 20°C, So = 200 mg/l is 0.83 hrs^{-1} which is equivalent to 2.3×10^{-4} sec^{-1}.

Now, comparing 1/kLA to $1/K^1$ as shown below, it is clearly seen that the effect of mass transfer is insignificant compared to the effect of the chemical reaction coefficient:

$$4337 \text{ seconds } (1/K^1) = 9.76 \text{ seconds } (1/kLA) + 4327.24 \ (1/kc) \qquad \text{Eq (31)}$$

Therefore, it can be concluded that liquid film resistance to interphase transport is insignificant and that the substrate concentration at the film surface is equal to that of the bulk liquid (i.e., no gradient occurs across the stagnant film layer). Therefore, the reaction rates can be meaningfully described by the biological kinetics, and one of the assumptions for using the modified Monod and specific substrate utilization rate equations is verified.

Of a more practical consequence of this is the fact that mass transfer limitations were not observed at the lower fluid velocities used in this study. Since no advantage would be realized by going to higher expansion rates (i.e., to overcome mass transfer resistances) and more energy would be needed to achieve higher expansions, the use of a low expansion in the design of the AAFEB seems justifiable.

6.7 Summary

The AAFEB process has been demonstrated to be effective for the treatment of low strength organic substrates (COD less than or equal to 600 mg/l) operating at low temperatures (10°, 20°C) at high organic loading rates (up to8.0 kg COD/m_3/day (0.5 lbs COD/ft^3/

day)) at short hydraulic retention times (several hours) anaerobically. In comparison to other anaerobic processes treating dilute wastes the AAFEB was found to be more efficient in terms of removal rate at lower temperatures and higher organic and hydraulic loading rates. The process was also found to compare favorably with conventional aerobic processes in terms of efficiency of treatment in the same range of operational variables.

Temperature was found to be an important variable affecting process efficiency, but the process was shown to be able to compensate well for changes in temperature. The kinetic parameters did not follow the approximate rule of Van Hoff, but increased with increasing temperature at a lesser rate. This was believed due to the fact that a very large mass was present in the system and larger biomass concentrations are less suceptible to temperature changes than small concentrations. Also, it was observed that larger biomass occurred at lower temperatures. This, too, is important for temperature compensation. The organic loading rate was found to be a more significant process control parameter than hydraulic retention time for the lab scale AAFEB.

The success of the process is believed due mostly to the large biomass concentrations and high SRT values achieved with thin films because of the large surface area to volume ratio realized by the small support particles. Also, because of the thin films and physical nature of the expanded bed hydraulics, mass transfer resistances were not found in the AAFEB process and the process was, thus, kinetically controlled by the biochemical reactions.

Modifications of the Monod and specific substrate utilization equations were found to describe the efficiency of the system for low influent substrate concentrations (COD = 200 to 600 mg/l) and for a temperature range of 10° to 30° and these equations showed the importance of the influent substrate concentration on process efficiency. These equations related the process efficiency to two widely used operational parameters, which are based on fundamental microbial concepts and are controllable parameters.

CHAPTER 7. ENGINEERING SIGNIFICANCE

Based on the results of this study, the AAFEB process appears to have potential as a secondary treatment unit process as well as in other applications. In terms of removal efficiencies, detention times, and organic loading rates, ranges achieved were comparable to those found with conventional aerobic systems for the treatment of dilute organic wastes. The high removal rates obtained with the AAFEB are attributed to the high surface area to volume ratio of the biologically coated support particles. This results in high solids retentions times with low hydraulic retention times.

The process also was stable at high loading rates, and washout did not occur even at HRT's as low as 0.33 hours and organic loading as high as 43.2 kg COD/m^3/day (2700 lbs COD/1000 ft^3/day) because of high SAT values. Effluent suspended solids concentrations were found to be relatively low even at the highest loading rates.

In comparison to other anaerobic process configurations, the AAFEB has been found to be superior in terms of higher removal efficiencies at lower temperatures, shorter retention times and at higher organic loading rates. The process was also shown to be able to successfully treat low strength wastes which has not commonly been found with anaerobic fermentations.

The major economic advantages of the AAFEB process over conventional aerobic processes are 1) elimination of aeration requirements, 2) lowering of sludge handling requirements, and

3) production of methane during waste treatment. Possible disadvantages are 1) high energy requirements for expanding the bed, and 2) high initial capital cost for the columns. These points are addressed in this section with a brief preliminary comparison of the AAFEB and two conventional treatment processes.

7.1 Economic considerations

At this point, there is insufficient engineering information regarding design parameters to optimally build a full scale AAFEB system. Thus it is difficult to compare directly the economics of an AAFEB system with conventional aerobic systems such as the trickling filter and activated sludge processes. However, certain aspects can be discussed.

First of all, because of the low yield associated with the anaerobic fermentation, less sludge is produced with the AAFEB than conventional aerobic processes treating dilute organic wastewaters. This is significant in terms of capital costs, but mainly in terms of operation and maintenance costs where sludge handling can make up a very large fraction of the total operational and maintenance costs. Also, because the column also acts as an upflow clarifier (the upflow rate is less than the settling velocity of the sludge) low effluent solids concentrations can be maintained without a secondary clarifier.

As the AAFEB is anaerobic, no oxygen requirement is necessary, thus higher loading rates can be used and no aeration equipment is

necessary. This provides a large savings on operation and maintenance costs as aeration is a high energy consumption operation.

Finally, methane is a byproduct of anaerobic waste treatment. For a flow of one million gallons per day (1 MGD) assuming a wastes strength of 100 mg/l soluble COD after primary settling, and an 80 percent removal efficiency by the AAFEB, 100,000 liters of methane would be produced daily which would have an energy value of 930,000 kcal. This is a significant portion of the pumping energy consumption needed by the AAFEB to expand the beds as will be shown in the next section.

