ANAEROBIC WASTEWATER TREATMENT BY AN EXPANDED BED REACTOR

by

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AN EXPANDED BED REACTOR

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ANAEROBIC WASTEWATER TREATMENT BY AN EXPANDED BED REACTOR

The use of an Anaerobic Expanded Bed for the treatment of waste, high in organic load,is a subject under investigation in the last years.

Due to the fact that microorganisms active for the decomposition are forming a film on the large surface area of the filter media, as well as trapped in the voids, the efficiency system is expected to be higher than other conventional processes.

To investigate the performance of this system, a model consisting of three reactors was prepared. Out of these reactors, one was used as a control unit (not any media present) while in the other two different substrates were fed.

The most important results obtained from this study are :

i) The Chemical Oxygen Demand removal efficiency increased both for batch and semi-continuous reactors when the organic load of the system increased.

ii) Similarly, the removal efficiency of total volatile solids was proportional to the load of total volatile solids applied.

iii) The performance of the Anaerobic Expanded Bed system was better than the performance of the simple anaerobic reactor.

iv) Substrates containing yeast showed a better treatability than similar substrates without yeast in Anaerobic Expanded Bed Reactors.

ANAEROBİK GENİŞLETİLMİŞ YATAK'LI REAKTOR KULLANIMI İLE ATIKSULARIN TEMİZLENMESİ

Organik yükü fazla olan atıkların temizleme işleminde Anaerobik Genişletilmiş Yatak kullanımı son yıllarda önemli bir araştırma konusu olmuştur.

Sistemin veriminin alışılagelmiş metotlardan daha fazla olması beklenmektedir. Bunun başlıca sebebi filtreyi meydana getiren taneli malzemenin yüzeylerini kaplıyan mikrobiyolojik filmden başka tanelerin arasındaki boşluklarda da organik maddenin çözülmesini sağlıyan mikroorganizma flokları birikmesidir.

Bu sistemin başarısını incelemek için üç reaktörden oluşan bir model hazırlanmıştır. Bu reaktörlerden biri kontrol ünitesi olarak kum ortamı olmadan kullanılırken diğer iki reaktörde iki değişik atık kullanılmıştır.

Bu çalışmanın sonunda elde edilen en önemli sonuçlar şunlardır:

a) Yarı sürekli reaktörlerde ve kesikli (batch) reaktörlerde Kimyasal Oksijen İhtiyacı yoketme verimi sisteme verilen atığın organik yükünün artması ile birlikte artmaktadır.

b) Benzer bir şekilde, Toplam Uçucu madde yoketme verimi de uygulanan Toplam Uçucu madde ile doğru orantılıdır.

c) Geliştirilmiş Anaerobik Genişletilmiş Yatak Sisteminin atıkları arıtma kapasitesi konvansiyonel anaerobik reaktörlerden daha yüksektir.

d) Anaerobik Genişletilmiş Yatak'lı reaktörlerde maya ihtiva eden atıklar mayasız atıklardan daha fazla başarılı olmuştur.

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

CHAPTER I

INTRODUCTION

Anaerobic waste treatment is one of the principal waste treatment processes in use today. The Anaerobic filter is a feasible type of anaerobic waste treatment. This filter basically consists of a rock-filled bed in which the wastewater flows upward. The waste as it passes through the filter comes in contact with a large active biological mass which is accumulated on the surface of the filter media or in the voids existing in the filter bed. The Anaerobic Attached Film Expanded Bed (AAFEB) reactor is an advanced type of anaerobic biofilter which contains small sized-light weight media and utilizes attached microbial films on a larger surface area per unit weight.

Anaerobic treatment has a great advantage over aerobic treatment as far as energy requirements are concerned. Aerobic biological treatment, particularly of strong wastes, requires considerable amounts of energy while the energy consumed in anaerobic treatment is almost negligable. Consequently, anaerobic treatment has become an attractive wastewater treatment alternative.

This study presents one of the advanced types of anaerobic treatment namely the Anaerobic Attached Film Expanded Bed Reactor, and examines its feasibility.

After the first chapter, which is a general introduction to the subject, a literature review can be found in chapter two. In chapter three details of the experimental set-up are explained. This chapter is followed by chapter four, where the experimental procedure is presented. The obtained results and a discussion are given in chapter five. The study terminates with chapters six and seven where the conclusions and recommendations for further research take place.

CHAPTER II

LITERATURE SURVEY

The literature survey presented in this chapter is divided into two subsections. In the first subsection a general review of anaerobic treatment processes is given. This is followed by a summary of studies conducted in relation to anaerobic filtration.

2.I. General Review of Anaerobic Treatment Processes

The biological treatment processes used for wastewater treatment can be subdivided into two major groups as aerobic processes and anaerobic processes. In anaerobic processes, the organic wastes breakdown to methane and carbondixide while in aerobic processes the organic matter is converted to water and carbondioxide using the oxygen present in the system. Aerobic processes need the use of large quantities of energy resulting in rapid cell growth.

In anaerobic conditions, the microorganisms convert the organic material to methane and carbondioxide in the absence of oxygen. In this process the microorganisms take up relatively little energy and their rate of growth is small. In anaerobic treatment a small portion of the waste is converted to new cell material where as the largest part is converted to methane and carbondioxide.

2.1.1 Anaerobic Treatment Systems

The principal anaerobic treatment systems can be classified as;

i) Conventional Anaerobic Systems

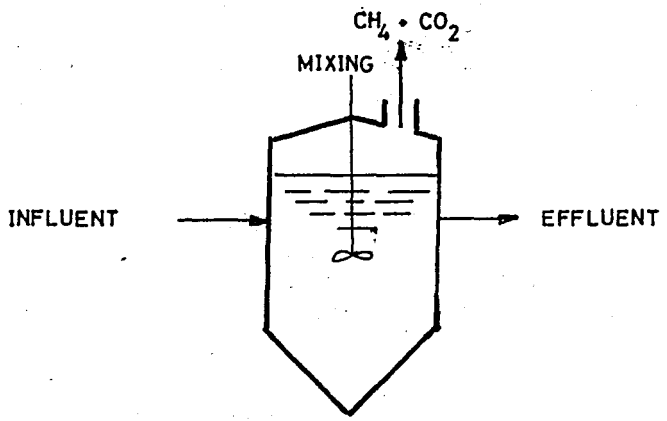
ii) Anaerobic Contact Systems

iii) Anaerobic Filter Systems

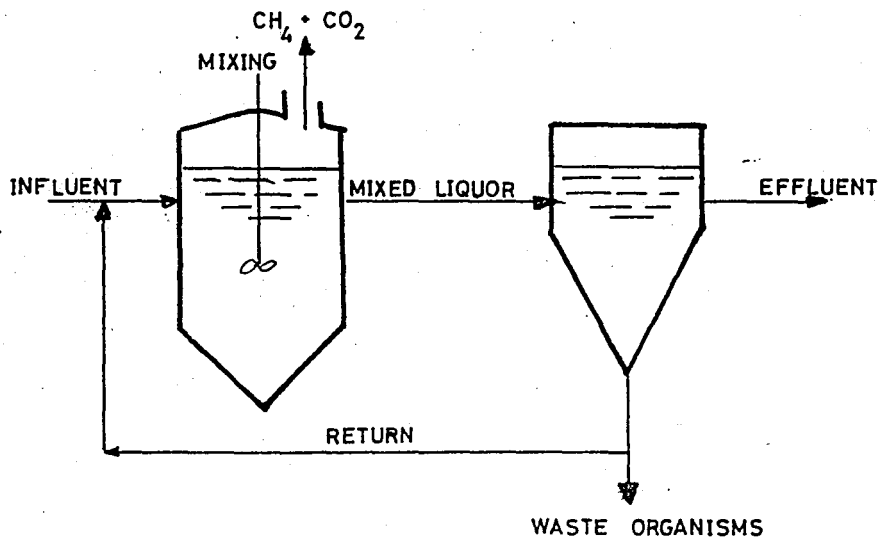
Figure 1 represents schematically the basic anaerobic systems.

2.1.2 Conventional Anaerobic Systems

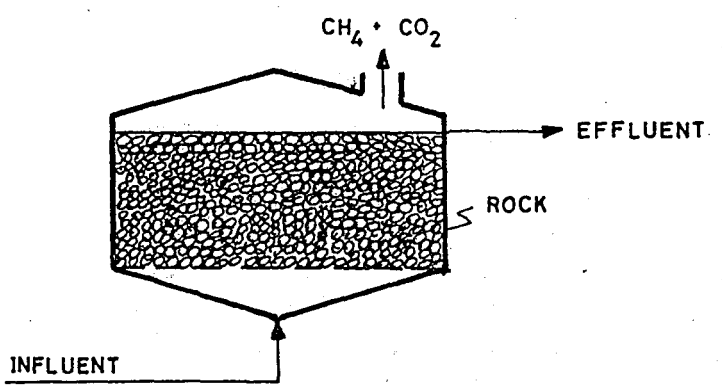
One of the oldest conventional anaerobic systems is the "septic



CONVENTIONAL PROCESS



ANAEROBIC CONTACT PROCESS



ANAEROBIC FILTER PROCESS

Figure: I. Basic Anaerobic Treatment Processes

tank." Septic tanks use the anaerobic decomposition process in the stabilization of domestic sludge. Septic tanks which have been in use for many years are rectangular chambers, usually cited just below ground level, in which sewage is retained for 1-3 days. During this time the solids settle to the bottom of the tank where they are digested anaerobically. A thick crust of scum is formed at the surface and this helps to maintain anaerobic conditions. In septic tanks, the settling and digesting solids are either kept in contact with the flowing wastewater in single-storied tanks or are separated from each other through a trapped slot into a digestion compartment in two-storied tanks.

In recent years, with the spread of wastewater treatment plants, conventional anaerobic treatment systems became to be known as simple digestrs where the sludge from the biological wastewater treatment units is treated before disposal. Usually this implies a completely mixed, flow-through system where the sludge detention time equals the hydraulic detention time. The main advantage of this process is its simplicity in design and operation. Conventional Anaerobic Systems require long detention times when compared with other systems. Mc Carty (1968) states that a detention time of about ten to thirty days at 35°C is necessary for practical control and reliable treatment. Mosey (1974) indicated 51 to 95 per cent removal efficiency in the mesophilic temperature range (25-38 C) for a wide range of different inlet concentrations and detention times.

2.1.3 Anaerobic Contact Systems

The anaerobic contact system differs from the conventional one by the return of biologically active sludge to the digester. This process has been successfully used for the stabilization of high strength so-

Table 1. Performance Data for the Anaerobic Contact Process , (Mosey, 1974)

Type of waste	Scale of Plant	Digestion temperature (C°)	Hydraulic retention time(d)	Parameter	Influent (mg/L)	Effluent (mg/L)	Percentage removal
Slaughterhouse	Laboratory	33	0.69	BOD	I500	I00	93
"	"	33	2.94	"	1500	84	94
"	Pilot	33	1.25	BOD	2I00	90	96
"	"	33	1.25	Org.C	940	92	90
Meat packing	Pilot	35	0.5	BOD	1600	80	95
"	Full	32	0.54	"	1380	180	9I
"	"	32	0.34	SS	990	200	80
Maize-starch	Full	Ambient	3.3	BOD	6280	755	88
"	"	"	3.3	Org.C	3250	3I7	90
"	"	"	3.3	VS	6556	623	90
Brewery	Pilot	not stated	2.23	BOD	3280	I30	96
Distillery	Full	30	7.2	COD	22400	540	98
"	"	30	5.3	"	12600	400	97
"	Laboratory	33	6.2	BOD	25000	986	96
"	"	33	6.2	Org.C	12000	18I2	85
"	Laboratory	35	0.92	BOD	845	60	93
"	"	35	0.92	TS	I820	850	53
Citrus	Laboratory	34	I.38	BOD	2670	I30	95
"	Pilot	34	2.32	"	3440	II00	68
Yeast	Laboratory	30	2.	BOD	3042	39I	87
"	Pilot	30	I.7	"	5076	76I	85
Chewing gum	Full	not stated	11.7	BOD	I840	740	60
Milk	Laboratory	3I	6	BOD	3300	I0-20	99.5
"	"	3I	6	"	380	20-40	90
"	"	3I	6	VS	3750	260	93
"	"	3I	6	"	3I0	I40	55

luble wastes. Mc Carty (1968) claims that such a system has been found economical with wastes having BOD concentrations of about 1000 mg/L. The detention time of such systems varies between 6 and 12 hours .The gas formed in the settling tank prevents the microbial particles to settle down and thus recycling of the settled sludge is impossible. To overcome this disadvantage a degasifier can be used between the digester and the settling tank. The settling tank can have the shape of a gravity thickener or vacuum floatation unit. A gravity thickener operates very much like a settling tank; the substrate entering in the middle is distributed radially. The sludge is withdrawn from the bottom of the tank and is pumped to the digester tank. In the vacuum flotation unit a partial vacuum is applied, which causes dissolved gasses to come out of the substrate as bubbles. The bubbles and the attached solid particles rise to the surface to form a scum blanket, which is removed by a skimming mechanism. Thus in anaerobic contact systems, it becomes possible to decrease the detention time by recycling the biologically active part by the methods mentioned above.

