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Design, laboratory verification, and mathematical modeling of an anaerobic system for the treatment of low-strength wastewater

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DESIGN, LABORATORY VERIFICATION, AND MATHEMATICAL MODELING OF AN ANAEROBIC SYSTEM FOR THE TREATMENT OF LOW STRENGTH WASTEWATER

BY

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A dissertation submitted to the Graduate Faculty in Engineering in partial fulfillment of the requirements for the degree of Doctor of Philosophy. The City University of New York.

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ABSTRACT

DESIGN, LABORATORY VERIFICATION, AND MATHEMATICAL MODELING OF AN ANAEROBIC SYSTEM FOR THE TREATMENT OF LOW STRENGTH WASTEWATER.

BY

Hassan E. Hussein

Advisor: Professor John Fillos

The goals of this research were (1) To develop a new modification of the anaerobic biofilter to treat low strength wastewater and experimentally investigate its effectiveness. The new modification has been designed to support both suspended and attached biological growth. (2) To develop a mathematical model that defines the dynamic behavior of the wastewater and the solids growth within the reactor. The model can be used to predict the process performance under different operating conditions. The model should also be helpful in the design of a full scale reactor. (3) To subject the model to a sensitivity analysis to identify the key parameters involved in the process, and to evaluate their impacts on the reactor as their values change.

The wastewater investigated has the short chain acetic fatty acid as its sole substrates. The acetic acid or acetate is probably the most important and rate limiting

methane precursor. The wastewater also contained sufficient minerals and vitamins needed by the fermentative bacteria.

The developed mathematical model uses Monod kinetics, the coefficients of which were collected from literature, or estimated either by calibration or based on theoretical considerations. The reactor performance was investigated over a temperature range from 10 to 30°C and an organic loading range of 0.65 to 0.95 Kg COD/m3/day, which are wide ranges for practical engineering applications. Two wastewater strengths (of 250 and 350 mg/L COD) were studied within the reactor temperature and organic loading ranges indicated above.

The experimental data suggest that the biofilter modification used is a successful application to the treatment of low strength soluble organic wastewater. The results showed that for temperatures above 20°C and organic loading below 0.85 Kg COD/m3/day the filter meets the NPDES requirements.

The mathematical model was found to be representative of the process. Sensitivity analysis indicates that, substrate utilization rate K, organic loading, influent substrate concentration So, temperature T, half-velocity coefficient Ks all have significant influence on process performance.

The experimental data suggest that the biofilter modi-

fication investigated is a significant improvement to the anaerobic biofilter.

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NOMENCLATURE

a	Specific surface area of media, area/volume
A	Cross sectional area of reactor, area
A_s	Surface area of the fixed media, area
b	Biomass decay rate, 1/time
COD	Chemical Oxygen Demand, mass/volume
C_g	Coefficient of methane production per unit of soluble COD removed, volume/mass
C_s	Percent of biosolids escaping the lower zone into the upper one, dimensionless
C^*	Variable order reaction coefficient, dimensionless
d	Diameter of a circle or a sphere, length
D	Molecular diffusivity of the substrate in the liquid, area/time
D_f	The molecular diffusion of substrates within the bio film, area/time
D_f^*	Dimensionless molecular diffusivity in biofilm
dH	Differential height, length
D_H	Hydrodynamic Dispersion Coefficient, area/time
dV	Differential volume, volume
E	Fraction of methane in biogas produced, a decimal
ϵ	Porosity of the packing media, percentage
f	Bypass flow fraction, dimensionless
F_{ac}	Biofilm mass density factor, dimensionless
F_s	Coefficient of sloughing for attached biomass, 1/time
G_g	Coefficient of methane produced per unit soluble COD removed, volume/mass
G_r	Total gas production rate, volume/time
H	Height, length

J	Flux of substrate into the biofilm surface, mass/time
K_s	Half-velocity coefficient. Concentration of substrate, at which substrate removal rate is one half the maximum substrate removal rate, mass/volume
K	Maximum specific substrate removal rate, 1/time
K_m	Liquid mass transfer coefficient, length/time
K_v	Fractional change in void volume due to biomass growth, dimensionless
l	Characteristic length, length
L	Depth of the effective diffusion layer, length
L^*	Diffusion-layer depth, dimensionless
L_f	Biofilm depth or thickness, length
m_g	Number of gas moles
m_w	Number of water moles
P	Coefficient of biomass change in the upper zone due to settling by gravity or lifting by biogas, 1/time.
P_e	Peclet number, dimensionless
P_r	General parameter symbol, dimension vary
q	Biogas flow rate at any reactor height, volume/time
Q	Volumetric flowrate, volume/time
Q_{bp}	Bypass flowrate to sludge blanket, volume/time
Q_o	Influent flowrate, volume/time
R_e	Reynolds number, dimensionless
R_s	Fractional change in void volume due to biogas production, dimensionless
S_b	Substrate concentration at biofilm surface, mass/volume
S_{bd}	Substrate concentration in the sludge bed, mass/volume
S_{bk}	Substrate concentration in the sludge blanket, mass/volume
S_c	Schmidt number, dimensionless

S_e Substrate concentration in effluent, mass/volume

S_f Substrate concentration in biofilm, mass/volume

S_i Substrate concentration at the bottom of active biofilm, mass/volume

S_{min} minimum concentration of substrate that is able to sustain a steady-state biofilm, mass/volume

S_o Substrate concentration in the influent, mass/volume

S_s Substrate concentration at the interface between the liquid film and the biofilm, mass/volume

S_u Substrate concentration in the upper zone, mass/volume

S_u^* Dimensionless bulk liquid substrate concentration

S_{se} The sensitivity of the Chemical Oxygen Demand Removal in the effluent, dimensionless

T Temperature, ^oC or ^oK

û Dynamic viscosity of the bulk liquid, mass/(length*time)

v Velocity, length/time

V_{bd} Volume of sludge bed, volume

V_{bk} Volume of sludge blanket, volume

V_L Volume of lower zone, volume

V_u Liquid volume of upper zone, volume

V_r Total liquid volume of reactor, volume

X_{bd} Concentration of suspended biomass in sludge bed, mass/volume

X_e Concentration of suspended biomass in the effluent, mass/volume

X_f Mass density of the attached biofilm, mass/volume

X_s Concentration of suspended biomass, mass/volume

X_{bk} Concentration of suspended biomass, in sludge blanket, mass/volume

θ_{bd} Hydraulic retention time in sludge bed, time

Θ_{bk} Hydraulic retention time in sludge blanket, time

Θ_l Hydraulic retention time in lower zone, time.

Z_g Equilibrium mole fraction of the dissolved gas, dimensionless

Greek symbols

Ω Variable order reaction order, dimensionless

β Logarithm of an adjusted dimensionless substrate concentration

τ Standard biofilm depth, length

σ Mass density of the bulk liquid, mass/volume

Φ Kinematic viscosity, area/time

μ Specific substrate removal rate, 1/time

CHAPTER ONE

INTRODUCTION

Anaerobic processes for wastewater treatment have gained considerable recognition in recent years. In anaerobic systems stabilization of organic matters is accomplished biologically using a variety of microorganisms of which the most prominent are bacteria. The microorganisms convert the colloidal and dissolved carbonaceous organic matter into various gases and into cell tissue. In the biochemical decomposition of organic matter basically three groups of bacteria work in harmony to accomplish the destruction of organic matter. First, hydrolyzing and fermenting bacteria of a wide variety convert the complex organic materials into fatty acids, alcohols, carbon dioxide, and small amounts of hydrogen. Next, a group of hydrogen producing acetonic bacteria transform the lower molecular weight compounds formed by the first group to hydrogen, carbon dioxide, and acetate. Finally, the products of the second group are transformed to methane and carbon dioxide. This is accomplished by two physiologically different groups of bacteria. One group uses hydrogen to reduce carbon dioxide to methane, and the other decarboxylates acetate to methane and carbon dioxide (see figure 1.1, after Klass (27)). For a steady-state condition, active populations from each one of the various bacteria groups must be present to utilize the organic material introduced into the reactor as well as the

products of each other.

Historically, anaerobic processes have been primarily used in the treatment of high strength wastewaters, and domestic wastewater sludges. Wastewater sludges are in general composed of raw solids removed in the primary sedimentation tank and excess biomass generated in the activated sludge process. Conventional anaerobic processes require long detention times for waste stabilization and usually have been uneconomical for the treatment of wastes containing less then approximately one percent (approximately 10,000 mg/L) biologically degradable organic material.

Recently, a modification of the anaerobic treatment process "The Anaerobic Biofilter" or, "ANBIOF" for short, has been found to be effective for the treatment of both medium and high strength soluble organic wastes(59). The anaerobic filter consists of a column filled with a solid media on whose surface bacteria grow. The waste usually flows up-ward through the column, contacting with bacteria attached to the medium and/or accumulated in the void spaces of the filter.

The carbonaceous organic matter from the wastewater is absorbed by the bacteria and stabilized through the three stages of anaerobic decomposition mentioned earlier, with a minimal amount of biomass production. Excess sludge is removed from the bottom of the reactor. The significantly

reduced production of sludge solids, when compared to aerobic treatment, represents a substantial benefit in terms of sludge solids for ultimate disposal. Furthermore, since the anaerobic filter processes are independent of oxygen transfer requirements and limitations, higher organic loadings are possible. Additional benefits are gained from the production of usable methane gas as a by product of a well operated anaerobic system. The increasing costs of energy and waste disposal over the past decade led to numerous attempts and claims for improvements and modifications of the anaerobic filter system for wastewater treatment. Yet, the number of "ANBIOF" applications is relatively small since design criteria have not been completely developed to a confidence level accepted by practicing engineers. Nevertheless, the available information can lead to several generalizations. These generalizations are summarized here as follows:

1- Type of waste:- widely differing wastes can be treated, and undoubtedly more types of waste will be tested.

2- Waste strength:- the anaerobic filter can be designed to treat wastes as low as 500 mg/L COD and produce high removal rates (90%).

3- Organic Loading rates: - a loading rate of 8 - 10 Kg COD/m^3/day (500 to 625 lb COD/ft^3/day) is probably the limit compared to 50-150 COD lb/1000 ft^3/day for aerobic activated sludge or aerobic trickling filter systems.

4- Treatment efficiency:- the efficiency of an anaerobic filter, based on COD removal, is slightly lower than that for aerobic systems (2 to 10% less).

5- Temperature:- the anaerobic filter process has been operated from 10°C to over 60°C

6- Nutrient requirements:- minimal nutrient requirements compared with aerobic treatment.

7- Tolerance to toxicants:- the filter was shown to be capable of treating many types of industrial wastewaters which contain toxic materials that were heretofore thought to be untreatable by methane fermentation. For example phenolic wastes up to a concentration of approximately 700 mg/L have been well treated anaerobically.

8- Start up period:- the start up period is one of the main limitations in the use of the anaerobic biofilter for wastewater treatment especially when toxic materials are expected in the wastewater. The start up process may also be of concern where toxic shock acts as a bacteriocide on the microorganisms inside the filter. The start up process may be affected by the following factors:

   a - Size, concentration and quality of the inoculum.

   b - Adaptation of inoculum to waste.

   c - The gradual increase in loading rate until the design loading is reached.

   d - Uniformity of the type of organic load being applied.

9- Upper influent concentration:- There are no upper influent organic concentration limitations reported.

10- Sludge production:- a very low excess sludge production is to be expected. Usually 0.02-0.1 gm of VSS are produced per gm COD removed. The low nutrient requirements experienced are usually due to low excess sludge growth.

11- Solid Retention Time (SRT):- the SRT, defined as the average time the individual cell spends inside the reactor before it is washed out, is usually in excess of 30 to 40 days. Such long retention times are easily achievable when using attached growth type reactors.

12- Attention required:- less operating effort required when compared to conventional aerobic and completely mixed anaerobic processes.

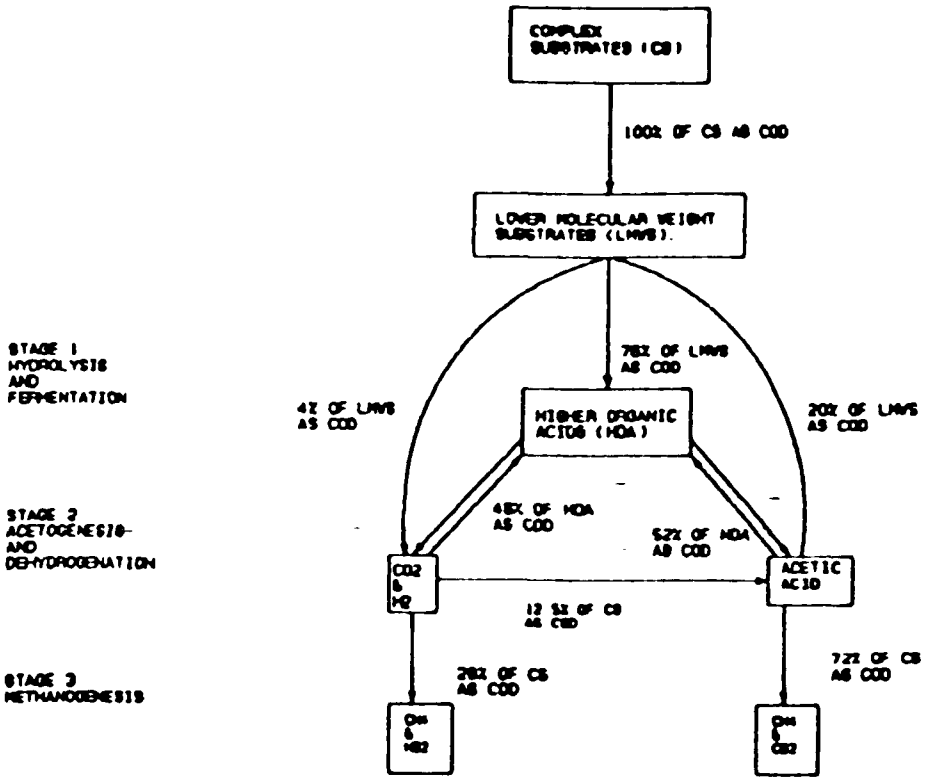
13- Lower influent concentrations:- Anaerobic filter

> systems have been shown to produce effluents of unacceptable quality, as required by EPA's Secondary Wastewater treatment Standards, with influent waste concentrations less than 500 mg/L. The main reason for this appears to be the washout of the biomass due to the high upward flow velocity which usually results when the applied organic loading reaches values similar to those applied to aerobic systems.

The major factor affecting the process efficiency is the Solid Retention Time (SRT). Unfortunately the role that "SRT" plays in process effectiveness was not adequately recognized by early researchers. The adequate control of SRT can overcome most of the process limitations, and in particular the treatment of low strength wastewaters such as domestic wastes.

The importance of SRT was recognized when the "Anaerobic Up-Flow Fixed Film system" or "ANFLOW", which is a modification of the original anaerobic filter, was developed. However, the ANFLOW system did not achieve acceptable removal efficiencies or effective biomass retention. Therefore, there is a need to modify the ANFLOW system to effectively control the SRT, while allowing adequate contact between the wastewater flow and the biomass and providing sufficient storage for biomass accumulation within the system. The proposed system, which is described in the next section, is designed to provide the requisite capabilities and operational flexibility.

FIGURE 1.1:
ROUTE TO METHANE BY ANAEROBIC FERMENTATION OF COMPLEX ORGANIC MATERIALS.



THE PERCENT CONVERSION IN EACH STEP IS FROM ZEHNDER ET AL.(81), EXCEPT THAT FATTY ACID CONVERSION TO HYDROGEN AND CARBON DIOXIDE WAS REVISED FROM 24 TO 48% AFTER MCCARTY(71).

Technology Description

In this section a brief description of the ANBIOF, and ANFLOW reactors along with the proposed reactor will be presented. The ANFLOW is a basic modification to the ANBIOF reactor that sought to increase the SRT.

1 - The Anaerobic Biofilter (ANBIOF)

A schematic diagram of a typical anaerobic filter is shown in figure 1.2 on page 29. The wastewater is introduced at the bottom of the unit and is distributed over the cross-section area by means of a perforated plate. The reactor contains porous filter media which is usually made of stone or plastic.

The tendency for the wastewater to short-circuit the filter media, at the walls, is usually minimized by placing dispersion rings along the depth of the filter. The number and the spacing of these rings depends on the size of the filter used, the type of media employed, as well as, the hydraulic loading applied. The treated wastewater discharges from the reactor through a U - tube that controls the liquid level in the tank and provides a seal preventing the gas from escaping. The gas produced during the treatment is collected and subsequently removed from the top of the reactor.

2 - The Anaerobic Up-Flow Fixed Film (ANFLOW)

A schematic diagram of a typical ANFLOW is shown in figure 1.3 on page 29. Basically, the ANFLOW reactor is an anaerobic biofilter that has a cone shaped zone at the bottom of the reactor to provide additional room for solid storage or entrapment for controlled wasting from the reactor.

3- The Employed Hybrid Reactor(Two Zone, Two Growth, Anaerobic Biofilter(TZ-TG-ANBIOF)

The system used in this study is a significant advancement of the basic design and operation of the ANFLOW reactor (see figure 1.4 on page 30). The proposed hybrid anaerobic reactor used is composed of two zones. The lower zone, approximately one fourth of the total reactor volume, will be the first stage of reaction and will contain completely mixed fluidized sludges. The upper zone is principally a combination suspended and attached film growth zone. It is filled with plastic media having high specific surface area and porosity. The biomass in the lower zone is distributed throughout a suspended sludge bed and suspended sludge blanket. The sludge bed is defined as the layer within which the concentration of solids is high (approximately 20 gm SS/L) and this concentration does not vary within the range of operating conditions.

The lower zone is designed to function as a fluidized

zone for three reasons:

1 - The first stage breakdown of most of the organic matter in the wastewater, to fatty acids and acetates by acid forming bacteria, and the initiation of fermentation by methane producing bacteria, are expected to occur in the lower foot or two of the anaerobic filters. Hence, complete fluidization, and the consequent mixing in this lower zone, would aid in initiating the fermentation of the volatile acids by methane formers, because as soon as the organic matter are broken down to the lower fatty acids they are placed in contact with the methane formers. In addition, the mixing prevents acid pockets developing and reduces conditions where the pH decreases to levels that can be harmful to the microorganisms.

2 - The complete fluidization in the lower zone provides a large mass that should attenuate any effects of toxic shocks to the reactor. Toxicity effects may be a factor affecting the performance of anaerobic biofilters, and limit their use and application.

3 - The lower part serves as a storage volume for excess solids to accumulate and also a convenient means of removing the digesting and digested solids from the reactor.

There is no need for special mechanical mixing devices

in the lower part of the reactor since the diameters of the solids are very small, less than 0.5 mm. These solids are completely fluidized by the upward velocities of the incoming wastewater at values as low as 0.5 ft/hr (0.25 cm/min). This velocity is much lower than the flow velocity expected inside the lower and upper zones of the reactor. This last effect (i.e. fluidization velocity vs. particle diameter) has been studied by the United States Environmental Protection Agency (64) and some of the results are shown in figure 1.5.

The upper zone of the reactor is filled with plastic media having a void ratio of approximately 87% . The plastic media will provide large surface areas for the bacteria to form an attached biological slime. The high porosity of the media will also provide considerable pore volume storage inside the second zone of the reactor. Consequently a higher actual Hydraulic Retention Time (HRT) and a much lower throughput velocity would be obtained for the same flow rate with this media than with a stone media. The higher HRT attained will increase the contact time between the organic waste and attached biological slime, improving the overall process efficiency. Further, the lower up-ward velocities that result from greater pore volumes and higher HRT, will help minimize washout of suspended solids. Consequently this should increase the SRT. Higher SRT should essentially result in even higher treatment efficiencies.

FIGURE 1.2:
SCHEMATIC DIAGRAM OF THE
ANAEROBIC BIOFILTER (ANBIOF).

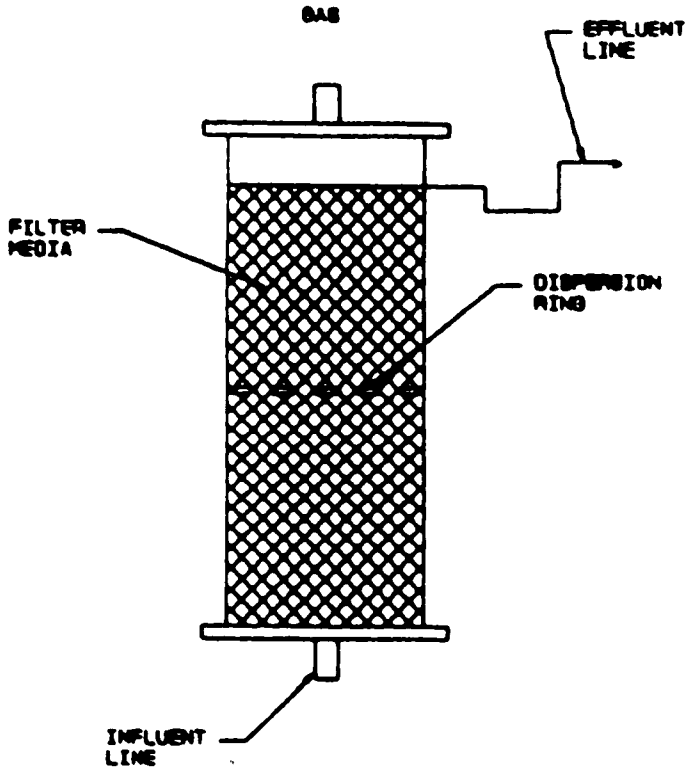


FIGURE 1.3:
SCHEMATIC DIAGRAM OF AN
ANFLOW REACTOR.

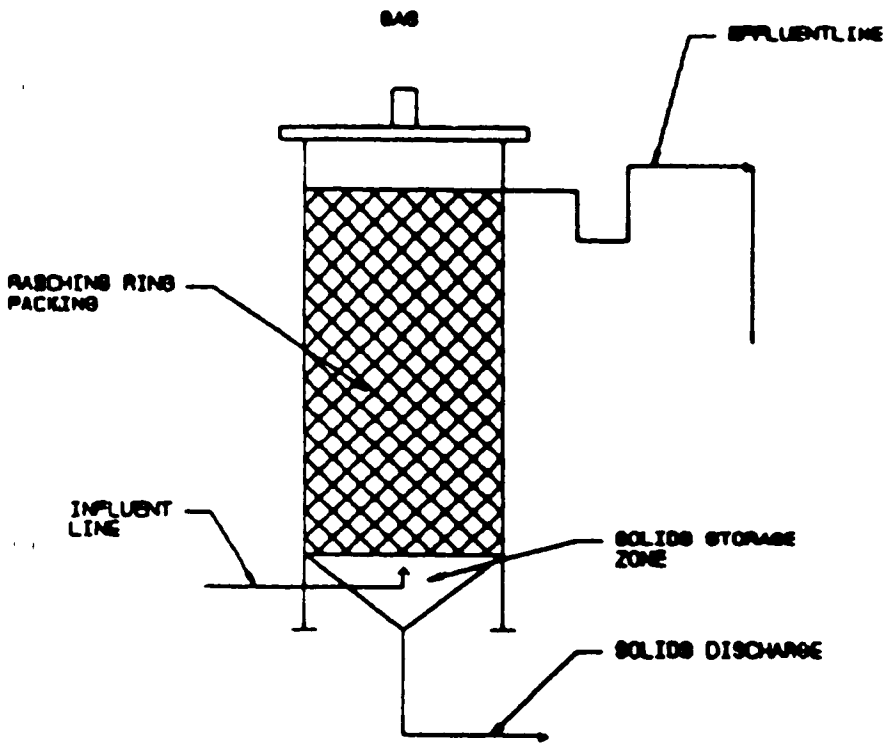
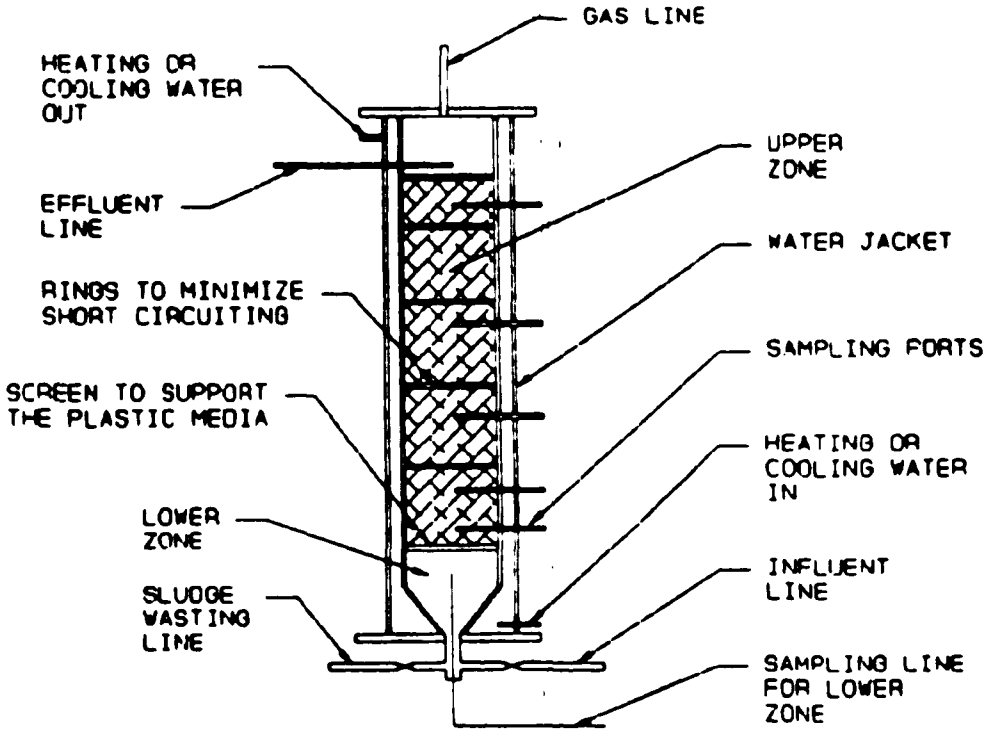


FIGURE 1.4:
SCHEMATIC DIAGRAM OF THE ANAEROBIC REACTOR USED IN THIS STUDY.



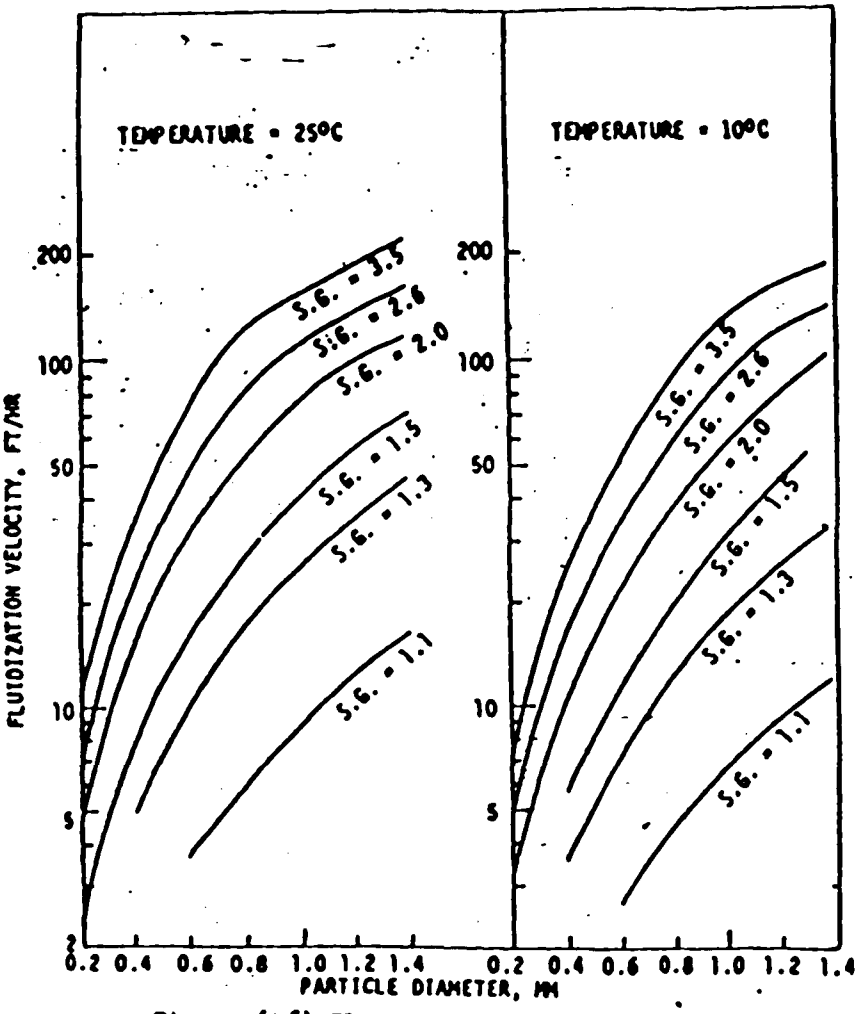


Figure (1-5) Fluidization velocity vs. Particle Diameter at different Specific Gravities.(18)

CHAPTER TWO

BACKGROUND INFORMATION

This chapter presents an overview of information, gathered from the literature, which is relevant in comprehending the various stages of anaerobic (methane) fermentation, anaerobic microbiology, kinetics of anaerobic biogrowth as well as anaerobic processes, and their mathematical modeling.

1) Stages of Methane Fermentation, and the Microbiology of Anaerobic Growth.

Information accumulated in the literature about methane fermentation indicates that at least three physiologically unique groups of bacteria are involved. In the first stage complex organic material (e.g. polysaccharides, proteins, lipids, carbohydrates) are hydrolyzed by fermentive bacteria to intermediate organics such as higher fatty and organic acids, hydrogen, carbon dioxide, and other lower molecular weight compounds. In this stage virtually no reduction in organic COD will occur.

In the second stage hydrogen-producing acetogenic bacteria convert the products of the first group into hydrogen, carbon dioxide, and acetate.

In the third stage the products of the second stage are converted by two different groups of methane forming bacte-

ria to methane, carbon dioxide, and water. Methanogenic bacteria yield methane and carbon dioxide from acetate by decarboxylation. Other methanogenic bacteria produce methane and water from hydrogen and carbon dioxide reduction. In this multistep digestion process (see figure 1.1 on page 24) the main volatile acid intermediates found during methane fermentation were acetic and propionic acids, and to a smaller extent, amounts of butyric, valeric, and caproic volatile acids can also be found. It has been reported by Jeris & McCarthy (22) that acetic and propionic acids are the precursors of approximately 85 percent of the methane produced from the digestion of the complex organic wastes. The methanogens are a distinctive group of bacteria. They are strict anaerobes. Naturally, they are only found in completely anaerobic environments like the bottom of lakes, rumen of cattle and the wet wood of trees (61). Early taxonomic studies included methanogens in one family called "methanobacteriaceae", which is divided into three genera on the basis of cell morphology. Recent studies of the methane bacteria (61, 62) have led microbiologists (56) to propose a classification as a third form of life more simple than the procaryotes. Methanogens differ from other bacteria in at least three major areas: (1) Their cell walls contain no muramic acids. (2) Their metabolism differs in many respects. For example, the methane bacteria have not been shown to contain the usual components of an electron transport chain that could be coupled to ATP formation by oxida-

tive phosphorylation using carbon dioxide as terminal electron acceptor for anaerobic respiration, and (3) their ribosomal RNA is different. Recently, Woese (57), proposed three taxonomic kingdoms: (1) eukaryotes, (2) eubacteria, and (3) archaebacteria.

Eubacteria and archaebacteria both are prokaryotes that lack nuclei and have simple cells. Woese also concluded that the three kingdoms proposed above have a common predecessor, a very simple cell, the progenate. Along with halophiles and thermoacidaphiles, methanogens belong to the proposed archaebacteria kingdom.

Seven species of methanogenic bacteria that have been observed to utilize acetate are reported in literature along with two species that utilize propionate and fourteen that use hydrogen (60, 61, 62, 63), see table 2.1.

TABLE 2.1: ACETATE, HYDROGEN, AND PROPIONATE UTILIZING BACTERIA, AFTER MONTGOMERY(35).