7.2 Energy consumption comparison

For the purpose of an energy consumption comparison, a full scale AAFEB will be briefly described. The design is simply a scale up of the bench units used in this study. For a one MGD flow, five towers 20 meters in height and 2 meters in diameter would be required assuming an organic loading rate of 4.8 $kg/m^3/day$ (300 lbs COD/1000 ft^3/day). This would result in a hydraulic retention time of one hour and this flow would be sufficient to expand the bed without recycle. As with the bench scale units, the towers would be operated at minimal expansion and only the bottom half would contain the support particles. The top half would be freeboard and act as a clarification zone.

It should be pointed out that this design represents a "worst case" basis since little data are available for scale up such as

optimal diameter to length ratio, depth of bed, and particle size. Clearly, a column 20 meters in height is unacceptable. One simple alternative would be to design a column with only the active reactor zone and no clarification zone. This column would be only 10 meters in height. Instead of using a clarification zone in the reactor, a polishing device such as a microscreen unit could be utilized to separate any excess amount of solids from the effluent.

The energy consumption for the AAFEB and conventional activated sludge and trickling are shown in Table 29.

The energy usage is expressed as kilowatt-hours per day and the requirements for all processes were based on an analysis by Mills and Tchobanoglous (90). All energy usages are calculated only for the unit processes and not for the rest of the treatment sequence.

The activated sludge energy requirements are based on aeration and mixing, intermediate pumping, and the sedimentation basin. The trickling filter is based on the intermediate pump and secondary clarifier.

The AAFEB is based on the energy needed for bed expansion and the amount of net energy derived from the methane produced if it were used to run the pumps. A total of 903 kwh/day would be needed for expansion of the 20 meter columns. However, 356 kwh/day could be recovered from the methane produced assuming a 0.33 conversion efficiency of the methane to mechanical energy. Thus the net

Table 29. Energy consumption comparison (for 1 MGO flow)

Process	Consumption (Kwh/day)
Activated sludge unit	2384
Trickling filter unit	405
AAFEB unit	546.7
AAFEB + Microscreen volt	135.7

energy as seen in Table 29 is 546.7 kwh/day. A lower requirement is found for the AAFEB-microscreen since the towers are only one half as tall, so that the pumping head is much less. This entry in Table 29 includes the energy for expansion pumping, microscreen operation and the energy recoverable from the methane produced.

Again a more detailed energy analysis would be necessary in order to draw a more definite conclusion regarding the energy consumption comparison, but this preliminary estimate demonstrates that the AAFEB process is competitive with conventional processes for the treatment of dilute organic wastes on an energy consumption basis. Again it should be kept in mind that the AAFEB will produce much less sludge than the conventional processes and therefore will need less energy for sludge handling.

7.3 Summary

In summary, the AAFEB process appears to have potential on an operational criteria basis, and energy consumption basis to be a secondary waste treatment process alternative. A more detailed analysis based on pilot scale or full scale operation is necessary to further analyze the potential and improve on this technology, and to provide a basis for capital cost comparison.

CHAPTER 8. CONCLUSIONS

Based on the results of this study, it can be concluded that:

1) The AAFEB process is a feasible organic carbon removal system for the treatment of low strength soluble organics.

2) The AAFEB process was shown to be a high efficiency process capable of achieving high organic removal percentages at low temperatures (10°C, 20°C) treating low strength wastes (COD $\leq$ 600 mg/l) at short hydraulic retention times (several hours) and at high organic loading rates greater than 4.8 kg/COD/m^3/day).

3) The unusually high effectiveness of the process is believed due to the large surface area to volume ratio created by the inert support media which enables a large active mass of attached microorganims to remain in the reactor at high liquid flow rates. Biomass concentrations exceeding 30 grams per liter were common in the AAFEB reactor. This enabled the AAFEB to maintain high SRT values.

4) A film process exists that is not limited by mass transfer because of the thin biofilms and absence of a stagnant liquid film layer due to the nature of the expanded bed.

5) The organic removal efficiency of the AAFEB is dependent upon the influent substrate concentration.

6) The efficiency of the AAFEB is shown to be dependent on temperature decreasing with decreasing temperature, but compensating in the decreasing efficiency did not follow a typical Arrhenius-Van Hoff relationship because of the low activation energy associated

with temperature dependence.

7) The fundamental process operational variables (net specific film growth rate) and (specific film substrate utilization rate) were found to be important process control variables.

8) For the bench scale reactor used in this study, the organic loading rate was shown to be a more significant operational variable than the hydraulic retention time.

9) Effluent suspended solids concentrations were found to be relatively low even at high hydraulic loadings.

10) Success of this demonstration warrants further investigation.

CHAPTER 9. SUGGESTIONS FOR FURTHER WORK

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

1) To look at the effects of diurnal fluctuation of temperature and influent substrate concentrations on process efficiency.

2) To examine the relationship between substrate removal and filter depth of the expanded bed (for a bed without recycle).

3) To further define the ability of the AAFEB as a methane producer using a substrate containing a more concentrated soluble waste.

4) To examine the efficiency of the AAFEB for treating particulate wastes.

5) To investigate other support media which are less dense and expensive than the aluminum oxides used in this study.

6) To determine the optimal design of the AAFEB including such factors as diameter to length ratio, depth of bed, size of clarification region, size of support particles, etc.

7) To conduct a pilot scale study of an AAFEB treating domestic sewage over a prolonged time period.

8) To conduct research to find ways of attaching viable organisms to support particles thus shortening start up time.

9) To evaluate the effect on effluent quality of generation organic matter of microbial origin in the expanded bed process.

10) The mechanisms of particulate and soluble carbon removal and relationships between the two should be further investigated.

11) A detailed kinetic model should be formulated and verified in order to understand the process in detail so that it can be improved even more and reach its potential.