Mosey (1974) tabulated the performance data of anaerobic contact processes for different types of wastes with a 60 to 99.5 per cent BOD removal range .(Table 1).

2.1.4 Anaerobic Filter

The anaerobic filter is one of the best anaerobic treatment processes for soluble wastes. In this process, the waste is forced to pass through a packed material. Anaerobic microorganisms attach themselves to the surface of the packed material or are in voids found in the media and are thus not carried out in the effluent stream.

The anaerobic filter is ideal for the treatment of soluble wastes while it is not recommended for wastewater having a high concentration

of suspended solids,since such wastes would clog the filter. For such a case either an upflow anaerobic filter or expanded bed-fluidized bed filter is recommended.

Another natural superiority of the upflow anaerobic filter is the rising of gas produced in the lower part providing turbulance. This mixes and helps to maintain clear passageways for the wastewater.

Table 2 summarizes some data reported by Mosey (1974) about laboratory scale anaerobic filters.

Generally,in expanded bed filters, wastewater passes through a filter media from bottom to top with a velocity sufficient to expand the bed. Once the filter bed is expanded, the media provides a vast surface area for biological growth, leading to the development of biomass concentration greater than that maintained in an activated sludge system. Further increase of the velocity of the wastewater results in a fluidized state. When biological growth occurs on the media,the particle diameter increases and the overall density of media is reduced. To prevent the overall density of the biofilm covered media (bio-particle) from decreasing to a level where bed carry-over occurs, it is necessary to limit the biofilm thickness.This can be achieved by controlling the bed expansion.

2.2 Microbiology and Biochemistry

2.2.1 Microbiology

Anaerobic degradation is performed mainly by two groups of bacteria a) the acid producing, and b) the methane producing bacteria. These two groups are subdivided into two subgroups each, as shown in Figure 2, (Henze et.al., 1982). The specific species of anaerobic bacteria have been discussed in details by Zehnder (1978) and (1981), Balch et. al. (1979).

Table 2. Performance Data for Laboratory Scale Anaerobic Filters (Mosey, 1974)

Type of waste	Digestion temperature (C)	Hydraulic retention time(d)	Parameter	Influent (mg/L)	Effluent (mg/L)	Percentage removal
Raw domestic sewage	4	1.5	BOD	180	40-40	67
"	25	1.5	"	180	10.35	82
"	25	1.5	SS		9	95
Protein/carbohydrate	25	0.75	COD	1500	122	91.5
"	25	0.375	"	1500	312	79.3
"	25	0.187	"	1500	950	36.7
"	25	3.00	"	3000	204	93.4
"	25	1.50	"	3000	247	88.4
"	25	0.375	"	3000	1100	63.0
Volatile acids	25	1.5	COD	1500	24	99.4
"	25	0.75	"	1500	139	90.5
"	25	0.375	"	1500	314	79.0
"	25	3.0	"	3000	42	98.6
"	25	1.5	"	3000	240	92.0
"	25	1.5	"	6000	139	97.7
"	25	0.75	"	6000	794	86.9
Food processing	35	3.56	COD	8475	546	93.5
"	35	3.56	BOD	5200	975	81.4
"	35	3.56	Org.C	2400	115	95.2
"	35	3.56	SS	1508	455	70.0
"	35	0.54	COD	8475	5000	41.0
"	35	0.54	BOD	5200	3890	25.2
"	35	0.54	Org.C	2400	1720	28.3
"	35	0.54	SS	1508	1855	-

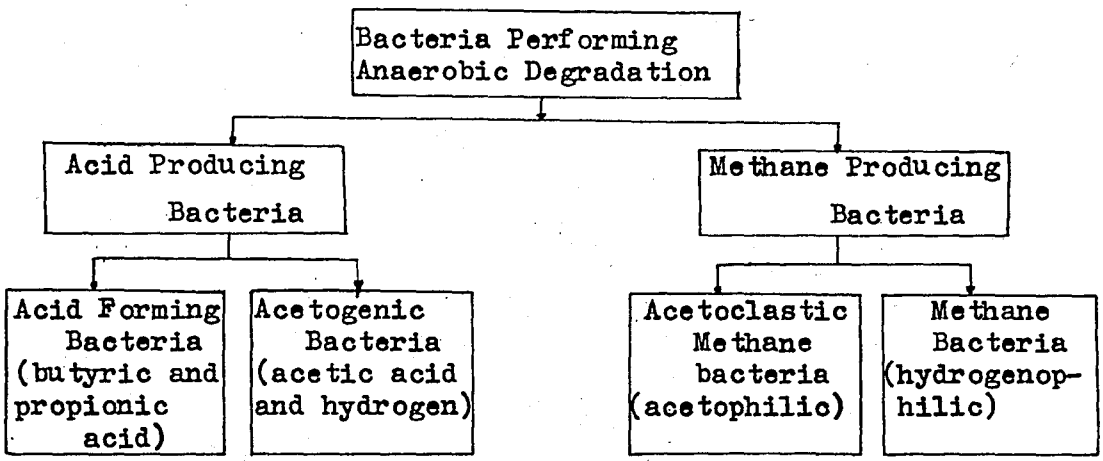


Figure 2. Major Groups of Anaerobic Microorganisms(Henze et.al.,1982)

2.2.2 Steps of Reaction

The anaerobic digestion of a complex substrate can be regarded as a three step process:

Step 1. Liquefaction,hydrolysis of suspended organics and soluble organics of high molecular weight.

Step 2. Acidification, degradation of small organic molecules to various fatty acids and ultimately to acetic acid.

Step 3. Gasification, production of methane,primarily from acetic acid and also from hydrogen and carbondioxide.

A schematic representation of the stages of anaerobic decomposition is given in Figure 3 (Edeline,1976).

Gujer and Zehnder (1982) state that of the three steps, the second one is rather quick, while the other two are slow.

Hydrolysis of organic matter is a rather slow process brought about by extracellular enzymes. Lipids are hydrolyzed very slowly.Consequently the hydrolysis step may be the overall rate limiting step for wastes containing a considerable amount of lipids and other slowly hydrolyzing compounds. Eastman and Ferguson (1981) have demonstrated that in a separate acid producing reactor, hydrolysis is always the rate limiting

step.

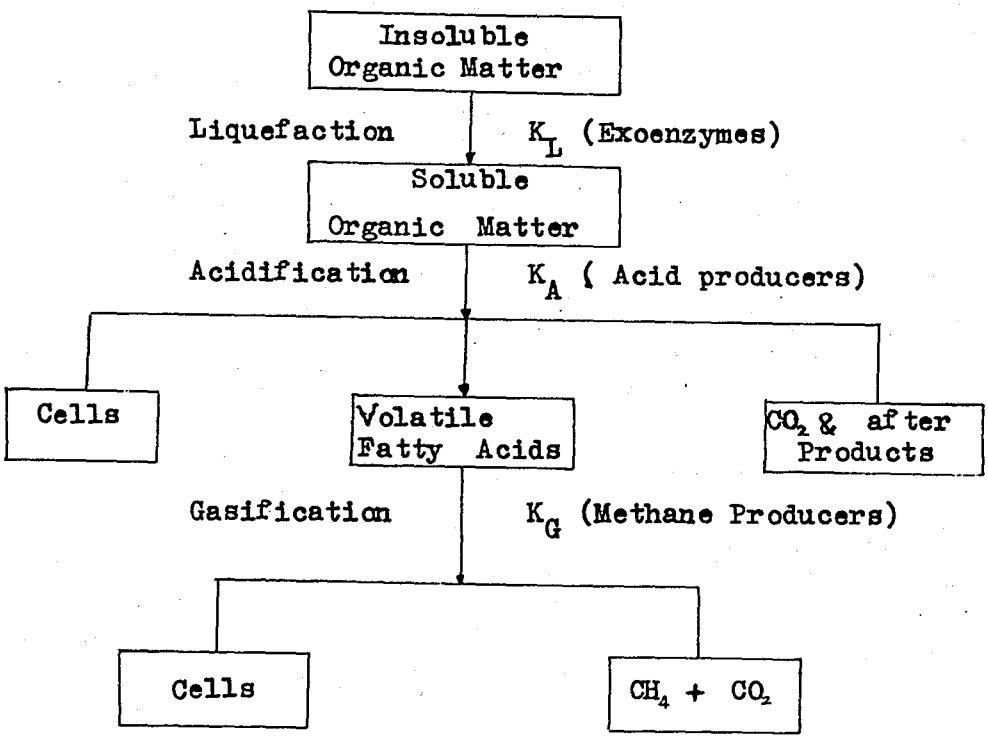


Figure 3. Stages of Anaerobic Decomposition (Edeline,1976).

The Acid Production step results in the formation of acetic acid or in case of instability, higher fatty acids such as propionic,butyric valeric and iso-valeric acid. In the acidification phase,energy is released for cell growth and a small proportion of the organic vaste is converted to cell material.In addition, a substantial proportion of the organic nitrogen is converted to ammonium ions,and organic sulphur appears as sulphide. Then the methane producing bacteria utilize the organic acids in the second stage, converting them into carbondioxide and methane. According to Mc Carty (1968),enzymatic hydrolysis of complex organics to organic acids coult be referred to as "waste conversion" and generation of methane and carbon dioxide from organic acids as "waste stabilization."

The acid production rate is high as compared to the methane production rate, which means that a sudden increase in easily degradable (soluble) organics will result in increased acid production with subsequent accumulation of the acids. This might inhibit the next step of the process, the methane step, because methane formers are very sensitive to pH. The accumulation of acids might lower the pH and necessary conditions for methane formation may not exist. Parallel to the acid production, ammonia is released by the degradation of proteins and amino acids (Mc Cready, 1978). The ammonia concentrations thus established will generally not be of a magnitude that will inhibit the anaerobic process, but for nitrogen rich wastes treated in highly loaded processes, ammonia inhibition could occur.

Methane production is a slow process and it is produced from acetic acid or from hydrogen and carbondioxide. About one third of the methane has its origin in molecular hydrogen (Gujer and Zehnder, 1982; Jeris and Mc Carty, 1965; Smith and Mah, 1966). Small amounts of methane can be produced from methanol and formic acid but these reactions have little practical importance (Smith and Mah, 1978). The bacteria producing methane from hydrogen and carbondioxide are fast growing ones as compared with the acetic acid utilizing bacteria. Generally, the methane producing reaction is the rate limiting factor, but hydrolysis may also play an important role (Gujer and Zehnder, 1982). The difference between the two is that the methane bacteria must exist in the reactor. The hydrolys of degradable suspended solids may be beneficial for the process, but is not essential for the process to function.

2.3 Environmental Factors Affecting Anaerobic Treatment

The following environmental factors can affect the efficiency of anaerobic treatment processes.

2.3.1 Temperature

Anaerobic decomposition may take place over a wide range of temperatures (5-60 °C), but the most common range for anaerobic process operation is 30-40°C. The three different temperature ranges for the anaerobic treatment process may be given as follows:

i. Psychrophilic: 5-25°C

ii. Mesophilic :25-38°C

iii. Thermophilic: 50-60°C

In general reactions in the psychrophilic range are very slow and this results in a residence time for microorganisms of about 100-300 days (Downing and Kell, 1980). In the mesophilic range, the reaction rate is higher and requires residence times in the range of 20 to 40 days. At temparatures of 40-45 C the microbial activity is still significant, but due to a high decay rate the observed yield coefficient of methane bacteria approaches zero, and this prevents continuous operation at that temperature (van den Berg 1977). Thermophilic processes have a rather constant methane production rate, independent of temperatures in the range of 50-70°C. The rate is about 25-50 % higher than the mesophilic rate at 35°C. The major disadvantage of the thermophilic processes is very slow start-up and very slow accomodation to loading variations, substrate changes or toxic substances. Another problem in the thermophilic range is that very few bacterial species are able to grow at high temperatures. Speece and Kam (1970) state that the response to quick temparature changes will be a temporary stop of activity. Henze and Harremoes (1982) noted that temperature changes in an anaerobic process in operation may be done stepwise; 1°C per day. Under these conditions, the microorganisms will adapt without halt in the metabolic processes, although the metabolic rate will change.

2.3.2 Nutrients

When waste with only small amounts of nutrients are subject to anaerobic treatment nutrient deficiency may occur. Often the COD/N-ratio or the COD/N/P-ratio is used to describe the nutrient requirements. The N/P ratio can be considered to be 7 (Speece and Mc Carty, 1964). The theoretical nutrient requirement as a function of organic load is given in Figure 4 (Henze and Harremoes,1982).