Acetate Utilizing Bacteria	Hydrogen Utilizing Bacteria	Propionate Utilizing Bacteria
1) Methanobrevibactor Arboriphilus	Methanobacterium Formicicum	Methanobacterium Propionicum
2) Methanobrevibactor Ruminantium	Methanobacterium Bryantii	Syntrophobacter Wolinii
3) Methanothrix Soehngenii	Methanobacterium Thermoautotrophicum	
4) Methanococcus Mazei	Methanobrevibactor Ruminantium	
5) Methanosarcina Barkeri	Methanobrevibactor Arboriphilus	
6) Methanobacterium Thermoautotrophicum	Methanobrevibactor Smithii	
7) Methanogenium Cariaci	Methanococcus Mazei	
8)	Methanococcus Vannielii	
9)	Methancoccus Voltae	
10)	Methanomicrobium Mobile	
11)	Methanogenium Cariaci	
12)	Methanogenium Marisnigri	
13)	Methanospirillium Hungatei	
14)	Methanosarcina Bakeri	

2) Kinetics of Anaerobic Biogrowth and Mathematical Modeling of Anaerobic Reactors

Environmental scientists, engineers, and microbiologists have developed a variety of mathematical expressions that describe various anaerobic processes (4, 9, 12, 35, 39, 40, 41). Mathematical models of anaerobic systems are very useful because they allow one to assess the effect on the performance and to anticipate the efficiency of the process considered, due to the changes in the various variables and biological constants involved in the process (e.g. temperature, biomass density, etc). The main objective in formulating a mathematical expression is to describe the rates of substrate utilization and biomass accumulation in the biological reactors as well as overall removal efficiency of the substrates. Such models are useful in understanding the mechanisms involved in biological growth. They are also employed in modern concepts of design and operation of biological wastewater treatment systems. This section presents a brief history of research efforts made to develop mathematical descriptions to several anaerobic biological systems. In addition, it will present a review of certain studies conducted to determine the growth kinetic parameter values of the microorganisms that utilize volatile fatty acids, and in particular the acetic and propionic acids.

Mathematical expressions that describe any anaerobic treatment process may be formulated mathematically if the

following are known:

a) The fluid flow pattern in the reactor

b) The distribution and behavior of sludge in the reactor and,

c) The kinetics of the biological conversion of substrates and subsequent formation of methane. Conversion of biological substrates is based on three relationships 1) biological growth rate, 2) biological growth yield, and 3) rate of consumption of essential nutrients.

3) Fluid flow patterns, and distribution and behavior of sludge in the reactor.

Several investigators studied the anaerobic filter process to determine their flow behavior and patterns. In developing a model for an up flow fixed media anaerobic filter, Young (58) assumed that an ideal plug flow exists in the reactor. To verify his assumption, Young conducted tracer studies and concluded that in the absence of mixing and channeling due to biogas, ideal plug-flow regime will prevail in the filter. Young, also, presented a mathematical formula to express reduced effective volume of reactor due to gas mixing which result in short-circuiting. Young's formula is as follows:

$$V_e = \epsilon\ V_o\ (1 - K_v\ M_T)\ (1 - r_s\ q) \qquad [2.1]$$

Where,

V_e = Reduced effective volume of the reactor, volume.

V_o = Initial empty-bed reactor volume, volume.

ϵ = porosity of packing media, dimensionless

K_v = Fractional change in void volume due to biomass growth, dimensionless.

M_T = Total biomass concentration, mass/volume

r_s = Fractional change in void volume due to biogas production, dimensionless.

q = Biogas flow rate at any reactor height, volume/time.

Equation 2.1 shows that for high bed porosity and under low gas flow rate the flow regime can be adequately modeled as a plug-flow.

Dague and Chiang (6), based on a theoretical mathematical model, stated that the physical characteristics of the packing media and the intensity of gas production, which is related to the process loading, are significant in terms of the intensity of mixing in the upflow fixed film anaerobic biofilter. They also added that this mixing affects both substrate and biomass distribution within the reactor and, in turn the kinetics of the process. Kuijvenhoven and Heertjes (16) studied the flow pattern in 30 and 200 cubic meter prototype completely mixed upflow anaerobic reactors. They stated that the fluid flow pattern in the two reactors were to a large extent analogous. For the pilot plant, three consecutive fully mixed zones appeared to be appropriate while for the prototype two a mixed zones model has been found. The lowest zone is called the sludge bed and contains most of the sludge solids. The following two zones called

the sludge blankets and contain much lower suspended solids concentrations than that which exists in the sludge bed. The solids concentration in the sludge bed was found to be constant over a wide range of process operating variables such as temperature and flowrate. On the other hand the solids concentration in the sludge blankets is very susceptible to operating condition. The difference between the flow patterns in the two reactors was attributed to the difference in the heights of the respective sludge beds. In the reactor: two to three meter in the pilot plant with only 0.4 m in the prototype reactor resulting in too small an amount of sludge. They recommended that suspended growth upflow reactors for anaerobic treatment of wastewater should contain enough sludge that the height of the sludge bed should be 1.5 - 2.5 m. At a bed height of 0.4 m most of the effluent bypasses the bed. The literature referred to above substantiates the assumptions made in the model introduced in this research. Namely, it is assumed in this model that the lower zone is completely mixed, and the upper zone, which contains the fixed media, is a plug flow.

Hudson, et al (19) studied the effects of packing media on the performance of upflow fixed film anaerobic reactors. Two bench-scale units, identical in size but different in the media they contained, were used to treat shellfish processing wastewater. One reactor was randomly packed with granite and the other with whole oyster shells. The porosity of the stone media was 53% and that of the oyster shells was

82% . Hudson, et al reported that a higher Chemical Oxygen Demand (COD) removal of 81% was observed in the filter containing the oyster shell media compared to 33% for the reactor containing the stone media. They suggested that the higher surface area of the oyster media was responsible for the higher COD removal of that reactor.

Young and Dahab (59) studied the effects of fixed media characteristics on process performance. They proposed that equivalent pore diameter (EPD), defined as four times the cross sectional area of the media pores divided by the wetted perimeter of the pores, is critical in determining the behavior of the process in terms of removal. They used the Equivalent Pore Diameter (EPD) criteria to evaluate the performance of four anaerobic filters using different media which had the same void ratio but different surface area values. They reported that the media with the highest EPD yielded the highest COD removal rate. They also speculated that a high specific surface area does not necessarily have to be coupled with the highest bed void ratio to guarantee the highest possible COD removal.

Muller and Manani (35) suggested that the use of low-weight, high-porosity plastic media, rather than the heavier and less porous rock media would improve the COD removal efficiency of the anaerobic reactor by allowing for a greater accumulation of biological solids per unit of reactor volume.

Flynn and Whitemore (11) studied methanogenic bacteria colonized on reticulated polyurethane foam. They reported that the binding forces between methanogens and particle surface were weak and imposed a limit on process operation in fixed or fluidized-bed methane digesters. They reported that the methanogenic bacteria washed out of the support matrix at a linear flowrate as low as 1.5 cm/min. They suggested that the rate of biomass accumulation at a given flowrate seemed to be a function of pore size.

Huysman, P., et al (3) examined various packing materials used in attached-film reactors to determine factors affecting microbial colonization. They concluded that surface roughness, porosity, and pore size are primary factors affecting colonization rate.

4) Anaerobic process kinetics

Mechanistic mathematical models of biochemical processes are frequently developed by engineers to describe the interrelationship of individual parameters and factors that are involved in the anaerobic process in a manner that allow the investigation to predict how the system might respond to certain changes in those parameters. In developing those mathematical models investigators quite often use formulas involving the famous Monod kinetics. Monod (34) in 1949 presented an expression which defines the kinetics of substrate removal in biological systems. The expression is as follows:

$$\mu = - \frac{dS/dt}{X} = \frac{K \ S}{K_S + S} \qquad [2.2]$$

μ = Specific substrate removal rate, 1/time

K = Maximum specific substrate removal rate, 1/time

S = Concentration of substrate surrounding the microorganisms, mass/volume

X = Concentration of microorganisms in the system, mass/volume

K_S = concentration of substrate at which substrate removal rate is one-half the maximum, mass/volume

Many researchers (13) used the Monod model to define the net growth rate with the following, commonly used expression:

$$\frac{dX}{dt} = - Y \frac{dS}{dt} - b X \qquad [2.3]$$

where,

$\frac{dX}{dt}$ = Net growth rate of microorganisms mass/mass

Y = Biomass growth yield coefficient, mass/mass

b = biomass decay rate, 1/time

X = microorganisms mass concentration mass/volume

Equations 2.2 and 2.3 are used in part to develop the model in this research.

5) Kinetic Constant of Methane Fermentation

Kinetic Constants in methane fermentation, (e.g Y, b, u) are functions of temperature, substrate concentration, and substrate type. Of importance to this research are the kinetic constants of the microorganisms that utilize acetic acid.

Montgomery (35) determined the utilization kinetics for the microorganisms that utilize acetates, as well as those that utilize propionates over a broad temperature range as shown in tables 2.2 and 2.3.

TABLE 2.2: SUMMARY OF ACETATE KINETIC PARAMETERS, AFTER MONTGOMERY (35)

Parameter							
T	^{o}C	10	15	20	25	30	35
K	$\frac{\text{mg acetate}}{\text{mg cells-day}}$	0.64	1.1	2.7	6.4	6.7	7.2
K_s	$\frac{\text{mg acetate}}{\text{liter}}$	110	55	24	10	7.5	3.9
Y	$\frac{\text{mg cells}}{\text{mg acetate}}$	0.034	0.033	0.034	0.035	0.036	0.036
b	$\frac{\text{mg acetate}}{\text{liter}}$				0.032		

TABLE 2.3: SUMMARY OF PROPIONATE KINETIC PARAMETER RESULTS AND METHODS, AFTER MONTGOMERY (35).

Parameter							
T	^{o}C	10	15	20	25	30	35
K	$\frac{\text{mg propionate}}{\text{mg cells-day}}$	0.85	1.8	4.0	7.1	7.5	7.9
K_s	$\frac{\text{mg propionate}}{\text{liter}}$	246	145	103	35	15	7.7
Y	$\frac{\text{mg cells}}{\text{mg propionate}}$	0.036	0.035	0.035	0.034	0.033	0.033
b	days^{-1}				0.007		

6) Biofilm kinetics

Several researchers have investigated the kinetics of anaerobic biofilms. This was done to develop mechanistic models that can be used to predict the reactor performance under various operating conditions. In certain cases, the model is used to optimize reactor's dimensions and configuration, or to predict the influence of system configuration upon the process performance.

La Motta (28, 29) reported that flux of substrate from the bulk liquid into a biofilm influences the rate of utilization of substrate within the biofilm. La Motta stated that four simultaneous processes control the flux of substrate from the bulk liquid into the biofilm as follows:

1) Mass transport of substrate from the bulk liquid and the biofilm surface (external diffusion)
2) Diffusion of substrate through the biofilm (internal diffusion)
3) Substrate utilization within the biofilm
4) Decay of the biofilm mass

Dague and Chiang (6) presented a theoretical development of a mathematical model of a two-culture submerged media upflow anaerobic reactor. Their model considers the suspended and attached biomass as two separate cultures. Four equations were used to describe the distribution of substrate, suspended biomass, attached biomass, and gas

production within the reactor.

Buijs, and Heertjes (3) developed and experimentally checked a mathematical model that describes the physical behavior of sludge in a completely mixed (suspended growth) upflow reactor for anaerobic treatment of wastewater. The model is based on the mass balance of the sludge in the reactor. The quantitative values of the transport factors that are a measure of the efficiency of the transportation of sludge by the fluid streams occurring in the reactor were determined experimentally. This was done for wastewater containing lower fatty acids as the main organic pollutants and for sludge with good settling characteristics.

Heertjes and Kuijvenhoven (16) experimentally studied the flow pattern in an up-flow suspended flow reactor and concluded, using a mathematical description of the reactor, that scaling-up of upflow reactors in the horizontal direction will not affect the process efficiency, if the influent distribution was uniform, and given special attention. They suggested that the optimum reactor height, in terms of substrate removal and sludge retention, seems to lie between 4 and 6 m.

Utilizing the principles of La Motta, McCarthy and his coworkers Williamson and Rittmann (41, 42, 43, 44, 54, 55) mathematically developed and experimentally verified a model that describes an upflow fixed film anaerobic filter process. They assumed that the density of active bacteria is

constant throughout the reactor biofilm. Using a mass balance for the substrate in the reactor; they took the decay and growth of microorganisms composing the biofilm into consideration in formulating the model. This was not considered in other models developed prior to the development of their model. Based on kinetic and energy constraints, they suggested that their model predicts for a single substrate, that a steady-state bulk concentration (S_{min}) develops, below which a steady-state biofilm cannot exist. Steady-state biofilm is defined as one that has neither net growth nor decay with time.

The Rittmann and McCarthy model (42, 43) will be employed as a part of the mathematical model developed for the anaerobic system investigated in this research (more information on the Rittmann and McCarty model can be found in chapter 4).

Grady (13) presented a comprehensive review of mathematical modeling techniques developed prior to 1982. The review is very useful in the understanding of the various aspects of modeling aerobic and anaerobic wastewater treatment processes.

CHAPTER THREE

LABORATORY EXPERIMENTAL APPARATUS AND RESULTS

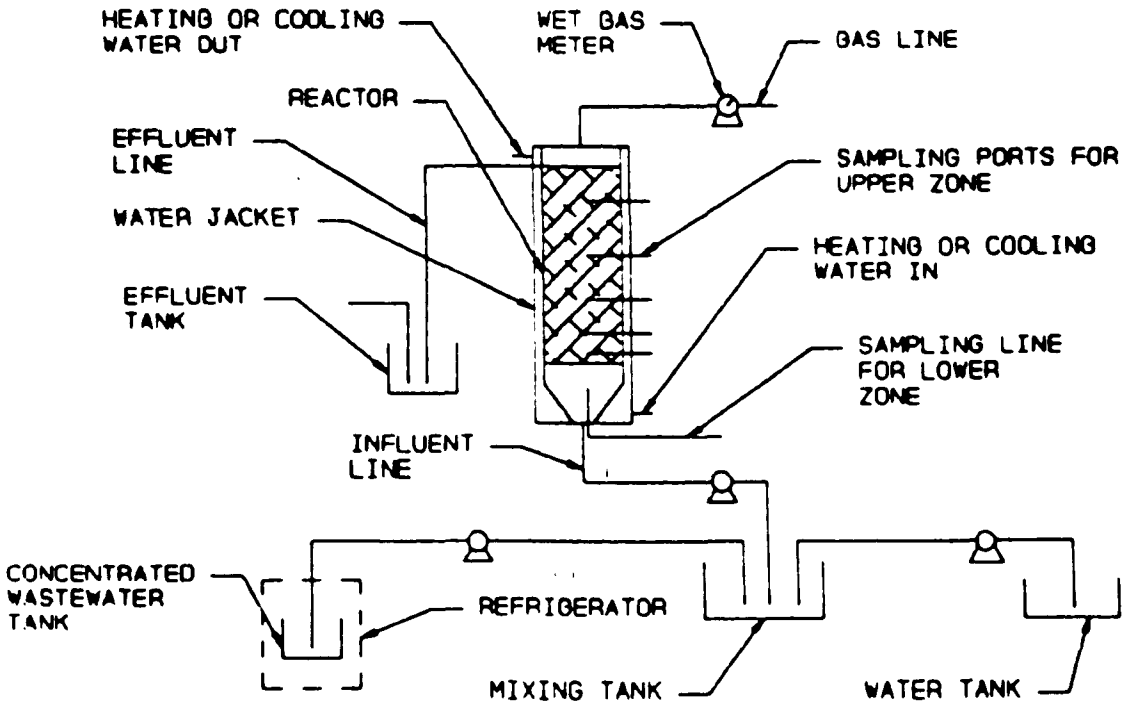
Experimental Apparatus

A schematic diagram of the experimental set-up used in this study, is shown in figure 3.1 on page 51. The COD concentration of the wastewater influent as well as the influent flowrate were controlled using three positive displacement pumps. To minimize biodegradation of the wastewater before entering the reactor, concentrated feed solutions were prepared and kept in a refrigerator. Tap water was mixed with the concentrated feed just before the waste was pumped into the reactor. By varying the dilution water flowrate and the COD values of the concentrated feed, as well as varying the concentrated feed flowrate, any combination of waste strength, hydraulic, and organic loading could be achieved.

A bench scale laboratory unit was constructed using a plexiglass column, 5.5 inches (14.0 cm) in diameter and 6.71 feet (203.8 cm) tall. The base of the column was constructed as a short cone to disperse the waste evenly into the lower zone of the reactor and to develop adequate mixing in the lower zone. The filter contains two consecutive zones: a lower zone, and an upper zone. The lower zone contains no media and contains the conical shaped filter base mentioned above. The upper zone is filled with actifill plastic media.

This plastic media has a specific gravity of about 1.15, a diameter of 5/8 inches, and specific surface area of 104 ft2/ft3 (3.41 cm2/cm3). Thin flat rings 0.5 in. (1.3 cm) in thickness, 0.25 inch (0.64 cm) in diameter, were placed approximately at one foot (0.3 m) intervals inside the filter to prevent short-circuiting of the waste through the large void spaces formed at the column's boundary. The liquid volume of the lower zone is 0.18 cu.ft (5100 cm3) and the liquid volume of the upper zone is 0.76 cu.ft (21600 cm3). The total liquid volume of the reactor is 0.94 cu.ft (26600 cm3). The gas produced in the reactor and the effluent are collected separately from the top of the reactor, as shown in figure 3.1. In addition, sampling ports were placed at distances 1.47, 1.92, 2.67, 3.67, and 5.67 feet respectively from the bottom of the filter.

FIGURE 3.1: SCHEMATIC OF THE EXPERIMENTAL SET-UP.



Synthetic Wastewater Composition

The synthetic wastewater used in this study contained acetate as its sole source of organic carbon. Vitamins and nutrients were added to the wastewater in amounts greater than those required for the anaerobic biodegradation. Sodium sulfide was added to provide a reducing environment. The final assay concentration as shown in table 3.1 for nitrogen, phosphorus, sulfide, and alkalinity are 31 mg/L as N, 7 mg/L as P, 10 mg/L as S, and 500mg/L as CaCo3.

The Consideration that led to the Waste Choice

As mentioned earlier anaerobic treatment requires a unique microbial balance between three different species of bacteria. The fermentive and hydrogen producing bacteria carry the degradation of the organic matter to the acid stage, and then the sensitive methane forming bacteria complete the conversion into methane and carbon dioxide. When a sufficient population of methane forming bacteria are present, and environmental conditions are favorable, they utilize the end products produced by the hydrolyzing and hydrogen producing microorganisms as fast as they are formed. The methane bacteria appear to be the most sensitive to changes in environmental conditions and food loading rates. They are affected much more radically by changes in pH and temperature than the acid producing bacteria. Inhibition caused by changes in either pH or temperature result in a decreased rate of destruction of volatile acids. Conse-

quently volatile acids begin to accumulate in the system. At pH value below 6.5, methane bacteria are seriously inhibited while the hydrolyzing and hydrogen producing bacteria are still able to function until pH levels fall to about 5. Active methane digestion may never develop in such a mixture unless neutralizing agents such as lime are added to produce favorable pH level for methane bacteria. Accumulation of volatile acids and inhibition of methane producing bacteria may also occur during unsteady state operation. Under increased food loading rates, volatile acids may be formed faster than what the slow-growing methane forming bacteria can utilize. That is because the hydrolyzing and hydrogen bacteria grow much more rapidly than the methane bacteria. From the foregoing discussion it can be seen that successful operation of an anaerobic treatment process depends upon three factors:

1) The presence of sufficient populations of both the acid and methane forming microorganisms inside the reactor to handle the organic load applied.

2) Maintaining a satisfactory balance between the methane and acid forming bacteria.

3) Maintaining, within the reactor, favorable environmental conditions as pH and temperature to support the growth of methane forming bacteria.

In any complex organic synthesis the overall kinetics is dependent on the slowest acting link. In methane fermentation, the overall kinetics of organic matter conversion to methane will also be governed by the kinetics of the slowest

step. In this case it is the reduction of fatty acids by methane forming bacteria to methane and carbon dioxide. Jeris and McCarthy (22) have reported that acetic and propionic acids are the precursors of approximately 85% of methane produced from the fermentation of a complex waste. Acetate alone is the precursor of 72% of the methane formed. Smith and Moh(47) demonstrated that in anaerobic sludge metabolism approximately 72 percent of methane comes from acetate. Kaspar and Wuhrmann (26) studied the kinetic parameters of some catabolic reactions in digesting sludge, then concluded that the acetate degradation alone was the rate-limiting reaction in methane fermentation.

From the forgoing discussion, it becomes clear that a successful anaerobic digestion is limited by the successful functioning of the methane bacteria and, in particular acetate utilizing bacteria. Therefore, in the study of any anaerobic system for wastewater treatment it is imperative to first investigate the behavior of the methane formers and to determine the most favorable conditions for them. These conditions include: pH levels, temperatures, waste concentrations and, organic loads, reactor configuration, and flow pattern inside the reactor. The synthetic wastewater chosen for this study will guarantee that the methane formers are the predominant species of bacteria inside the reactor.

TABLE 3.1: WASTEWATER COMPOSITION

1- ORGANIC SUBSTRATE		
Sodium acetate (Na CH_3COO), or acetic acid ($CH_3.COOH$)		
2- NUTRIENTS		
Ammonium Chloride ($NH_4.Cl$)	167	mg/L
Phosphoric acid (H_3PO_4)	19	mg/L
3- MINERALS		
$CoCl_2.6H_2O$	10	mg/L
$MnCl_2.4H_2O$	1	mg/L
$FeCl_2.4H_2O$	10	mg/L
$CaCl_2$	14	mg/L
$MgCl_2.6H_2O$	80	mg/L
KCl	80	mg/L
$Na_2S.9H_2O$	50	mg/L
Cu	00.002	mg/L
MoO_3	00.12	mg/L
$Ni(NH_4)_2(SO_4).6H_2O$	00.25	mg/L
$Zn.Cl_2$	00.15	mg/L
4- VITAMINS		
C	0.06	mg/L
B-1	0.0015	mg/L
B-2	0.0017	mg/L
B-6	0.002	mg/L
B-12	0.0006	mg/L
Niacin	0.020	mg/L
Folic acid	0.02	mg/L
Pantothenic acid	0.0004	mg/L
Biotin	0.01	mg/L
thiotic acid	0.005	mg/L
P-aminobenzoic acid	0.005	mg/L

Research Time Table and Modes of Operation

An objective of this study was to determine the performance of the anaerobic filter under a range of organic loading conditions. To determine the potential of the process to treat low strength wastewaters, waste concentrations of 350 mg/L as COD or less, were chosen for the study. (see table 3.2 on page 57). These low wastewater strengths, in terms of COD, are typical of domestic wastewaters. Treatment of these wastewater strengths have not been feasible by conventional anaerobic processes.

Organic loadings ranging from 40.5 to 59.2 lb COD /day/1000 cu ft of total filter liquid volume (0.65 to 0.95 $Kg/m^3/day$) were evaluated. These are ranges normally used with other biological processes such as the aerobic activated sludge, trickling filters, and anaerobic contact processes. The resulting hydraulic retention times, shown in table 3.3, are calculated on the basis of the liquid volume of the filter.

TABLE 3.2: RESEARCH TIME TABLE.

Start up	Phase (1)						Phase (2)						Phase (3)					
O. L. = 0.55	O. L. = 0.65						O. L. = 0.85						O. L. = 0.95					
T = 30 °C	30 °C		20 °C		10 °C		30 °C		20 °C		10 °C		30 °C		20 °C		10 °C	
COD = 350 Mg/L	350	250	350	250	350	250	350	250	350	250	350	250	350	250	350	250	350	250
	Run # 1	Run # 2	Run # 3	Run # 4	Run # 5	Run # 6	Run # 1	Run # 2	Run # 3	Run # 4	Run # 5	Run # 6	Run # 1	Run # 2	Run # 3	Run # 4	Run # 5	Run # 6
0	7	10	14	18	22	26	30	34	38	42	45	48	52	56	60	64	68	72	76

WEEK

O. L. = Organic Loading in Kg COD/m3/day

TABLE 3.3: HYDRAULIC RETENTION TIMES (HRT) USED IN THE DIFFERENT RUNS OF THE EXPERIMENT.

PHASE No.	RUN No.	ORGANIC LOADING Kg COD/m3/day	INFLUENT COD CONCENTRATION mg/L	HRT HOUR
1	1	0.65	350	12.9
1	2	0.65	250	9.3
1	3	0.65	350	12.9
1	4	0.65	250	9.3
1	5	0.65	350	12.9
1	6	0.65	250	9.3
2	1	0.85	350	9.9
2	2	0.85	250	7.1
2	3	0.85	350	9.9
2	4	0.85	250	7.1
2	5	0.85	350	9.9
2	6	0.85	250	7.1
3	1	0.95	350	8.9
3	2	0.95	250	6.4
3	3	0.95	350	8.9
3	4	0.95	250	6.4
3	5	0.95	350	8.9
3	6	0.95	250	6.4

ANALYTICAL METHODS

This section describes the methods that were employed in the laboratory research study. Alkalinity, total and volatile non-filterable residue, total residue, total volatile residue, ammonia nitrogen, and Chemical Oxygen Demand (COD) were measured according to the procedures described in "Standard Methods for the Examination of Water and Wastewater" (43).

Gas Chromatography

A) Volatile Acids

The presence of volatile fatty acids (in particular acetates) in the wastewater was periodically confirmed using gas chromatography. In this respect, a Supelco column (10% SP - 1000/ 1% H PO on 100/120 chromosorb WAW, 6ft x 4 mm ID glass column) was used for separation. Two identical columns were connected to a Hot Wire Detector (HWD) in a Sigma 2000 Perkin-Elmer Gas Chromatograph. The column was maintained isothermally at 155^{o}C. The temperature of both the injection port and the detector was maintained at 205^{o}C. The carrier gas used was helium. A data management integrator was used to analyze the signals from the gas chromatograph. The column was calibrated before and after each use. The calibration standards used were made by Supelco Inc. (catalog number 4-6975). The wastewater samples were filtered using a glass fiber filter and the filtrate

was taken for extraction before injection into the gas chromatograph. The extraction procedure which is described in Supelco Inc. publication number 7484 is as follows:

1 6 ml filtered samples are acidified with 0.1 ml of 50% aqueous $H_2 S O_4$.

2 2 ml of acidified samples are pipetted into a centrifuge tube.

3 1 ml of ethyl ether is added, the tube is sealed, and contents are mixed.

4 Tube is then briefly centrifuged.

5 The ether layer is removed with a pipette and placed in a test tube.

6 An amount of anhydrous Mg So_4 equal to about one half the volume of ether in tube is add to it.

7 The tube is allowed to stand for 10 minutes to remove all traces of water from ether.

8 An appropriate quantity of extract is injected into the column (approx. 14 μl).

B) Gas Analyses

The constituents of the biogas produced in the reactor were determined using the Sigma 2000 gas chromatograph and the integrator described above. Two identical Supelco Columns (100/120 carbosieve S-II, 10ft x 1/8" stainless steel column) were used with HWD. The carrier gas was helium. A temperature program method adapted from Supelco publication number 760 D was used. This method is outlined as Follows: Column was held for 7 minutes at temperature of 40^{o}C then the temperature was raised to 225^{o}C at a rate of 30^{o}C/min. The helium flowrate was kept at 30 ml/min. The

test is performed as follows:

A 0.5 ml gas sample is injected in the column at the beginning of the chromatograph run. The data are collected and analyzed by the integrator. The integrator plots a diagram showing the gas constituents and the percentage of each in the total volume. The integrator also prints a summary report which provide the same data as the plot.

Statistical Considerations

All the experimental results reported in this study were statistically examined as follows:

For each set of data, the average and the standard deviation were determined, then the individual data points in the set were examined separately. If the value of any data point lies outside the range of the average plus or minus 3 times the standard deviation, the point is discarded as an unacceptable. This was done to provide at least a 67% level of confidence in all the data reported in the study.

EXPERIMENTAL RESULTS

Start Up

The time required to establish a sufficient biological population in the filter was approximately seven weeks. The lower zone of the filter was initially seeded with approximately 70 gm (dry basis) of volatile solids. The seed was collected from anaerobic digesters at the Wards Island Wastewater Treatment Plant. In the start up period the filter was operated using a synthetic wastewater feed whose COD concentration was 350 mg/L. The hydraulic retention time of the filter was adjusted to 12.9 hours to provide an organic loading rate of 0.65 Kg/m3/day. The filter temperature was maintained at 30^{o}C.

Very little COD removal (approximately 3-4%) was achieved during the first two weeks of operation. On the fifteenth day the filter was reseeded with another 60 gm(total dry basis) of digesting sludge solids. The filter reacted positively and rapidly to this seed. COD removal improved steadily in the following four weeks. At the end of the sixth week, measured from the beginning of the start-up period, the filter reached a steady state COD removal of 97%. The gas production was also steady. The percent methane in the biogas was steady and equaled 98%.

The filter was operated at this steady state for 10 days before the first phase of the experiment was initiated.

COD Removal

The overall soluble COD removal by the anaerobic filter, for all the different runs, is plotted in figure 3.2. These results as well as the COD removal along the filter height are given in table 3.4. Table 3.5 shows the actual soluble COD values remaining in the filter. Considering the overall COD removal the following can be reported:

a) In general the overall COD removal efficiency decreased as the organic loading increased, as can be seen in figures 3.3, 3.4, and 3.5. In addition, the rate at which the COD removal efficiency decreased between the lower and higher organic loading (as indicated by the slope of the COD removal curve) was found to increase when the influent COD concentration decreases. Furthermore, this rate of decrease of the removal efficiency was found to climb as temperature increases. Decrease in the COD removal as the organic loading increases (as indicated by the slope of the COD removal curve) is higher for the lower influent COD concentration. Also, this rate of decrease of removal efficiency was found to grow at lower temperatures.

b) The filter response to changes in run conditions, such as temperature and organic loading changes, was swift and smooth. In all runs, less than a week from the time the previous run conditions were changed was required to reach a new steady state substrate removal condition.

c) The temperature effect on COD removal, when all other variables are kept constant, was relatively small between the 30°C and 20°C runs when compared to the difference in the removal between 20°C and 10°C runs, as can be seen in figures 3.6, 3.7, 3.8, 3.9, 3.10, 3.11.

Effect of Filter Height

During the periods of steady-state removal efficiency, samples were taken three times a week from the filter at different heights (datum is the bottom of the reactor). In the lower zone, samples were taken at the lower 2/3 of the

height of this zone, as well as at a location 0.03 ft outside the zone, at a height of 1.47 ft. Within the upper zone the locations of the sampling ports are as indicated in tables 3.4 and 3.5.

Inspection of the resulting profiles of soluble COD removal vs. filter height (figure 3.12 through 3.29) as well as the corresponding COD values remaining in the wastewater (table 3.5), indicates a very high COD removal rate, more than 85% of the overall removal, occurring in the lower zone of the reactor for all runs. This high removal happened regardless of the fact that the hydraulic retention time in the upper zone is approximately four times greater than that in the lower zone. The reason for this may be explained as follows:

The COD remaining in the bulk solution in the upper zone is too diluted to support an effective biological growth or to assure adequate contact between the bacteria and the waste. In the lower 1/4 height of the upper zone, the COD removal rate was higher than the removal rate in the remaining height. In the top 3/4 of the upper zone's height, the lower COD removal rate closely approximated a straight line. The total alkalinity of the influent was approximately 500 mg/L CaCo3. This level of alkalinity was sufficient to buffer the filter to maintain pH values between 7.0 and 7.6 at all filter levels, and for all the different run conditions.

The soluble COD profiles throughout the filter (figures

3.12 through 3.29) clearly shows the effect of substrate concentration on the removal rate. With an influent COD concentration of 350 mg/L the fraction of COD removed at any filter height was greater than that removed for the same temperature, same organic loading and height but with an influent COD concentration of 250 mg/L.

In the range of the run conditions used in this research, it was found that the overall COD removal rate for the lower influent COD value (i.e. 250 mg/L) was 3 to 12 percent less than those for the higher influent COD concentration (i.e 350 mg/L). One possible explanation for this is that in order to maintain the same organic loading for the 350 and 250 mg/L COD influent concentration, the hydraulic retention time for the lower concentration would have to be smaller in magnitude than that for the higher COD value. (see table 3.4). This will result in less contact between the bacteria and the waste.

TABLE 3.4: COD REMOVAL PROFILES IN REACTOR FOR ALL THE DIFFERENT RUNS OF THE EXPERIMENT.

T °C	Organic Loading Kg COD/m3/d	0.65		0.85		0.95	
	HRT hours	12.9	9.3	9.9	7.1	8.9	6.4
	INF COD	350	250	350	250	350	250
	* Reactor Height (ft)	COD remov. %	COD remov. %	COD remov. %	COD remov. %	COD remov. %	COD remov. %
30	0 - 1.42	73	69	69	60	64	52
	1.47	89	84	85	79	79	71
	1.92	92	87	87	80	82	72
	2.67	90	87	87	80	83	72
	3.67	93	89	88	80	84	73
	5.67	95	90	89	81	86	73
	6.71	97	94	92	83	87	75
20	0 - 1.42	70	62	61	52	53	41
	1.47	83	77	76	72	68	59
	1.92	86	80	79	74	71	61
	2.67	86	83	78	74	72	61
	3.67	87	85	80	75	73	62
	5.67	90	86	83	76	74	63
	6.71	92	87	84	77	76	64
10	0 - 1.42	52	46	42	32	35	27
	1.47	68	62	57	51	51	46
	1.92	71	64	60	52	53	48
	2.67	71	64	59	53	53	48
	3.67	73	66	62	54	55	49
	5.67	75	68	64	54	57	49
	6.71	76	70	65	55	58	50

* Measured from the bottom of the filter.