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GLOSSARY

SRT	solids retention time (time)
HRT	hydraulic retention time (time)
BOD	biochemical oxygen demand (mass/volume)
AFEB	attached film expanded bed
AAFEB	anaerobic attached film expanded bed
COD	chemical oxygen demand (mass/volume)
Em	Minimum porosity for fluidization (dimensionless)
L	height of unexpanded fluidized bed (length)
L_E	height of expanded fluidized bed (length)
E_E	porosity of expanded bed (dimensionless)
E	porosity of unexpanded bed (dimensionless)
ρp	density of support particles (mass/volume)
S	cross sectional area of column ($length^2$)
ΔP	head loss in fluidized bed ($mass/length^2$)
N_{Re}	Reynold's number (dimensionless)
D_p	diameter of particles (length)
ρ	density of liquid (mass/volume)
$\bar{V}o$	superficial or empty tower velocity (length/time)
g_c	acceleration due to gravity ($mass/time^2$)
x_i	fraction of particles of size i (dimensionless)
μ'	absolute viscosity (mass/length/time)
g	gravitational acceleration constant ($mass/time^2$)
L_M	bed height at incipient fluidization (length)
$\bar{V}_{OM}$	minimum superficial velocity for fluidization (length/time)

V_{OL} limiting superficial velocity (length/time)

U_t terminal settling velocity of particle (length/time)

ϕs shape factor (dimensionless)

S substrate concentration (mass/volume)

k maximum substrate utilization coefficient (mass/time)

X concentration of active microorganisms (mass/volume)

K_s the "half velocity coefficient", the substrate concentration at which the rate of reaction is one half the maximum rate (mass/volume)

U specific utilization, the rate of removal of substrate per unit weight of microorganisms ($time^{-1}$)

Y growth yield coefficient (mass/mass)

b microorganisms decay coefficient ($time^{-1}$)

μ the net specific growth rate of microorganisms ($time^{-1}$)

μ_{MAX} maximum net specific growth rate of microorganisms ($time^{-1}$)

S_o influent substrate concentration (mass/volume)

S_e effluent substrate concentration (mass/volume)

K' substrate removal coefficient ($time^{-1}$)

A specific film utilization rate ($time^{-1}$)

B net specific film growth rate ($time^{-1}$)

B_{MAX} the maximum net specific film growth rate ($time^{-1}$)

r^2 correlation coefficient (dimensionless)

K_1 equal to B_{MAX} divided by K_s, kinetic coefficient (volume/mass/time)

K_2 equal to k divided by K_s, kinetic coefficient (volume/time)

$\bar{v}$ velocity of reaction

A' a constant

ΔH	the activation energy (energy/mass)
R	gas constant (1.99 calories/mole °K)
T	the absolute temperature (°K)
N_u	mass transfer Nusselt number (dimensionless)
S_c	Schmidt number (dimensionless)
$\bar{k}$	mass transfer coefficient (length/time)
v	kinematic viscosity of bulk fluid (length^2/time)
D_f	diffusivity of substrate in bulk fluid (length^2/time)
k_{LA}	overall mass transfer coefficient (time^{-1})
X_o	amount of substrate present at time = 0 (mass)
X_T	amount of substrate present at time = Z (mass)

APPENDICES

APPENDIX 1. ACCLIMATION STUDIES

In order to ascertain the effects of shock loadings on the AAFEB, batch runs with an influent substrate concentration two to three times the normal influent substrate concentration were performed on the AAFEB reactors and the time for recovery was monitored.

Figure 37 shows some typical results of these tests. In Figure 37, the fraction of substrate removed is plotted against time. The fraction of substrate removed is expressed as

$$X_o - X_T / X_T$$

where: X_o = amount of substrate present at time = 0 (i.e., beginning of run, mg COD)

X_T = amount of substrate present at time = T, mg COD

Time is expressed as hours after the beginning of the run.

It should be noted that the AAFEB reactors showed no difficulty in responding to influent substrate concentrations two to three times higher than their normal ones and within one hour the removal curves begin to level off. Also, the fraction of removal is generally lower than the 80% removal efficiencies experienced with continuous flows through the expanded beds. This may be explained by cellular products generated by the films in the system which accumulate in batch runs and are measured as COD, thus causing the removal fraction to appear to be lower.