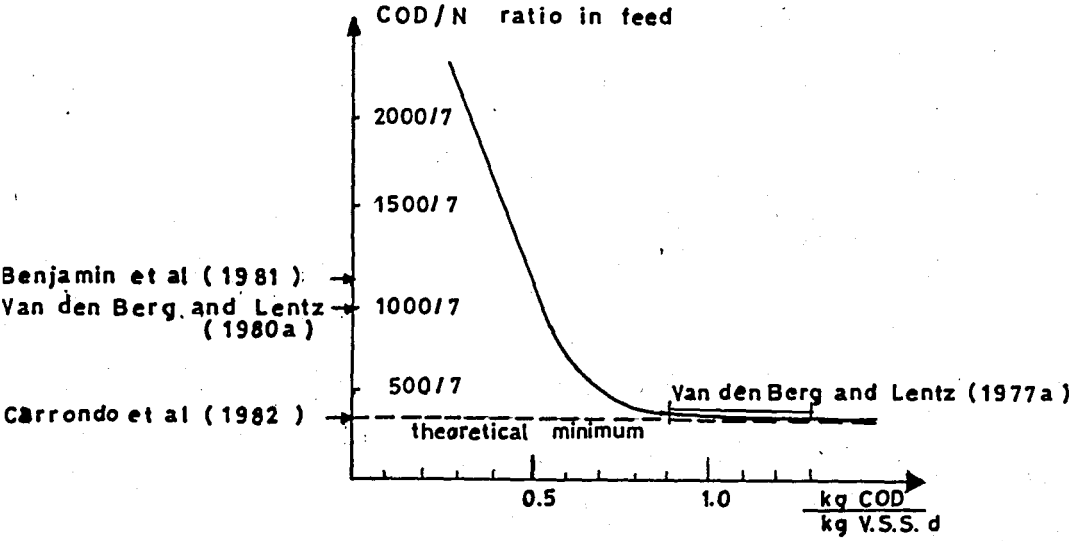


Figure 4. Theoretical nutrient requirement as a function of organic load (Henze and Harremoes,1982)

The theoretical minimum COD/N-ratio is observed to be 350/7. A value around 400/7 must be regarded reasonable for high loaded processes. For low loaded processes the COD/N-ratio is observed to values of 1000/7 or more. Many observations gave COD/N ratios of 200/7-300/7 which are too small for normal process operation (Benjam et.al,1981; Martensson and Frostell,1982; van den Berg and Lentz,1980

Other than the nitrogen and phosphorous,nutrients given in Tabl are also essential for anaerobic processes. For high loaded anaerobic industrial waste processes the possibility of nutrient deficiency mus

be considered. Addition of standard nutrient salts or yeast extract might be necessary.

Table 3. Micro-Nutrients/Compounds Improving Process Performance of Anaerobic Treatment Systems.

Compound	Beneficial concentration in the study (g/m^3)	Effect	Literature
Fe^{++}	0.2	Prepicitation of sulphide Flocculation /Biofilm structure	Speece and Mc Carty (1964)
Ni^{++}	0.01	Part of F_{430} cofactor in metonogens	Thauer (1981)
Mg^{++}	0.01-0.02	Flocculation	Lettinga et. al. (1980-a)
Ca^{++}	0.01-0.04	Flocculation	Lettinga et. al. (1980-a)
Ba^{++}	0.01-0.1	Flocculation	Lettinga et. al.(1980-a)
Co^{++}	0.01	Vitamin B_{12}	Speece and Mc Carty (1964)
SO^{--}-S	0.02	-	van den Berg and Lentz (1980-a)

2.3.3 Inhibition/Toxicty

Thera are many organic and inorganic materials which may be toxic or inhibitory to the anaerobic waste treatment process. At low dosages a toxic material causes a stimulatory effect by which a maximum rate of biological activity is reached. The most important toxic,inhibitory or stimulatory materials are heavy metals,sodium,calcium, ammonium, and sulphide ions.

Sulphide toxicity is of particular interest because the non-toxic

sulphide and sulphate may be converted to the toxic sulphide under anaerobic conditions. The toxicity of sulphide is closely related to free hydrogen sulphide concentration. Low pH (lessthan 6.5) increases toxicity, where as the presence of iron reduces toxicity due to precipitation of ferrous sulphide.

Ammonia is another potential inhibitor. Free ammonia is the most toxic compound (inhibition at 0.1-0.2 kg N/m^3). Total ammonium-ammonia concentrations of 5-8 kg N/m^3 can be tolerated, if the pH of the reactor is low enough (Henze and Harremoes, 1982). Table 4 presents the optimum and inhibitory concentrations of inorganic ions.

Volatile acids are potential inhibitors (Mc Carty and Mc Kinney, 1969). According to Mc Carty and Mc Kinney (1965), volatile acids are not toxic to methane producers up to 6000mg/L, so there is no problem for a typical anaerobic treatment process due to volatile acid concentration. It is a good idea however to neutralize the acid with non-toxic concentrations of alkaline substances before being fed to the reactor inorder to prevent adverse changes in the pH of the system.

pH is one of the most important environmental factors affecting the anaerobic treatment. The methane bacteria have an optimum pH range between 6 and 8 whereas the acid producing bacteria have an optimum range between 5 and 6. As the methane step is the rate limiting one, the pH should be kept above 6. Clark and Speece (1971), demonstrated an optimum pH range of 6-8 for anaerobic filters.

Methane bacteria are strictly anaerobes and even low concentrations of dissolved oxygen are toxic to them. This is rarely a practical problem in treatment processes, since generally the facultative organisms rapidly remove any traces of dissolved oxygen. Anaerobic treatment should be carried out in complete absence of oxygen.

Table 4. Optimum and Inhibitory Concentrations of Inorganic Ions in Anaerobic Reactors

	Optimum Concentration Range	Concentration Causing Moderate Inhibition	Concentration Causing Strong Inhibition	Reference
Sodium (mg/L)	100 - 200	3500-5500	8000	
Potassium (mg/L)	200 - 400	2500-4500	12000	
Calcium (mg/L)	100 - 200	2500-4500	8000	McCarty (1964)
Magnesium (mg/L)	75 - 150	1000-1500	3000	
Ammonia (mg N/L)	50 - 1000	1500	3000	
Sulphide in solution (mg S/L)	0.1 - 10	100	200	McCarty (1964) & Mosey & Hughes (1975)
Chromium (percent of total solids)	Not known	2	3	Mosey & Hughes (1975)
Cobalt (mg/L)	20	Not known	Not known	Speece & McCarty (1964)

2.4 Kinetics

The literature dealing with biofilm kinetics in anaerobic processes is very limited. Table 5 lists some investigations related to anaerobic biofilms.

Table 5. Investigations Related to Anaerobic Biofilms.

Authors	Model
De Walle and Chian (1976)	Partially penatrated biofilm, intrinsic 0' Order 1' - order at low subs trate concentrations
Rittman and Mc Carty (1980)	General anaerobic process model
Lindgren (1982)	Completely mixed reactors in series Monod Kinetics
Shieh and Mulcahy (1982)	Biofilm model, 0' -order

The anaerobic degradation of organic matter, mentioned in previous sections, can be looked upon as a sequence of only three processes:

Hydrolysis of particulates to soluble substrates

Degradation of soluble substrates to acetic acid

Reduction of acetic acid to methane

Because of the complexity of these processes, the above three bulk reactions attempt to simulate adequately the basic anaerobic concept by covering their main features.

Although the processes involved in anaerobic reactors are very complex, the traditional anaerobic process concept has been worked on for many years. Henze and Harremoes (1982) claim that the anaerobic systems operated properly as long as the loading is within the conventionally experienced limits (Table 6).

Table 6. Typical Kinetic Coefficients for the Anaerobic Digestion of Various Substrates (Metcalf & Eddy, 1979).

	Coefficient	Basis	Value Range	Typical
Domestic Sludge	Y	mg VSS/mg BOD	0.040-0.100	0.06
	k_d	d^{-1}	0.020-0.040	0.03
Fatty Acid	Y	mg VSS/mg BOD	0.040-0.070	0.050
	k_d	d^{-1}	0.030-0.050	0.040
Carbohydrate	Y	mg VSS/mg BOD	0.020-0.040	0.024
	k_d	d^{-1}	0.025-0.035	0.030
Protein	Y	mg VSS/mg BOD	0.050-0.090	0.075
	k_d	d^{-1}	0.010-0.020	0.014

It is reasonable to assume that microorganisms use the substrate for growth, maintanance and multiplication. Substrate refers to the concentration of biodegradable organics in the waste stream. According to Downing and Kell(1979) within any system under steady state conditions, a material balance for the net rate of change of microbial mass can be written as:

$$\begin{bmatrix}\text{The rate of change}\\ \text{of microbial mass}\end{bmatrix} = \begin{bmatrix}\text{The growth rate}\\ \text{of microorganisms}\end{bmatrix} - \begin{bmatrix}\text{Washout or decay}\\ \text{rate of micreorganisms}\end{bmatrix}$$

Lawrence and Mc Carty (1969) expressed the above expression for the net growth rate of microorganisms as follows;

$$\frac{dX}{dt} = Y\frac{dS}{dt} - k_d X \qquad \text{(Equation 1)}$$

where,

dX/dt = net growth rate of microorganisms per unit volume of reactor, (mass/volume-time)

Y = maximum growth yield coefficient, defined as the ratio of the mass of cells formed to the mass of substrate consumed, (mass/mass).

of cells formed to the mass of substrate consumed,(mass/mass).

dS/dt = rate of microbial substrate utilization per unit volume,(mass/ volume-time)

k_d = microbial decay coefficient, ($time^{-1}$)

X = microbial mass concentration, (mass/ volume)

Eqn. 1 can be written as,

$$\frac{dX/dt}{X} = Y \frac{dS}{X.dt} - k_d X$$

The mean cell residence time or sludge retention time, θ_c, is defined as the ratio of the total cell mass in the system to the wasted cell mass per unit time, and is given as:

$$\theta_c = \frac{X}{(\Delta X/\Delta t)_T} \qquad \text{(Equation 2)}$$

where,

X_T = total active microbial mass in the treatment system, (mass).

total quantity of active microbial mass withdrawn daily, including those solids purposely wasted as well as thase lost in the effluent, (mass/time).

Lawrence and Mc Carty (1970) introduced a new parameter, namely the specific utilization, U,for the design and operation of biological systems,which is expressed as:

$$U = \frac{(\Delta S/\Delta t)}{X} \qquad \text{(Equation 3)}$$

where, $(\Delta S/\Delta t)_T$ is the rate of microbial substrate utilization per unit volume, mass per volume-time.

By inserting the new parameters defined above, θc and U, into equation 1, the following relation is obtained;

$$\frac{1}{\theta_c} = YU - k_d \qquad \text{(Equation 4)}$$

Monod (1950) on the other hand proposed the following equation

Table 7. Kinetic Coefficients for Anaerobic Digestion

Temperature	Substrate	Growth Coeffs. Y(mg/mg)	K(day)	Waste Removel Coeffs. K(mg/mg-day)	K(mg/L)	Reactor type	References
20	Domestic Sludge	0.040	0.015	3.5	4620	Cont.fed anaerobic digesters	O'Rourke(1968)
"	Acetic A.	0.040	0.015	3.6	2130	"	"
"	Propionic Acid	0.040	0.015	-	3860	"	"
"	Stearic/ Palmitic	0.040	0.015	3.85	4620	"	"
"	Mixed A.	0.040	0.015	3.85	10620	"	"
25	Acetic A.	0.054	0.011	4.70	869	Anaerobic chemostat	Lawrence/Mc Carty(1969)
"	Propionic A	0.041	0.040	9.8	613	"	"
"	Stearic/ Palmitic	0.040	0.015	4.65	3720	Cont.fed anaerobic digesters	O'Rourke(1968)
"	Mixed A.	0.040	0.015	4.65	5790	"	"
"	Volatile Acids	-	-	2.5	950	Anaerobic Filter	Switzen baum Danskin(1982)
30	Acetic A.	0.040	0.015	8.1	154	Anaerobic chemostat	Lawrence/Mc Carty(1969)
35	Acetic A.	0.044	0.015	8.5	105	"	"
"	Acetic A.	0.050	-	8.8	250	"	Smith Mah(1980)
"	Propionic Acid	0.042	0.010	7.38	60	"	O'Rourke(1968)
"	Mixed A.	0.040	0.015	6.67	2000	"	Huser(1981)
"	Volatile A.	-	-	1.4	1000	Expanded Bed	Switzenbaum Danskin(1982)
37	Acetic Acid	0.023	-	4.8	28	Anaerobic chemostat	Huser(1981) Zehnder et.al.(1980)
"	Stearic A.	0.110	0.01	0.91	417	"	Novak and Carlson (1970)
"	Oleic Acid	0.110	0.01	1.04	3180	"	"
"	Linoelic A.	0.110	0.01	5.09	1816	"	"

which relates the rate of substrate utilization to the concentration of microorganisms in the reactor and to the concentration of substrate surrounding the organisms,

$$\frac{dS}{dT} = \frac{kSX}{K_s + S} \qquad \text{(Equation 5)}$$

where,

k = maximum rate of substrate utilization per unit weight of microorganisms, ($time^{-1}$)

S = concentration of substrate surrounding the microorganisms, (mass/volume)

K_s = half-velocity constant, substrate concentration at one-half the maximum growth rate, mass/unit volume.