TABLE 3.5: COD PROFILES IN REACTOR FOR ALL THE DIFFERENT RUNS OF THE EXPERIMENT.

T °C	Organic Loading Kg COD/m3/d	0.65		0.85		0.95	
	HRT hours	12.9	9.3	9.9	7.1	8.9	6.4
	INF COD	350	250	350	250	350	250
	Reactor Depth (ft)	COD mg/L	COD mg/L	COD mg/L	COD mg/L	COD mg/L	COD mg/L
30	0 - 1.42	95	78	109	100	126	120
	1.47	39	40	53	53	74	73
	1.92	28	33	46	50	63	70
	2.67	35	33	46	50	60	70
	3.67	25	28	42	50	56	68
	5.67	18	25	39	48	49	68
	6.71	11	15	28	43	46	63
20	0 - 1.42	105	95	132	120	165	148
	1.47	60	58	84	70	112	103
	1.92	49	50	74	65	102	98
	2.67	49	43	77	65	98	98
	3.67	46	38	70	63	95	95
	5.67	35	35	60	60	91	93
	6.71	28	33	56	58	84	90
10	0 - 1.42	168	135	203	170	228	183
	1.47	112	95	151	123	172	135
	1.92	102	90	140	120	165	130
	2.67	102	90	144	118	165	130
	3.67	95	85	133	115	158	128
	5.67	88	80	126	115	151	128
	6.71	84	75	123	113	147	125

FIGURE 3.2:
OVERALL SOLUBLE COD REMOVAL BY THE REACTOR FOR ALL THE DIFFERENT RUNS.

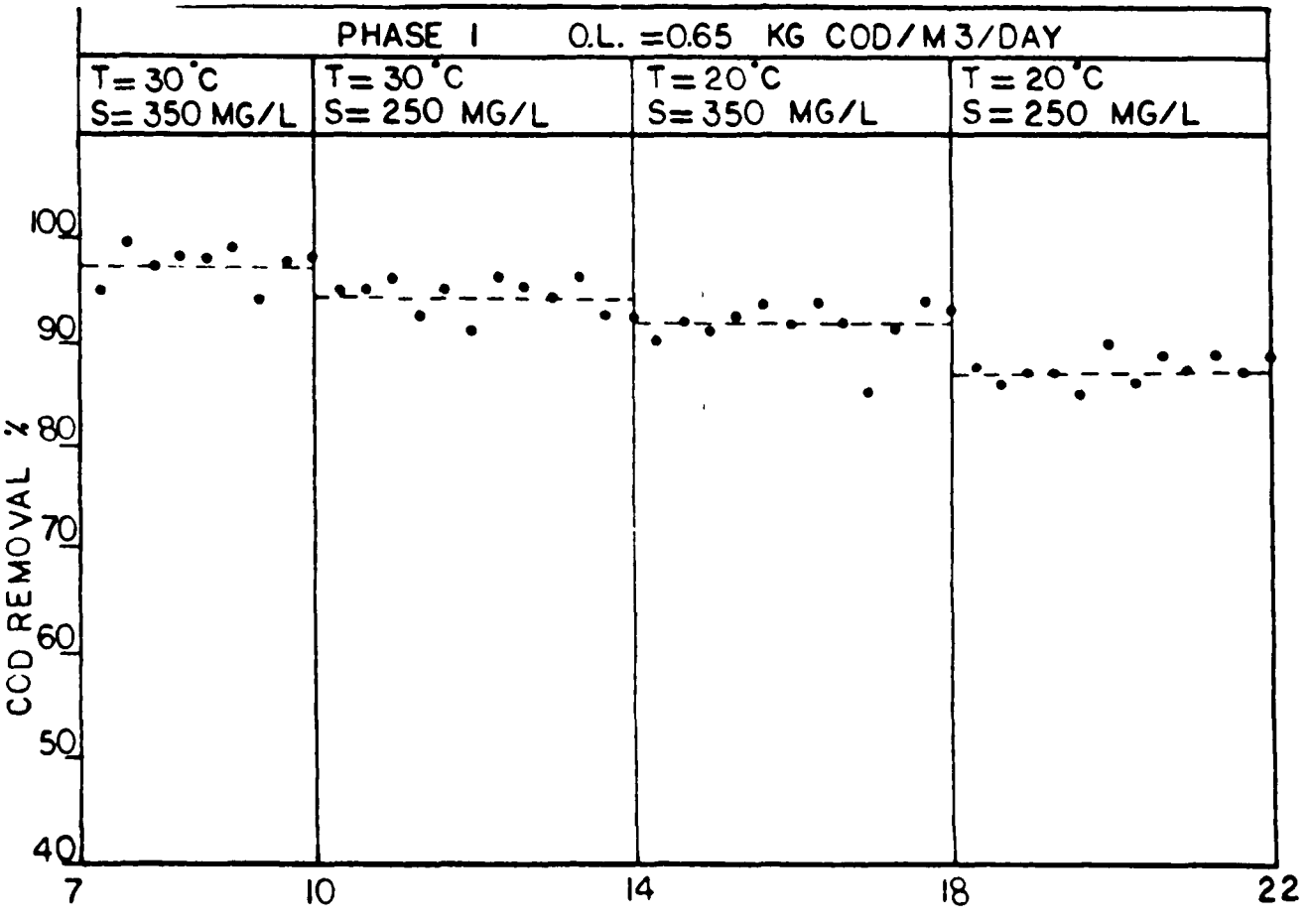


FIGURE 3.2 CONT.

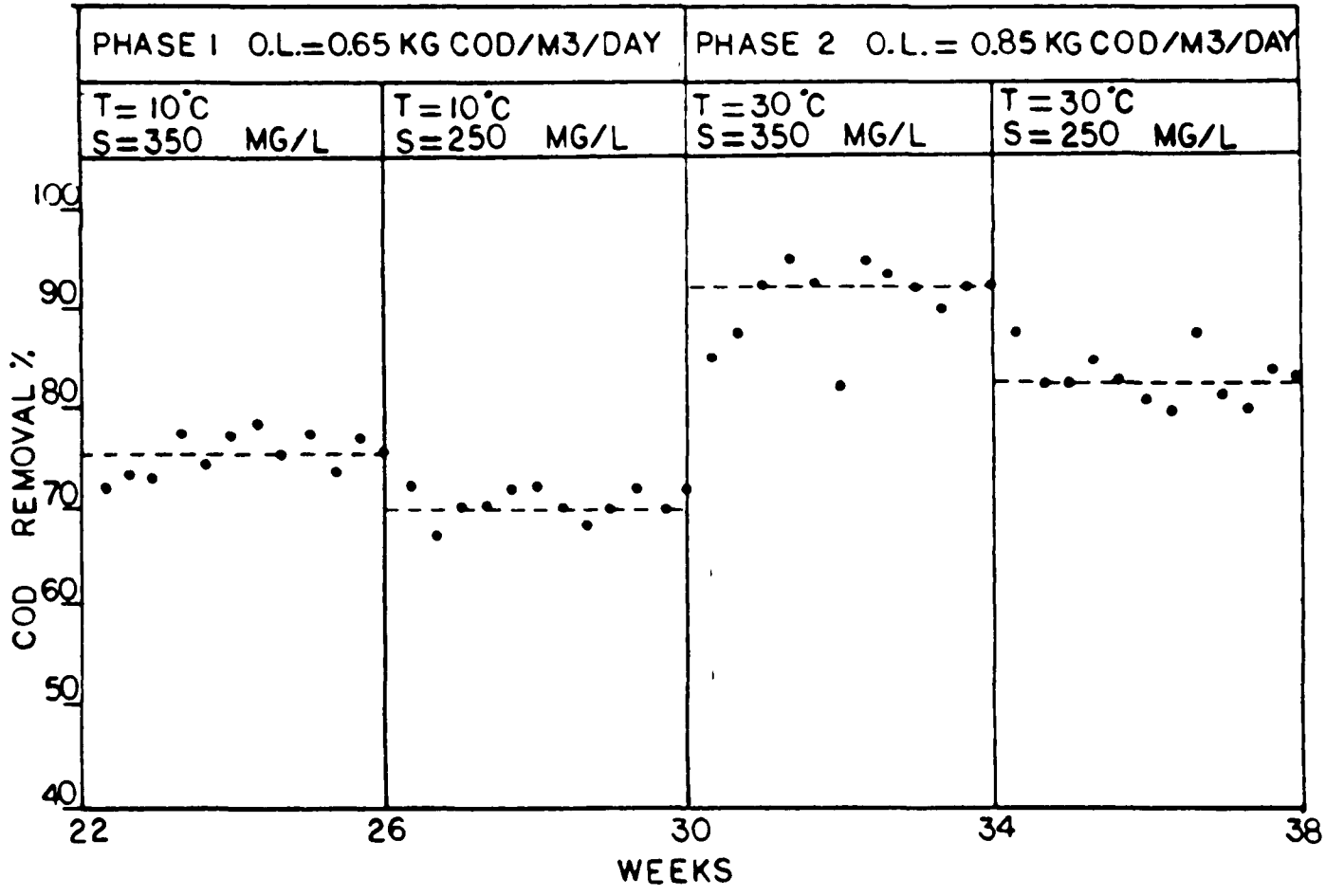


FIGURE 3.2 CONT.

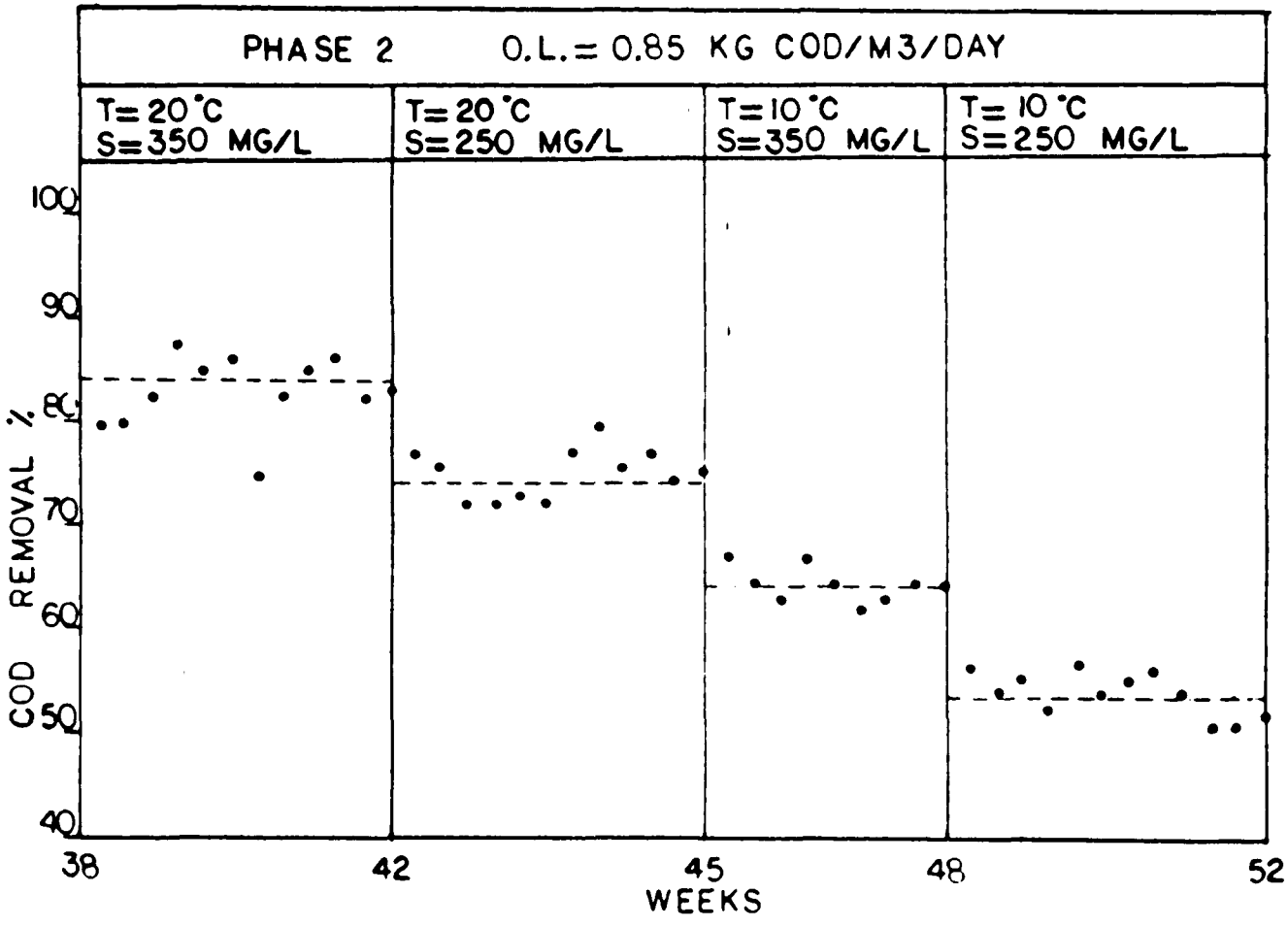


FIGURE 3.2 CONT.

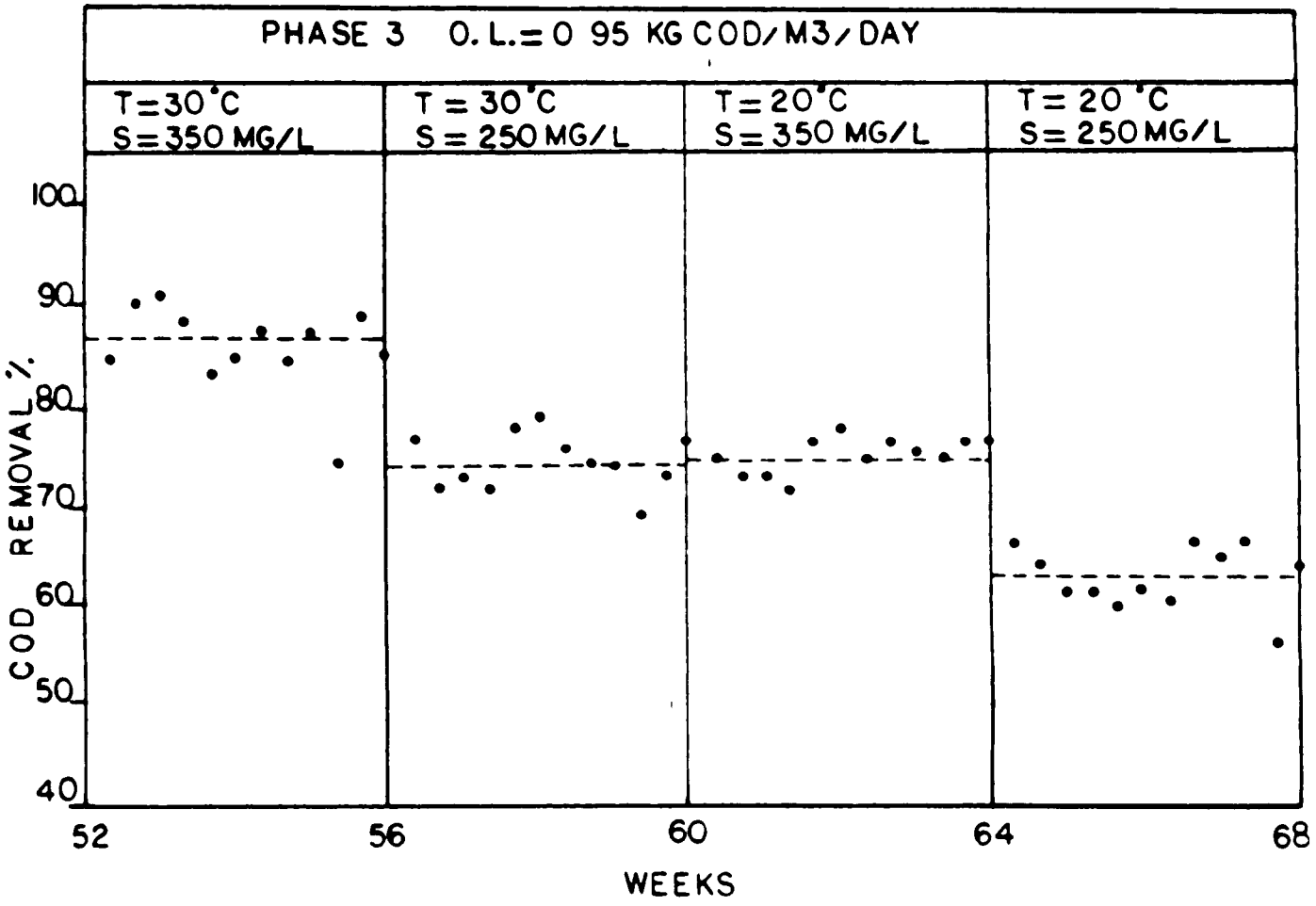


FIGURE 3.2 CONT.

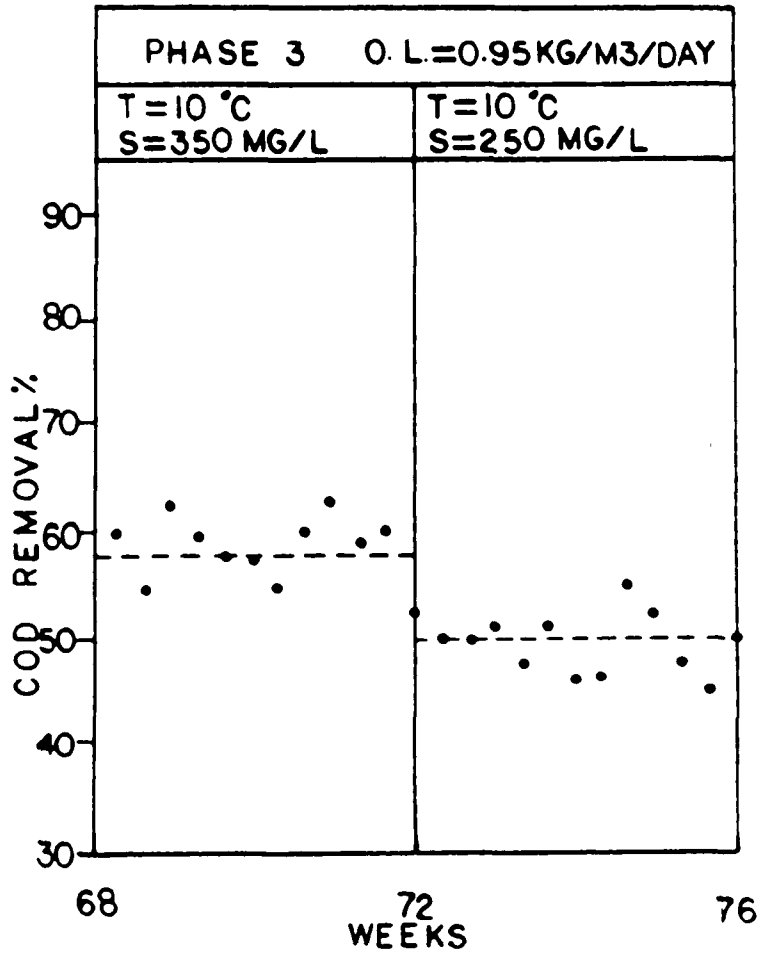


FIGURE 3.3:

COD REMOVAL VERSUS ORGANIC LOADING RATE.

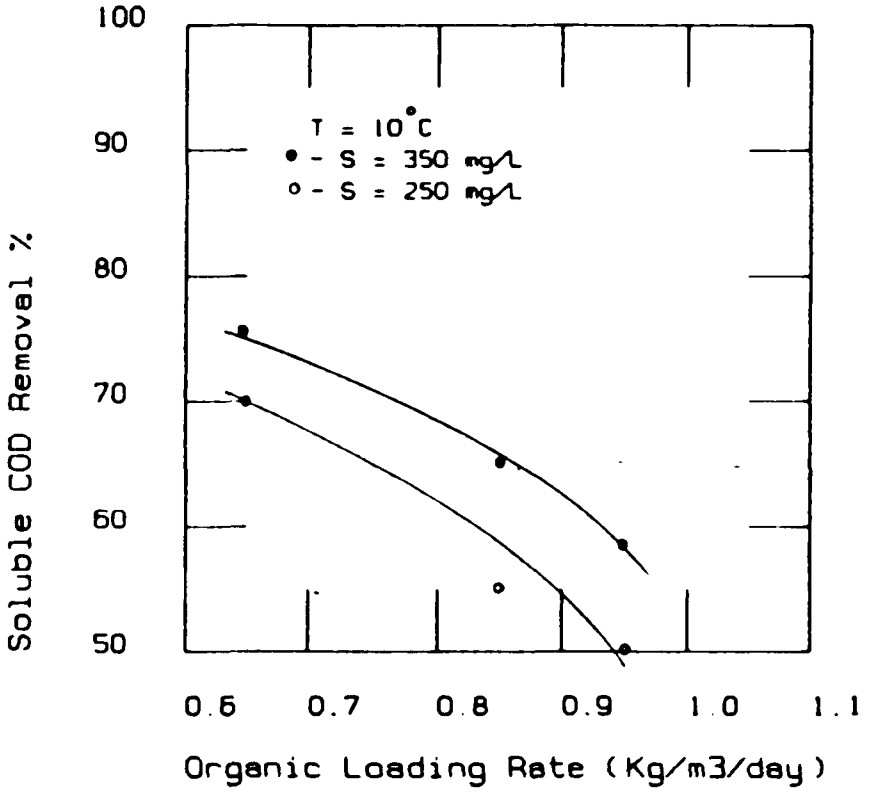


FIGURE 3.4:

COD REMOVAL VERSUS ORGANIC LOADING RATE.

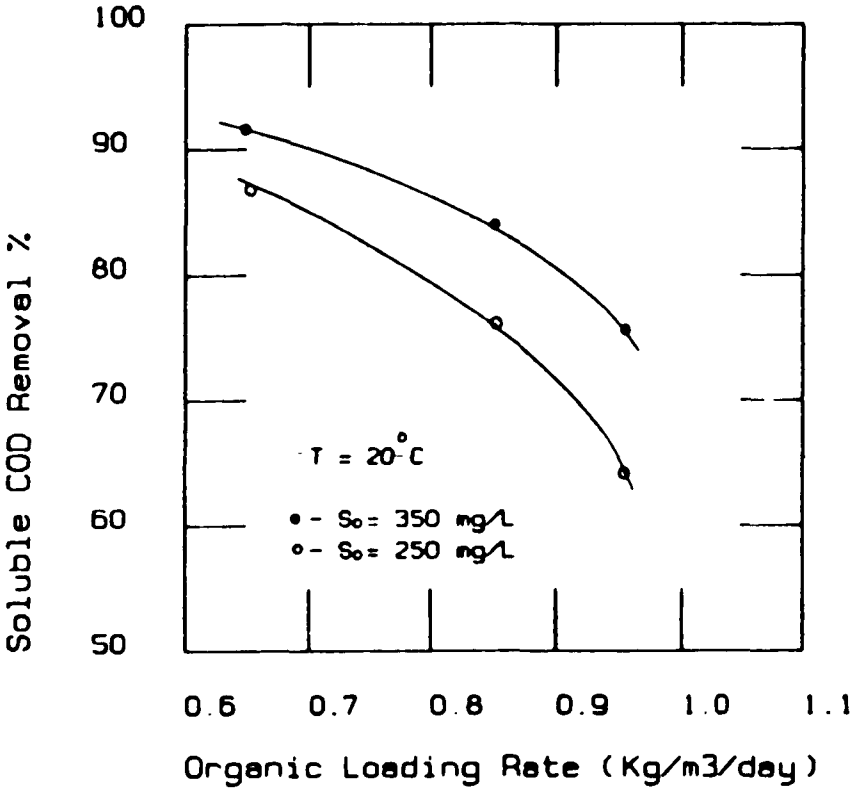


FIGURE 3.5:

COD REMOVAL VERSUS ORGANIC LOADING RATE.

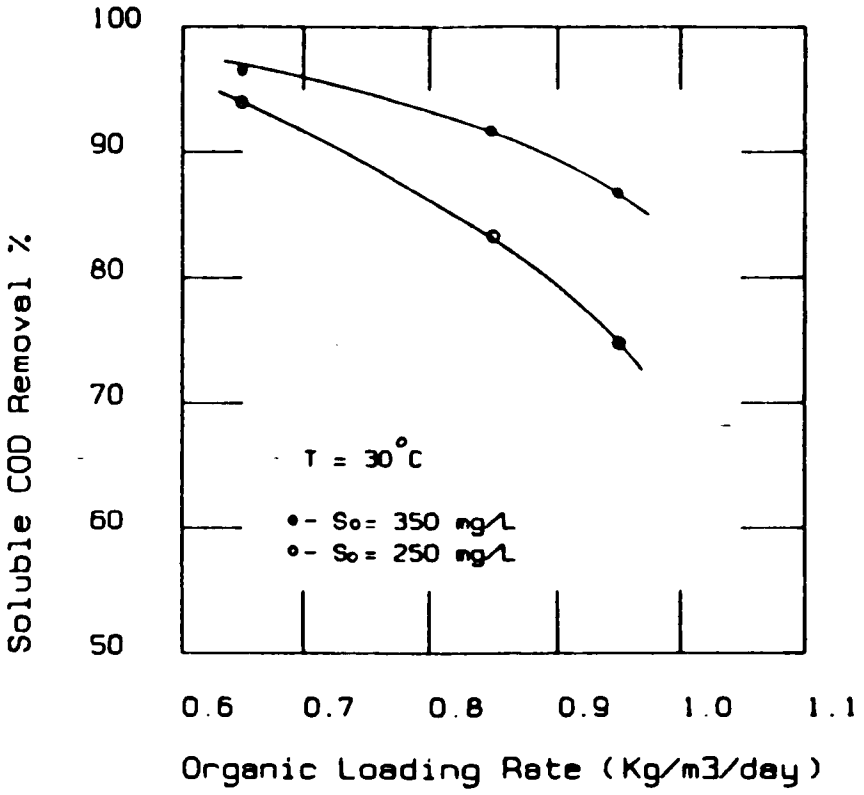


FIGURE 3.6:

COD REMOVAL VERSUS TEMPERATURE.

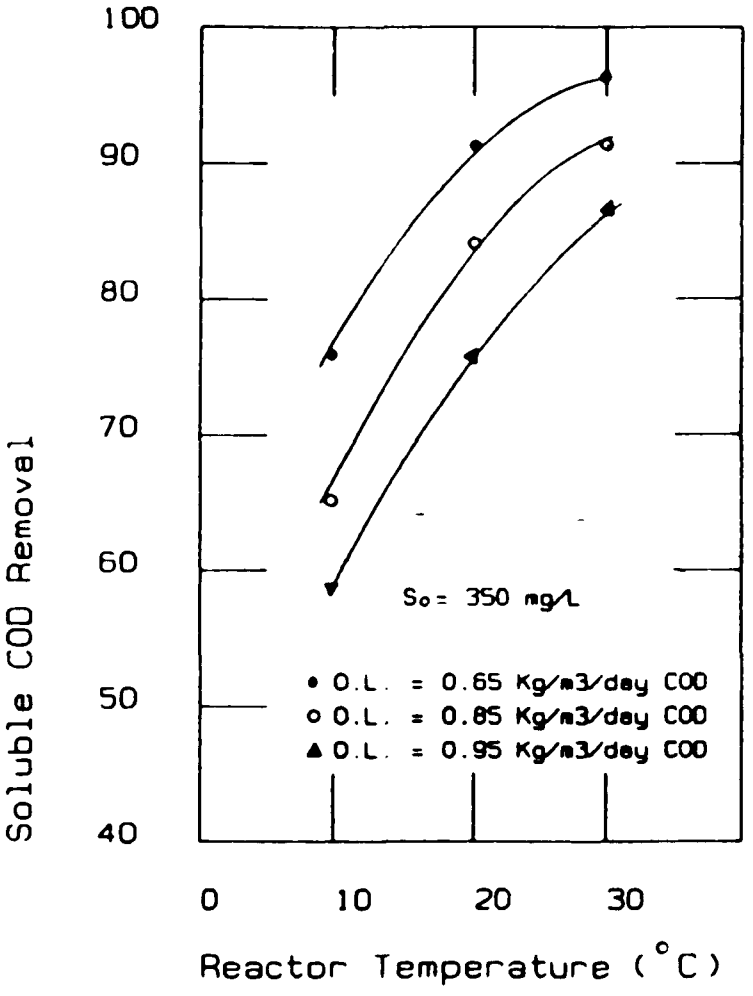


FIGURE 3.7:

COD REMOVAL VERSUS TEMPERATURE.

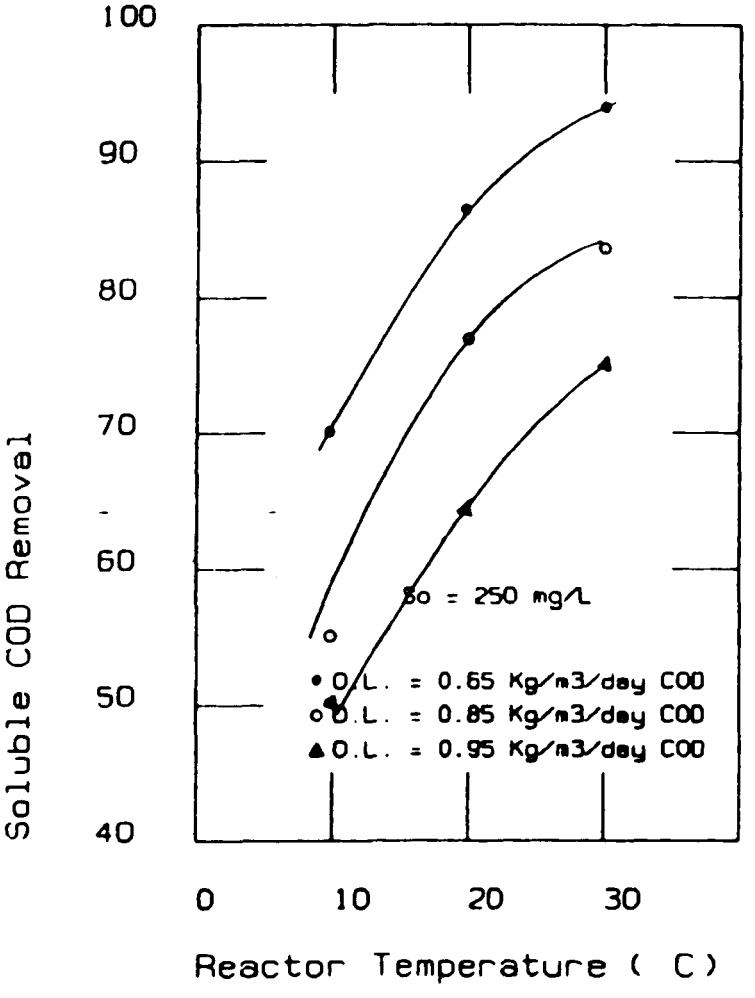


FIGURE 3.8:

COD REMOVAL VERSUS REACTOR HEIGHT.

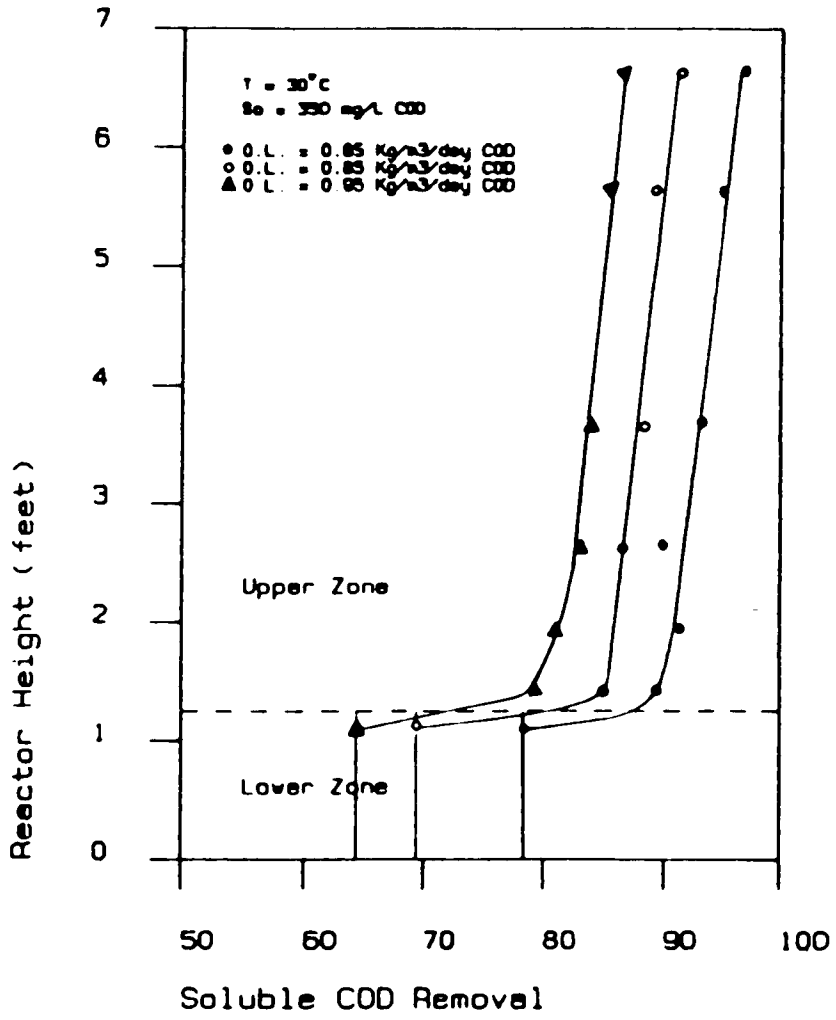


FIGURE 3.9:

COD REMOVAL VERSUS REACTOR HEIGHT.