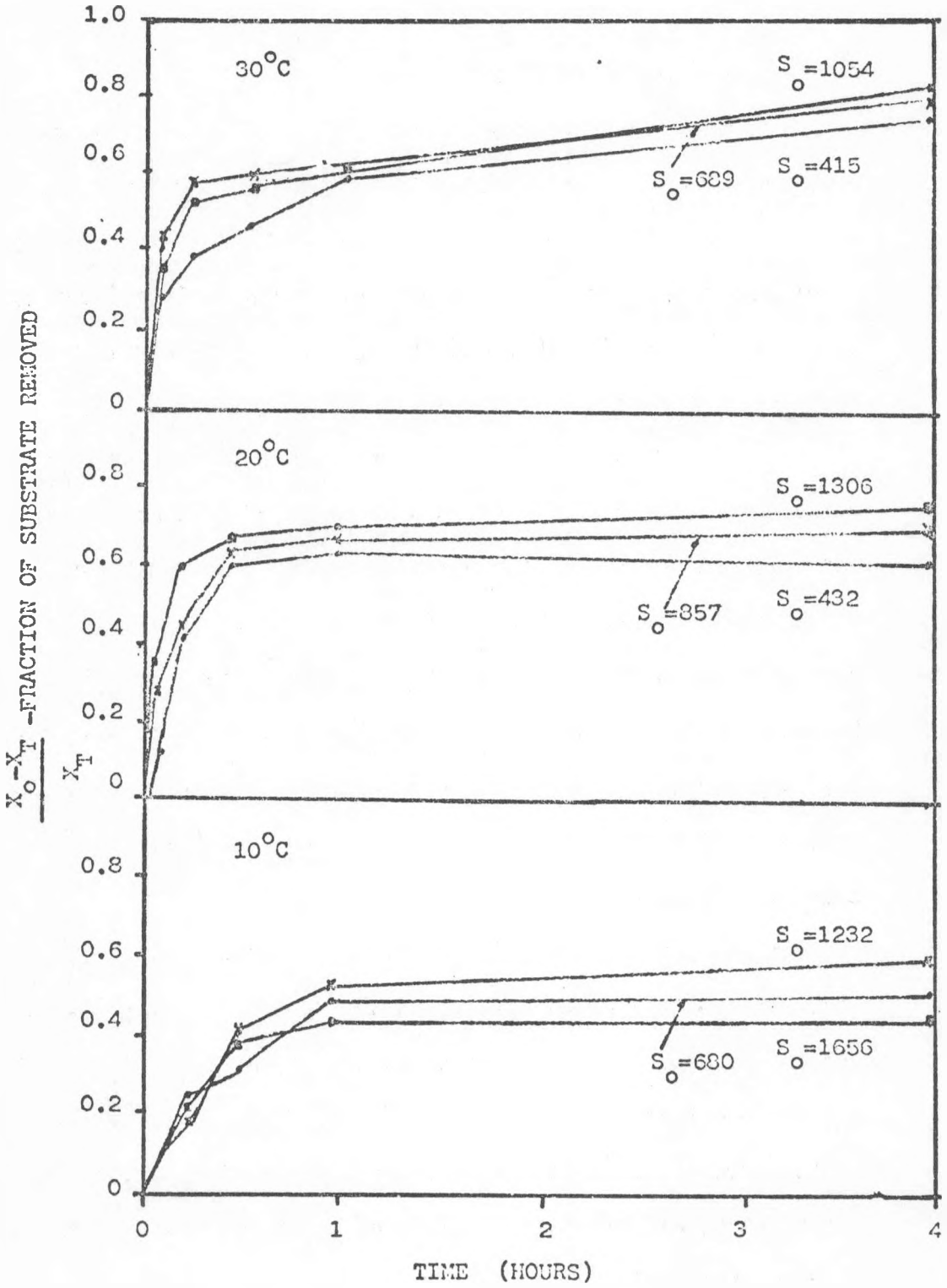

FIGURE 37. ACCLIMATION STUDIES-EFFECT OF SHOCK LOADINGS

APPENDIX 2. DILUTE STUDIES

The three AAFEB reactors used in the laboratory investigation were operated continuously at 200, 400, and 600 mg/l COD, as has been previously mentioned. In addition for each temperature studied, dilute waste side studies were performed with influent substrate concentrations of 50, 100, and 150 mg/l COD at hydraulic retention times of 6 hours and 0.33 hours.

Figures 38 and 39 show the relationship between organic removal efficiency expressed as percentage of initial COD removed and influent substrate concentration. Figure 38 shows the influence of temperature while Figure 39 shows the influence of hydraulic retention time. Figures 40 and 41 show the relationship between specific utilization A (mg COD removed per mg mass per day) and influent substrate concentration. Figure 40 shows the influence of temperature while Figure 41 shows the influence of hydraulic retention time.

In general COD removals are higher at higher temperatures and lower hydraulic loading rates, and with higher influent substrate concentrations. It should be noted that even at the lowest influent substrate concentrations of 50 mg/l COD, some removal still occurred. Higher specific utilization rates were observed at higher temperatues, higher hydraulic loading rates and with higher influent substrate concentration.