The rate of substrate utilization is defined as follows,

$$r_{su} = \frac{kXS}{K_s + S} = -\frac{S_o - S}{\theta} \qquad \text{(Equation 6)}$$

where,

r_{su} = substrate utilization rate, (mass/unit volume.time)

S_o = influent concentration of substrate

θ = hydraulic detention time [total volume of reactor(V)] / [flowrate of the reacter (Q)]

The arrangement of the equation 6 becomes as,

$$\frac{X\theta}{S_o - S} = \frac{K_s}{k}\cdot\frac{1}{S} + \frac{1}{k} \qquad \text{(Equation 7)}$$

The kinetic coefficients Y, k, Ks and kd for a specific waste can be obtained by using equations 4 and 7 after conducting the related experiments.

The kinetic coefficients for different substrates in different treatment systems encountered in the literature in Engineering Index and Dissertation abstracts are given in Table 7.

2.5 Summary of Studies Conducted in Relation to Anaerobic Filtration:

It has been known for decades that anaerobic fermentation processes can be used to convert organics to a combustible gas composed of methane, carbondioxide and other trace impurities. Numerous investigators have attempted to introduce some variation to the conventional anaerobic fermentation process in order to increase its efficiency. However, nobody was successful in developing a more economical alternative. Winslow and Phelps (1911) used a "Biolytic" tank consisting of an inverted conical tank containing a blanket of digesting biomass to treat domestic wastewater anaerobically.

The anaerobic filter was another variation of anaerobic treatment. At first, this filter was a submerged column packed with rocks, which provided a support surface for the attachment of microorganisms (Figure 4) . Coulter et.al.(1957) developed a two stage system for the treatment of sewage. This system consisted of a sludge contact chamber and a packed rock contact chamber. The total hydraulic retention time (HRT) within the system was approximately 35 hours with only a 2.5 hours hydraulic retention time required for the filter portion of the system. This unit was not supplied with any external heat, but rather was operated at simulated winter and summer ambient temperature conditions. Results from this study showed that during winter conditions (4 °C) approximately 67 percent of the BOD and more than 90 percent of the suspended solids were removed. Operation at summer conditions (20 ° C) showed a significant increase in BOD removal efficiency (approximately 81 per cent) , while suspended solids removal remained the same.

Pretorius (1969) developed a similar system to the one proposed by Coulter et.al.(1957) for the treatment of raw sewage. This system was also a two-stage process consisting of a sludge contact chamber and a

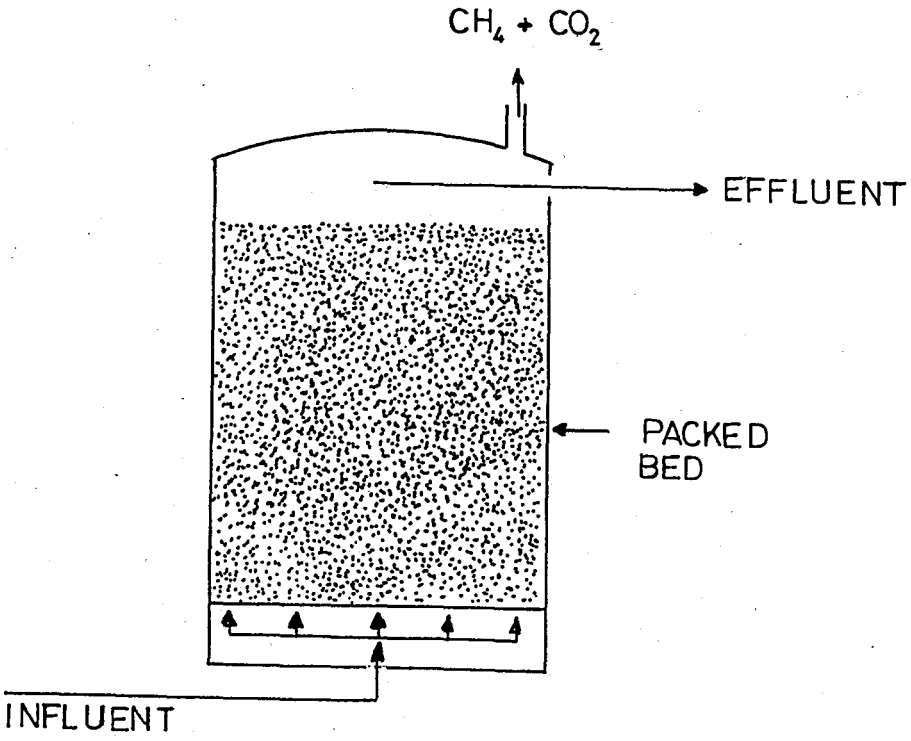


Figure:5. The anaerobic submerged packed column (Mc Carty, 1968).

packed anaerobic filter which was termed a biophysical filter. This unit was operated at 20°C and the H R T of the entire system was varie between one and two days. The sludge contact chamber provided the conversion of organics to volatile fatty acids while in the biophysical filter the acids were converted to carbondioxide and methane gas.

In the early experiments conducted by Mc Carty (1968), four particular soluble wastes were used as the substrate for the anaerobic filter: a) methanol, b) methanol plus acetate, c) methanol plus acetate plus propionate and d) acetate plus propionate. The waste strength varied between 2140 and 2650mg/L COD and hydraulic retention times of 6 and 12 hours were used. COD removal efficiencies were found to vary between 74 and 88 percent for all conditions. In addition, analysis of

effluent solids and solids accumulated within the system showed that the solids retention time (SRT) of the unit was greater than 100 days.

Young and Mc Carty (1969) utilized a protein carbohydrate synthetic substrate. Influent COD concentrations varied between 1500 to 3000 mg/L and HRT's of between 4.5 and 72 hours.At the lowest loadings COD removal efficiencies of more than 90 per cent were observed, whereas at the highest loadings the efficiency dropped to values between 36.7 and 63.0 per cent.

The most obvious disadvantage of the submerged filter is the probability that as biological sludge accumulates on the static surfaces, clogging may occur. Also the large static packing material limits the quantity of viable organisms to that which could exist in suspension or on the surface of the filter media.

Lawrence et. al. (1969) noted that control of the solids retention time (SRT)is essential for process stability. Pretorius (1971) considered anaerobic digestion with concentrated wastewaters showing that even highly diluted ones can be treated with high efficiencies provided the SRT is carefully controlled.

In practice, the control of the SRT is difficult, as the gas produced tends to adhare to the sludge and results in poor sedimentation. Lettinga (1975), (1976) and Heertjes et . al. (1978) proposed many solutions to the problem and the most promising ones are the anaerobic filter and the upflow reactor.

The anaerobic filter has given excellent results in many laboratory investigations and it combines high loading capacity with high efficiency without the need for external sludge separation (Young et.al.,1969 ; Plummer et.al.,1969 ; Jennet et.al.,1975 ;Mueller et.al.,1977 ; Norman et.al.,1978 . Frostell (1981) studied anaerobic treatment by comparing a

sludge bed system with a filter system. Although the sludge bed reactor may combine the advantages of the filter process (high loads,low effluent suspended solids concentration) with those of the anaerobic contact process (low construction cost,no need for an influent free from suspended solids),there is a large risk of the possibility of further bed expansion. Unless very sophisticated sludge separation systems are installed,heavy sludge losses may occur with disastrous results from the high loads applied. Furthermore,the factors regulating the formation of a stable sludge bed are not known with certainty (Lettinga, 1975,1976)

Frostell (1981) studied the Chemical Oxygen Demand (COD)removal rate at different organic loadings in sludge bed reactor and in anaerobic filter. The COD removal rates increased linearly with organic load in both reactors (Figure 6).

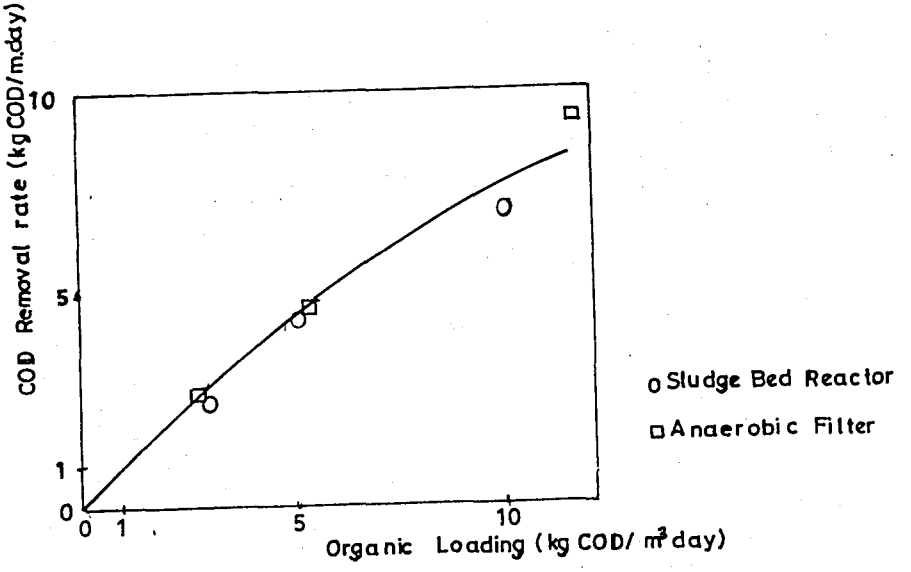


Figure: 6. Chemical Oxygen Demand (COD) removal rate at different organic loadings (Frostell,1981)

Anaerobic biofilters offer an alternative to the suspended sludge system and are a form of contact system that provides supportive mdia for anaerobic organisms and thus effectively increases the SRT over the usually short HRT . Oleszkiewicz et. al. (1982) observed that anaerobic biofiltration can be applied to pretreatment of strong organics and polishing of dilute effluents at ambient temperatures.

The success of the anaerobic packed bed has led other investigators to use fluidized or expanded bed technology for the removal of contaminants from wastewaters. Attached microbial films in trickling filters are krown to be easily managed and the microbial attached film expanded bed process was a natural development of trickling filters (Figure 7). Jewell and Mc Kenzie (1972) showed that static attached films enabled accumulation of up to 10 times the mass per unit volume over the suspended microbial systems. The thick biomass accumulated in the static void spaces caused clogging problems and substrate diffusion limitations. Thus it is necessary to optimize the biological process for utilizing attached microbial films on a large surface area per unit volume inorder to minimize diffusion limitation. Also the process should be designed so that media does not clog. One method of achieving these characteristics was to utulize small, inert particles that would encourge microbial attachment. To eliminate clogging , a process utilizing an upflow design was proposed. Jeris et.al. (1974) cited four advantages in using expanded/fluidized bed technology over the packed bed systems: "a) greater surface area available for growth per unit of reactor volume, b) very small headloss, c) no danger of clogging,and d) easier carrier removal procedure."The primary difference between the term "fluidized " and "expanded" refers to the change in reactor volume when the unit is in operation. Fluidization

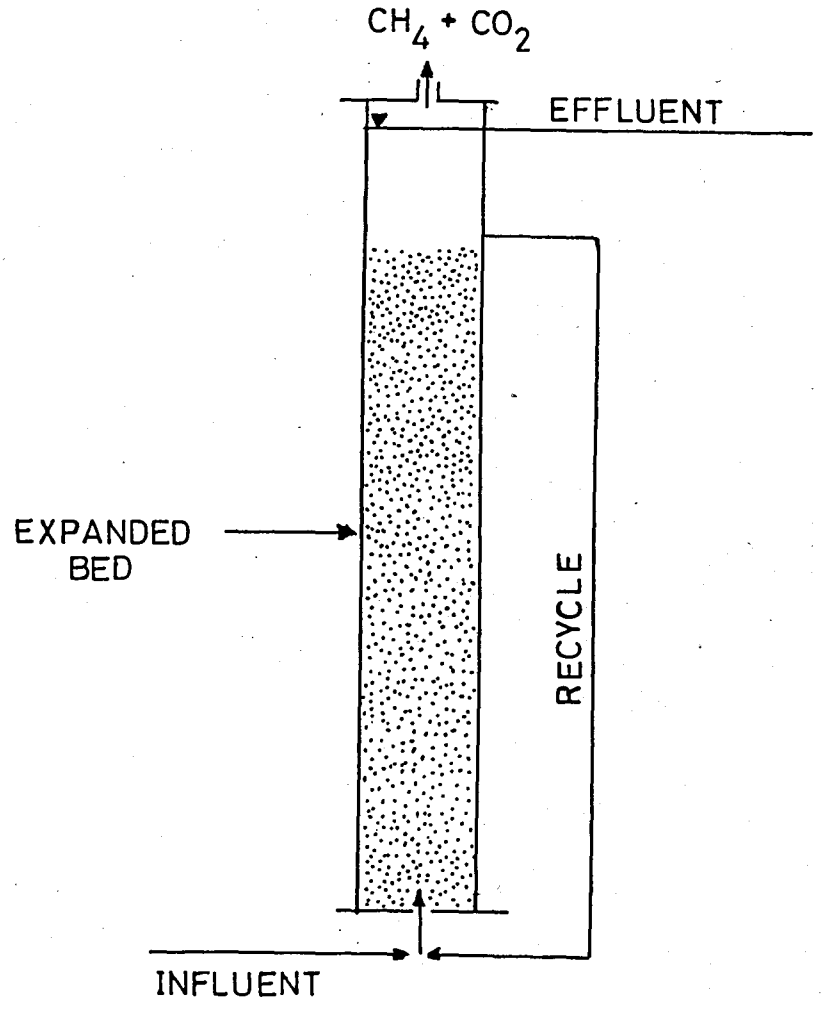


Figure:7. Schematic representation of the expanded bed reactor

refers to a large increase in reactor volume (up to 100 percent),wheras in the expanded beds only 10 to 20 per cent of the static (non-flowing) volume is obtained.