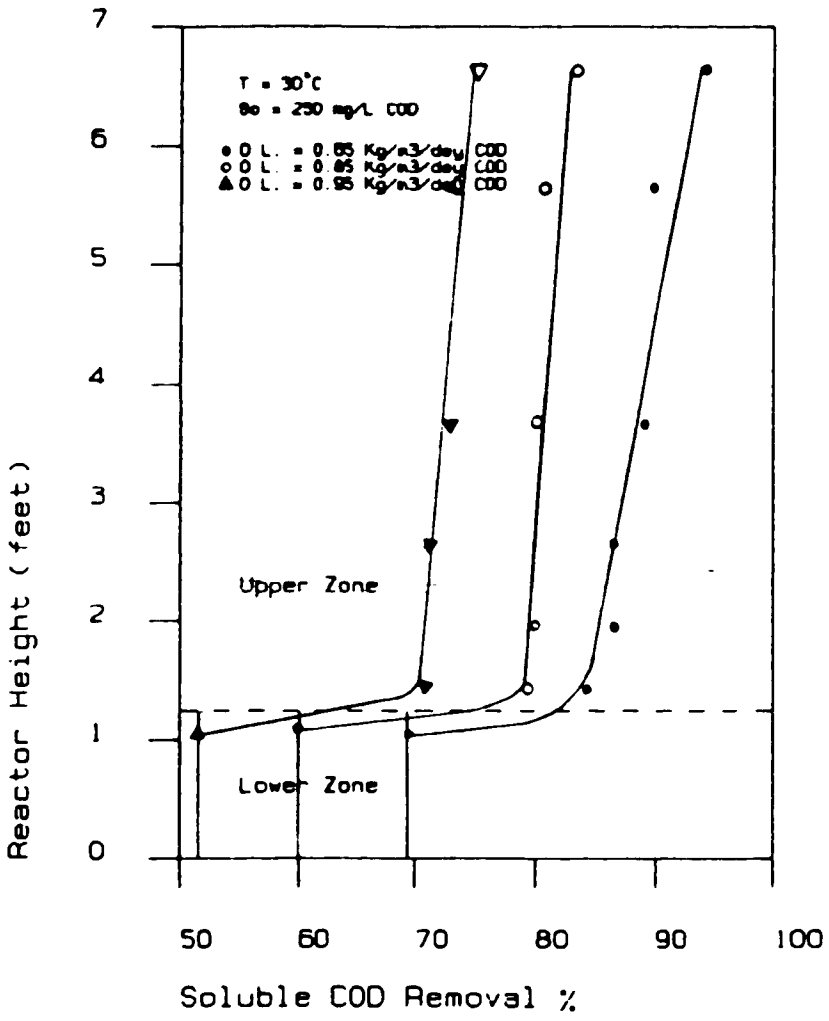


FIGURE 3.10:
COD REMOVAL VERSUS REACTOR HEIGHT.

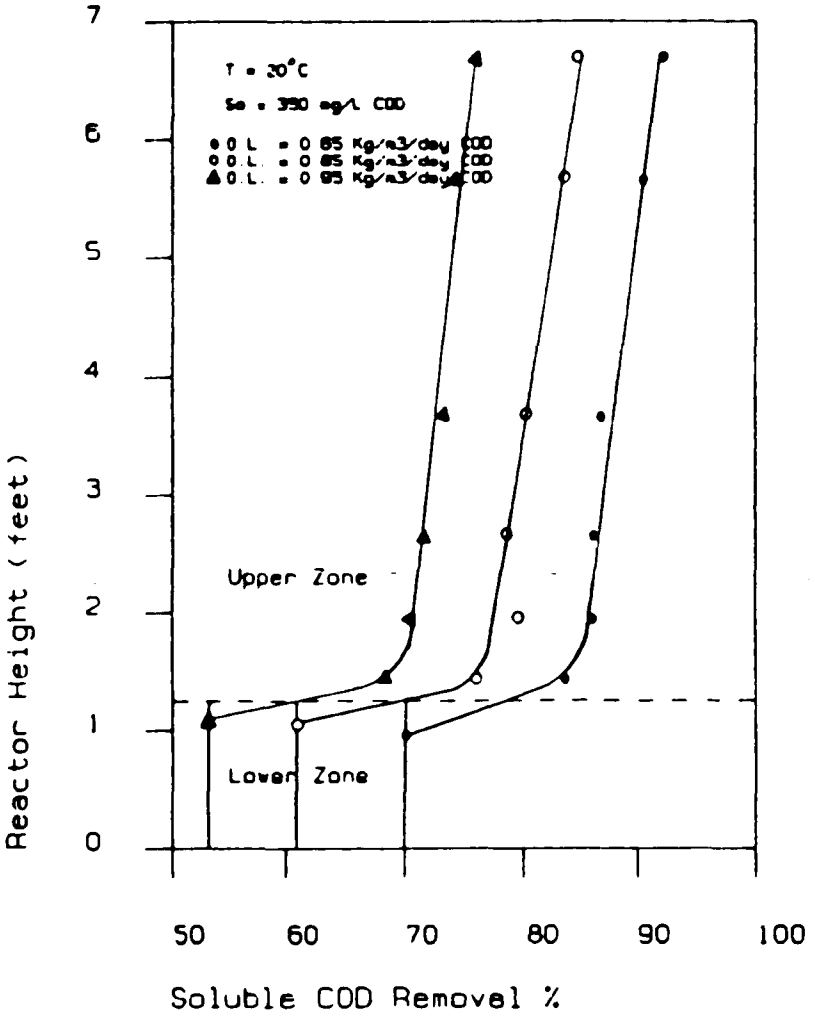


FIGURE 3.11:
COD REMOVAL VERSUS REACTOR HEIGHT.

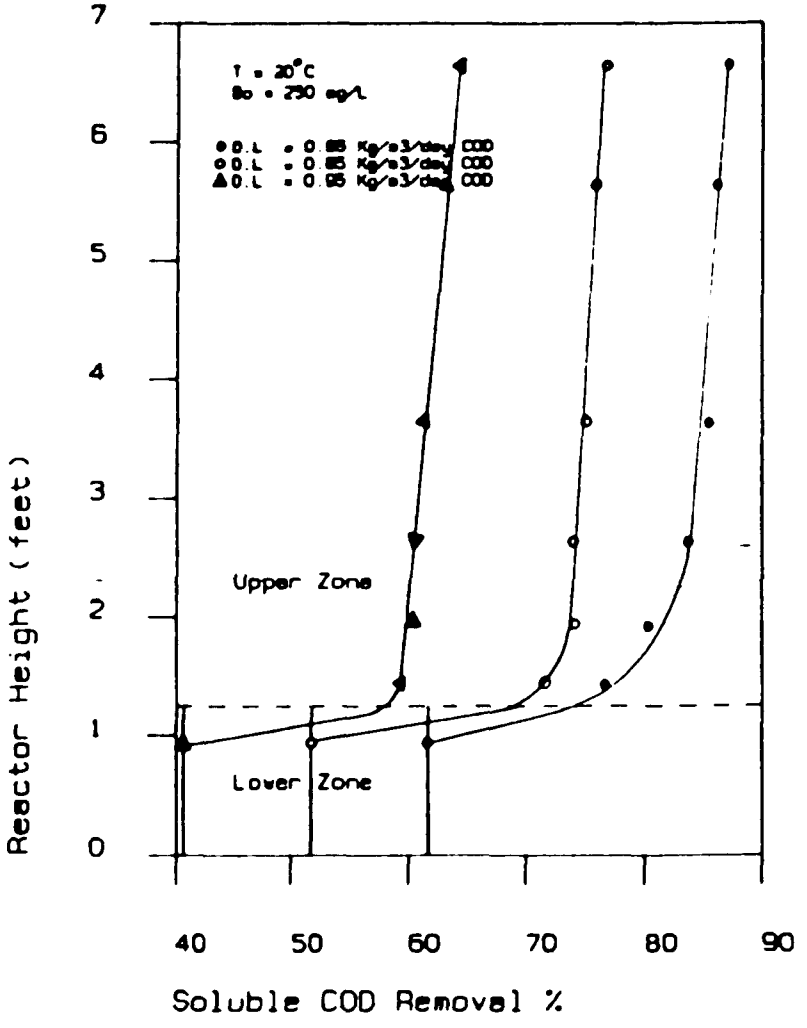


FIGURE 3.12:
COD REMOVAL VERSUS REACTOR HEIGHT.

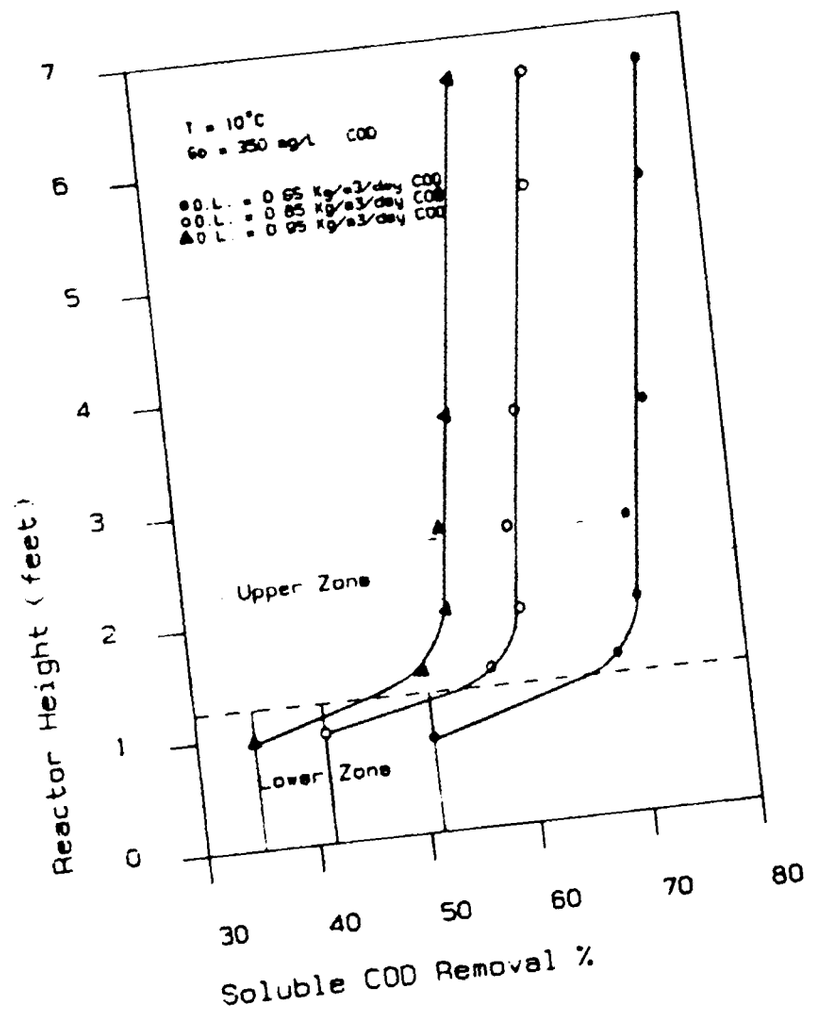
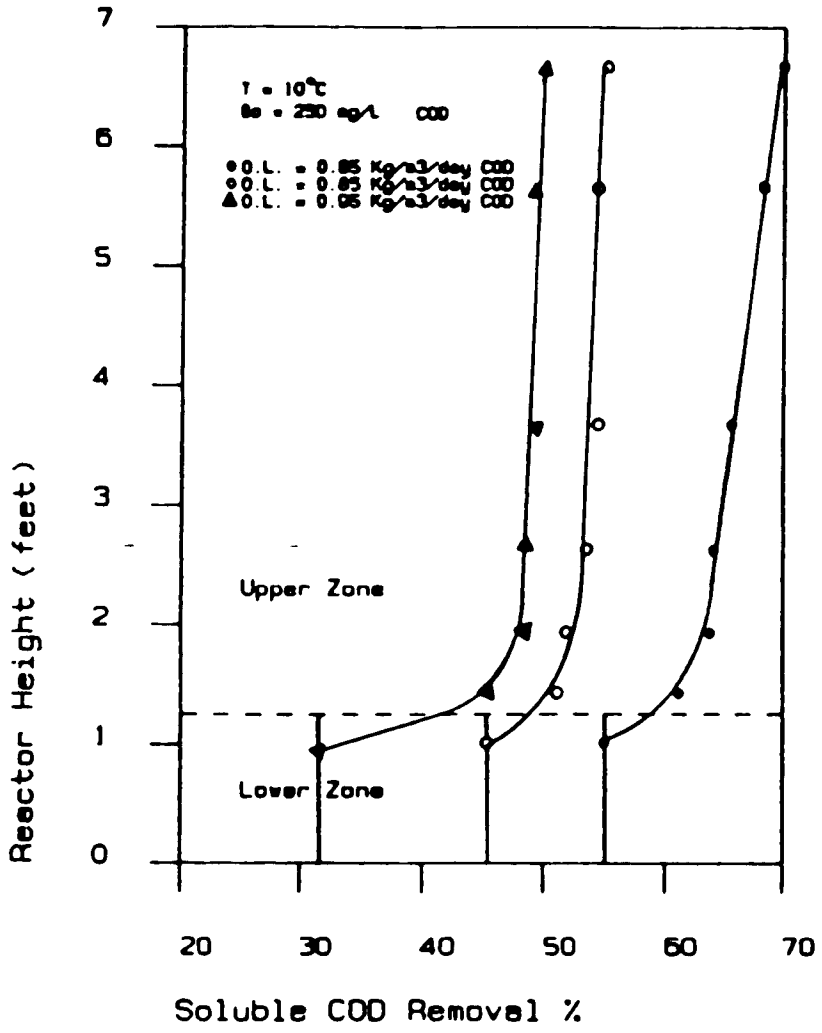


FIGURE 3.13:
COD REMOVAL VERSUS REACTOR HEIGHT.



BIOLOGICAL SOLIDS PRODUCTION

A main design consideration of this reactor is to use the lower zone to contain the bulk of the solids within the reactor and the upper zone as a polishing mechanism for the effluent leaving the lower one. These mechanisms result in a build up of solids in the filter resulting from biological synthesis. Appreciable loss of biosolids were found to occur when the level of the sludge bed reaches the first port in the upper zone. This appreciable biosolids loss may be explained as follows. When an appreciable amount of the sludge bed, which contains higher biosolids concentration, enters the upper zone fluidization of the solids would occur. This fluidization causes the noticeable solids washout reported. The higher upward flow velocity in the upper zone is due to two factors: 1) The zone is packed with plastic media. In other words , the cross section area is smaller than that in the lower zone (continuity principle). 2) The flow pattern is assumed to be plug flow. This caused the flow direction to be principally upward. In addition to the above, another factor which may also contribute to the appreciable biosolids loss is the shrinking of the polishing depth of the reactor. When the sludge bed occupies a part of the upper zone, which is usually used to polish and trap the solids flowing upward from the lower zone, this leaves a smaller depth to perform the polishing. The first port in the upper zone is located at a distance of 1.47 ft measured from the bottom of reactor, and about 0.27 ft measured from

the beginning of the upper zone. Visual observations were made, during the various runs, of the solids build up in reactor. In addition, solid analysis were performed at all sampling locations. When the concentration of solids increased significantly at the lowest sampling port in the upper zone, enough solids were withdrawn from the lower zone of the filter to bring the solid level just up to the top of the lower zone. This procedure was found to decrease the concentration of the solids in the effluent, with no apparent effect on the removal efficiency. For example, when the filter was operated at 0.65 Kg/m3/day, at temperature of 30^{o}C, and an HRT of 12.9 hours, it was found that the effluent total solids concentration was 36 mg/L when the top of the sludge bed was at a level just about the first sampling part in the upper zone. The removal of solids in the filter to bring the sludge bed level to the desired level was done twice, once at the beginning of the first phase of operation (end of week 7) and again at the beginning of the third phase (beginning of week 52). The first time solids were wasted was to remove the excess solids due to the heavy seeding. The second time was to remove excess solids that were built-up due to substrate synthesis. The long period of operation between the first and the second time solids were wasted, is indicative of low biological synthesis, which is a major advantage of anaerobic treatment. The volatile solids profiles in the filter are shown in figures 3.30 through 3.47. The total and volatile solids concentration

profiles are also reported in tables 3.6, 3.7, 3.8, 3.9. From the solids analysis results the following can be reported:

1) When the sludge bed was contained in the lower zone, very low total and volatile solid values were found in the effluent. This confirms the fact that the lower zone is a trap area for the biosolids.

2) The concentration of solids in the lower zone was high and varied slightly during the different runs. The variation of the total solids concentration ranged from 1.5 % for the first run in phase one to 2.4 % for run 6 in phase (3). The reason for the low solids concentration value in run one phase one, is the high gas production that bubbles and fluidizes the solids in the lower zone. For run 6, phase 3, the gas production was low and little fluidization occurred. In general, the solid concentrations in the lower zone increased as temperature decreased (the lower gas production at the lower temperature reduces the fluidization of the solids). The solids concentrations also increased as organic loading was increased (filter was being gradually overloaded.

3) In the upper zone, the solid concentrations were much lower than those in the lower zone. In general they followed the trend of the solids concentrations in the lower zone. They increased when gas production was high, and decreased otherwise.

4) It appears that the more biogas produced the more the solids are fluidized in the reactor. This usually results in smaller mass concentration of suspended solids per unit volume.

5) The filter effluent was normally clear with very light grayish colored liquid, and contained very low concentration of suspended solids. Samples removed from various filter heights were low in suspended solids at all organic loadings.

In summary, the filter was effective in digesting and trapping biosolids, as indicated by the low volatile suspended solids concentrations in the effluent. In fact the SS

in the effluent never exceeded those of secondary effluent criteria for publicly owned treatment facilities as required by EPA.

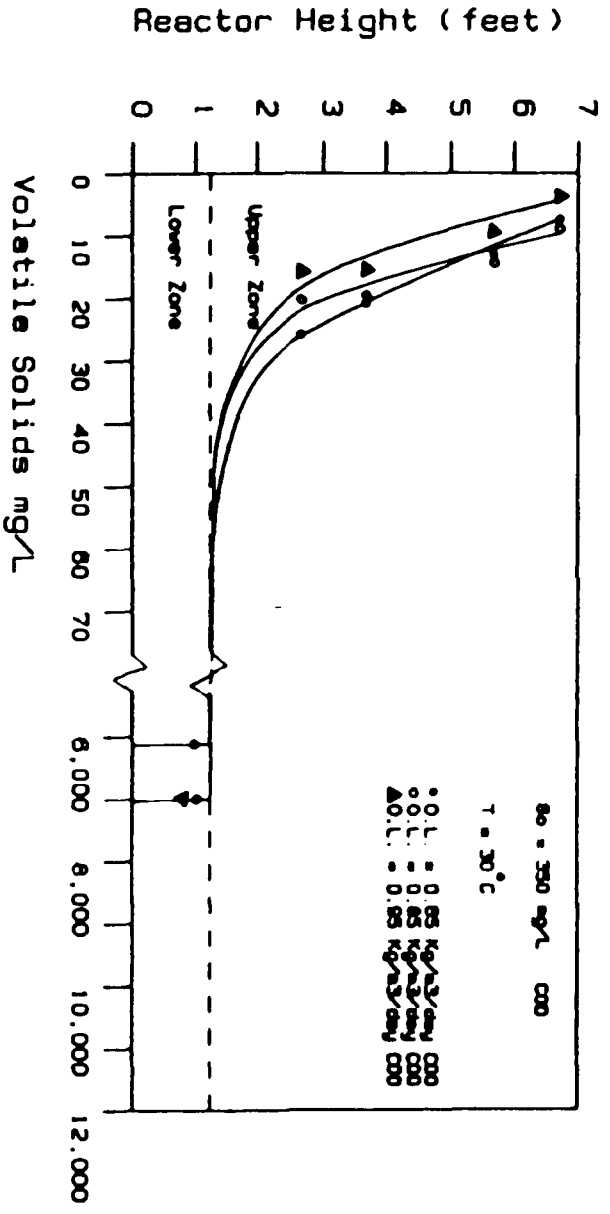


FIGURE 3.14:
VOLATILE SOLIDS DISTRIBUTION IN REACTOR.

FIGURE 3.15:

VOLATILE SOLIDS DISTRIBUTION IN REACTOR.

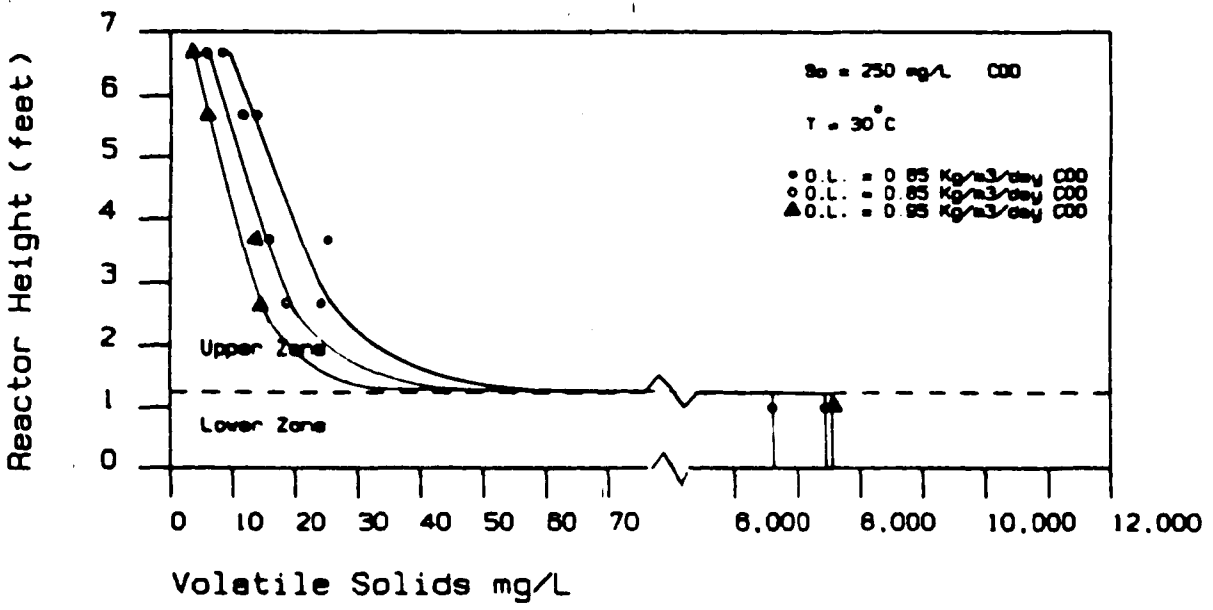


FIGURE 3.16:

VOLATILE SOLIDS DISTRIBUTION IN REACTOR.

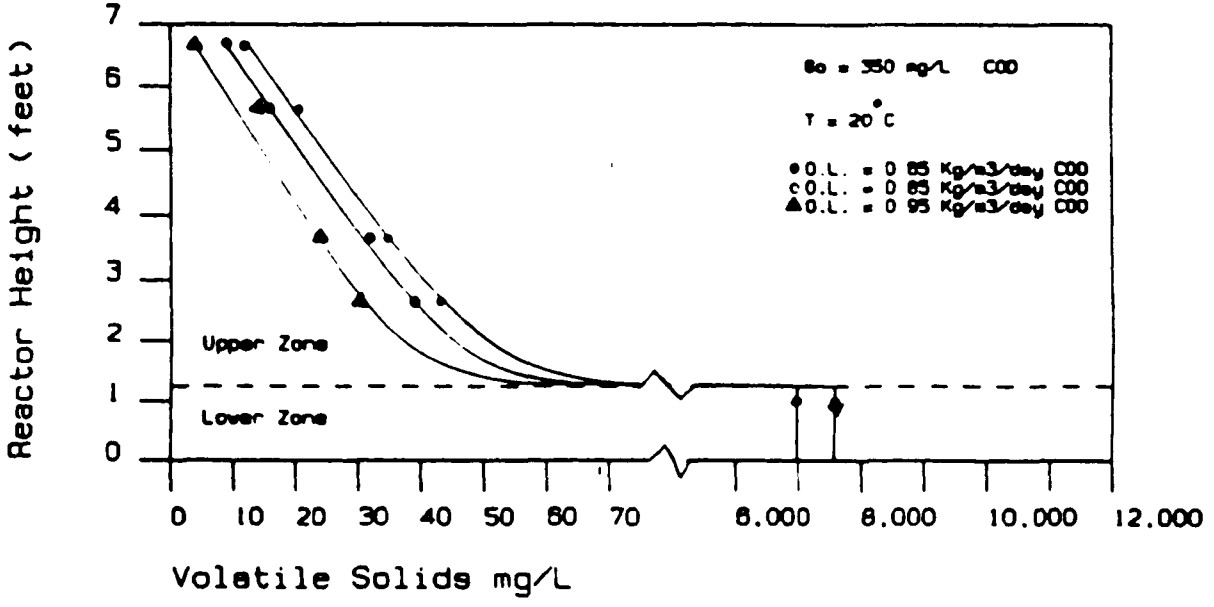


FIGURE 3.17:

VOLATILE SOLIDS DISTRIBUTION IN REACTOR.

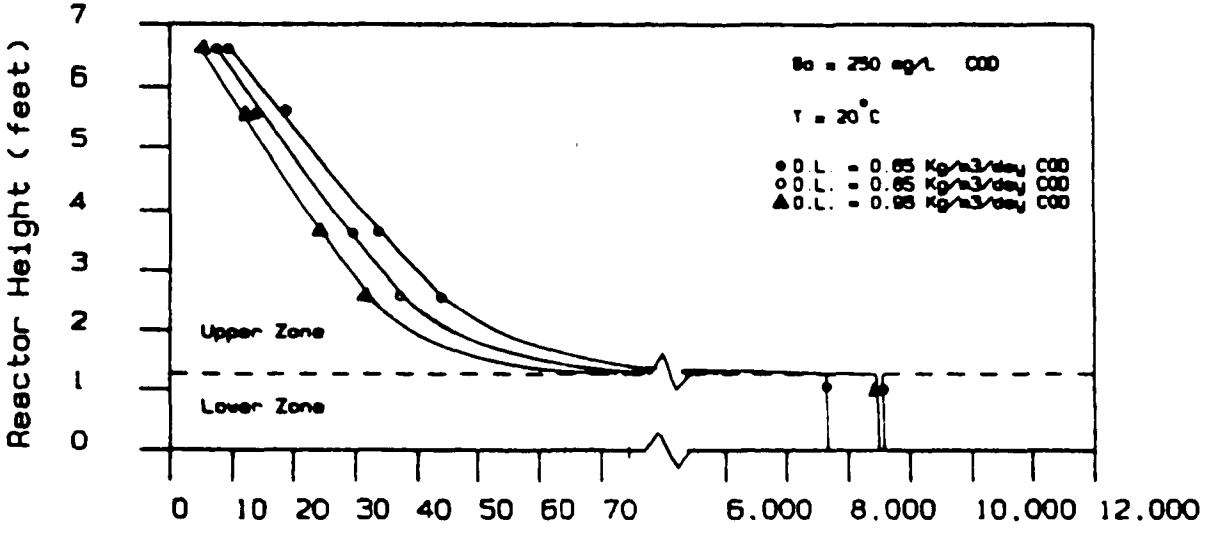


FIGURE 3.18:

VOLATILE SOLIDS DISTRIBUTION IN REACTOR.

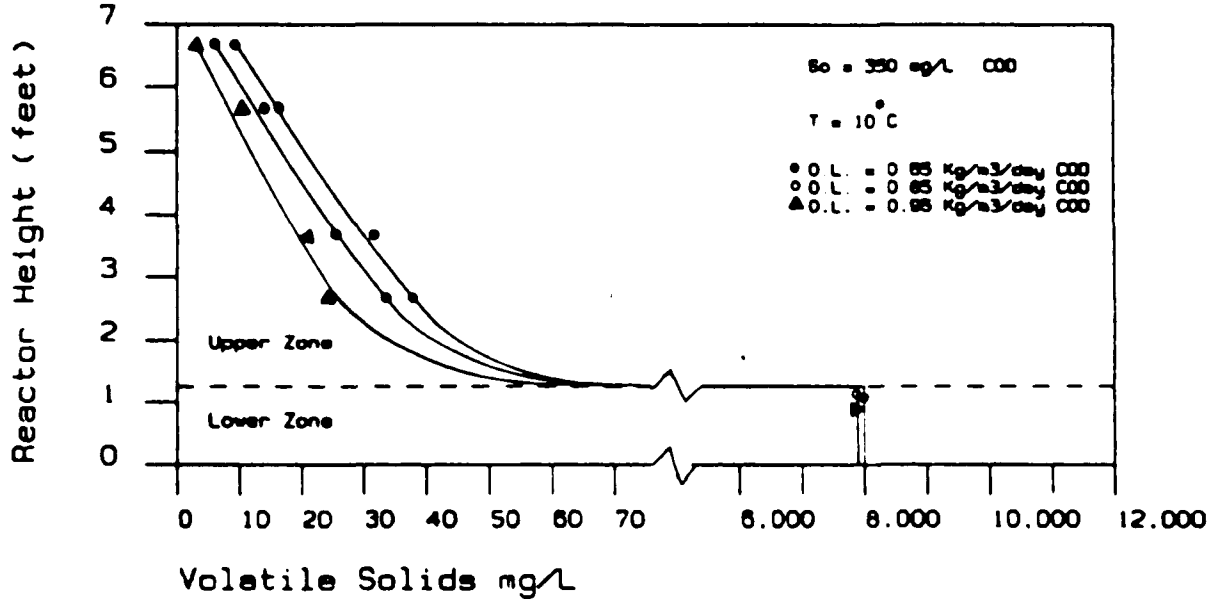


FIGURE 3.19:
VOLATILE SOLIDS DISTRIBUTION IN REACTOR.

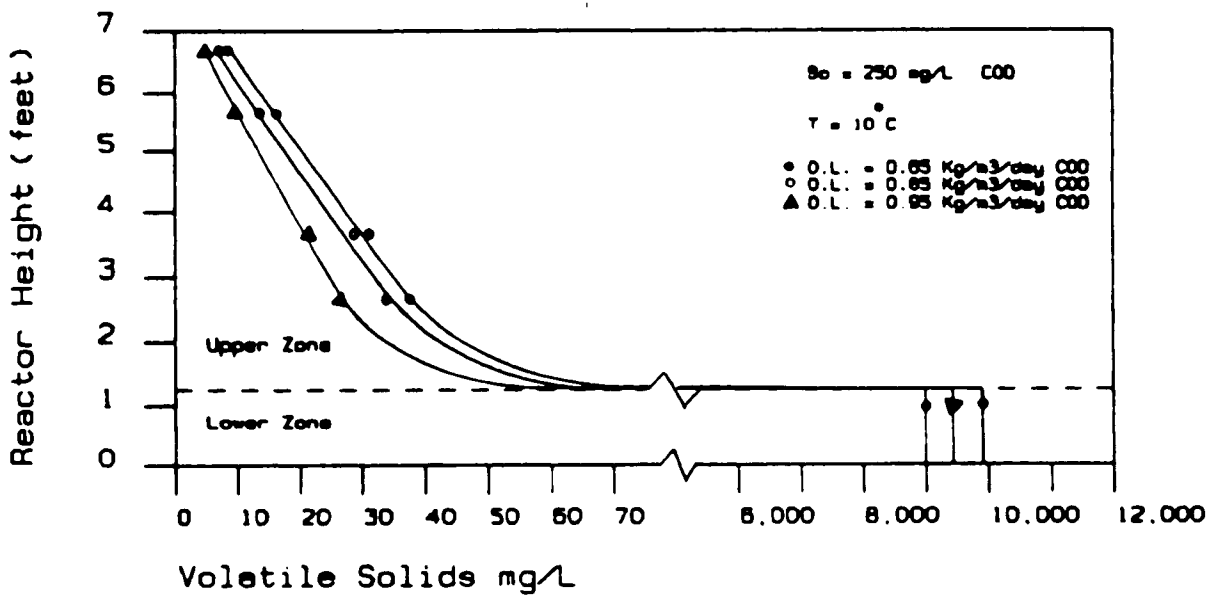


TABLE 3.6: PERCENT TOTAL SUSPENDED SOLIDS IN LOWER ZONE, AS WELL AS, PERCENT VOLATILE SUSPENDED SOLIDS IN THE TOTAL.