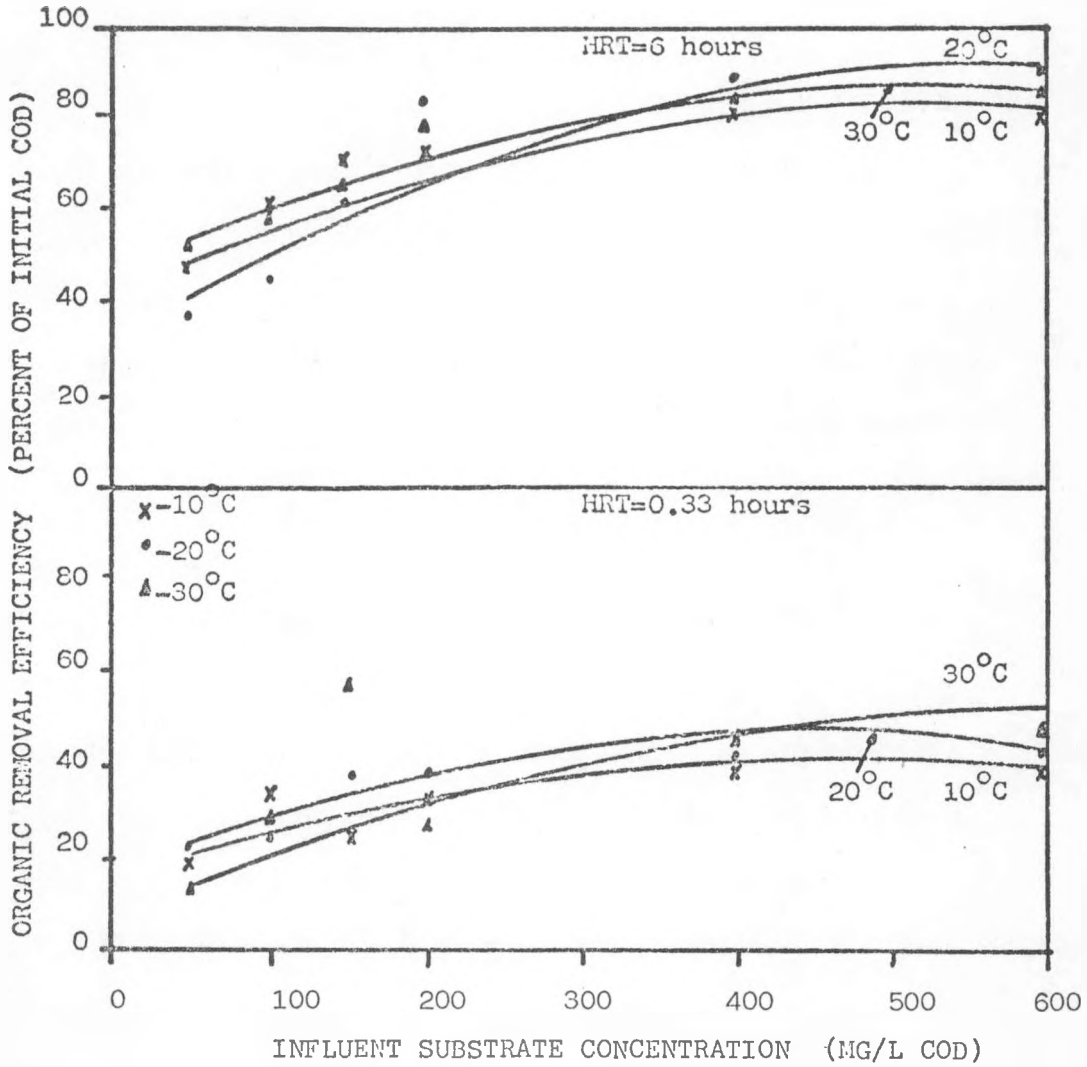

FIGURE 38. ORGANIC REMOVAL EFFICIENCY VS. INFLUENT SUBSTRATE CONCENTRATION-INFLUENCE OF TEMPERATURE

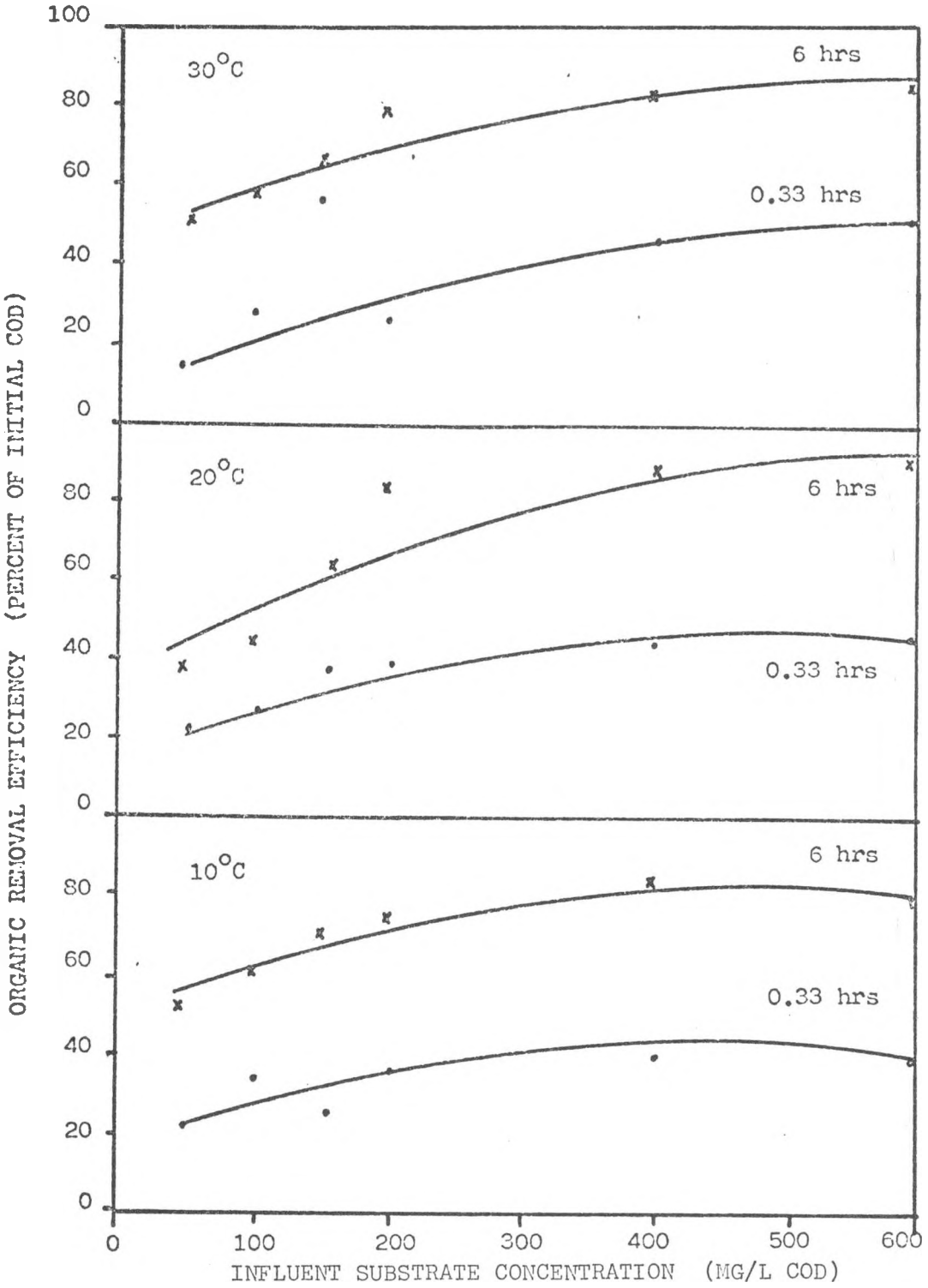

FIGURE 39. ORGANIC REMOVAL EFFICIENCY VS. INFLUENT SUBSTRATE CONCENTRATION-INFLUENCE OF HYDRAULIC RETENTION TIME

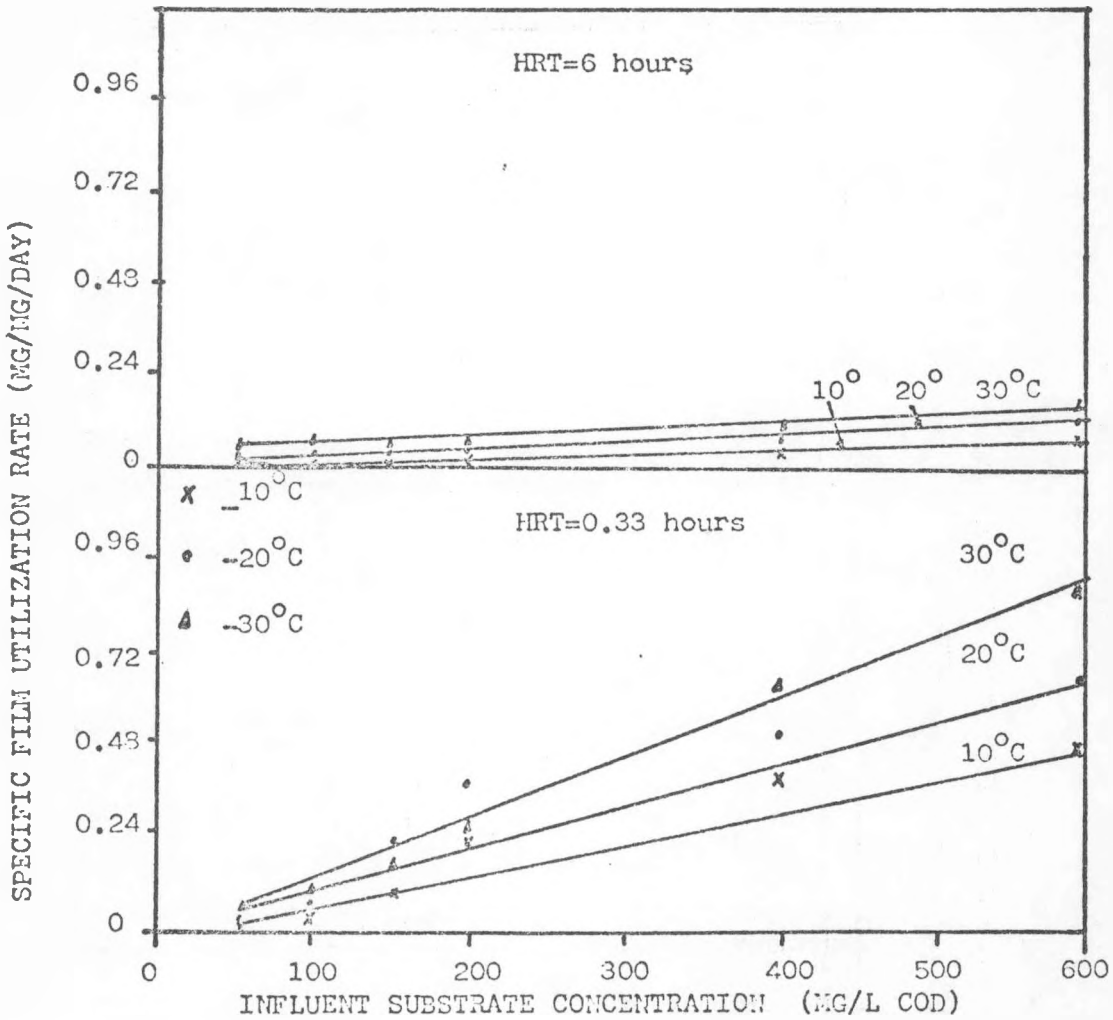

FIGURE 40. SPECIFIC FILM UTILIZATION RATE VS. INFLUENT SUBSTRATE CONCENTRATION-INFLUENCE OF TEMPERATURE

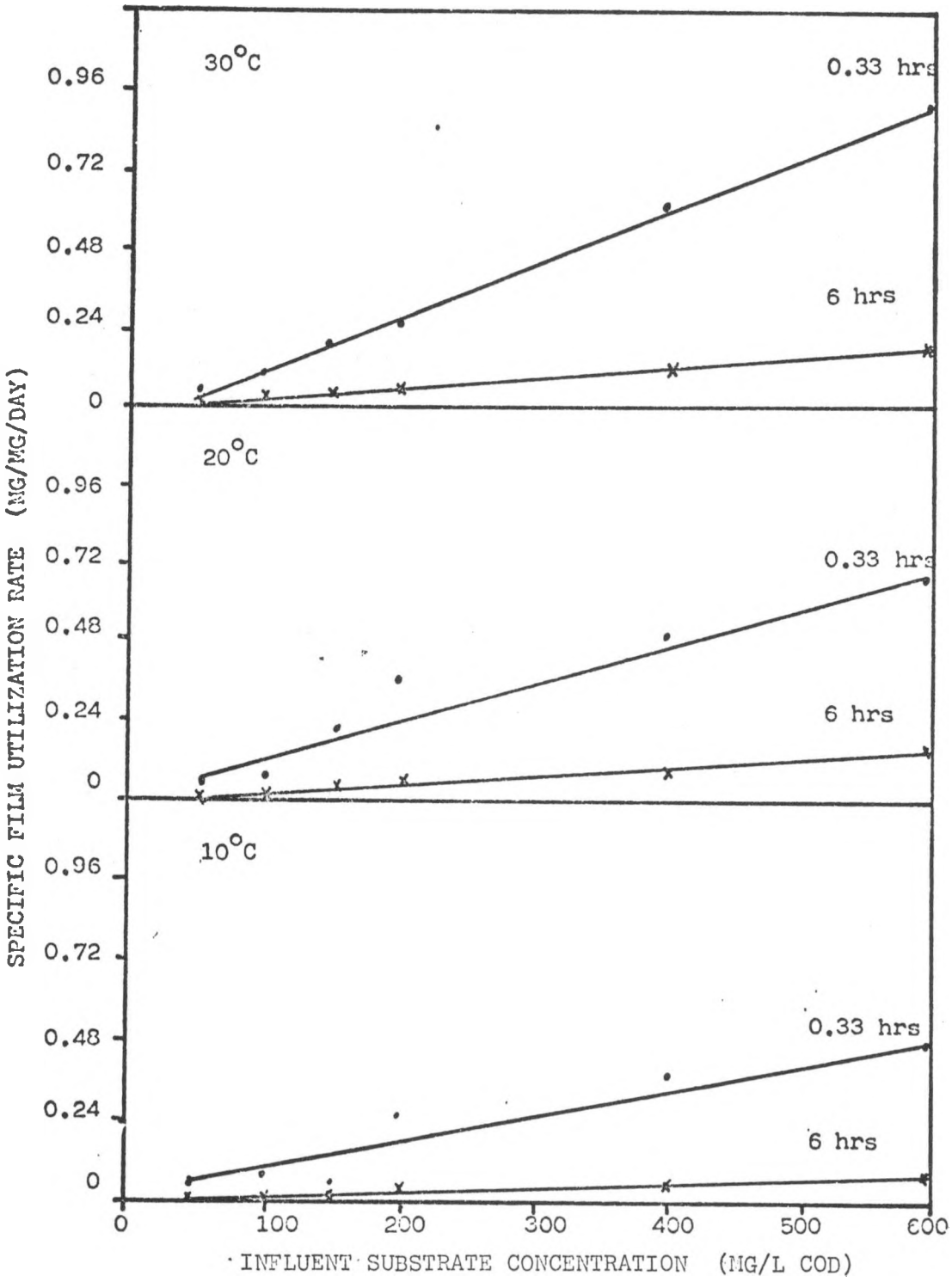

FIGURE 41. SPECIFIC FILM UTILIZATION RATE VS. INFLUENT SUBSTRATE CONCENTRATION-INFLUENCE OF HYDRAULIC RETENTION TIME

APPENDIX 3. SAMPLE ORGANIC CARBON BALANCE CALCULATION

Organic carbon balance equation:

Carbon in = soluble carbon in effluent + gaseous methane production + methane dissolved in effluent + effluent volatile suspended solids + accumulated carbon in the AAFEB (biomass).

Example: data for 30°C, HRT = 4 hrs, So = 200 mg/l COD

1. Carbon in: multiply influent substrate concentrations by flow rate 200 mg/l COD x 3 l/day = 600 mg COD/day.
2. Soluble carbon in effluent: multiply effluent substrate concentration by flow rate 56.0 mg/l COD x 3 l/day = 168.0 mg COD/day.
3. Gaseous methane production: multiply amount of gas produced by percent of CH_4; then convert to COD equivalent.

   133 ml gas/day x 59.6% CH_4 = 68.8 ml CH_4/day

   68.8 ml CH_4 = 174.3 mg COD (assume 1 gram COD = 393 ml CH_4 at 30°C) → 174.3 mg COD/day
4. Methane dissolved in effluent: multiply dissolved methane concentration by flow rate, assume effluent is saturated with dissolved methane at a concentration of 32.8 ml CH_4/l

   32.8 ml CH_4/l x 3 l/day = 98.4 ml CH_4/day

   98.4 ml CH_4/day = 249.9 mg COD/day
5. Effluent volatile suspended solids: multiply effluent VSS concentration by flow rate then convert to COD equivalent

   12.0 mg/l VSS x 3 l/day = 36.0 mg VSS/day

   Assume 1 mg solids = 1.42 mg COD

   36.0 mg VSS/day x 1.42 = 51.1 mg COD/day

6. Accumulated carbon in the AAFEB:

Difference between two successive biomass measurements, divided by number of days between the measurements; then convert to COD equivalent; at HRT = 6 hrs, So = 200 mg/l, biomass = 14600 mg/l TVS; at HRT = 4 hrs, So = 200 mg/l, biomass = 15400 TVS