Atkinson, et.al. (1974,1979) have reported on a number of different types of microbial film fermentors,mathematical models, and applications to industrial fermantations. The first application of the expanded bed process to organic carbon removal from wastewater was initiated by Jewell and Cummings in 1972 and reported in 1974 using an aerobic system

with pure oxygen. Jeris et.al. (1974, 1975, 1977) used the fluidized bed concept with sand as the media to obtain high rates of denitrification and BOD removal. The denitrifying sewage effluent unit was columnar in shape and both activated carbon and sand (Jeris et.al.1974 1975)were used independently as supporting media for the microorganisms. The hydraulic retention times within the reactor varied from 3 to 10 minutes. Investigations on a pilot scale unit where a HRT of 6.5 minutes was employed resulted in 99 per cent of the nitrate and nitrate nitrogen removal (Jeris and Owens,1975). The addition of methanol in stoichiometric amounts as a carbon source also contributed to this removal efficiency. High removal efficiency and short HRT's in the fluidized bed denitrification process were mostly obtained due to the large masses of microorganisms (20000 to 40000mg/L),which could be concentrated within the fluidized bed and to the use of methanol at proper amounts (Jeris and Owens,1975). Jeris et.al.(1977) used the fluidized bed technology for carbonaceous BOD and nitrogen removal in municipal wastewater treatment,by using aerobic, anaerobic and denitrification systems connected in series. The main results of this study are: a) Fluidized bed technology combines best features of activated sludge and trickling filtration into one process, b) Fluidized bed systems require less than 5 per cent of the reactor space required for conventional treatment, c) In fluidized bed systems 93 per cent BOD removal in 16 minutes, 99 per cent NO_3-N removal in 11 minutes,and 99 per cent NO_2-N removal in 6.5 minutes were obtained. d) Intermediate clarification between processess is not necessary. e)Fluidized bed technology should be significantly less expensive than conventional treatment technology, based on the savings in area and rapid time for treatment " (Jeris et.al.1977).

Leuschner et.al.(1976) have demonstrated that an expanded bed type of process using an anaerobic attached film was capable of treating dilute wastes at relatively short retention times and at ambient temperatures. In his study he used a similar attached film expanded bed unit as Jewell (1974). The substrates used were a) a synthetic waste with a strength similar to domestic wastes, and b) primary effluent.The COD removal efficiencies varied from 50 to 90 per cent with hydraulic retention times ranging from 3 to 12 hours when synthetic substrate was loaded. Leuschner and Jewell (1978) used attached film expanded bed processes for the anaerobic fermentation of dairy cow manure diluted to 1 to 2 per cent total solids. The anaerobic attached film expanded bed reactor was successfully operated at five different conditions with HRT as low as 0.15 days. The total volatile solids destruction efficiencies of 39.5 per cent were achieved at an HRT of 1.2 days,using a 2 per cent total solids feedstock.where as equal efficiencies occured at an HRT of 10 days with the same substrate in a conventional anaerobic fermentor.

Leuscher et.al.(1976) figured the rate of total volatile solids destruction with different organic loadings. Total volatile solids destruction increased with increase in organic loading in the reactor. Figure 8 demonstrates the rates of solids destruction resulting from varying organic loading rates for the anaerobic attached film fermentor operated at 35 C .

Switzenbaum (1980) and Jewell et.al.(1981) studied the anaerobic attached film expanded bed process for municipal wastewater treatment by minimizing energy input while producing energy, and minimizing excess biological solids production. From these studies it is concluded that the Anaerobic Attached Film Expanded Bed can be operated at reten-

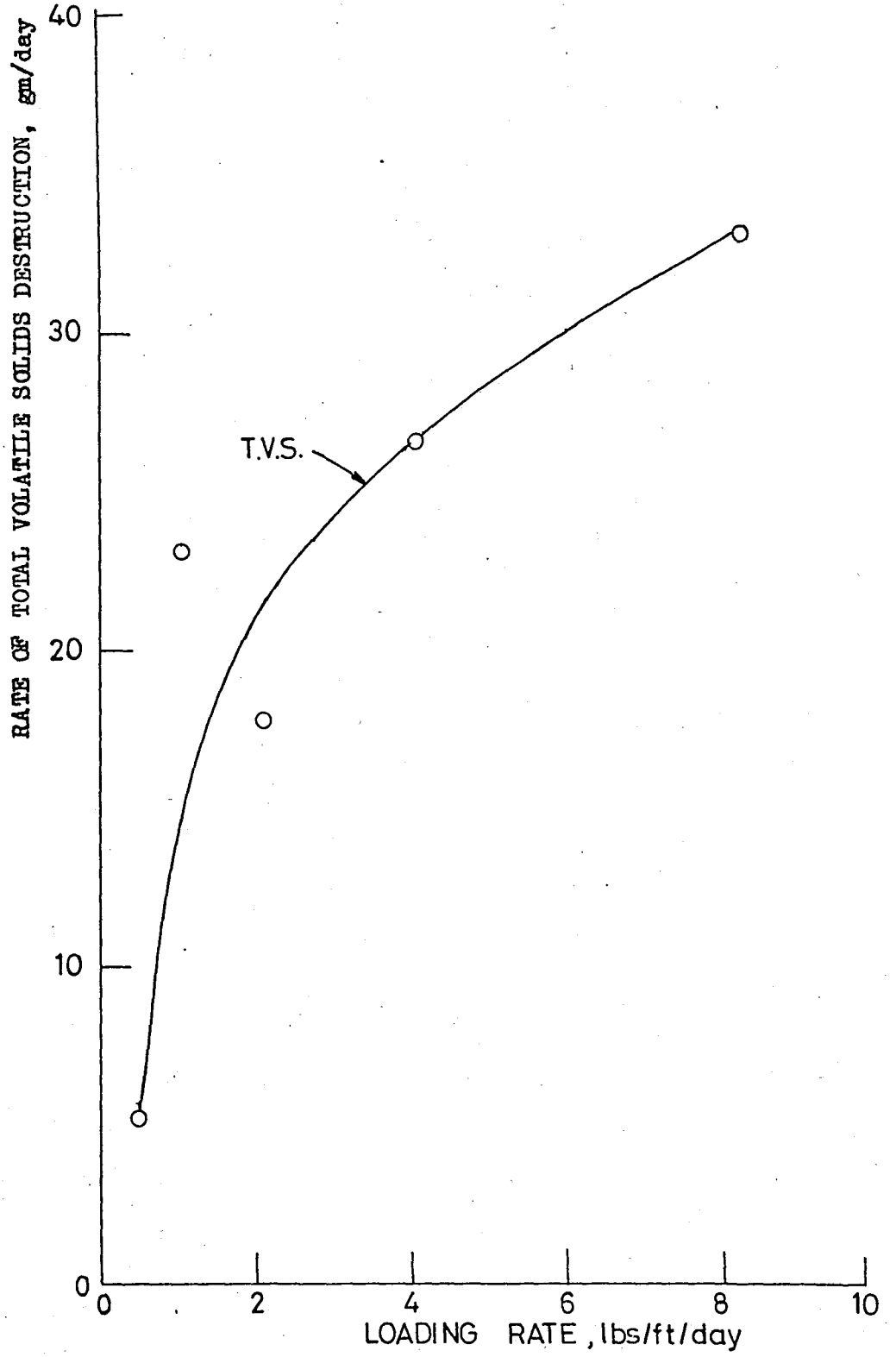


Figure:8.Rates of solids destruction resulting from varying organic loading for the anaerobic attached film fermentor operated at 35 C with and 2 per cent Total Solids Hydraulic Retention Time varying from 0.15 - 12 days. (Frostall,1981).

tion periods two to three orders of magnitude shorter than most other anaerobic processes. Also it became possible to treat dilute organic wastewaters at low temperatures and offered energy production and minimal biological sludge production.

The literature clearly shows that the anaerobic attached film expanded bed and fluidized bed systems are highly promising and advantagous processes. It is for this reason that the Anaerobic Attached Film Expanded Bed Reactor was studied in this thesis.

CHAPTER 3

EXPERIMENTAL SET-UP and MATERIALS

3.1 Selection of the Experimental System

The main purpose of the present study was to determine the performance of an anaerobic attached film exparded bed reactor and to compare the results with those obtained using an anaerobic control reactor. To achieve this,an experimental system consisting of two anaerobic attached film expanded bed reactors and an anaerobic sludge bed reactor was prepared and investigated.

3.2 Experimental set-up

The anaerobic treatment system used in this study consisted of the following parts:

a) Reactor

b) Heater

c) Pump

d) Liquid/gas seperation chamber

e) Substrate container

f) Gas collector

The experimental equipment is depicted in Figure 9 in detail and the general pictures are given in Figures 10,11 and 12. Details of the different units are given in sections which follow.

3.2.1 Reactors:

The reactor columns used in this study were made of standard hollow glass columns,each 1 m. in length with 4.1 cm.internal and 4.8 cm external diameters. The internal volume of the reactors were 1.6 Liters each as to support the filter media,and to ensure a uniform distribution of the wastewater within the reactor. A circular plexiglass plate having a diameter of 4.0 cm. and a thickness of 0.6 cm. was located at a dis-

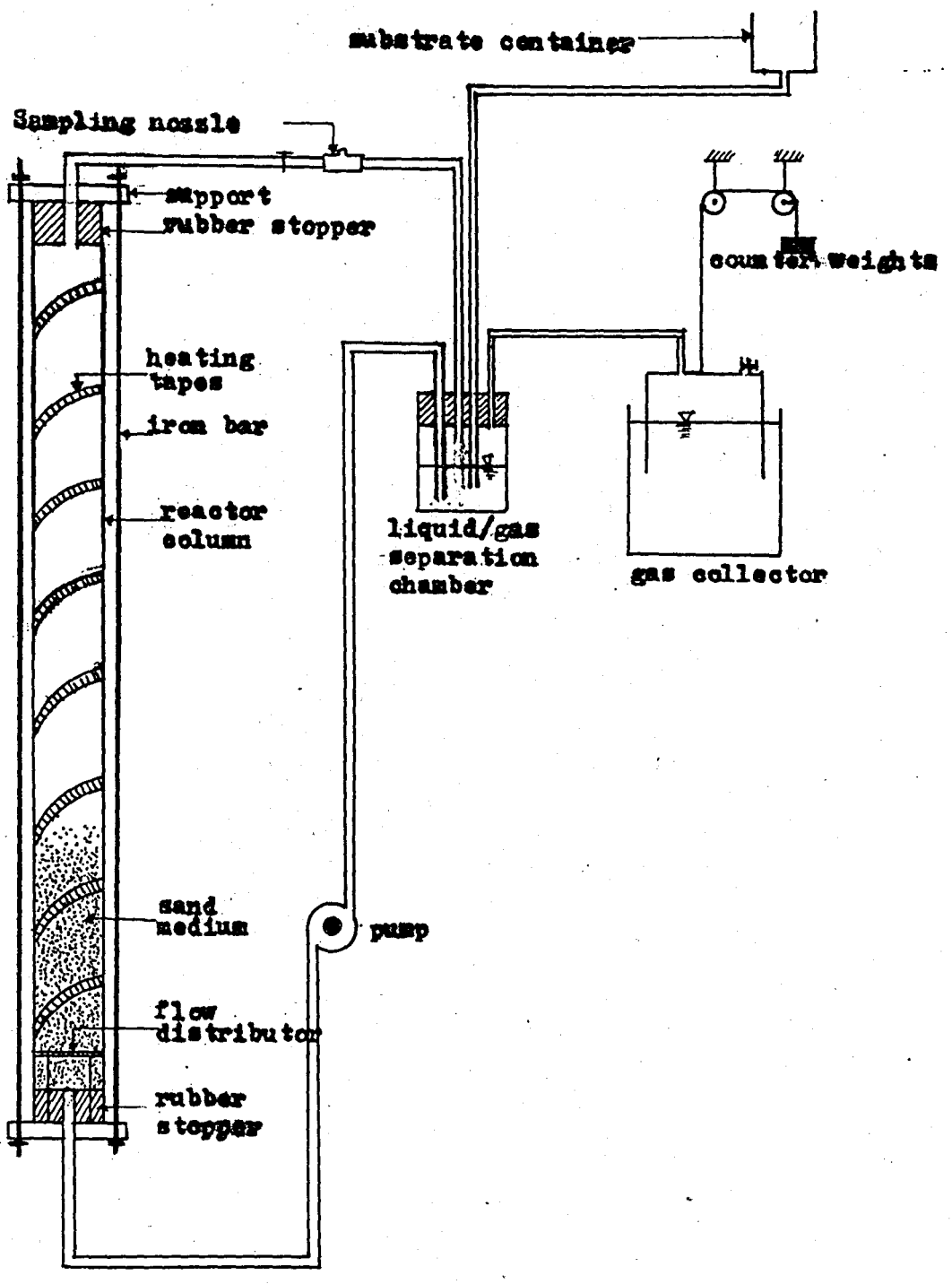


Figure 9. General view of the experimental set-up (The system consists of three similar set-ups)