Phase No.	Run No.	Total Suspended Solids %	Percent Volatile Solids
1	1	1.5	41
1	2	1.6	42
1	3	1.9	37
1	4	2.0	37.5
1	5	2.0	35
1	6	2.1	36
2	1	1.7	41
2	2	1.8	43
2	3	2.0	38
2	4	2.2	39.5
2	5	2.1	36
2	6	2.3	37
3	1	1.9	42
3	2	2.0	45
3	3	2.0	39
3	4	2.3	43
3	5	2.1	37
3	6	2.4	39

TABLE 3.7: PROFILES OF THE TOTAL RESIDUE, AND THE PERCENT VOLATILE SOLIDS IN THE TOTAL RESIDUE IN THE UPPER ZONE OF THE REACTOR, FOR THE DIFFERENT RUNS OF PHASE ONE.

REACTOR HEIGHT FEET	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6
			TOTAL SUSPENDED SOLIDS mg/L / % VOLATILE SUSPENDED SOLIDS			
2.67	67 / 38	61 / 39	55 / 37	48 / 38	48 / 34	40 / 36
3.67	53 / 40	57 / 43	47 / 42	39 / 41	41 / 39	35 / 38
5.67	32 / 41	29 / 48	29 / 50	25 / 46	20 / 50	15 / 40
6.71	18 / 42	16 / 50	16 / 57	12 / 47	13 / 60	9 / 42

TABLE 3.8: PROFILES OF THE TOTAL RESIDUE, AND THE PERCENT VOLATILE SOLIDS IN THE TOTAL RESIDUE IN THE UPPER ZONE OF THE REACTOR, FOR THE DIFFERENT RUNS OF PHASE TWO.

REACTOR HEIGHT FEET	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6
	TOTAL SUSPENDED SOLIDS mg/L					
	% VOLATILE SUSPENDED SOLIDS					
2.67	95 / 45	77 / 56	92 / 42	74 / 51	86 / 35	65 / 49
3.67	75 / 47	60 / 57	75 / 43	58 / 52	64 / 37	50 / 50
5.67	41 / 51	35 / 58	35 / 46	29 / 52	33 / 41	28 / 50
6.71	22 / 55	19 / 59	18 / 48	17 / 53	12 / 43	13 / 49

TABLE 3.9: PROFILES OF THE TOTAL RESIDUE, AND THE PERCENT VOLATILE SOLIDS IN THE TOTAL RESIDUE IN THE UPPER ZONE OF THE REACTOR, FOR THE DIFFERENT RUNS OF PHASE THREE.

REACTOR HEIGHT FEET	RUN 1	RUN 2	RUN 3	RUN 4	RUN 5	RUN 6
			TOTAL SUSPENDED SOLIDS mg/L			
			% VOLATILE SUSPENDED SOLIDS			
2.67	92 / 41	73 / 52	86 / 38	70 / 49	74 / 32	62 / 44
3.67	75 / 43	58 / 53	66 / 40	58 / 50	59 / 35	51 / 43
5.67	37 / 47	32 / 54	35 / 44	28 / 50	28 / 38	24 / 43
6.71	20 / 50	17 / 55	15 / 46	16 / 49	11 / 40	13 / 47

Gas Production

a) Actual Gas Production

Generally speaking the gas production rates found in this study (shown in table 3.10) were relatively small, due to the small size of the reactor, as well as the relatively low organic loadings used. The fact that the gas production was low, it was difficult to find a gas meter that is specifically designed to measure the small gas flow rates anticipated in the study. Two methods were tried to measure the gas accurately : 1) A displacement method, and 2) Using a wet gas meter below its specified range. The first method proved to be cumbersome as well as inaccurate. The second method was found to be more flexible in terms of ease of operation and measurement. However, by using the meter below its minimum recommended flowrate, it is anticipated that the gas values, reported here, are not very accurate. At best, they show the relative variations in quantities or rates of biogas produced in the different runs of the experiment. On the other hand, the constituents of the biogas produced in the reactor were measured much more accurately using gas chromatography.

b) Theoretical Gas Production

The quantity of methane produced can be estimated at Standard Temperature and Pressure (zero ^{o}C & 1 atm. pressure) using the following stoichiometric relationship:

$$CH_4 + 2O_2 \longrightarrow CO_2 + 2H_2O \qquad [3.1]$$

At Standard Temperature and Pressure (STP) the law of ideal gases states that one mole of gas occupies a volume of 22.414 liters. Hence, equation (3.1) shows that at STP the methane produced by anaerobic digestion is 0.382 Liter/gm of soluble COD removed. HASSAN (P93)

To estimate the methane production at conditions other than STP, the combined Charles-Boyle equation may be used:

$$\frac{P_1 V_1}{T_1} = \frac{P_2 V_2}{T_2} \qquad [3.2]$$

where:

P = Absolute pressure, Force/Area

V = Gas volume, volume

T = Absolute temperature, degree Kelvin

Assuming that the pressure inside the reactor is near atmospheric, and changes slightly in the range of operating conditions used, P1 can be considered approximately equal to P2. Hence, equation (3.2) can be written as:

$$\frac{V_1}{T_1} = \frac{V_2}{T_2} \qquad [3.3]$$

Volume of the total gas produced can be estimated from the following equation:

$$G_r = \frac{C_g\ Q_o\ (S_o - S_e)}{E} \qquad [3.4]$$

where:

G_r = Total gas production rate, volume/time

C_g = Coefficient of the methane produced per unit soluble COD removed, volume/mass

S_o = Soluble COD concentration in influent, mass/volume.

S_e = Soluble COD concentration in the effluent, mass/volume.

Q_o = Volumetric flowrate of substrate into the filter, volume/time.

E = Fraction of methane in the total gas produced, volume/volume.

Using equations 3.1 and 3.3 the theoretical methane production for all run conditions was estimated and presented in table 3.11. In addition, considering the results of the gas chromatography analyses (table 3.10) and using equation 3.4 an estimate of the total gas productions for the different runs was made and presented in table 3.12. Table 3.12, also, shows the measured values for comparison.

TABLE 3.10: THE GAS PRODUCTION AND CONSTITUENTS IN THE REACTOR.

T °C		S_o = 350 COD mg/L					S_o = 250 COD mg/L				
	Organic Loading COD Kg/m3/day	Soluble COD Removal %	Gas Produced Liters per gm soluble COD rem.	CH_4 %	CO_2 + N_2 %	H_2S %	Soluble COD Removal %	Gas produced Liters per gm soluble COD rem.	CH_4 %	CO_2 + N_2 %	H_2S %
30	0.65	97	0.44	98	1	T	94	0.26	96	4	T
	0.85	92	0.50	95	4	T	83	0.30	92	8	T
	0.95	87	0.53	87	12	T	77	0.35	83	16	T
20	0.65	90	0.32	92	7	T	85	N/A	88	10	T
	0.85	84	0.44	90	9	T	75	N/A	85	14	T
	0.95	76	0.46	78	20	T	64	N/A	75	24	T
10	0.65	83	0.15	78	21	T	71	N/A	74	24	T
	0.85	74	0.22	71	28	T	63	N/A	66	33	T
	0.95	69	0.25	67	32	T	55	N/A	60	38	T

NOTE: T ≡ Trace amount.

TABLE 3.11: THEORETICAL METHANE PRODUCTION FOR ALL RUN CONDITIONS.

T °C		S_e = 350 mg/L COD	S_e = 250 mg/L COD	S_e = 350 or 250 mg/L
	Organic Loading COD Kg/m3/day	methane Produced liters per day	methane Produced liters per day	Liter methane per gm COD removed
30	0.65	6.519	6.327	0.389
	0.85	8.098	7.640	0.389
	0.95	8.544	6.886	0.389
20	0.65	5.989	5.787	0.376
	0.85	7.147	6.534	0.376
	0.95	7.227	5.211	0.376
10	0.65	4.774	4.698	0.363
	0.85	5.327	5.257	0.363
	0.95	5.324	4.590	0.363

TABLE 3.12: THEORETICAL TOTAL GAS PRODUCTION FOR ALL RUN CONDITIONS.

T ^{o}C	Organic Loading COD Kg/m3/day	S_e = 350 mg/L COD theoretic. gas Production liters per day	S_e = 250 mg/L COD theoretic. gas Production liters per day
30	0.65	6.652	6.591
	0.85	8.524	8.304
	0.95	9.821	8.296
20	0.65	6.510	6.576
	0.85	7.260	7.687
	0.95	9.265	6.948
10	0.65	6.121	6.349
	0.85	7.503	7.965
	0.95	7.946	7.650

TABLE 3.13: TOTAL GAS PRODUCTION PER GRAM COD REMOVED ESTIMATED VS. MEASURED.

T °C	Organic Loading COD Kg/m3/day	S_e = 350 mg/L COD		S_e = 250 mg/L COD	
		total gas produced liters per gm COD removed		total gas produced liters per gm COD removed	
		estimated	measured	estimated	measured
30	0.65	0.397	0.325	o.405	0.324
	0.85	0.409	0.280	0.423	0.276
	0.95	0.447	0.270	0.469	0.260
20	0.65	0.409	0.355	0.427	0.365
	0.85	0.418	0.308	0.442	0.310
	0.95	0.482	0.297	0.501	0.292
10	0.65	0.465	0.426	0.491	0.445
	0.85	0.511	0.354	0.550	0.365
	0.95	0.542	0.350	0.605	0.361

The effect of varying the organic loading rate, as well as COD concentration on the gas production for the different temperature values are shown in figures 3.20, 3.21, 3.22. The percent of methane in the gas produced in the reactor, corresponding to the different organic loadings and COD concentrations for the three temperature values used in this study, are shown in figures 3.23 through 3.24. Considering the gas production results the following may be reported:

1) The constituents of the gas produced were measured with high accuracy using gas chromatography. It was found that in all runs and phases methane constitutes the highest fraction of biogas (by volume), as seen in table 3.9. The biogas was found to contain the following components,listed in increasing order of magnitude by volume.

    a) hydrogen sulfide (trace amount)
    b) carbon dioxide
    c) nitrogen
    d) methane

2) In general the methane percent in the biogas produced decreased as:

    a) organic loading increased
    b) influent COD value decreased
    c) temperature decreased

It can be seen that the percent methane decreased at an accelerating rate as the organic loading increases, while the temperature and the COD concentration in the influent were kept constant. For example,the rate of change of the slope of the curve between organic loadings of 0.65 and 0.85 kg/m3/day was slower than that between 0.85 and 0.95. It can also be seen that the rate of decrease in the methane percentage varies as organic loading changes and is higher for the lower influent COD value. Temperature decrease appears

to decelerate the rate of decrease of the percent methane in the total gas as organic loadings increase.

FIGURE 3.20:
GAS PRODUCTION VERSUS ORGANIC LOADING RATE.

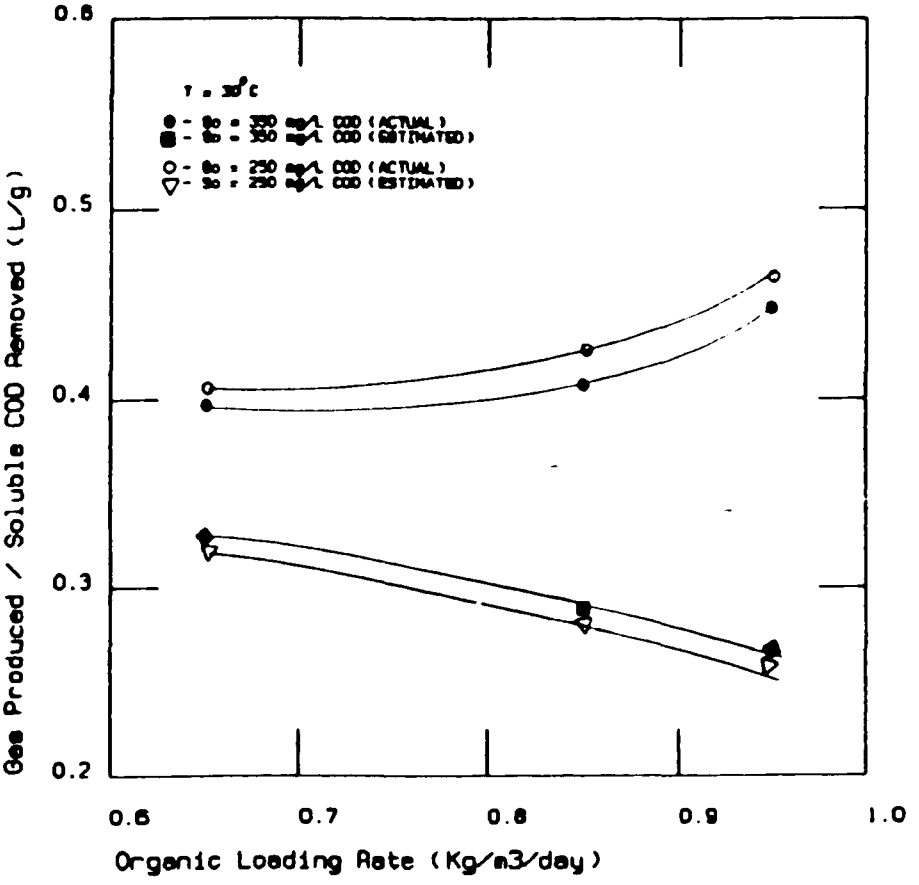


FIGURE 3.21:
GAS PRODUCTION VERSUS ORGANIC LOADING RATE.

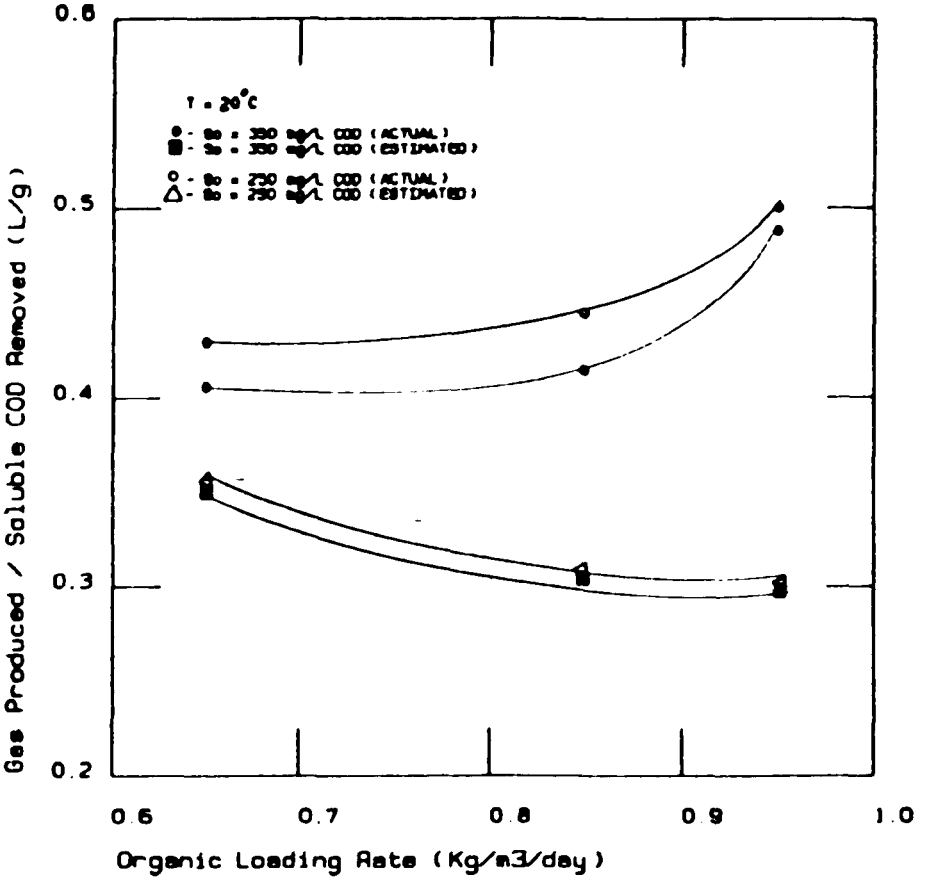


FIGURE 3.22:
GAS PRODUCTION VERSUS ORGANIC LOADING RATE.

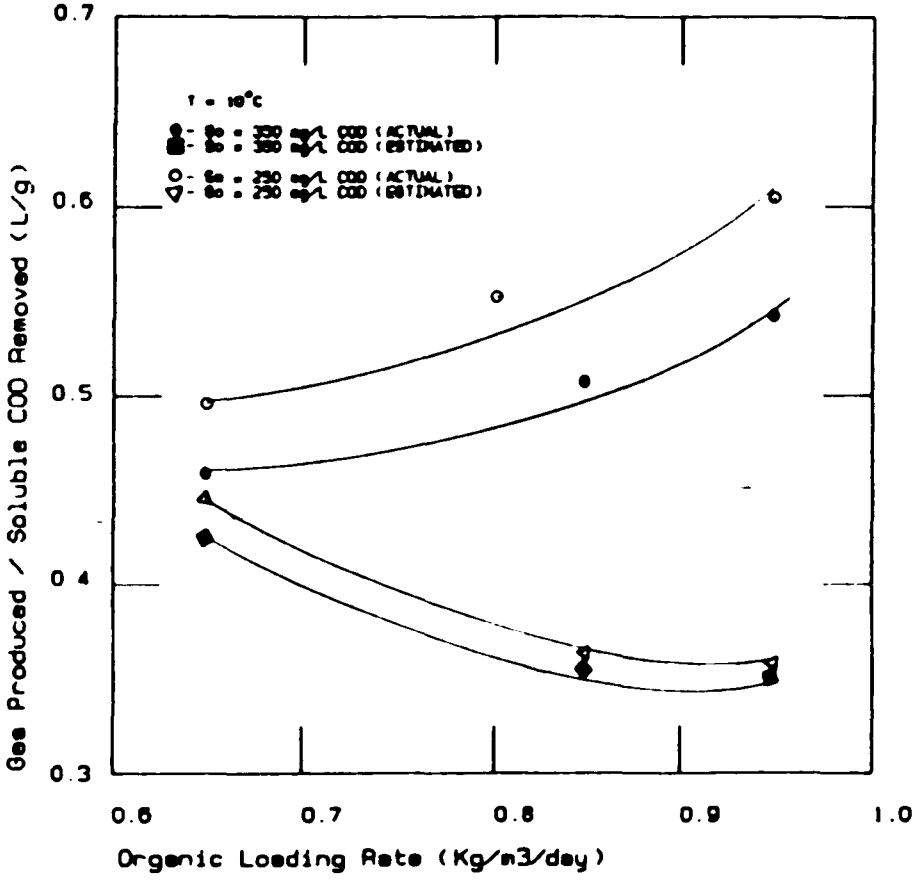


FIGURE 3.23:

ORGANIC LOADING VERSUS METHANE PRODUCTION.

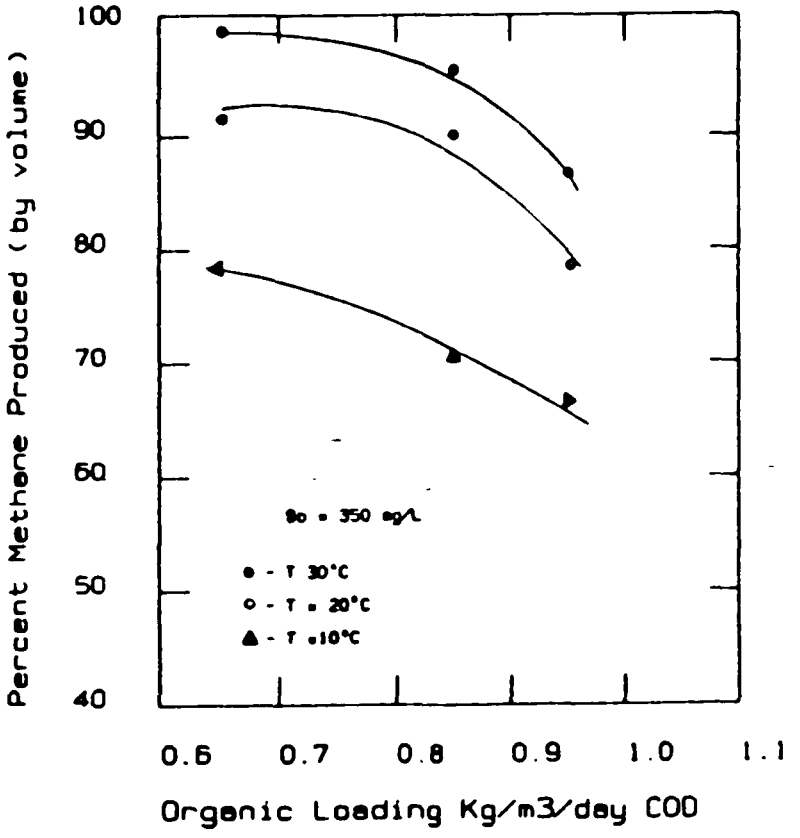
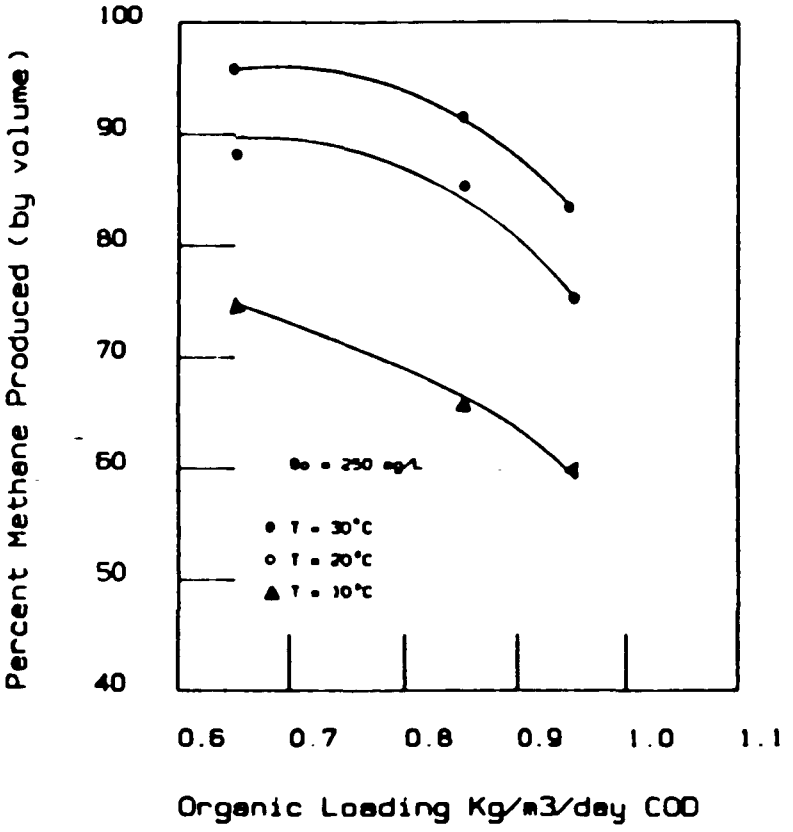


FIGURE 3.24:

ORGANIC LOADING VERSUS METHANE PRODUCTION.



Methane Solubility

Dalton's law for partial pressures and Henry's law for gas solubilities were used to calculate the amount of methane lost in the effluent due to its solubility in water. Dalton's law states that "In a mixture of gases, each gas exerts pressure independently of the others. The partial pressure of each gas is proportional to the percent by volume of that gas in the mixture."

Henry's law states that "In a given volume of a liquid, the weight of any gas that will dissolve in it is directly proportional to the partial pressure of that gas on the surface of the liquid."

In symbolic form Henry's law can be written as,

$$P_g = \alpha Z_g$$

Where,

Z_g = equilibrium mole fraction of the dissolved gas.

$$= \frac{\text{mole gas } (m_g)}{\text{mole gas } (m_g) + \text{mole water}(m_w)}$$

α = Henry's law constant, mass/[volume/(force/area)]

P_g = the partial pressure of gas above the liquid, mass/[length(time/time)]

Based on these two laws the amounts of methane lost in the effluent, as well as the total methane produced in the reac-

tor for all the different runs of the experiment were calculated, (see tables 3.15 & 3.16). For example, for the first run of Phase one, and assuming that the total gas pressure inside the reactor was one atmosphere, the percent methane lost in the effluent is calculated as follows:

$P_g = 0.97$ (The biogas contained 97 percent methane, see table 3.10)

α for methane at $30^oC = 4.49 \times 10^4$ atm/mol fraction

and,

$$Z_g = \frac{P_g}{\alpha} = \frac{0.97}{4.49 \times 10^4} = 2.16 \times 10^{-5}$$

Each liter of water contains 1000/18 or m_m = 55.56 gm mole H_2O

Thus,

$$\frac{\text{No. of methane moles in water}}{\text{No. of methane moles in water + No. of water moles}}$$

= equilibrium mole fraction of methane gas.

or

$$\frac{m_m}{m_m + m_w} = 2.16 \times 10^{-5}$$

or

$$m_m = m_m \times (2.16 \times 10^{-5}) + m_w \times (2.16 \times 10^{-5})$$

By iteration

$m_m \approx 1.2 \times 10^{-3}$ mole methane/L

At 30°C and one atmosphere pressure, one mole of gas will occupy a volume of 24.88 Liters. Hence, volume of methane in each liter of effluent is,

$$1.2 \times 10^{-3} \times 24.88 = 0.03 \text{ Liter}$$

The volumetric flowrate (Q_o) was 49.437 L/day, hence, the total volume of methane lost in effluent is,

$$0.03 \times 49.437 = 1.48 \text{ L/day.}$$

The estimated total volume of methane produced in the reactor is 8.544 Liters (table 3.11). Therefore, the percent methane lost in the effluent is,

$$(1.48 \div 8.54) \times 100\% = 17\%$$

Henry's law constant α, varies with temperature and type of liquid. Values for Henry's constant for methane in water, which are used in this research, are referenced in standard chemical handbooks. The values are shown in table 3.14.

TABLE 3.14: HENRY'S LAW CONSTANTS FOR METHANE IN WATER.

T ^{o}C	α atm/mol fraction
30	4.49×10^4
20	3.79×10^4
10	2.97×10^4

TABLE 3.15: EVALUATION OF METHANE SOLUBILITY FOR S_e = 350 mg/L.

T °C		S_e = COD 350 mg/L		
	Organic Loading Kg COD/m3/day	methane produced L/day	methane lost in the effluent L/day	percent methane lost
30	0.65	6.519	1.48	22.7
	0.85	8.098	1.83	22.6
	0.95	8.544	1.94	22.7
20	0.65	5.989	1.58	26.4
	0.85	7.147	1.93	27.0
	0.95	7.227	1.95	27.0
10	0.65	4.774	1.78	37.3
	0.85	5.327	2.08	39.0
	0.95	5.324	2.17	40.8

TABLE 3.16: EVALUATION OF METHANE SOLUBILITY FOR S_e = 250 mg/L.

T °C		S_e = 250 mg/L COD		
	Organic Loading Kg COD/m3/day	methane produced L/day	methane lost in the effluent L/day	percent methane lost
30	0.65	6.327	2.00	31.6
	0.85	7.640	2.31	30.2
	0.95	6.886	2.40	34.9
20	0.65	5.787	2.09	36.1
	0.85	6.534	2.41	36.9
	0.95	5.211	2.30	44.1
10	0.65	4.698	2.14	45.6
	0.85	5.257	2.48	47.2
	0.95	4.590	2.42	52.7

DISCUSSION OF EXPERIMENTAL RESULTS

The Two-Zone Two-Growth Up-Flow Anaerobic Biofilter, used in this study, demonstrated high COD removal efficiencies for low wastewater strengths, and for different combinations of organic loading and temperature values. The filter has demonstrated that it is capable of removing organic pollutants to levels below those required by NPDES. The filter achieved removal standards as required by NPDES under the following conditions, (see table 3.5):

1- Organic loading of 0.65 COD kg/m^3/day and temperature of 30^{o}C with influent COD concentration of 350 mg/L (phase 1, run 1)

2- Organic loading of 0.65 COD kg/m^3/day and temperature of 30^{o}C with influent COD concentration of 250 mg/L (phase 1, run 2)

3- Organic loading of 0.65 kg COD/m^3/day and temperature of 20^{o}C with influent COD concentration of 350 mg/L (phase 1, run 3)

4- Organic loading of 0.65 kg COD/m^3/day and temperature of 20^{o}C with influent COD concentration of 250 mg/L (phase 1, run 4)

5- Organic loading of 0.85 kg COD/m^3/day and temperature of 30^{o}C with influent COD concentration of 350 mg/L (phase 2, run 1)

For all other run conditions, the filter still achieved high removal efficiencies. In general, the filter removal efficiency increases when a) temperature increases b) the influent COD concentration increases c) organic loading de-

creases. In fact, the removal efficiency never dropped below 50%. The methane produced (see tables 3.10, 3.11, 3.12 & 3.13) may insure enough energy to heat up the filter in order to achieve higher removal efficiencies. The filter has a simple design and requires no effluent or solid recycling. Sludge build up was very small. The filter was operated for more than a year before sludge removal was needed. An accurate figure of solids build up is very difficult to estimate due to the numerous samples withdrawn from the reactor for laboratory analysis as the experiment proceeded.

Reactor response to changes in run conditions such as organic loading and temperatures were rapid. Less than a week was required to reach a new steady state COD removal every time the run conditions were changed.

These results suggest that the filter is suitable for treatment of shock loads as well as intermittent wastewater discharges.

A significant design advantage in the filter is its high void volume. This can be utilized in the treatment of colloidal suspension such as starches and dilute milk without fear of clogging. The filter would also treat wastewaters with low concentrations of degradable suspended solids without clogging problem.

Although Biochemical Oxygen Demand (BOD) was not measured in this experiment, and since acetate is a simple organic

compound, it can reasonably be said that the BOD of acetates wastes is fairly close to its COD values. This assumption can be used to demonstrate that the filter compares favorably with other anaerobic treatment processes in terms of organic loading applied and removal achieved. For example, for common activated sludge processes, Stewart, referred to by McCarty (60), reported organic loadings ranged from 0.32 to 2.40 kg BOD5/m3/day (20 to 150 lb BOD5/1000 ft3/day) based on aeration tank capacity with a removal from 60 to 95 percent.

CHAPTER FOUR

Reactor Mathematical Modeling

The second goal of this research is to develop a mathematical model that can be used to provide a better understanding of the factors affecting the treatment kinetics of the reactor. The model may be used to predict the performance of the reactor under various operating conditions. The predicted performance is especially important when a full scale reactor is being designed. For effective purification of wastewaters, and good process reliability, the reactor should ultimately be designed for optimum loading and realistic operating conditions. A summary of the quantitative mathematical descriptors needed to develop the model is presented in this chapter.

The model contains, as its principle elements, a mathematical description of the dynamic behavior and distribution of both the fluid and the anaerobic sludge in the reactor, and a quantification of the kinetics of the anaerobic conversion of the organic waste into methane gas. The model as presented here might be modified to represent the data that would be collected from a proposed pilot plant anaerobic system study. The model may then be used to predict the optimum configuration, in terms of removal, of a full scale reactor.

Set up of the Mathematical Description of the Process

A mass balance for the substrates, gases and biosolids that are involved in the process, has been formulated mathematically based on the following assumptions:

a) the fluid flow pattern in the reactor,

b) the distribution and behavior of sludge in the reactor, and

c) the kinetics of the biological conversion of substrate and subsequent formulation of methane.

From earlier studies reported in the literature on fluid-flow patterns in upflow reactors (17), it can be concluded that the flow regime in the upper zone of the filter, where the plastic media is located, can be adequately described as ideal plug-flow. On the other hand, two sequential fully mixed reactors can be used to model the flow-pattern in the lower zone of the reactor. In the general case the biomass in the lower zone is distributed through a suspended sludge bed and a suspended sludge blanket. In the sludge bed the solids concentrations is high (ca 20 kg SS/m^3) and does not vary over a large range of process conditions such as flow-rate and organic loading rates. In the blanket, which is located above the bed, the sludge concentration is lower and depends on the process operating conditions. In the studies on the fluid-flow pattern in upflow reactors mentioned above, it was found that both the sludge bed and the sludge blanket could be described as separate and independently well mixed flow regions. A portion of the influent bypassing

the sludge bed and flowing into the sludge blanket may be assumed. The bypassing flow occurs essentially along the reactor walls. The reactor used in this study was operated in such a way that a zone of high solid concentration would completely fill the lower zone. That is, the lower zone would be fully filled with the sludge bed. Hence, all calculations for the sludge blanket were eliminated from the mathematical model. The original model for the fluid flow pattern is shown in figure (4.1) and is in agreement with these earlier studies (17 & 18). Figure 4.1 also shows the simplified model that represents more closely the experimental results in this research effort.