Difference = 800 mg/l TVS

800 mg/l x 0.4 l (volume of particles) = 320 mg TVS accumulated

320 mg TVS ÷ 7 days = 45.7 mg TVS/day

45.7 x 1.42 = 64.9 COD/day

Total in = #1 = 600 mg COD/day

Total out = sum of #2 + #3 + #4 + #5 + #6

= 168 + 174.3 + 249.9 + 51.1 + 64.9 = 708.2

$$\text{Ration In/Out} = \frac{600}{708.2} = 0.847$$

APPENDIX 4. TIME SERIES PLOT OF REACTOR PERFORMANCE

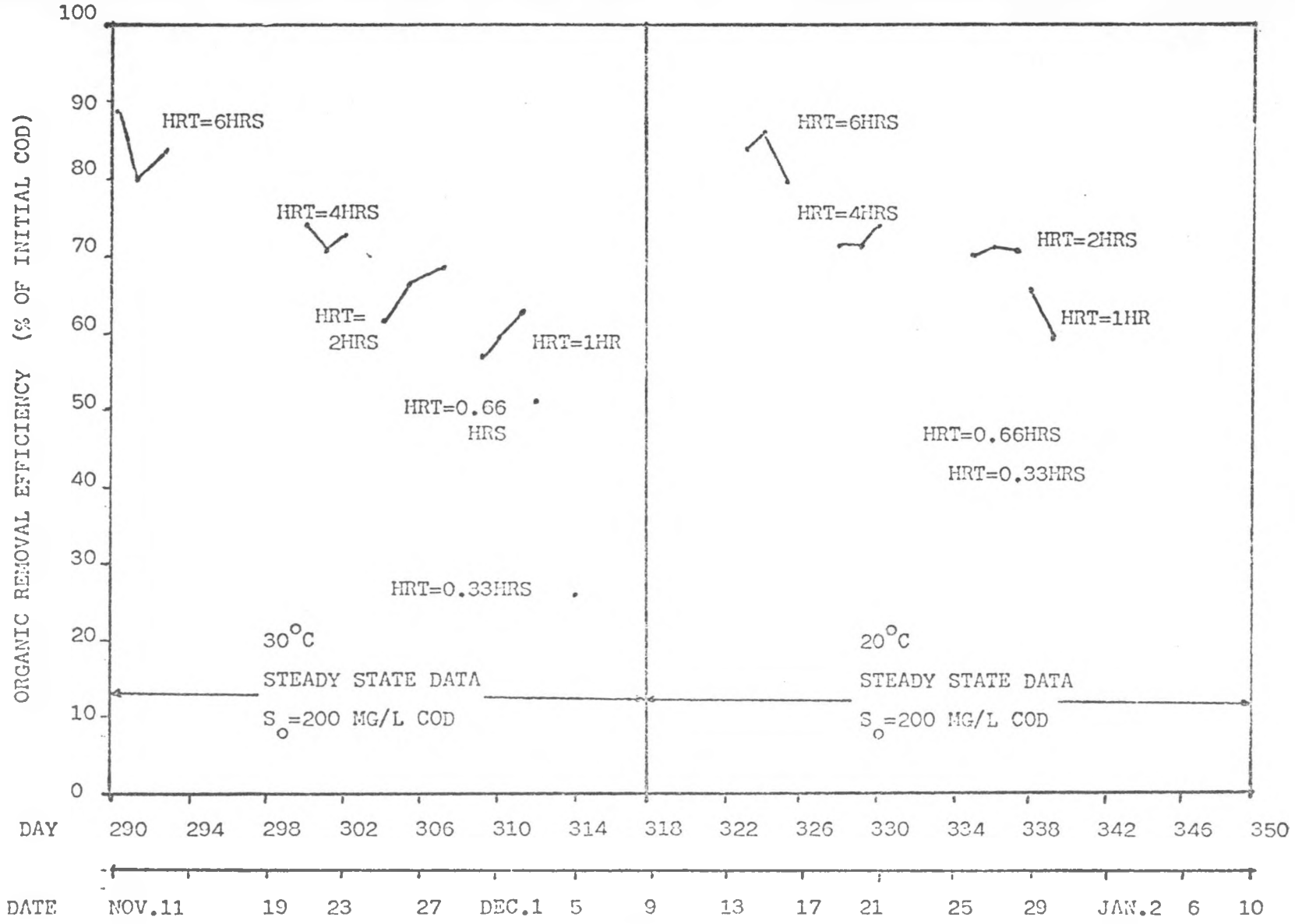

FIGURE 42. TIME SERIES PLOT OF REACTOR PERFORMANCE (REACTOR #1)

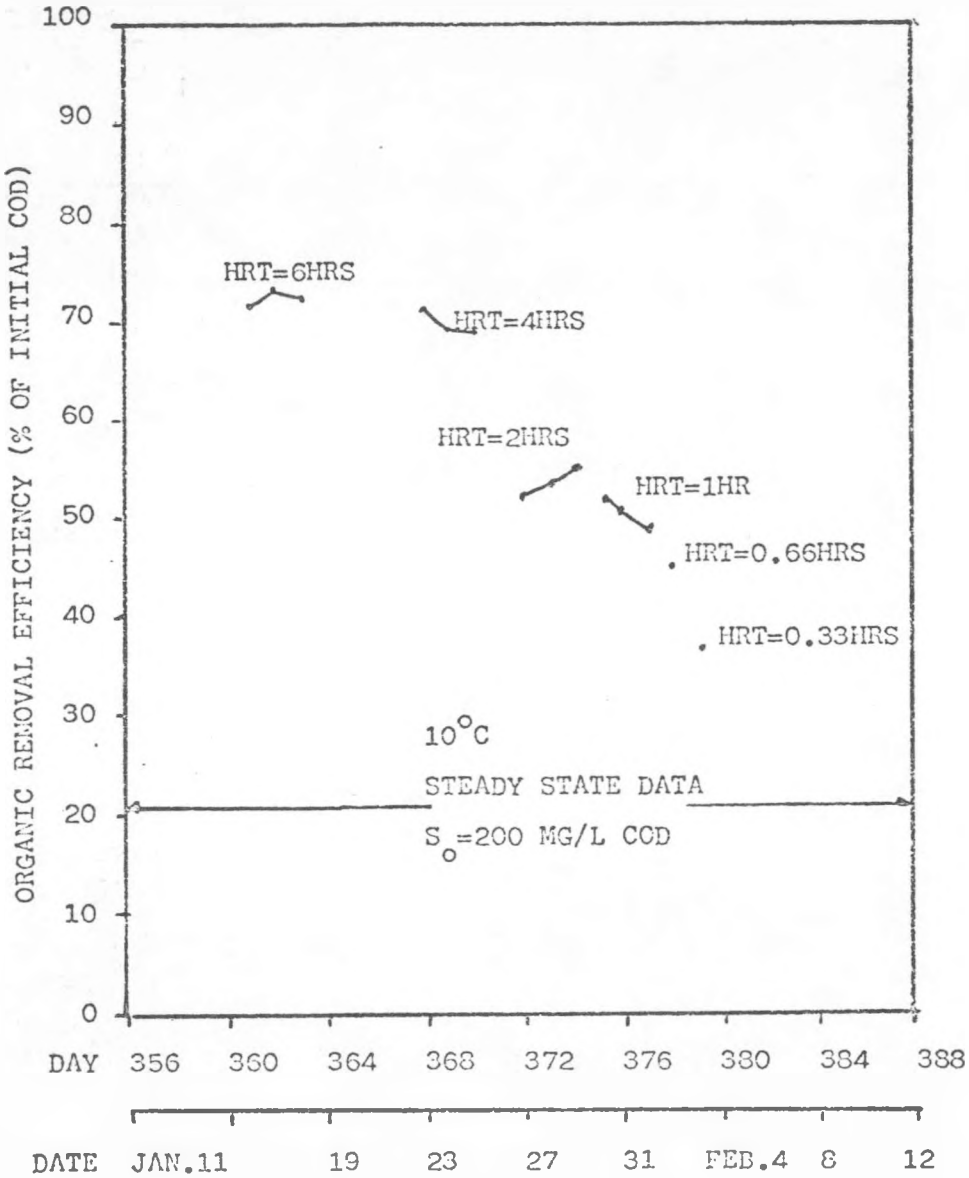

FIGURE 42. CONTINUED

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