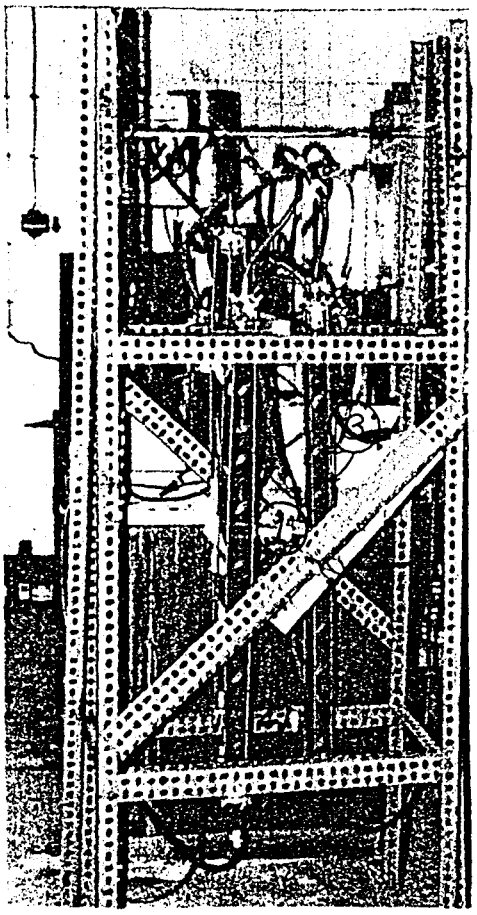


Figure 10. The experimental set-up

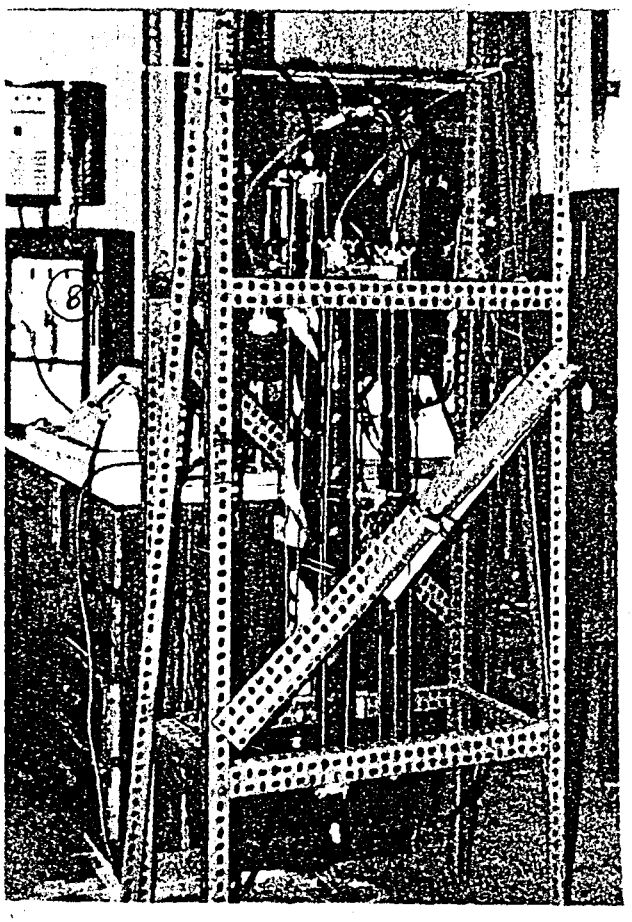


Figure 11. The experimental set-up

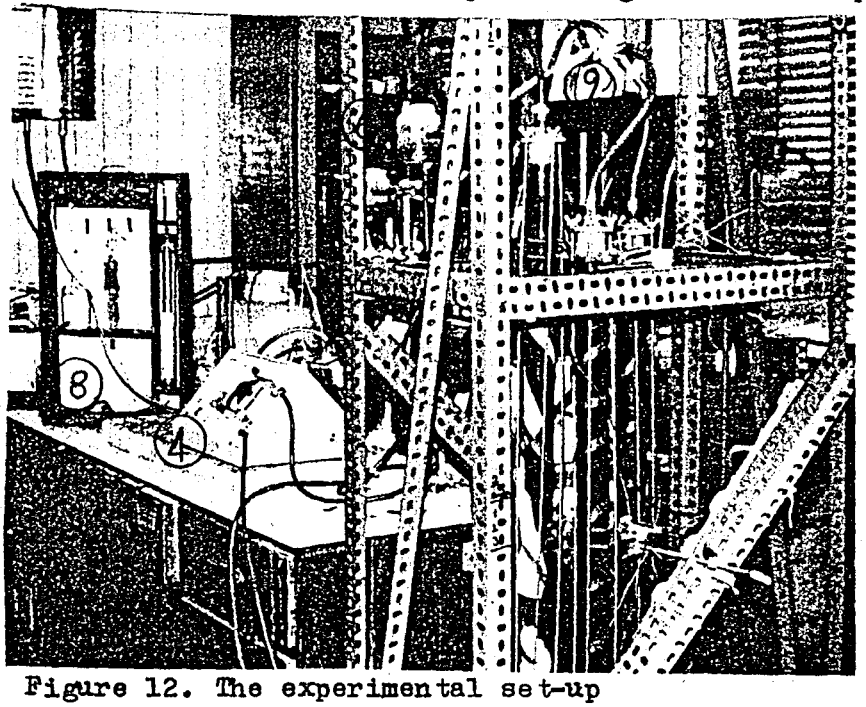


Figure 12. The experimental set-up

1)Columns, 2)Filter Media, 3)Heating Tapes, 4)Recycle Pump, 5)Liquid/ Gas seperation chamber, 6)Substrate Container, 7)Gas Collector, 8)Orsat Analyzer, 9)Sampling Nozzle

tance of approximately 4.0 cm from the base of the column (Fig.13).31 small holes of 0.3 cm diameter were opened in this plate. Sand was placed as the filter media in two of the three reactor columns used in this study while, the third one used as a control reactor was empty.

The sand used, at the beginning of the study had a size distribution specified as maximum sand size of number 40 ASTM Standard Sieving (sieve diameter of 0.420 mm) and minimum of number 70 ASTM Standard Sieving (Sieve diameter of 0.210 mm). The particle density of the support medium was determined to be 1.39 g/cm , specific gravity 2.64 and porosity 0.47 in the unexpanded form of the expanded bed reactor. The support medium was estimated to displace a liquid volume of 110 mL in both of the reactors.

The inlet and outlet of each reactor consisted of 0.8 cm diameter glass tubes inserted through a rubber stopper, clogging hermitically the end of the column. Two metallic plates located infront of the inlet and outlet rubber stoppers were held together by four Ø 0.6 cm steel bars. This plate-bare system gave the necessary strength to prevent opening of the covers by the internal pressure created within the reactor(Fig.14). The outlet of the column was connected by a 0.8 cm rubber tubing to the liquid/gas seperation chamber while the inlet was connected to the recycle pump (Fig.13).

3.2.2 Heaters:

Temperature variations have an important effect on the efficiency of wastewater treatment. This effect is more noticable in anaerobic processes. Therefore; it was necessary to maintain a constant temperature in all of the reactors. The necessary heat was provided by heating tapes[1] twisted around the reactor columns (Figure 9).

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[1] Briskeat flexible heating tape MGF

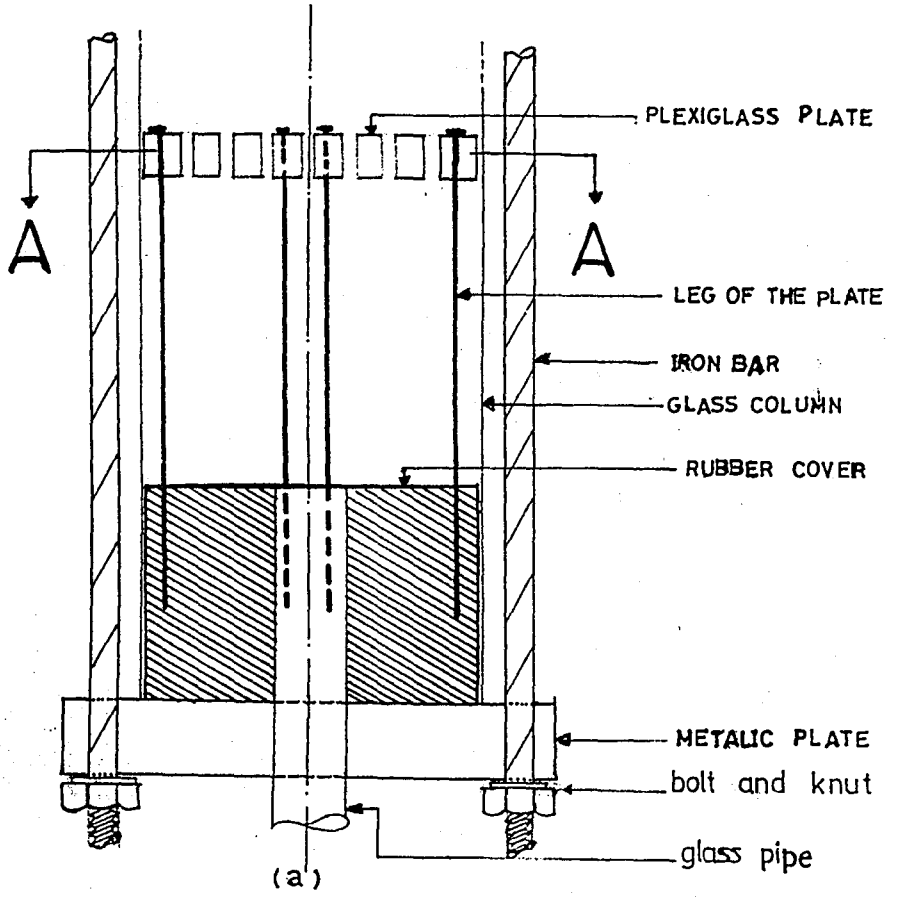
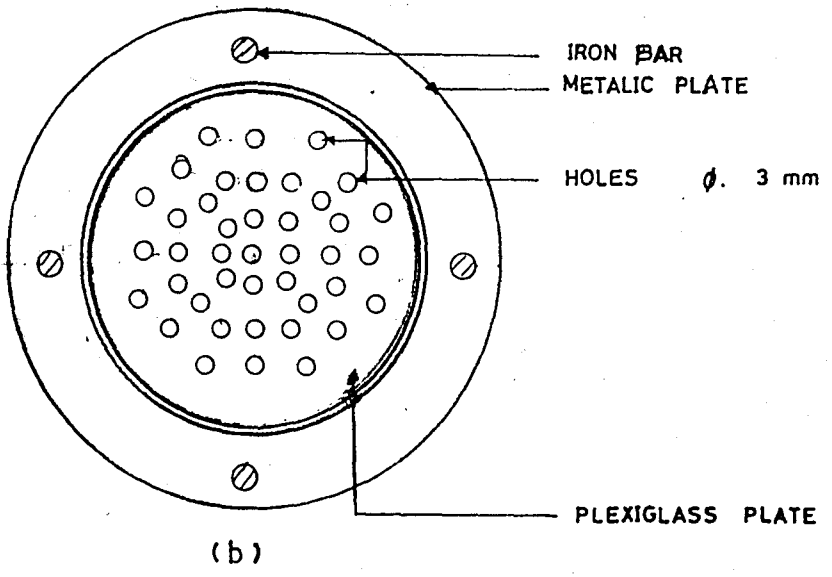


Figure:13. Details of flow distributor a)Cross section b)Top view



3.2.3 Recycle Pump:

The circulation of wastewater within the system was achieved with a peristaltic pump[1] which pumped the substrate from the liquid/gas separation unit to the inlet of the reactor (Figure 15).

3.2.4 Liquid/gas separation chamber:

A 200 mL glass cylindrical jar, closed hermitically with a rubber stopper so that no air could enter, was used as the liquid/gas separation chamber. The rubber stopper contained four glass tube connections. One, acting as the gas outlet tube,had an internal diameter of 0.2 cm and terminated at the inner surface of the rubber stopper. The 0.8cm internal Diameter feeding tube, which was immersed in the substrate, terminated 6cm below the bottom of the rubber stopper. The inlet tube lying 1 cm and the outlet tube 7 cm below the bottom of the rubber stopper were both of 0.6 cm Internal Diameter. Details of the liquid/gas separation chamber are given in Figures 16 and 17.

3.2.5 Substrate Container:

A 250 mL Plastic jar acted as the substrate feeding container. This container was clamped at a point higher than the liquid/gas separation unit. The tubes connecting the feed container to the liquid/gas separation chamber were kept full with wastewater during the whole experimentation period (Figure 18).

3.2.6 Gas Collector:

In order to collect the gases produced during the anaerobic decomposition, a gas collector, consisting of two cylindrical containers with the smaller container inserted upside down into the larger,was used. The generated gas was collected by the system as shown in Figure 19(a).