For the model used in this research the following relation holds:

$$V_r = V_{bd} + V_u$$

where,

V_r = Liquid volume of reactor, volume

V_{bd} = Volume of sludge bed, volume

V_u = Voids or liquid volume of upper zone, volume

Q_o = Influent flowrate, volume/time

Substrate removal in the reactor

Substrate removal in the reactor was modeled by considering the mass balance of substrate over each of the two reactor zones as explained in the following sections:

1) Substrate Removal in the Lower Zone

With reference to the model presented in figure (4.2), the substrate removal within the lower zone may be obtained by considering the mass balance of the substrate within the sludge bed as follows,

1-General word statement:

Rate of removal of substrate within the sludge bed. = Rate of flow of substrate into the sludge bed.

- Rate of flow of substrate out of the sludge bed.
- Rate of disappearance (utilization) of substrate within the sludge bed.

+ Rate of generation of substrate within the sludge bed.

The substrate in this model is defined as the soluble COD.

2- Simplified word statement:

(a) Removal = Inflow - Outflow + Generation - Utilization

or

(b) Removal = Inflow - Outflow + Net mass change due to transformation

3- Symbolic representation for 2(b):

$$V_{bd} \frac{ds}{dt} = Q_o S_o - Q_o S_{bd} + V_{bd} \left(- \frac{K X_{bd} S_{bd}}{K_s + S_{bd}}\right) \quad [4.1]$$

where the net mass change due to transformation per unit volume

is approximated by the expression $- \frac{K X_{bd} S_{bd}}{K_s + S_{bd}}$ which is used by many researchers (3, 13).

where:

S_o = Substrate concentration in influent, mass/volume

S_{bd} = Substrate concentration in bulk solution in the sludge bed, mass/volume

Q_o = influent flowrate, volume/time

V_{bd} = Volume of sludge bed, volume

X_{bd} = Concentration of suspended biomass, in sludge bed mass/volume.

K = Maximum specific rate of substrate utilization, 1/time

K_s = Concentration of substrate at which substrate removal is one half the maximum, mass/volume.

At steady state conditions,

$$\frac{dS}{dt} = 0$$

Hence, equation (4.1) reduces to

$$0 = Q_o S_o - Q_o S_{bd} - V_{bd} \left(\frac{K X_{bd} S_{bd}}{K_s + S_{bd}}\right) \quad [4.2]$$

Rearranging equation 4.2,

or

$$\frac{K_S + S_{bd}}{K\ S_{bd}} = \left(\frac{V_{bd}}{Q_o} \right) \frac{X_{bd}}{(S_o - S_{bd})} \qquad [4.3]$$

Knowing the values of S_o, X_{bd},-V_{bd}, K, K_s, Q_o the equation can be iterated to obtain the value of S_{bd}.

Substrate Removal Efficiency In the sludge bed:

$$E_{bd} = \frac{S_o - S_{bd}}{S_o} \times 100\%$$

From equation 4.3 we can see that

$$E_{bd} = \frac{\theta_L}{S_o} \left(\frac{K\ X_{bd}\ S_{bd}}{K_s + S_{bd}} \right) * 100\% \qquad [4.4]$$

or

$$E_{bd} = \frac{V_{bd}}{S_o\ Q_o} \left(\frac{K\ X_{bd}\ S_{bd}}{K_s + S_{bd}} \right) * 100\% \qquad [4.5]$$

It can be seen from equations 4.4 and 4.5 that the removal efficiency (E_{bd}) is directly proportional to the hydraulic retention time (θ_L), increasing when θ_L increases and decreasing when θ_L decreases. In addition, E_{bd} is inversely proportional to the influent COD concentration (S_o), increasing as S_o decreases and vice versa.

FIGURE 4.1:
GENERAL MODEL OF FLUID FLOW PATTERN IN REACTOR.

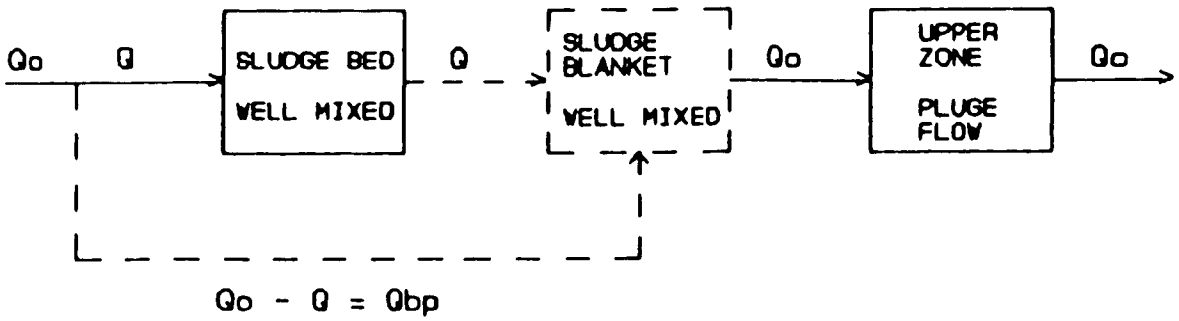
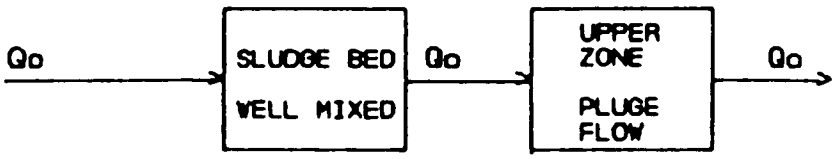


FIGURE 4.2:
MODEL OF FLUID FLOW PATTERN USED IN THIS STUDY.



2) Substrate Removal in the Upper Zone

Substrate removal relationships, at steady state conditions, can be derived through a mass balance across a differential height of the upper zone of the reactor, as illustrated in figure (4.3) on page 131. Note, also, that inside the differential volume dV (i.e. A dH) the substrate concentration is approximately equal to $S_u + dS_u/2$. At any rate, $(dV * S_u)$ has a much greater value than $(dV * dS_u/2)$ and the term $dV * dS_u/2$ is, therefore, neglected in following mass balance equations.

- Rate of removal of substrate within a differential volume dV in upper zone.
- Rate of flow of substrate out of dV

= Rate of flow of substrate into dV

- Net mass change due to transformation

or

dV

Rate of removal of substrate within a differential volume dV in upper zone

= Rate of flow of substrate into the differential volume

- Rate of flow of substrates out of dV
- Substrate removed by attached biomass in dV

- Substrate removed by suspended biomass in dV

or

$$\frac{dS_u}{dt}\, dV = Q_o\, S_u - Q_o\, (S_u + \frac{dS_u}{dH}\, dH) -$$

$$\epsilon\, dV\, \left(\frac{K\, X_{su}\, S_u}{K_s + S_u} \right) - a\, j\, dV \qquad [4.6]$$

where:

a = Specific surface area of media, area/volume

A = Cross sectional area of reactor, area

H = Height (depth) of reactor, length

j = Flux of substrate into the bioflim per unit surface area , (mass/area)/time

K = Maximum specific substrate utilization rate by suspended biomass, 1/time

K_S = Half velocity coefficient of suspended biomass, mass/volume

S_u = Substrate concentration in bulk solution in the upper zone, mass/volume

ϵ = Porosity

X_{su} = Concentration of suspended biomass, mass/volume

Based on a biofilm model developed by Williamson and McCarty (54,55), Rittmann and McCarty (41, 42, 43 & 44) formulated a variable order model for the substrate removal by attached biomass in the anaerobic filter process. Their model provides a mathematical means to evaluate the term (j) in equation 4.6 and was, therefore, used here. Incorporating the expression developed by McCarty and Rittmann for(j) in equation 4.6, we obtain:

$$\frac{dS_u}{dt} = Q_o S_u - Q_o \left(S_u + \frac{dS_u}{dH} dH\right) -$$

$$\epsilon \, dV \left(\frac{K X_{su} S_u}{K_s + S_u}\right) -$$

$$a \, C^* \, (S_u^*)^q \left(\frac{D K_s}{\tau}\right) dV \qquad [4.7]$$

Where,

τ = Standard biofilm depth, length

= $[2 K_s D_f/(K X_f)]^{\frac{1}{2}}$

C^* = Variable order reaction coefficient, dimensionless

S_u^* = Dimensionless bulk liquid substrate concentration

= S_u/K_s

q = Variable order reaction order, dimensionless

Rittmann and McCarty presented the following expressions for C^* and q :

$$C^* = \frac{2 D_f^* [(2)^{\frac{1}{2}} + 2 L^* D_f^*]^{(1-2q)}}{1 + (0.54)[1 + 0.0121 \ln(1 + 2L^*)][1 - 8.325 (\ln q/0.707)^2]}$$

and,

$$q = 0.75 - 0.25 \tanh(0.477 \beta)$$

Where,

D_f^* = Dimensionless molecular diffusivity in biofilm

= D_f/D

D_f = Molecular diffusivity in biofilm, area/time

D = Molecular diffusivity of the substrate in the liquid, area/time

L^* = Dimensionless diffusion-Layer depth

= L/τ

L = Depth of effective diffusion layer, length

β = Logarithm of an adjusted dimensionless substrate concentration

= $\ln(S_u^*) - \ln[2 + (\ln D_f^*)/2.303] -$

$1.8 \ln(1 + 2L^* D_f^*) + 0.353$

At steady state conditions, $\frac{dS}{dt} = 0$

Hence equation 4.7 becomes:

$$0 = Q_o S_u - Q_o (S_u + \frac{dS_u}{dH} dH) -$$

$$\epsilon \, dV \left(\frac{K X_{su} S_u}{K_s + S_u} \right) -$$

$$a \, C^* (S_u^*)^q \left(\frac{D K_s}{\tau}\right) dV$$

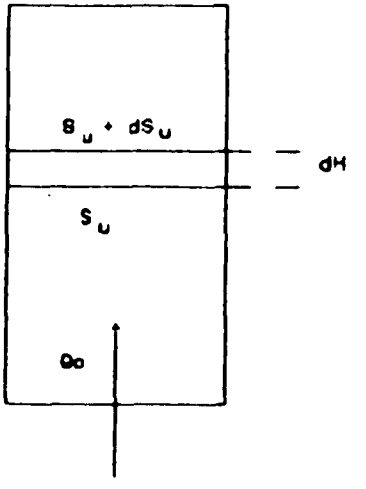


FIGURE 4.3

or

$$0 = Q_o \frac{dS_u}{dH} dH - \epsilon A \, dH \left(\frac{K X_{su} S_u}{K_s + S_u} \right) -$$

$$a \, C^* (S_u^*)^q \left(\frac{D K_s}{\tau}\right) A \, dH \qquad [4.8]$$

Rearranging equation 4.8,

$$\frac{dS_u}{dH} = \epsilon \left(\frac{A}{Q_o}\right) \left(\frac{K\ X_{su}\ S_u}{K_s + S_u}\right) -$$

$$a\ C^*\ (S_u^*)^q \left(\frac{D\ K_s}{\tau}\right) \left(\frac{A}{Q_o}\right) \qquad [4.9]$$

Equation 4.9 is used in the mathematical model developed in this research to account for the removal of the substrate in the upper zone of the reactor. A more detailed discussion about how equation 4.9 is used in the model and how the parameters used in this equation are evaluated will be presented later in this chapter.

The biofilm model for (j) as it was developed by McCarty and his coworkers does not need information about the biofilm thickness (D_f) as long as the biofilm is deep and the operating conditions are steady. A deep biofilm is a film in which the substrate concentration and the concentration gradient approach zero. In this study all the runs were operated at steady state conditions. In addition the biofilm developed in this study was found to be deep, hence no L_f values were needed. To determine whether or not the biofilm is deep, the procedure introduced by McCarty and Rittmann (41) was used. This procedure uses a formula developed by Rittmann and McCarty to calculate mathematically the lowest substrate concentration (S_{deep}) to sustain a deep biofilm, (see table 4.2 on page 149). In this study the values of

substrate concentrations (as mg/L COD) throughout the reactor, for the different run conditions, were calculate using McCarty and Rittmann's formula and found to be higher than those needed to support a deep biofilm. The formula used to calculate the S_{deep} is:

$$S_{deep} = K_s \, [4.6 \, S^*_{min\text{-}s} \, (1 + 1.414 \, D_f^* \, L^*)] \qquad [4.10]$$

For more description of the biofilm model developed by McCarty and his coworkers, the reader is referred to papers by Williamson and McCarty (54 & 55) and Rittmann and McCarty (41, 42, 43 & 44).

where,

$S^*_{min\text{-}s}$ = Minimum dimensionless substrate concentration at the biofilm surface to support a steady state biofilm
= $b^*/(2 - b^*)$

b^* = Dimensionless decay rate coefficient
$2b/(Y\,K)$

Substrate Removal Efficiency in the Upper Zone:

$$E_u = \frac{S_e - S_{bd}}{S_{bd}} \times 100\% \qquad [4.11]$$

The solution for Se requires a numerical analysis as explained earlier. Therefore, no explicit solution is available for equation 4.11 similar to that of equation 4.5 for the lower zone.

Biomass Distribution in the Reactor

To determine a model for the biomass distribution in the reactor a mass balance of the biomass was taken over each one of the two reactor zones as explained in the following sections:

1) Biomass Distribution in the Lower Zone

The suspended biomass concentration within the sludge bed varies slightly as the operating conditions change. This variation is a result of the fluidization by the upward flow velocity, and biogas lifting as well as biomass synthesis. With reference to figure (4.1) the change in suspended biomass in the lower zone may be formulated as follows:

Change in suspended biomass in the sludge bed.	= Rate of flow of microorganisms into the sludge bed.
- Rate of flow of microorganisms out of the sludge bed.	+ Net biomass growth in the sludge bed.

Net biomass growth in bed can be approximated by an equation used by a number of researchers (12) similar to:

$$\frac{dX}{dt} = Y \left(- \frac{dS}{dt} \right) - b\,X$$

where

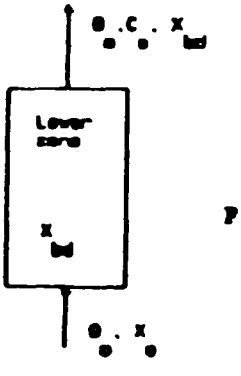


FIGURE 4.4:

$\frac{dX}{dt}$ = Growth rate of microorganisms, mass/time

Y = Growth yield coefficient, mass/mass

$-\frac{dS}{dt}$ = Rate of substrate utilization, mass/time

X = Microorganisms concentration, mass/volume

b = Decay rate, 1/time

Hence, for the sludge bed we have

$$V_{bd} \frac{dX_{bd}}{dt} = X_o Q_o - C_s Q_o X_{bd} + V_{bd} \left(\frac{X_{bd} Y K S_{bd}}{K_s + S_{bd}} - b X \right) \quad [4.12]$$

At steady state $\frac{dX_{bd}}{dt} = 0$, and assuming no active microorganisms in the influent (i.e. $X_o = 0$), we have:

$$0 = C_s Q_o X_{bd} + V_{bd} \left(\frac{Y K X_{bd} S_{bd}}{K_s + S_{bd}} - b X_{bd} \right)$$

or

$$Q_o C_s X_{bd} = - V_{bd} \left(\frac{Y K X_{bd} S_{bd}}{K_s + S_{bd}} - b X_{bd} \right) \quad [4.13]$$

or

$$C_s = - V_{bd} \left(\frac{Q_o}{X_{bd}} \right) \left(\frac{Y K X_{bd} S_{bd}}{K_s + S_{bd}} - b X_{bd} \right) \quad [4.14]$$

where,

X_{bd} = Suspended biomass concentration in the sludge bed, mass/volume

C_s = Percent of biosolids escaping the lower zoned into the upper one.

Other terms are as defined before.

Knowing the values of Q_o, X_{bd}, Y, X_{bd}, S_{bd}, K_s, b, and V_{bd} equation 4.10 can be solved for C_s. Then the value of the suspended solids concentration moving into the upper zone can be determined from:

$$X_{su} \text{ (at the beginning of the upper zone)} = C_s X_{bd} \qquad [4.15]$$

Where

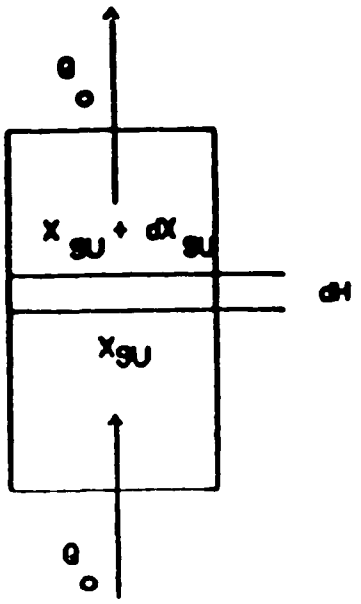
X_{su} = The concentration of the suspended solids in the bulk liquid in the upper zone.

2) Biomass Distribution in the Upper Zone

The change in suspended biomass with respect to time in the upper zone may be formulated through a mass balance across a differential height dH as follows:

Consider the differential height dH shown in figure 4.5. Note, also, that inside the differential volume dV (i.e. A dH) the concentration of X_{su} is approximately equal to $X_{su} + dX_{su}/2$. At any rate, $(dV * X_{su})$ has a much greater value than $(dV * dX_{su}/2)$ and the term $dV * dX_{su}/2$ is, therefore, neglected in following derivations.

FIGURE 4.6:



The change in the suspended biomass can be stated as:

Rate of accumulation of suspended microorganisms within the differential volume.

= Rate of flow of microorganisms into the differential volume.

- Rate of flow of microorganisms out of the differential volume.

+ Rate of suspended biomass gain due to the sloughing of the attached biomass.

+ Rate of biomass change due to settling or lifting.

+ Net rate of growth of suspended biomass.

or

$$dV \frac{dX_{su}}{dt} = Q_o X_{su} - Q_o \left(X_{su} + \frac{dX_{su}}{dH} dH \right) + F_s a L_f X_f dV +$$

$$P X_{su} dV + \epsilon A X_{su} dH \left(\frac{Y K S_u}{K_s + S_u} - b \right) \qquad [4.16]$$

F_s = Coefficient of sloughing for attached biomass, 1/time

X_f = Mass density of attached biomass, mass/volume

L_f = Average thickness of attached biofilm within the differential volume considered, length.

a = Specific surface area of media, area/volume.

P = The coefficient of biomass change in the upper zone due to settling by gravity and/or lifting by the biogas, 1/time.

All other terms are as defined before. At steady state,

$$\frac{dX_{su}}{dt} = 0$$

Hence, the above equation reduces to:

$$\frac{dX_{su}}{dH} Q_o \, dH = a \, F_s \, L_f \, X_f \, A \, dH + \epsilon \, A \, X_{su} \, dH \left(\frac{Y \, K \, S_u}{K_s + S_u} - b\right)$$

$$+ P \, X_{su} \, A \, dH$$

dividing through by Q_o dH we obtain:

$$\frac{dX_{su}}{dH} = \frac{F_s}{Q_o} a \, L_f \, X_f \, A + \frac{\epsilon}{Q_o} A \, X_{su} \left(\frac{Y \, K \, S_u}{K_s + S_u} - b \right) +$$

$$P \, X_{su} \left(\frac{A}{Q_o} \right) \qquad [4.17]$$

Equation 4.17 is used in the mathematical model to calculate the change in the concentration of the biosolids along the

reactor height. The first term in equation 4.17 which accounts for the sloughing of the attached biofilm was neglected in the calculations. The reason is: it is assumed in this model that the biofilm has a constant density through out the height of the upper zone, hence, at steady state conditions the amount of sloughed active bacteria will be minimal and can be neglected. More discussion on the biofilm density in the upper zone will be presented later in this chapter.

A computer algorithm which includes all the mathematical principles discussed above for both the lower and the upper zones of the reactor was prepared in FORTRAN , see appendix A. The algorithm incorporated "DVERK" a differential equation solver computer subroutine, based on a code designed by T. E. Hull, W. H. Enright, and K. R. Jackson(20).

The DVERK subroutine was used to solve sequentially equations 4.7 and 4.16 over the height of the upper zone of the reactor. The solutions give the values of S_u and X_u along the upper zone height corresponding to a selected incremental heights.

The computer algorithm or model requires that the values of K, K_s, Y, D, b, D_f, V_{bd}, S_o, A, X_{bd}, H, dH, ϵ, Q_o, P, a, the equivalent diameter of the media (d), the dynamic viscosity of the bulk liquid ($\hat{u}$) and, the mass density of the bulk liquid (σ) be supplied by the user. Then the model will calculate the value of S_{bd} in the lower zone as well as the

values of S_u and X_{su} at the preselected incremental heights in the upper zone. If the "H" specified is equal the total height of the reactor. Then the last two values of S_u and X_{su} will correspond to S_e and X_e respectively. The smaller the incremental height value is , the more precise the calculated values of X_{bd}, S_u and, X_{su} will be.

The following sections present brief discussions intended to explain how the values of the parameters needed by the model were selected or evaluated for this research.

Coefficient of Molecular Diffusion D

The coefficient of molecular diffusion D of acetate in water, at 25 ^{o}C, can be found in most standard chemical hand books. Wilke and Chang (66) equation may be used to find D at temperatures other than 25 ^{o}C:

$$D = 7.4 \times 10^{-8} \times \frac{(Z_m)^{0.5} \quad T}{M_v \ (V_m)^{0.6}} \qquad [4.18]$$

where:

Z_m = Association parameter (2.6 for water)

M = Molecular weight of solvent, mass/mole

T = Absolute temperature, degree Kelvin

M_v = Viscosity of solvent, centipoise

V_m = Molecular volume of solute at normal boiling point, volume/mole

Montgomery, M. (35) Calculated the D values for temperature values that range from 25 to 10 ^{o}C using Wilke and Chang

equation, see table (4.1). The D values at 10 and 20°C used by Montgomery were also used in this study. The other D value at 30 °C needed for this research was interpreted from the other values as well as calculated using Wilke and Chang equation. Both the interpreted and the calculated values were in agreement with each other.

The Effective Diffusion Layer Depth (L)

McCarty and Rittmann (41) define "L" as the equivalent depth of liquid through which the actual mass transport is described by molecular diffusion alone.

In mass transportation L is commonly expressed as:

$$L = \frac{D}{K_m}$$

where

D = Molecular diffusivity of the substrate in the liquid, Area/Time

K_m = The coefficient of mass transportation, Length/Time

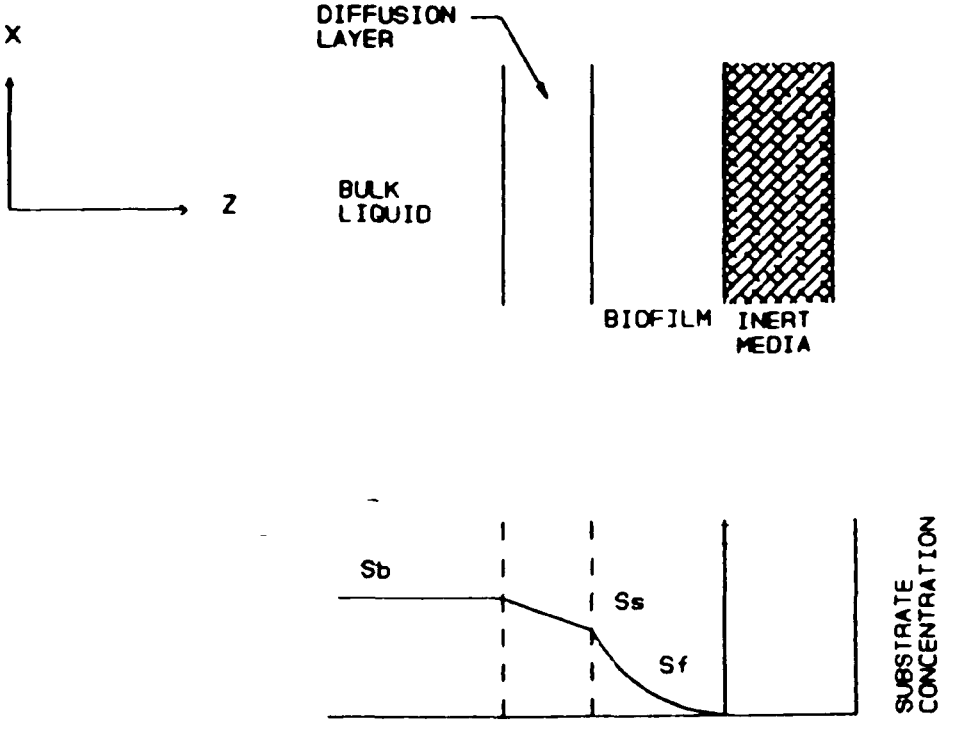
Figure (4.6) shows the concept of the biofilm model, as well as substrate profile from the bulk liquid through the biofilm. The figure shows a deep biofilm. A deep biofilm is a film in which the substrate concentration and the concentration gradient approach zero. The substrate concentration profile varies linearly from S_b in the bulk to S_s at the liquid/biofilm interface. However, inside the biofilm the substrate concentration varies asymptotically to zero within

the biofilm, down from S_s at the liquid/biofilm interface. In a shallow biofilm the substrate concentration does not approach zero.

TABLE 4.1: DIFFUSION LAYER PARAMETER VALUES.

T	Volumetric Flowrate	Upward Flow Velocity	Reynolds Number	Schmidt Number	K_m	D	L
°C	L/day	cm/day	dimensionless	dimensionless	Cm/day	Cm/day	Cm
30	49.437	354.1	0.538	552.4	10.0	1.25	0.126
30	69.212	495.7	0.754	552.4	11.1	1.25	0.112
30	64.649	463.0	0.704	552.4	10.9	1.25	0.115
30	90.508	648.3	0.985	552.4	12.2	1.25	0.103
30	72.254	517.5	0.787	552.4	11.30	1.25	0.111
30	101.156	724.5	1.098	552.4	12.87	1.25	0.099
20	49.437	354.1	0.430	901.8	8.34	0.98	0.115
20	69.212	495.7	0.601	901.8	9.38	0.98	0.102
20	64.649	463.0	0.562	901.8	9.13	0.98	0.105
20	90.508	648.3	0.786	901.8	10.22	0.98	0.094
20	72.254	517.5	0.628	901.8	9.47	0.98	0.101
20	101.156	724.5	0.879	901.8	10.60	0.98	0.091
10	49.437	354.1	0.328	1594.1	6.8	0.71	0.104
10	69.212	495.7	0.459	1594.1	7.7	0.71	0.093
10	64.649	463.0	0.430	1594.1	7.5	0.71	0.095
10	90.508	648.3	0.601	1594.1	8.38	0.71	0.085
10	72.254	517.5	0.480	1594.1	7.75	0.71	0.092
10	101.156	724.5	0.672	1594.1	8.65	0.71	0.082

FIGURE 4.6:
CONCEPTUAL BASIS FOR THE BIOFILM MODEL.



In the diffusion layer Fick's first law of diffusion applies to give:

$$j = D \frac{S_b - S_s}{L} \qquad [4.19]$$

where:

j = Flux of substrate into the biofilm surface per unit surface area, (mass/area)

S_b = Concentration of substrate in the bulk liquid, mass/volume

S_s = Substrate concentration at biofilm surface, mass/volume

D and L are as defined before.

In order to determine L, K_m must be known. K_m is usually calculated from empirical relationships. These empirical correlations are usually function of Reynolds number (Re), and Schmidt number (Sc). Schmidt number is used as a measure that expresses the ratio of momentum diffusion caused by viscous actions to molecular diffusion of substrate:

$$Sc = \hat{u}/(\sigma\ d)$$

where:

$\hat{u}$ = Dynamic viscosity, mass. area^{3}/time

σ = Mass density of the bulk liquid, mass/volume.

A display of the ratio of inertial forces to viscous forces is the Reynolds number:

$$Re = \frac{v\ l}{\blacklozenge}$$

where,

v = Exterior flow velocity, length/time

l = A characteristic length of media, length

Φ = Kinematic Viscosity, area/time

Using benzoic acid spheres and propylene glycol solutions, Wilson and Geankoplis (56) formulated an empirical correlation to estimate Km at Reynolds numbers in the range of 0.0016 to 55 :

$$K_m = \frac{1.09}{\epsilon} (Re)^{-2/3} (Sc)^{-2/3} \qquad [4.20]$$

where:

ϵ = Porosity of media

In deriving the empirical equation for Km, Wilson and Geankoplis used spherical benzoic acid media. In this research the plastic media has a cylindrical configuration as was described earlier. Therefore, an estimate of an equivalent diameter corresponding to the characteristic length of the plastic media used is necessary. The equivalent diameter was obtained as follows:

The diameter of a sphere that has the same volume as a unit of the plastic media was computed, that is

$$0.61 \text{ cm}^3 = \frac{1}{6} \pi d^3$$

or

d_2 = 1.05 cm

Table (4.1) shows the values of the parameters used in equation 4.19 and the derived L values that are pertinent to this research. The Reynolds Number values reported in this research are within the values for which equation 4.12 was found to be valid.

The Molecular Diffusion of Substrates within the biofilm D_f

D_f was evaluated using a value of $D_f/D = 0.80$ found by Williamson (53) and used by Montgomery (35), refer to table (4.1). Montgomery stated that the diffusion coefficient of a solute in a biofilm (D_f) is less than (D) in water because bacteria and their slime polymers make up a significant part of the biofilm.

The Porosity of Media

The porosity of the media used in the packed part of the reactor (upper zone) was determined experimentally by water displacement method and from the manufacture's literature and found to be 87.5%.

Monod Kinetics

Montgomery, M. (35) experimentally determined the Monod Kinetics for acetates utilizing bacteria. These Kinetic values are used in this research, refer to table 2.2 in chapter two. Also, see table 4.4 for estimation of the specific decay coefficient b.

Biofilm Density

Rittmann (41) reported that biofilm density X_f may vary from 20 mg/cm3 to 110 mg/cm3 as total suspended solids (TSS). The biomass portion of the TSS varies from 50 to 100% . In addition, the biomass portion include both active and dead biomass. At steady state the active biomass can be estimated using an equation presented by Bouwer (2):

active biomass =

$$\frac{\text{(yield coefficient) (substrate loading)}}{\text{(decay coefficient)}} \qquad [4.21]$$

Applying these principles to the lower zone the active biomass and the corresponding suspended biomass density for the different runs are shown in table (4.3).

In the upper zone attached biofilm as well as suspended solids exist. The active portion of the suspended solids is predicted using equation (4.21) which is incorporated directly into the model. As for the attached biofilm a value of 0.81 mg/cm^3 for acetate utilizing biofilm used by Montgomery (35) was employed.

TABLE 4.2: MINIMUM SUBSTRATE CONCENTRATION(S_{deep}) NEEDED TO SUPPORT A DEEP BIOFILM AND RELATED PARAMETERS FOR THE DIFFERENT RUNS OF THE EXPERIMENT.

Phase #	Run #	T °C	Q L/d	Organic Loading Kg/m³/day COD	Influent Concentration mg/L COD	Effluent Concentration mg/L COD	S_{deep} mg/L
1	1	30	49.43	0.65	350	10.5	3.9
1	2	30	69.21	0.65	250	15.0	3.6
1	3	20	49.43	0.65	350	28.0	3.1
1	4	20	69.21	0.65	250	32.5	2.9
1	5	10	49.43	0.65	350	84.0	5.8
1	6	10	69.21	0.65	250	75.0	5.7
2	1	30	64.64	0.85	350	28.0	2.4
2	2	30	90.50	0.85	250	42.5	3.4
2	3	20	64.64	0.85	350	58.0	2.9
2	4	20	90.50	0.85	250	57.5	2.8
2	5	10	64.64	0.85	350	122.5	5.7
2	6	10	90.50	0.85	250	112.5	5.6
3	1	30	72.25	0.95	350	45.0	3.8
3	2	30	101.1	0.95	250	62.5	3.3
3	3	20	72.25	0.95	350	84.0	2.9
3	4	20	101.1	0.95	250	80.0	2.8
3	5	10	72.25	0.95	350	147.0	5.1
3	6	10	101.1	0.95	250	125.0	5.8

TABLE 4.3: ESTIMATES OF BIOMASS DENSITY IN THE LOWER ZONE OF THE REACTOR FOR THE DIFFERENT RUNS.

PHASE No.	RUN No.	TOTAL SUSPENDED SOLIDS mg/cm^3	X_{bd} mg/cm3	BIOMASS mg/cm3	ACTIVE BIOMASS / VOLATILE SUSPENDED SOLIDS
1	1	15	6.2	1.8	0.288
1	2	16	6.7	1.9	0.281
1	3	19	7.0	3.5	0.503
1	4	20	7.5	3.1	0.418
1	5	20	7.0	5.5	0.785
1	6	21	7.6	4.9	0.648
2	1	17	7.0	2.5	0.354
2	2	18	7.7	2.1	0.277
2	3	20	7.6	4.0	0.530
2	4	22	8.7	3.4	0.395
2	5	21	7.6	5.8	0.768
2	6	23	8.5	4.4	0.519
3	1	19	8.0	2.6	0.324
3	2	20	9.0	2.1	0.231
3	3	20	7.8	3.9	0.502
3	4	23	9.9	3.0	0.307
3	5	21	7.8	5.4	0.696
3	6	24	9.4	4.2	0.445

Results and Discussion of the Mathematical Model

The experimental results were used to calibrate the model, essentially to determine the settling and lifting coefficient P. For simplicity P was assumed to constant for the range of operating conditioned investigated and was found from calibration to be equal to (-5 per day). To check for the validity of the model, the model was calibrated using the results of runs of the first phase. Then, the model was used to predict the results of the runs of the second and the third phase.

Figure 4.7 through figure 4.24 show the filter model results. For convenience two plots are presented in each figure to show the actual COD values and the predicted ones. Considering these figures the following observations can be made:

a) In phase 2 for run 1 through 4, (figures 4.13 & 4.16), the model closely predicted the filter performance. In fact the variation between the measured and predicted values never exceeded 25 mg/L as COD. As for run 5 and 6 (figures 4.17 & 4.18) the model under predicted the COD removal in the filter in the lower part of the upper filter zone, by about 35%. However, in the rest of the upper zone the predicted values were in agreement with the measured values.

b) In regard to phase 3 the following can be noted: In run one (figure 4.19), the model predictions were very close to the measured values. COD removal was within 25 mg/L from the actual values. However, in the upper zone the model over estimated the measured values. For run 3 (figure 4.21) the model predicted the correct values of COD removal in the lower zone of reactor, but under estimated the removal in the upper zone. In run 5 (figure 4.23) the model prediction was very close to the measured COD removal values in the upper zone but under estimated the removal in the lower parts of the

filter. However, this variation was off by about 24% or 40 mg/L. In run 2 (figure 4.20) the model predicted the correct COD values in the lower zone and lower levels of upper zone. Similar results were obtained for run 4 (figure 4.22), where the maximum variation was noted at the effluent exit point and was found to be approximately 45 mg/L less than the actual ones. In run 6 (figure 4.24) the model predicted values were close to the measured ones in the upper zone. In the lower zone and beginning section of upper zone the model slightly under estimated the filter removal by 25 mg/L .

The model for all purposes is a practical presentation of the filter. The variations in the measured and predicted values are due to the following modeling limitation:

1- The model assumes a constant value for the Settling Coefficient Cs.

2- The exact biofilm density and the way it varies with the different operating parameter as well as along the filter height is not known.

3- The mass transport coefficient "Km" was determined using an empirical equation.

4- A perfect plug flow regime was assumed for the upper zone.

5- An ideal fully mixed reactor was assumed for the lower zone of reactor.

6- No flow was assumed to bypass the lower zone.

7- The fixed film model used after McCarty and Rittmann does not consider the gradual increase in inert cell mass resulting from decay. Besides, the biofilm model lacks any constraints on the maximum allowable biofilm thickness.

8- The values of the decay rates used in modeling were estimated. Previous research work on decay rates of methanogens is scarce. Montgomery (35) stated that a difference between the bacterial decay coefficients exists, when substrate was present or absent. The two decay values may represent limits over which decay can vary. If decay can vary between these limits, then different COD values may have been predicted by the model than these reported in this study. Montgomery measured the decay coefficient for acetate at temperature of 25^{o}C. To arrive at the decay rate values used

in this study three different possibilities were considered, following the procedure used by Montogomery(35):

Firstly, the decay coefficient "b" was assumed to vary with temperature in the same manner as the substrate utilization rate "K".

Secondly, b was estimated assuming an exponential temperature dependence as suggested by O'Rourke (38) in the following formula:

$$b_T = b_{25^oC} \, e^{0.07(T - 25)}$$

Where,

b_T = The value of the decay coefficient at a given temperature T, 1/day.

b_{25^oC} = The value of the decay coefficient at a temperature of 25^oC, 1/day.

Thirdly, b assumed not to change with temperature.

The, b values estimated using O'Rourke equation were found to produce the best model fits. Therefore, they were the b values of choice in this study (see table 4.4).

The model may be used to predict the optimum configuration of the filter in terms of removal. This prediction may be achieved by trying different values for V_u, H, and/or A to find those that will cause the filter to achieve a maximum COD removal. The trial values for V_{bd}, H and/or A should

be used against known values of S_u, T, Q_o, K, K_s, Y, b, P. In order for the model to yield more reliable values for the optimum configurations, it is recommended that P and Km should be found by independent research work that determines their true behavior with the different environmental conditions such as Q_o and T.

The model presented here provided a practical means to mathematically describe the modified Anaerobic Filter Treatment Process presented in this research work. Future modeling endeavors need to consider the limitations of the existing model.

TABLE 4.4 : ESTIMATED DECAY RATE VALUES USED IN THIS STUDY. AFTER MONTGOMERY (35)

Temperature ^{o}C	A b^{A}	B b^{B}	C b^{C}
30	0.034	0.045	0.032
20	0.015	0.023	0.032
10	0.004	0.011	0.032

Notes:

1) b^{A} : The decay rate estimated at temperatures other than 25 C assuming temperature effect on decay rate is the same as on substrate utilization rate (K).

2) b^{B} : The decay rate estimated at temperatures other than 25 C using O'Rourke equation:

$$b_T = b_{25^{o}C} e^{0.07(T - 25)}$$

3) b^{C} : The decay rate assumed independent of temperature and constant.