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1 Monostat Solid State Model-Peristaltic Pump

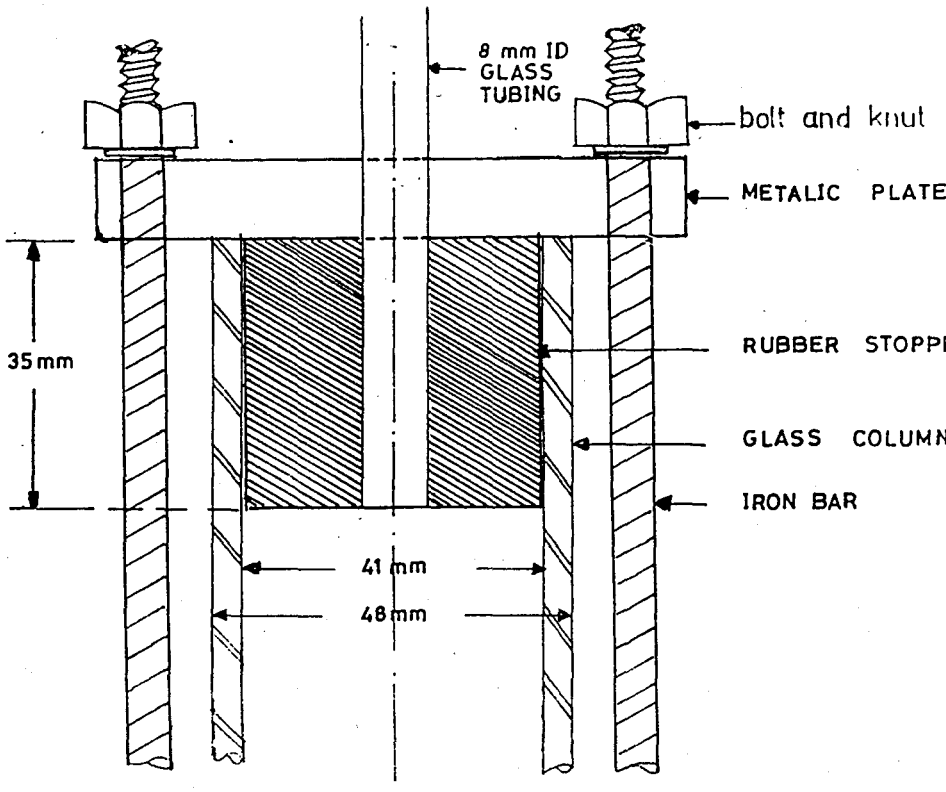


Figure:I4. The details of an outlet rubber cover of reactor

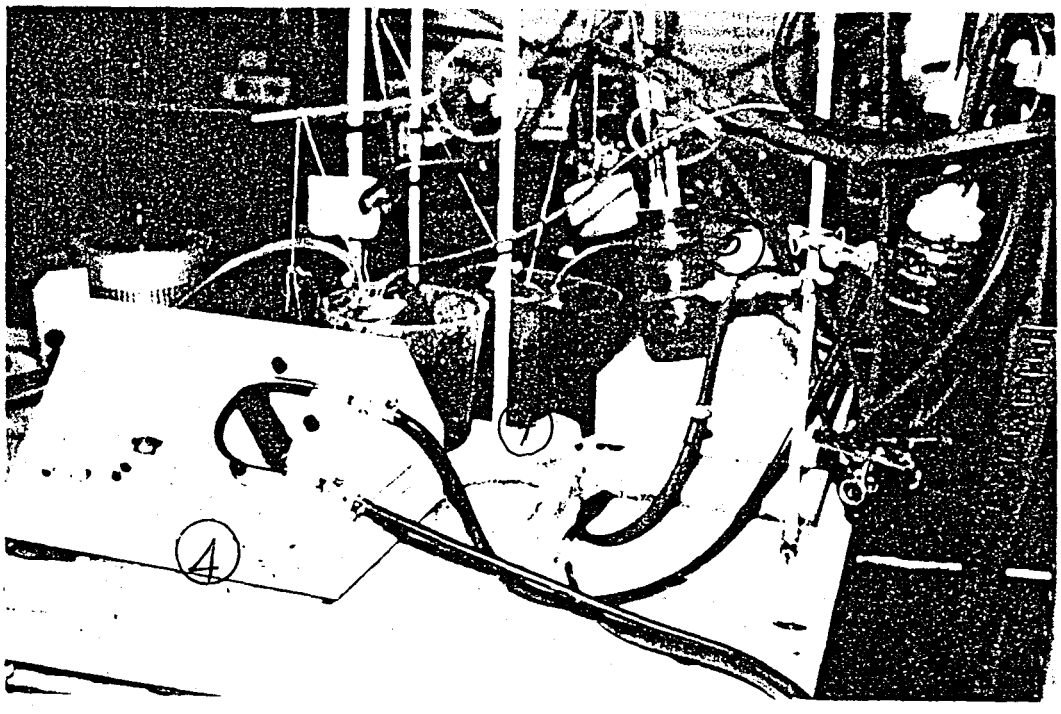


Figure 15.Recycle Pump(4), Liquid gas seperation chamber(5) and gas collectors(7).

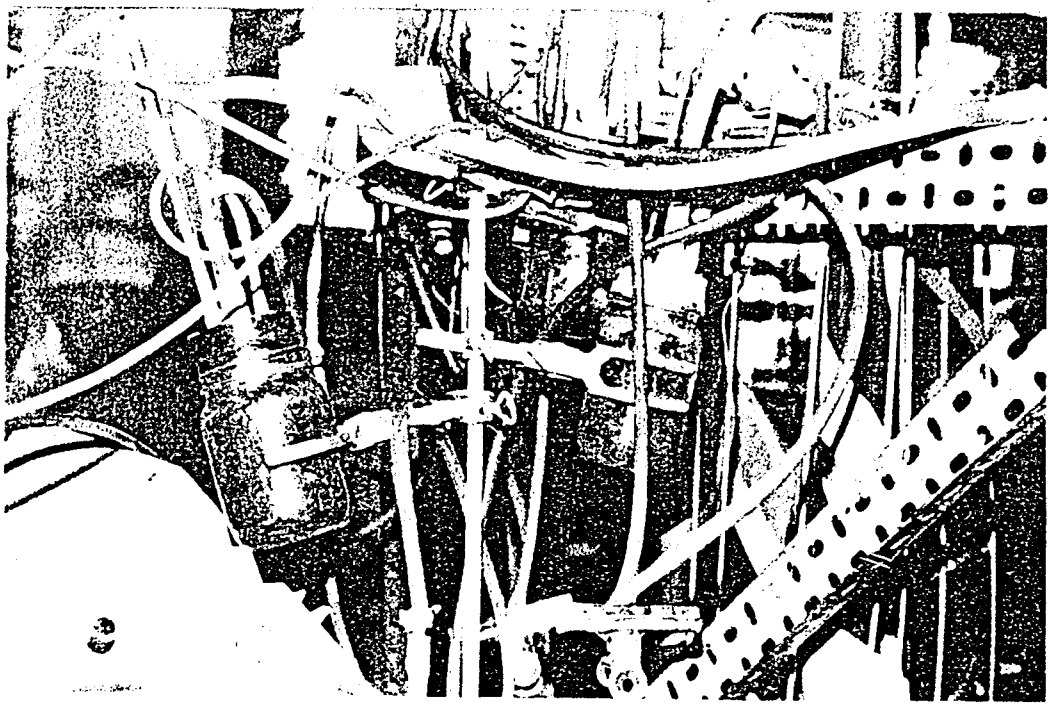


Figure 16. The liquid/gas seperation chamber with inlet and outlet tubings.