FIGURE 4.7

COD Removal Versus Reactor Depth

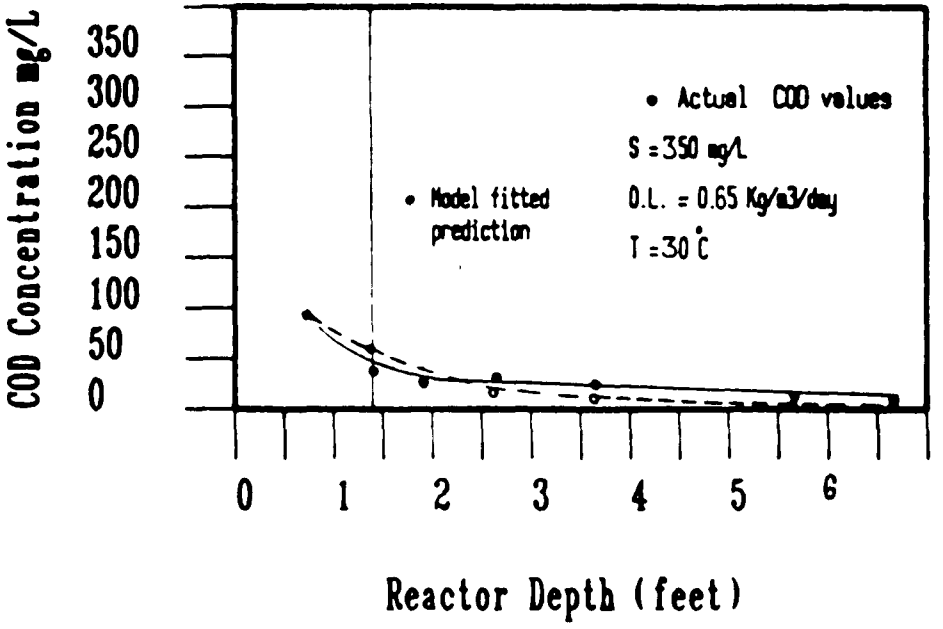


FIGURE 4.8

COD Removal Versus Reactor Depth

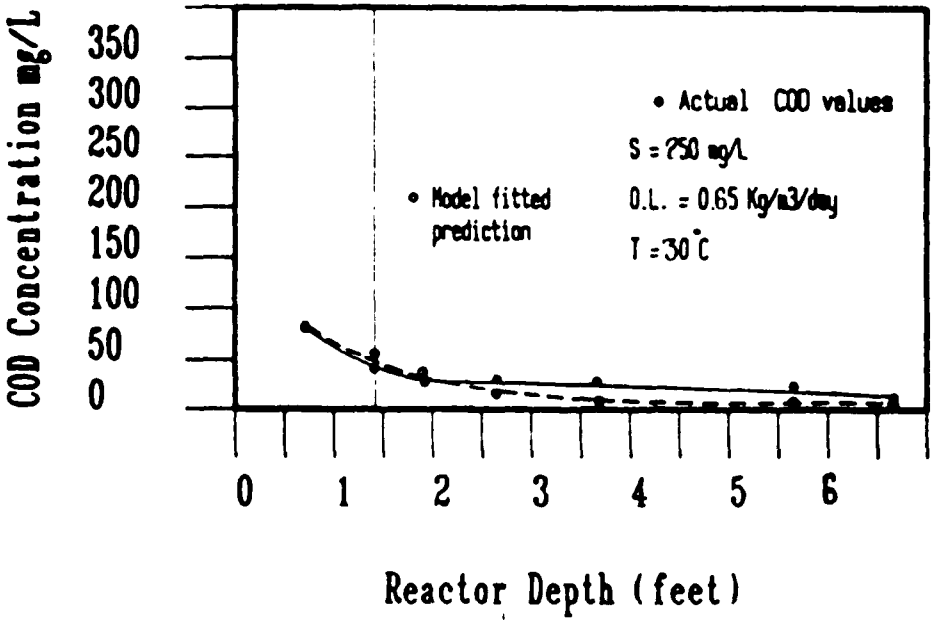


FIGURE 4.9

COD Removal Versus Reactor Depth

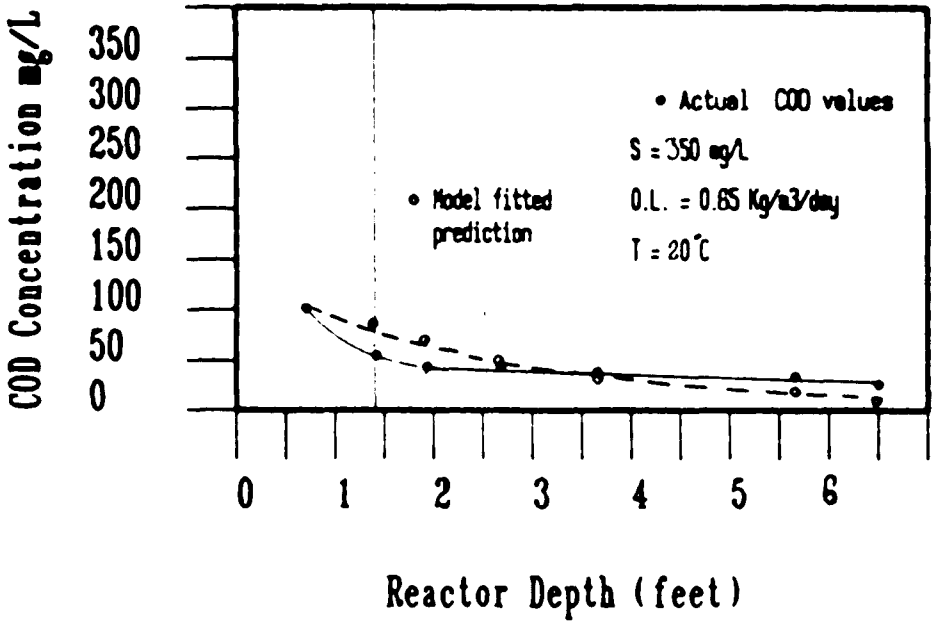


FIGURE 4.10

COD Removal Versus Reactor Depth

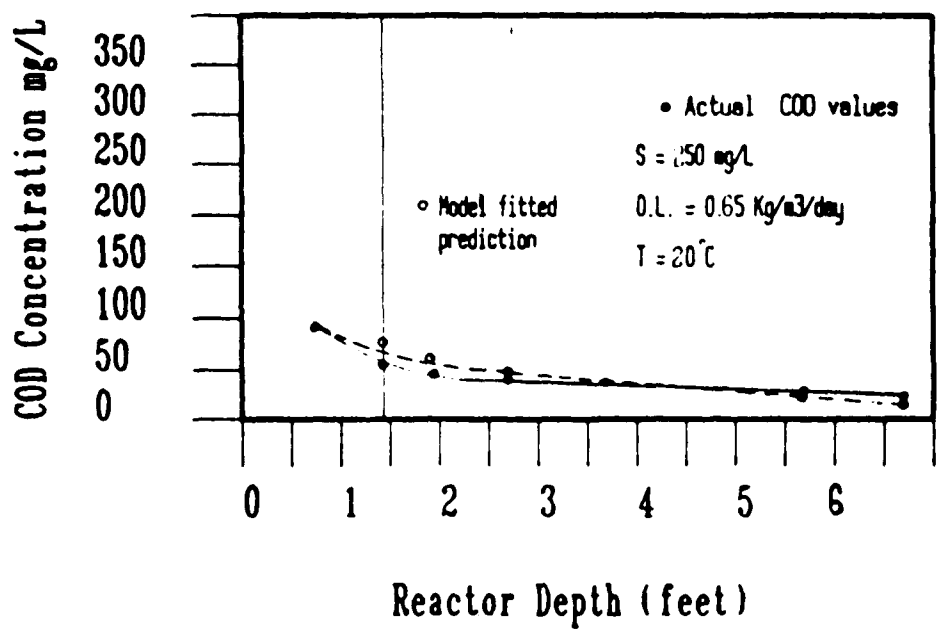


FIGURE 4.11

COD Removal Versus Reactor Depth

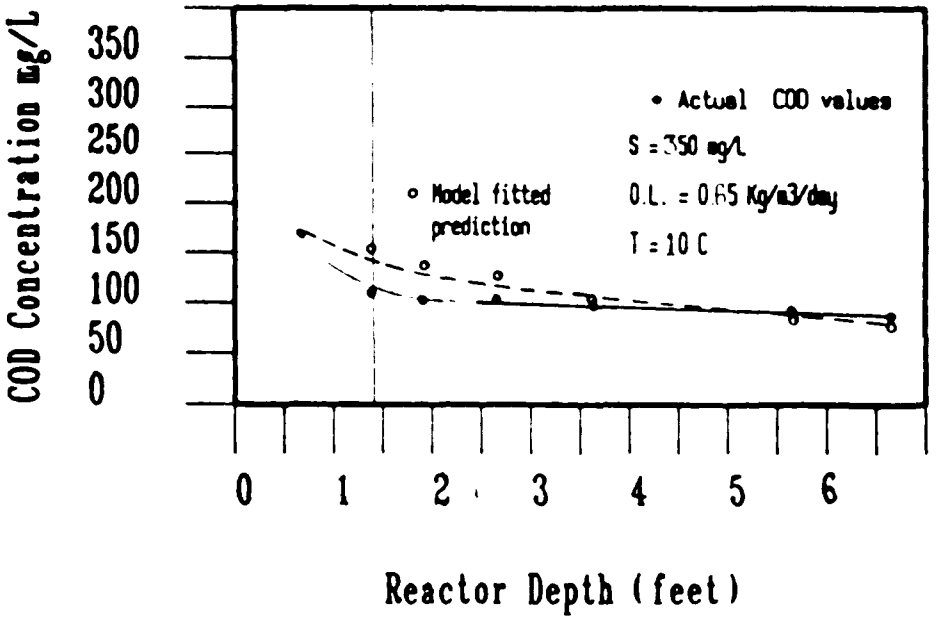


FIGURE 4.12

COD Removal Versus Reactor Depth

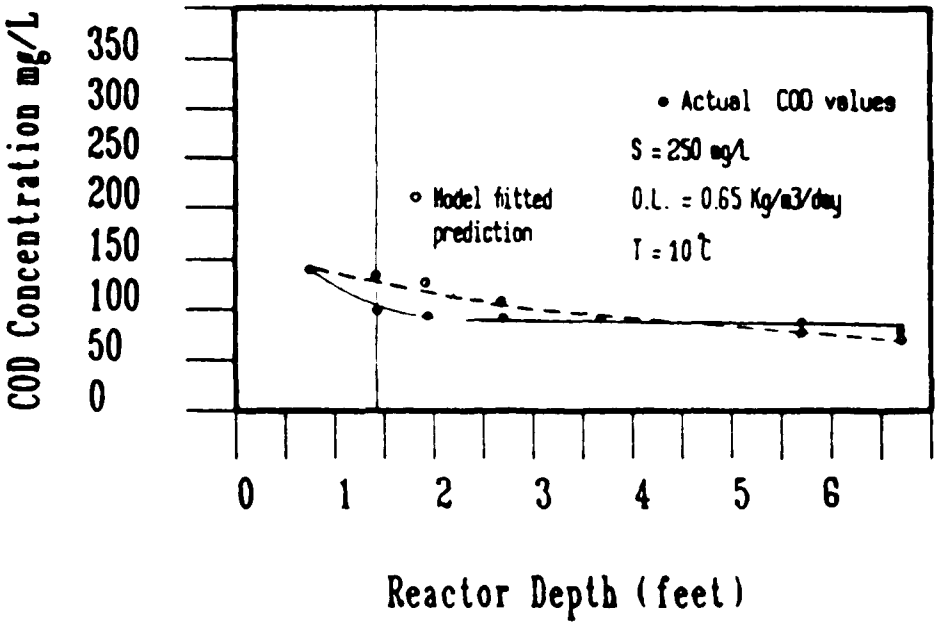


FIGURE 4.13

COD Removal Versus Reactor Depth

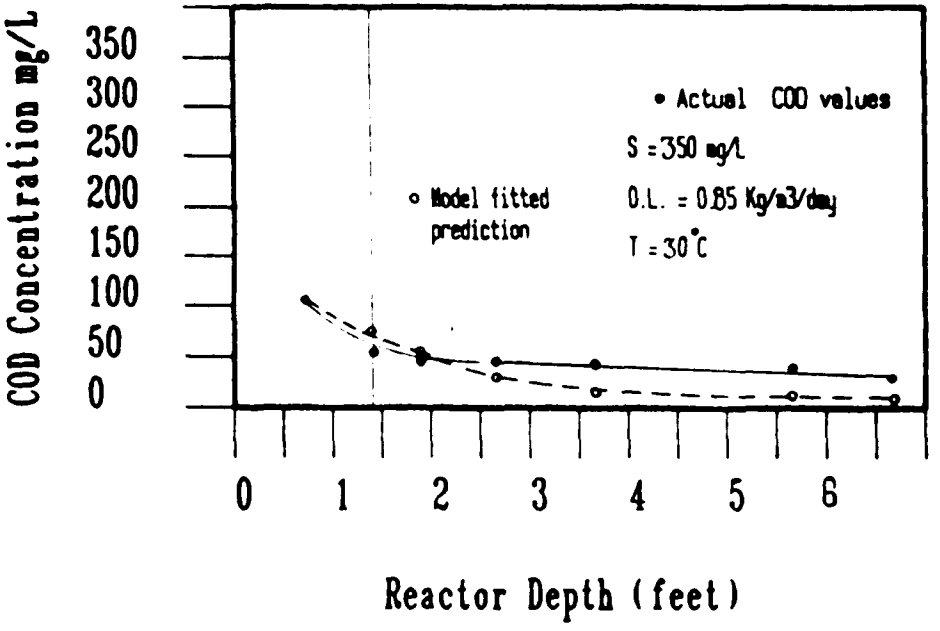


FIGURE 4.14

COD Removal Versus Reactor Depth

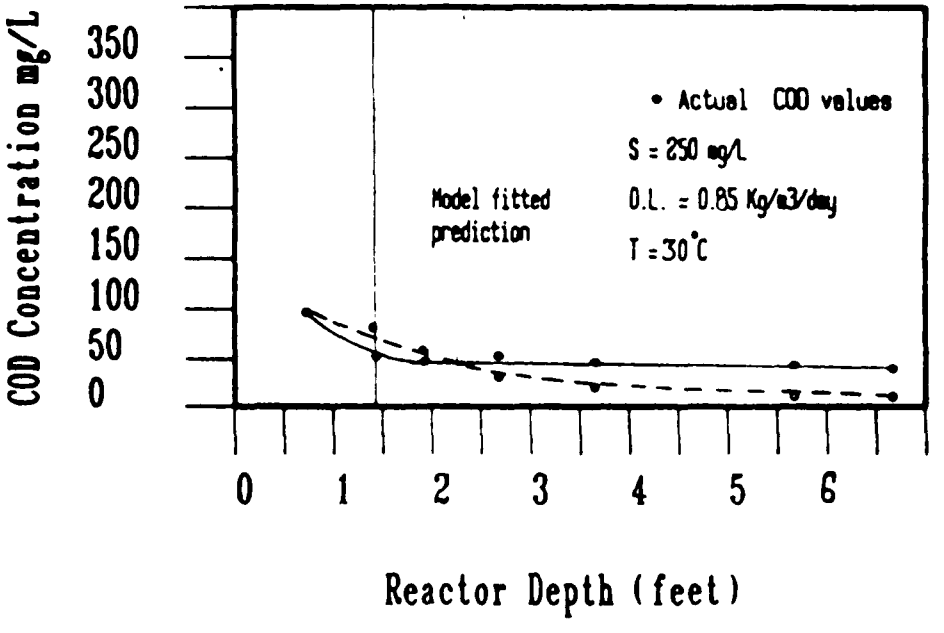


FIGURE 4.15

COD Removal Versus Reactor Depth

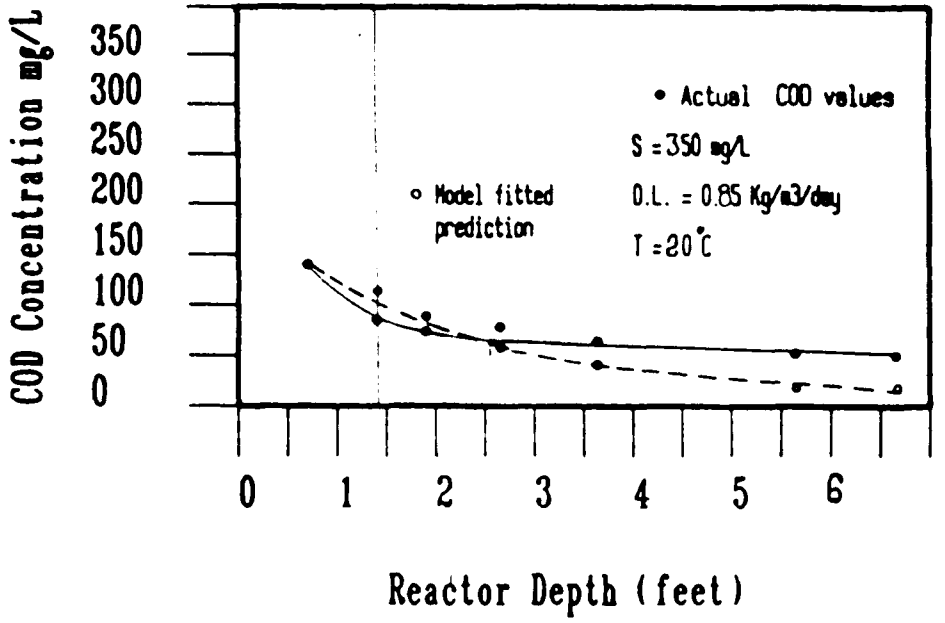


FIGURE 4.16

COD Removal Versus Reactor Depth

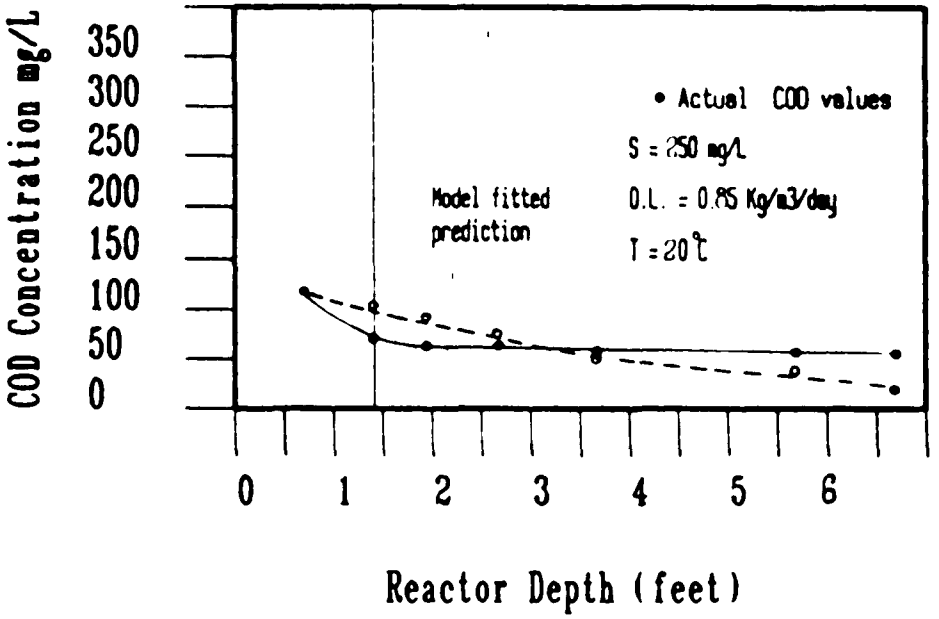


FIGURE 4.17

COD Removal Versus Reactor Depth

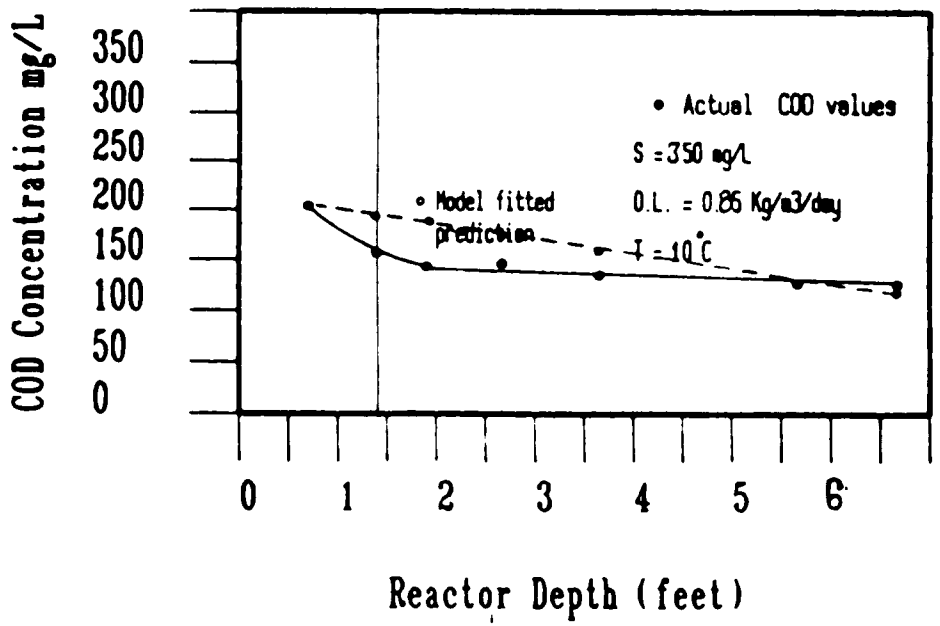


FIGURE4.18

COD Removal Versus Reactor Depth

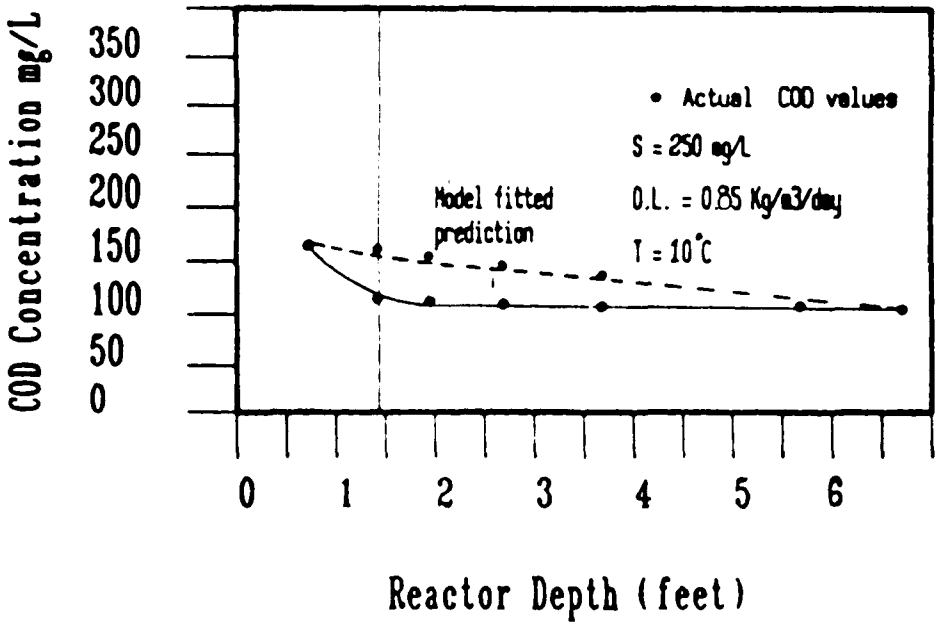


FIGURE 4.19

COD Removal Versus Reactor Depth

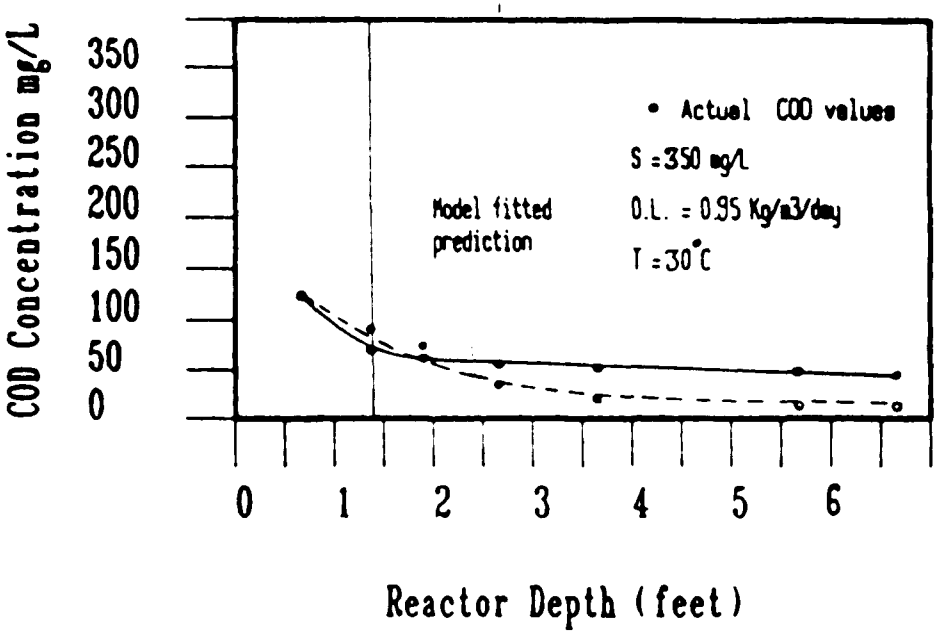


FIGURE 4.20

COD Removal Versus Reactor Depth

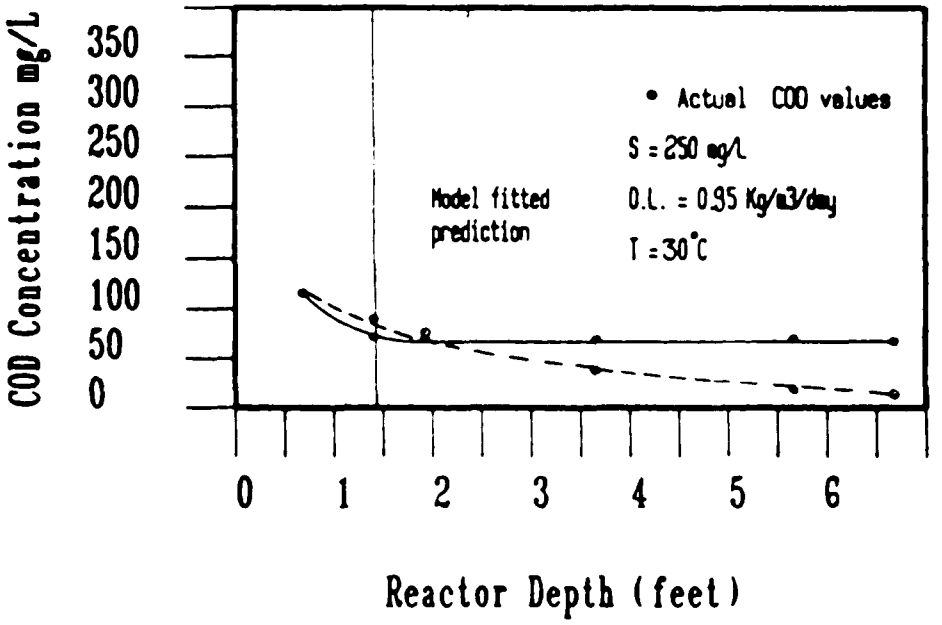


FIGURE 4.21

COD Removal Versus Reactor Depth

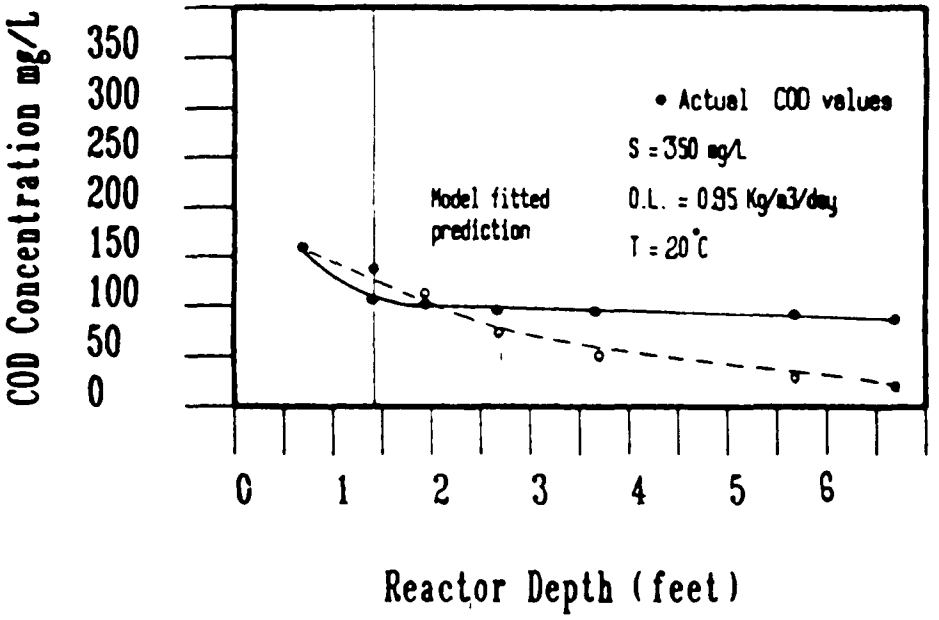


FIGURE4.22

COD Removal Versus Reactor Depth

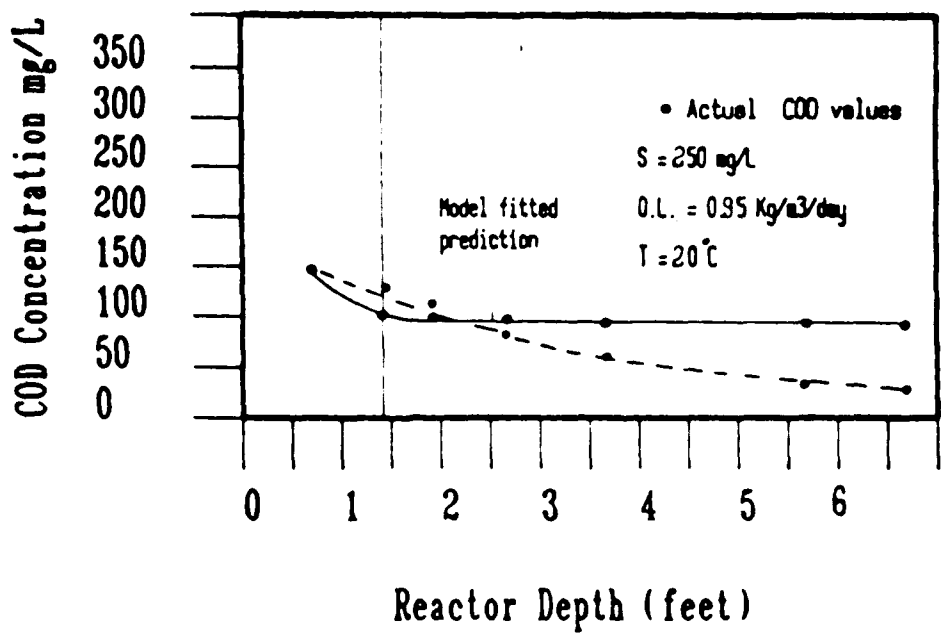


FIGURE 4.23

COD Removal Versus Reactor Depth

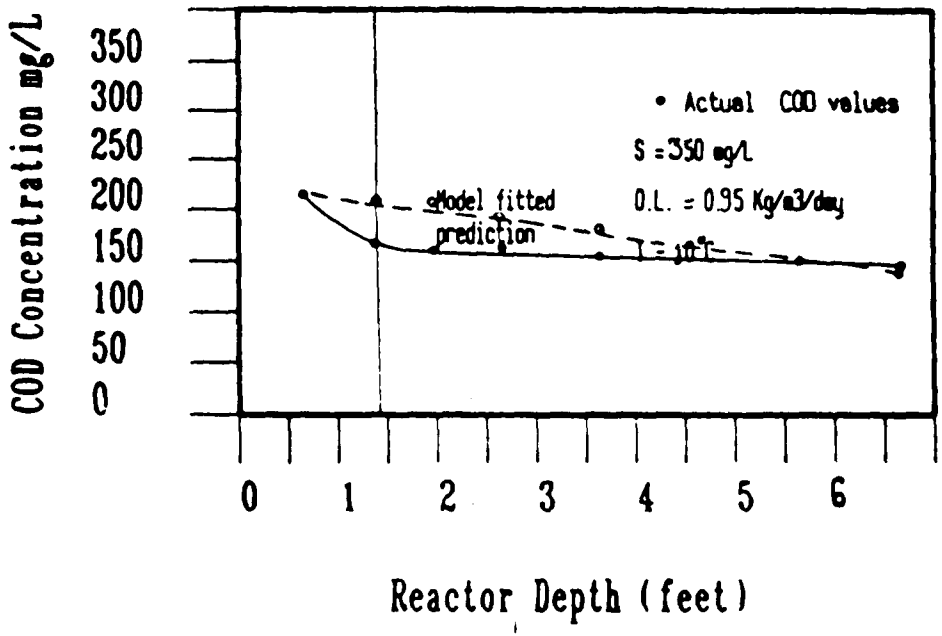
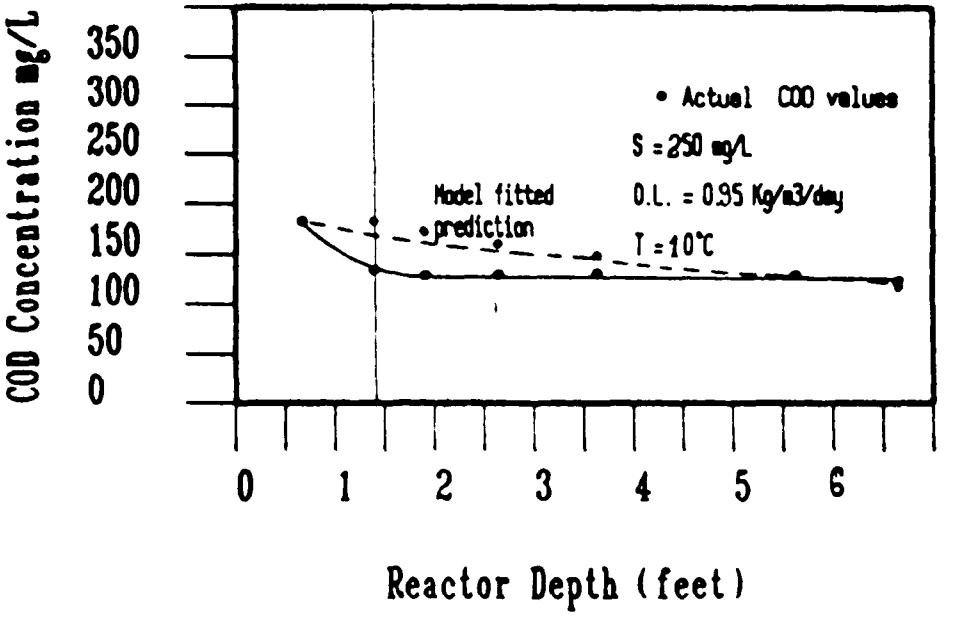


FIGURE 4.24

COD Removal Versus Reactor Depth



CHAPTER FIVE

Sensitivity Analysis

In order to determine which of the parameters has the largest influence on the results of the calculations using the developed mathematical model, and which therefore should be most accurately measured a sensitivity analysis was made. The sensitivity of the soluble COD in the reactor effluent, S_{Se}, to a change of a parameter P_r is expressed as:

$$S_{Se} = \frac{S_e/S_e}{P_r/P_r}$$

The results of these analyses have been collected in table 5.1 and presented graphically in figures 5.1 and 5.2. It appears from the results of the sensitivity study that, of the parameters studied, substrate utilization rate K, influent flowrate Q_o, substrate concentration in influent So, temperature T, half velocity coefficient Ks, and the molecular diffusion coefficient have the largest influence on the COD removal. It is, therefore, of great importance that the values of K, K_s, and D are accurately known. Besides, in the operation of the process, special attention must be paid to keep the influent flowrate Q_o, the concentration S_o, and temperature T steady. The variations that may take place in the values of the other parameters apparently influence the outcome of the calculations less strongly. In conclusion, it is clear that Q_o and S_o or organic

loading must be used as the main design parameter. The calculations are most sensitive to changes of these parameters. The next design parameter should be the temperature.

As was mentioned in chapter four a constant value for the biofilm density was assumed in the upper zone regardless of the possibility that this density may vary with the substrate concentration in the bulk liquid. To determine the sensitivity of this assumption (i.e X_f varies with S_u the biofilm mass density factor F_{ac} was created which is expressed as follows:

$$F_{ac} = X_f/S_u$$

or

$$X_f = F_{ac}\ S_u$$

A one to one linear correlation between X_f and S_u was then established in such a way that X_f = 0.8 mg/L when S_u = 100 mg/L COD and X_f = 0 when $S_u = S_{min}$, where S_{min} is the minimum substrate concentration to support a biofilm. For simplicity S_{min} was assumed to be 1 mg/L. The results of the sensitivity analyses of F_{ac} demonstrate that

1- The sensitivity of F_{ac} increases as temperature decreases and vice versa.

2- The assumption that X_f is constant throughout the height of the filter is reasonable at temperatures above 20°C. This may be reasonable only for the S_u concentrations encountered in this study.

T=30℃

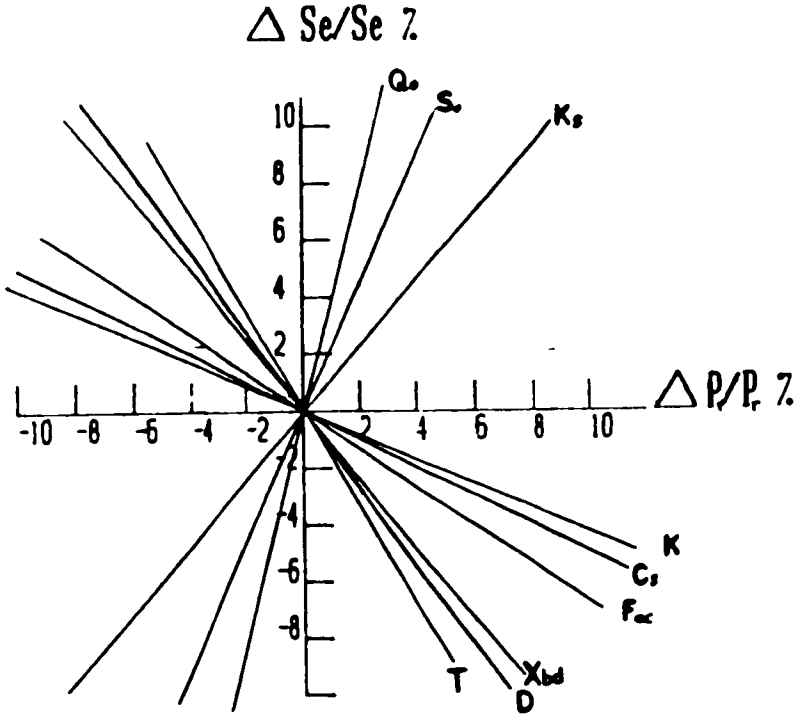


Fig.5.1 Parameter sensitivity

T = 20°C

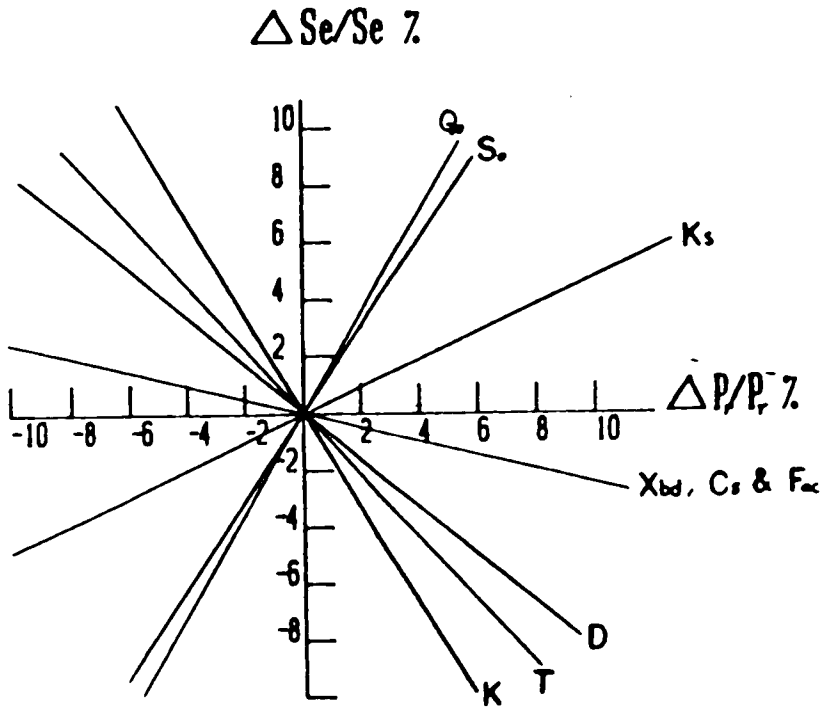


Fig.5.2 Parameter sensitivity

TABLE 5.1: RESULTS OF THE SENSITIVITY ANALYSIS.

Parameter (P_r)	S_{se} at 30°C	S_{se} at 20°C
K	- 4.6	- 1.75
K_s	+ 1.1	+ 0.461
Y	0.00	0.00
b^b	0.00	0.00
D	-1.4	- 0.856
Q_o	+ 3.83	+ 1.79
S_o	+ 2.2	+ 1.53
T	- 1.86	- 1.03
C_s	+ 0.49	+ 0.217
F_{ac}	+ 0.014	+ 0.217
X_{bd}	- 1.30	+ 0.217

Conclusion

1- The modified anaerobic filter investigated in this study seems to be effective in the treatment of low strength wastewater as indicated by the high COD removal achieved under a wide range of practical operating conditions. The removal achieved met the national secondary treatment standards, as mandated by the USEPA, in a number of runs. No effluent recycling was needed to maintain the high removal efficiency.

2- The filter exhibited high percentage methane production (72% to 98% of biogas by volume). The gas produced provides a source of useful energy.

3- The experimental data showed that the filter is able to respond very quickly to the changes in the run conditions (e.g. temperature, and organic loading), because the active biomass remained in the filter at all times. Therefore, the filter appears to be suited for the treatment of intermittent waste discharges. In that respect, the filter is suited for the treatment of domestic wastewater. In fact, the experimental results also demonstrated that less than a week is needed in order for the filter to establish a new steady state condition when a change is made in run conditions.

4- The filter achieved a very low biosolids production. The

reactor was operated for almost a year before sludge wasting was required. The sludge wasting interval could become longer if it is not necessary to meet NPDES secondary wastewater treatment standards.

5- The filter configuration proved to be an effective trap for the biosolids, as demonstrated by the very low solids washout achieved. The solids concentration in the effluent met the national secondary standards in all the runs.

6- The experimental results indicate that, in order for the solid wash out to be minimal, the depth of sludge bed must not exceed 110% of the depth of the lower zone. The results also indicate that the volume of the lower zone, which contains the sludge bed seems to be in appropriate proportion to the total volume of the reactor.

7- There is virtually no fear of plugging due to the biomass production because of the very high void ratio in the reactor. The filter is expected to be able to treat colloidal suspensions such as starches and dilute milk waste without problems of clogging.

8- Generally speaking, the biosolids produced at all heights of the reactor were found to settle rapidly. This may provide a simple mains for recycling of the

solids back into the reactor as an alternative way to improve the removal efficiency.

9- The performance of the filter suggests that the gas production in the lower zone serves to maintain the lower zone completely mixed. Therefore, no mixing device may be needed inside the reactor.

10- The experimental data suggest that the short circuiting in the lower zone would be minimized if the lower zone is kept full of a sludge bed at all times. In other words the lower zone should be carefully monitored so that the sludge blanket does not appear above the sludge bed inside the lower zone. This should guarantee against significant short circuiting which will dramatically reduce filter efficiency.

11- As was anticipated no ammonia reduction has been achieved in the reactor.

12- The mathematical model developed simulated well the COD removal in the lower zone as well as the upper zone for the lower organic loading rates(0.85 Kg/m3/day). The agreement between the model predicted results and the experimental results were close (within = +/- 40 mg/L maximum) except at the bottom two inches of the upper zone (i.e. at the level between the upper zone and the

lower one) where the mass density of the biofilm (X_f) is likely not to be constant due to the possible fluctuation of the sludge bed height and the consequent variation of substrate concentration near this level.

At the higher loading rate (i.e. 0.95 Kg COD/m^3/day) the model less successfully predicted the removal. This may be because at this organic loading the filter was being over loaded and therefore the assumptions that were made to calculate K_m and X_f values may not be accurate .

In general, the model underestimated the removal in the lower levels of the reactor and overestimated the removal in the upper zone. This variation may be in part due to the use of biofilm density values taken from the literature and not by experimental verification. In addition, the biofilm density was assumed to be constant throughout the upper zone of the reactor. The X_f value used was 0.81 mg/cm^3, which may be different from the actual value in the filter. No information was available in the literature on how the density of biomass varies with environmental conditions such as the COD concentration. Besides this limitation, the model did not consider the gradual increase in inert cell mass resulting from decay. Further, the liquid mass transfer coefficient K_m was estimated using an empirical equation, and the settling and lifting coefficient P was estimated from calibration of the mathematical model and

not experimentally.

13- The results of the mathematical modeling suggest that any attempt to reduce the organic loading to treat very low influent concentrations should be done by either reducing the organic loading, the volumetric flowrate, or increasing the size of the filter horizontally keeping the height of the lower zones and the total height constant.

14- The model calibration indicates that the exponential change of the biomass decay rate (b) yields the best simulation.

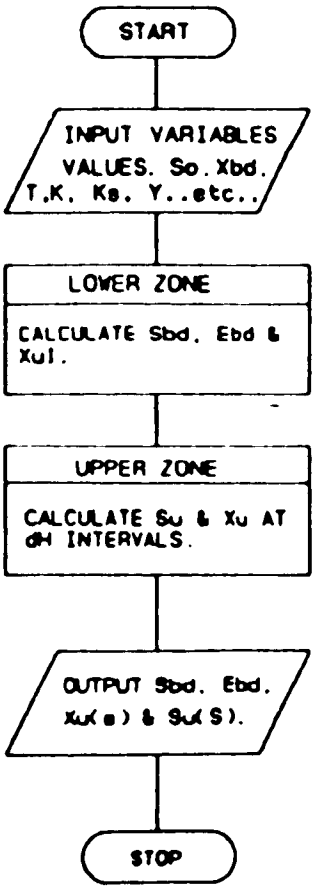
15- It appear from the results of this research that, using the developed anaerobic filter system, the treatment of low strength domestic wastewater is now feasible.

16- In conclusion, at least in the temperature range of 20^{o}C and higher, and with organic loadings of 0.85 kg/m3/day and lower, this modified anaerobic filter studied may be a viable alternative to aerobic processes for the treatment of low strength domestic wastewaters.

17- Future research work should concentrate on the determination of all the estimated kinetic parameters (e.g. biofilm density, and liquid mass transfer coefficient).

APPENDIX A: COMPUTER ALGORITHM OF THE MATHEMATICAL MODEL

FIGURE A.1: FLOW CHART OF THE MATHEMATICAL MODEL.



```
C*************************<<<LEGEND>>>*************************
C    SO = INFLUENT COD CONCENTRATION, MG/CM3
C    QO = INFLUENT FLOWRATE, CM3/DAY
C    VL = VOLUME OF THE LOWER ZONE, CM3
C    XBD= VOLATILE SOLIDS CONCENTRATION IN LOWER ZONE, MG/CM3
C    T  = TEMPERATURE, C
C    K  = MAX. SPECIFIC RATE FOR SUBSTRATE UTILIZATION, 1/DAY
C    KS = HALF-VELOCITY COEfficient, MG/CM3
C    B  = DECAY COEFFICIENT, 1/DAY
C    YIELD = GROWTH YIELD COEFFICIENT, MG/MG
C    AS = SURFACE AREA NORMAL TO BIOFILM SURFACE, CM2
C    AA = CROSS-SECTION AREA OF REACTOR, CM2
C    E  = PLASTIC MEDIA VOID RATIO, dimensionless
C    D  = MOLECULAR DIFFUSIVITY IN WATER, CM2/DAY
C    DF = MOLECULAR DIFFUSIVITY IN BIOFILM, CM2/DAY
C    RE = REYNOLDS NUMBER, DIMENSIONLESS
C    MUD = VISCOSITY OF THE BULK LIQUID, MG/CM/DAY
C    KM = MASS-TRANSPORT COEFFICIENT, CM/DAY
C    L  = DEPTH OF "EFFECTIVE" DIFFUSION LAYER, CM
C    LF = DEPTH OF BIOFILM, CM
C    SU = BULK-LIQUID SUBSTRATE IN UPPER ZONE, MG/CM3
C    ROU = MASS DENSITY OF THE BULK LIQUID, MG/CM3
C    TOTVOL = TOTAL LIQUID VOLUME OF THE REACTOR, CM3
C    ORGLOD = ORGANIC LOADING, KG COD/M3/DAY
C    TEMP = FILTER OPERATING TEMPERATURE, DEGREE CENT.
C    VEL = UP-WARD VELOCITY,
C    RE = REYNOLDS NUMBER, DIMENSIONLESS
C    FAC  = PROPORTIONAL FACTOR BETWEEN BIOMASS  DENSITY  AND
C SUBSTRATE CONCENTRATION, DIMENSIONLESS
C*************************************************************
      REAL*8 LM, DFM, SUM, L, D, KM, XF, DF, J, CM, QO, SBD,
     +VL,K, XBD, KS, SU(2), C(24), W(2,9), H, TOL, HEND, F,
     +EBD, AS,A, AA, Q, B, YIELD, P, XUG(180), SUG(180),
     +HENDG(180),ORGLOD,TEMP, TOTVOL, DIAM
      INTEGER N, IND, NW, TER, I, ISO, NN, HMAX, PHASE, RUN
      COMMON K, KS, B, QO, YIELD
      EXTERNAL FCN1
      NW        = 2
      N         = 2
      NN        = 1
      H         = DBLE(0.0)
      HENDG(1)  = DBLE(0.0)
      TOL       = DBLE(0.001)
      IND       = 1
      VL        = DBLE(5060.0)
      SO        = DBLE(?.??)
      HMAX      = 168
      XBD       = DBLE(?.???)
      K         = DBLE(?.???)
      KS        = DBLE(?.????)
```

```
      B         = DBLE(?.???)
      QO        = DBLE(49437.17)
      YIELD     = DBLE(?.???)
      TEMP      = ??.?
      TOTVOL    = DBLE(26617.84)
      ORGLOD    = (SO*QO/TOTVOL)
      PHASE     = ?
      RUN       = ?
      WRITE(6,1000) PHASE, RUN
 1000 FORMAT(6X,'PHASE ', I2,10X,'RUN',I2//)
      WRITE(6.1)
1     FORMAT(6X,'***********<<<INPUT DATA>>>**************'//)
      WRITE(6.2) SO, QO, TEMP, ORGLOD, K, KS, B, XBD
2     FORMAT(6X,'SO= ',F6.4,2X,'MG/CM3',6X,'QO= ',F8.1,2X,
     +'CM3/DAY',//6X,'TEMP= ',F6,2,2X,'C',9X,'ORGANIC LOADING = ',
     +F5.3,2X,'KG/M3/DAY',//6X,'K= ', F6.2, 1X,'1/DAY',9X,'KS= ',
     +F9.5,1X,'MG/CM3',9X,'B= ',F8.4,1X,'1/DAY',//6X,'XBD= ',
     +F8.4,1X,'MG/CM3'//6X,
     +'**********************<<< OUTPUT >>>*********'//)
C
C
C*********************<<<< LOWER ZONE >>>>************************
C
C
      ISO   = IFIX(SO * 1000000.0)
      DO 10 KKK = 10,ISO
      SBD = DFLOAT(KKK)/ DBLE(100000.0)
      F = QO* SO - QO*SBD - VL * (K*XBD*SBD)/(KS + SBD)
      IF(DABS(F). LT. DBLE(0.20) GOTO 30
C     IF(F. LT. 0.0) GOTO 20
C     IF(F. LT. 0.001) GOTO 30
 10   CONTINUE
      STOP
 20   SBD = SBD - DBLE(0.001)
 30   EBD = ((SO - SBD)/SO)* DBLE(100.0)
      P = YIELD*(VL/QO)*((K * SBD)/(KS + SBD)) - B*(VL/QO)
      WRITE(6,3)SBD, EBD, P,F
 3    FORMAT(6X,'SBD= ',F5.3,2X,'EBD= ',F4.1,2X,'%',5X,'P=    ',
     +F6.4,2X,'F= ', F6.4//)
C
C************************<<<<UPPER ZONE>>>>******************
C
      SU(1)     = SBD
      SUG(1)    = SU(1)
      SU(2)     = P * XBD
      XUG(1)    = SU(2)
      DO 50 KK = 5, HMAX, 5
      HEND = DFLOAT (KK)
      CALL DVERK(N, FCN1, H, SU, HEND, TOL, IND, C, NW, W, IER)
      IF(IND .LT. O .OR. IER .GT. O) THEN
      WRITE (6,35) KK
 35   FORMAT(1X,' ERROR IN IND OR IER AT KK = ', 13)
      GOTO 60
      ENDIF
```

```
      IF(SU(1). LT. 0.012DO) SU(1) = 0.012DO
 40   WRITE(6,4) H, SU(1), SU(2)
 4    FORMAT(6X,'H = ', F6.1, 3X,' SU = ',F6.4, 3X,'XU = ', F6.4/)
      IF( KK .LT. HMAX+1 ) NN = NN+1
      SUG(NN) = SU(1)
      XUG(NN) = SU(2)
      HENDG(NN) = HEND
 50   CONTINUE
      CALL GRAPHS(  HENDG, SUG, NN)
      CALL GRAPHX(  HENDG, XUG, NN)
 60   STOP
      END
C***************<<<< SUBROUTINE FCN1 >>>>**********************
      SUBROUTINE FCN1(N, H, SU, SUPRIM)
      REAL*8 DFM, DF, D, L, JJ, ROU, MUO, K, KS, XF, TAO,
     +KM, SUM, AA, AS, QO, E, CM, DELT, Q, Z, LM, VEL,
     +SUPRIM(N), SU(N), RE, T1, B, H, SCHEM, FAC, DIAM, P
      INTEGER I,N
      COMMON K, KS, B, QO, YIELD
      P      = DBLE(?.????)
      E      = DBLE(0.870)
      D      = DBLE(?.???)
      DIAM   = DBLE(?.??)
      MUO    = DBLE(?.??E-??)
      ROU    = DBLE(???.??)
      DF     = DBLE(0.80)*D
C     FAC    = DBLE(?.???)
C     XF     = FAC * SU(1)
      XF     = DBLE(?.???)
      AA     = DBLE(153.2)
      AS     = DBLE(3.41)
      VEL    = QO/((E**(2/3)*AA)*DBLE(100))
      RE     = (ROU*VEL*DIAM)/(MUO *DBLE(24*60*60*100))
      SCHEM = MUO/(ROU*D)*DBLE(10000*24*60*60)
      KM     = DBLE(1.10)*VEL/(E*RE**(2.0000/3.000)
     +*SCHEM**(2.000/3.000))*DBLE(100)
      DFM    = DF/D
      SUM    = SU(1)/KS
      L      = D / KM
      T1     = DBLE(2.0) * KS * DF/(K * XF)
      TAO    = DSQRT(T1)
      LM     = L/TAO
      DELT   = DLOG(SUM) - DLOG(DBLE(2.0) + DLOG(DFM)
     +/DBLE(2.303)) - DBLE(1.8) * DLOG(DBLE(1.0) + DBLE(2.0)
     +* LM * DFM) + DBLE(0.353)
      Z      = DELT * 0.4700
      Q      = DBLE(0.75) - DBLE(0.25) * DTANH(Z)
      CM     = (DBLE(2.0)* DFM *(DSQRT(DBLE(2.0)) + DBLE(2.0)
     +* LM * DFM)
     + **(DBLE(1.0) - DBLE(2.0) *  Q))/(DBLE(1.0) +
     +DBLE(0.54) * (DBLE(1.0) + DBLE(0.0121) *
     +DLOG(DBLE(1.0) +DBLE(2.0) * LM)) * (DBLE(1.0)
     +- DBLE(8.325) * DLOG( Q/ DBLE(0.707) ** DBLE(2.0)))
      SUPRIM(1)= - E * (AA/QO) * (( K * SU(2) *SU(1))
```

```
     +/(KS + SU(1)))-(AA/QO) * AS * CM * (SUM ** Q) *
     +( D * KS/TAO)
      SUPRIM(2) = (E/QO) * AA * SU(2) * ((YIELD * K * SU(1))
     +/(KS + SU(1))- B) - P * ((SU(2)* AA)/QO)
      RETURN
      END
C
C*******************<<<< GRAPH  COD>>>>************************
C
      SUBROUTINE GRAPHS (H, SUG, N)
      INTEGER M, K, I
      CHARACTER BLANK, DOT, CROSS, LINY(100)
      REAL*8 SUG(N), SCALE, H(N)
      DATA  BLANK, DOT, CROSS, LINY/ ' ', '.', '*', 100*' '/
      SCALE = DBLE(200.0)
      WRITE(6,40)
 40   FORMAT('1', 30X, 'GRAPH OF COD PROFILE'// 4X, 'H', 5X,
     +'COD'/4X,'CM',4X,'MG/CM3'//)
      DO 50 I = 1, N
      M = INT ( SUG (I)*SCALE + 0.5 ) + 1
      DO 55 K = 1, M
      LINY(K) = DOT
55    CONTINUE
      LINY(M) = CROSS
      WRITE(6,60) H(I), SUG(I), LINY
 60   FORMAT(1X, F6.2, 2X, F6.4, 2X, 100A1)
      DO 50 KK = 1, 100
      LINY(KK) = BLANK
 50   CONTINUE
      RETURN
      END
C
C*******************<<<<<  GRAPH  XU  >>>>>*********************
C
      SUBROUTINE GRAPHX(H,XUG,N)
      INTEGER M, K, I
      CHARACTER BLANK, DOT, CROSS, LINY(110)
      REAL*8 XUG(N), SCALE, H(N)
      DATA BLANK, DOT, CROSS, LINY/ ' ', '.', '*', 110*'  '/
      SCALE = DBLE( 9000.)
      WRITE(6,40)
 40   FORMAT('1', 30X, 'GRAPH OF XU PROFILE '// 4X, 'H', 6X,
     +'XU'/4X,'CM',4X,'MG/CM3'//)
      DO 50 I = 1, N
      M = INT( XUG(I)*SCALE + 0.5 ) + 1
      DO 55 K = 1, M
      LINY(K) = DOT
 55   CONTINUE
      LINY(M) = CROSS
      WRITE(6,60) H(I), XUG(I), LINY
 60   FORMAT(1X, F6.2, 2X, F6.4, 2X, 110A1)
      DO 50 KK = 1, 110
      LINY(KK) = BLANK
 50   CONTINUE
```

```
RETURN
END
```

APPENDIX B

WASTE WATER EFFLUENT STANDARDS

In response to the Water Pollution Control Act of 1972 (Public Law 92-500), the Environmental Protection Agency (EPA) established standards for wastewater discharges. Currently, the EPA requires that as a minimum all municipal wastewaters must be treated to the standards shown in table B-1. These standards are commonly referred to a secondary wastewater treatment standards. The States have the right to set more stringent effluent requirements than those shown in table B-1. EPA requires industries to treat their discharges to the level obtainable by the "best available technology" for the particular wastewater treated. EPA requires that all industrious discharges to municipal wastewater collections systems must be pretreated to become compatible with the untreated domestic wastewater. receiving streams are defined in the federal regulations as either "effluent-limited" or "water-quality-limited". If a stream meets its in-stream standards when all the discharges to that stream meet the secondary treatment and best available technology standards the stream is an effluent limited. Under the National Pollution elimination discharge System (NPDES), municipalities and industries discharging to effluent-limited streams are assigned discharge permits reflecting the secondary treatment and the best available technology standards. A water-quality stream is a stream which will not meet its

in-stream standard even when the wastewater discharges to it meet the secondary treatment standards shown in table B-1.

TABLE B-1: SECONDARY EFFLUENT CRITERIA FOR PUBLICLY OWNED TREATMENT FACILITIES.

Parameter	Monthly Average	Weekly Average
BOD, mg/L	30	45
SS, mg/L	30	45
Fecal Coliform bacteria, number/100 ml	200	400
pH	Within the range of 6.0 to 9.0	

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