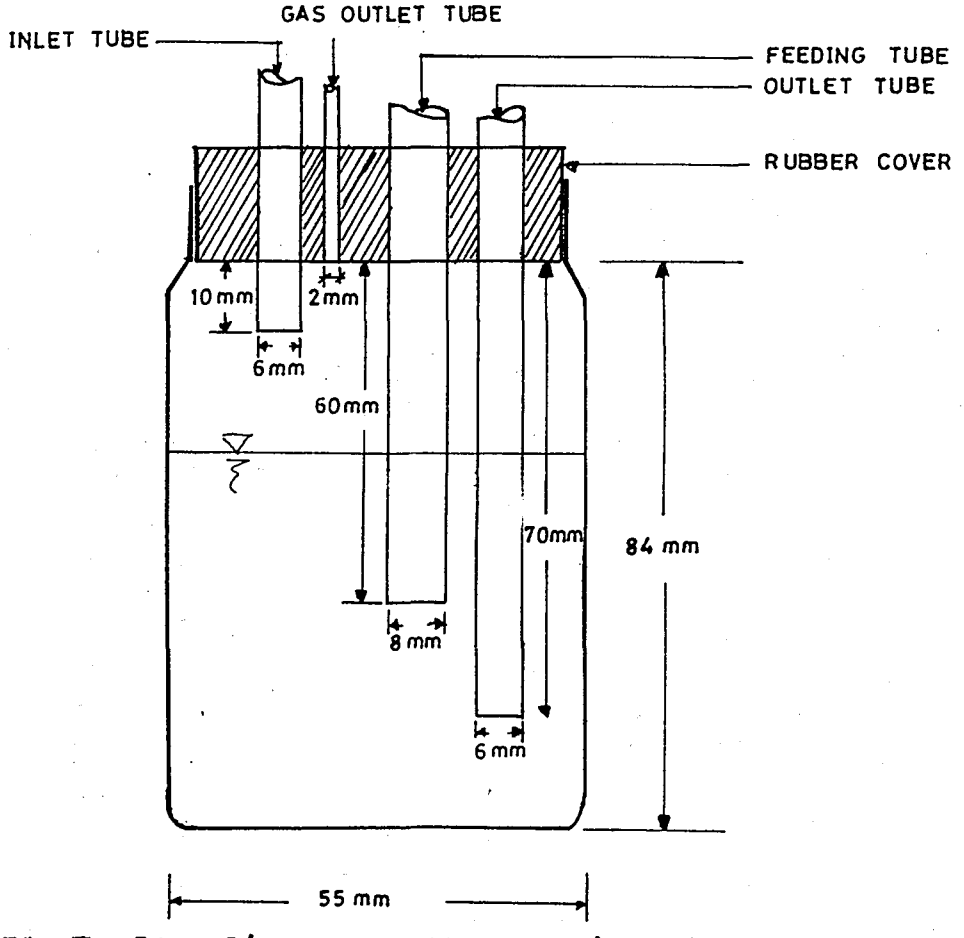


Figure: 17. The Liquid/gas separation chamber

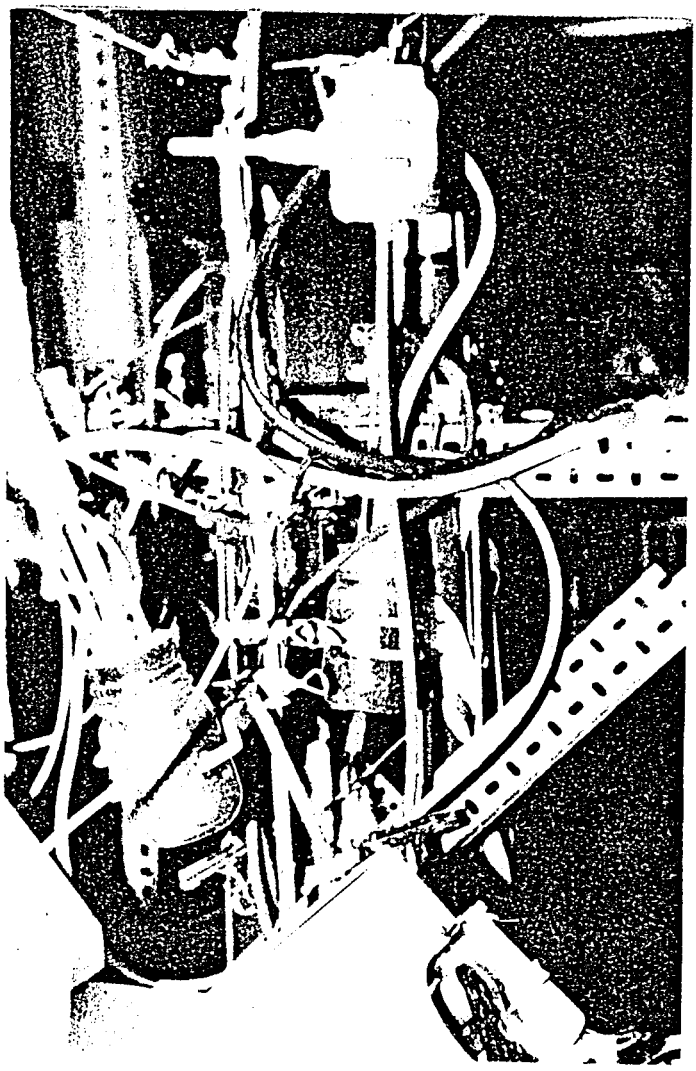
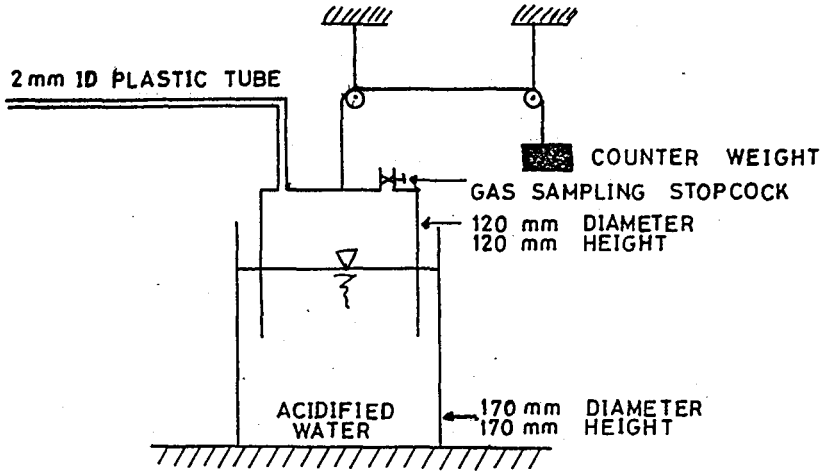
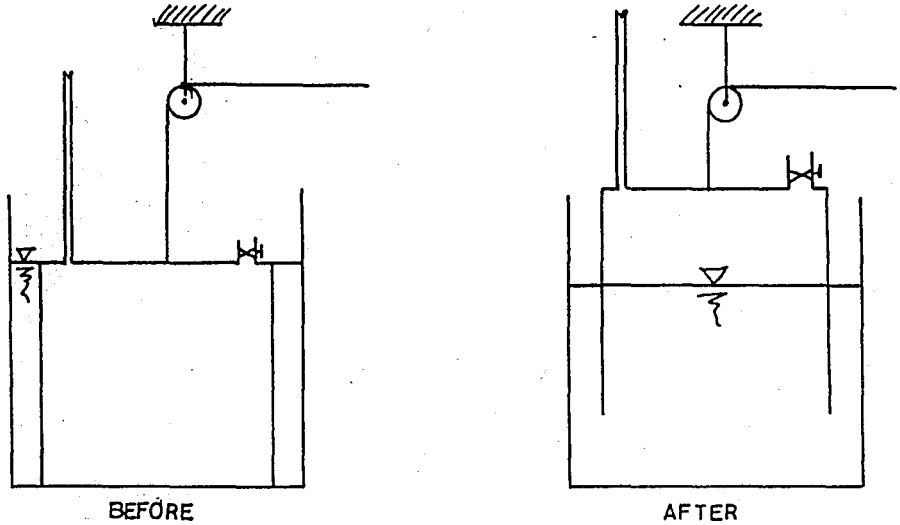


Figure 18. Substrate container and liquid/ gas seperation chamber.

The diameter of the outer cylindrical container was 17 cm while the inner was 12 cm. The outer container was filled with acidified water to prevent the adsorption of carbondioxide. Methane has no adsorption problems by virtue of its insolubility (Geisser and Pfeffer, 1977). The gas which was generated in the reactor unit was carried by a 2 mm ID plastic tube to the inner cylinder. The generated gas caused the inner cylinder to ascend,as shown in Figure 19(b). Connecting the



(a)



(b)

Figure: I9. Details of gas collector.

inner cylinder with a string to a counter-weight,with a weight equal to the immersed weight of the inner cylinder,kept the collected gas almost at zero gauge pressure. The volume of generated gas was measured directly by determining the height of rise of the inner cylinder. The gas collector was equipped with an outlet to allow for sampling and subsequent gas composition.

The details of the gas collectors are shown in Figures 20 and 21 .

3.3 Substrate:

Two different substrates were used in the present study. The base of the substrates was made of dung to which water was added so that the total solid content fell in the range of 4000-10000 mg/L. To this, the chemicals listed in Table 8 were added in order to provide the necessary nutrients for the decomposition,composing the substrates of reactor number one and the control reactor.

Table 8. Concentrations of Substrate Components in 1 liter of Diluted Dung Waste.

Component	Concentrations
CH_3OH	50 mL.
NH_4cl-N	3 mg
KH_2PO_4 -P	2 mg
Na_2HPO_4 -$7H_2O$-P	2 mg

The second substrate was similar to the first, with the only difference being an addition of 5 g/L bear yeast. Both substrates were stored in 2 liter tightly closed glass bottles and kept in an incubator at 4 °C.

3.4 Sampling Nozzle :

A sampling nozzle was placed on the tube carrying the substrate from the reactor column to the liquid/gas separation chamber (Fig 22).

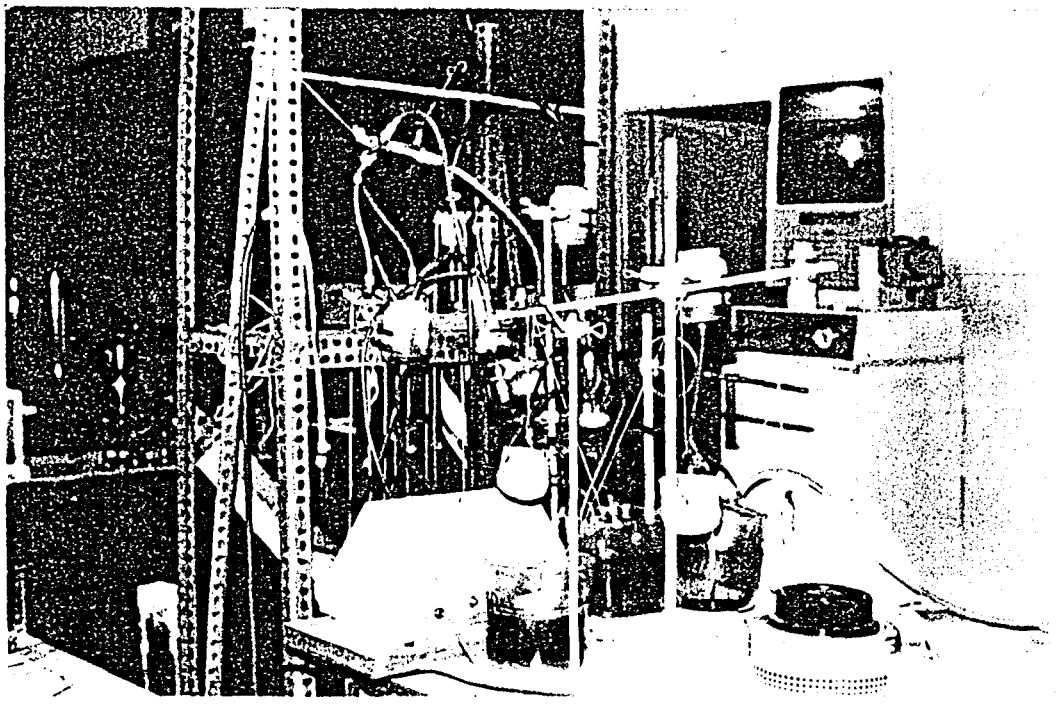


Figure 20. General view of gas collectors

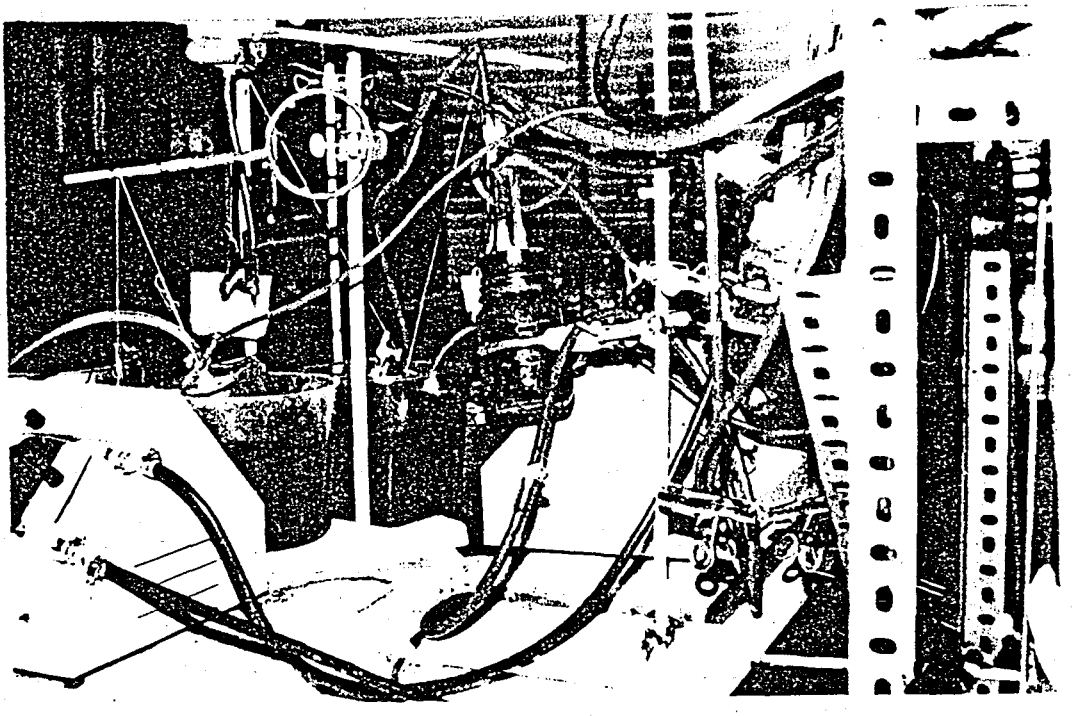


Figure 21. General view of gas collectors

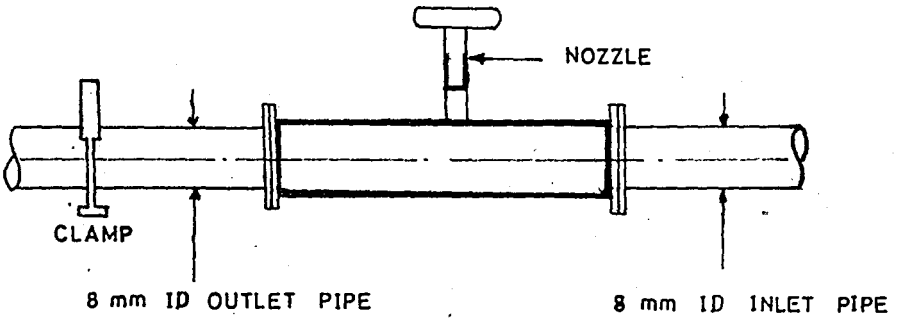


Figure 22. Sampling Nozzle

CHAPTER 4

EXPERIMENTAL PROCEDURE

The procedure followed in this study can be summarized in four groups:

a) Preliminary work before starting the experiments,

b) Procedure followed during the start-up period,

c) Procedure followed during steady-state in the semi-continuously fed reactors,

d) Procedure followed in the batch reactors.

4.1 Preliminary Work:

The preliminary preparation before starting the experiments can be summarized as follows:

a) After several trials a sand having effective size of 315 μm was chosen to be used in this study. This selection is based on the fact that it was not possible to expand larger particles with the given flowrate.

b) The reactors were filled with the filter media. The necessary amount of sand was determined by trial and error until the amount providing the necessary expansion under a given flowrate was obtained,

c) Complete air tightening of the reactors was provided by insulating the necessary points,

d) The system was controlled so that no seepage of substrate out of the system was possible,

e) The temperature of the system was adjusted to the desired value.

f) The flow rate of all the reactors was adjusted to 0.0028 L/sec and maintained at this value during the study period. The area of the columns was $1.225 \times 10^{-3} m^2$ and consequently the flux was 137 $L/min\text{-}m^2$. The initial media height was 17 cm and became 20.5 cm (with 20 per

cent expansion) after the expansion.

4.2 Procedure Followed During the Start-up Period:

a) The two expanded bed reactors and the control reactor were seeded with bovine rumen fluid and dung waste. Then, one of the expanded bed reactors was fed with substrate I (see section 3.3), the second with substrate II, and the control reactor with substrate 1. It was allowed 5 months of operation at 35 ± 1 °C for necessary biomass accumulation and during this period monitoring was performed and experimental techniques were evaluated.

b) Flowrate of substrate was adjusted to obtain the desired hydraulic detention time .

c) Samples were collected at different time intervals and tests (COD, pH, Solids, Nitrogen and Volatile Acids) were performed inorder to detect the initiation of steady-state.

4.3 Procedure followed during the steady-state in semi-continuously fed-reactors

a) Samples were collected and substrate was added periodically to the reactors as explained below.

First, the substrate container was filled with 250 mL substrate by syphoning. At the time of sampling, the clamp between the substrate container and the liquid/gas separation chamber was opened and the tube between the sampling nozzle and the liquid/gas separation chamber was clamped. The reason of clamping the tube between the sampling nozzle and liquid/gas separation chamber was to prevent the entrance of air from the sampling nozzle and to push the effluent out of the sampling nozzle.

During sampling the influent substrate was filled in to the liquid/gas separation chamber from the substrate container by gravity.

After sampling, the clamp between the nozzle and liquid/gas separation chamber was opened and the clamp between the substrate container and the liquid/gas separation chamber was closed. A picture taken during the sampling operation is given in Figure 23.

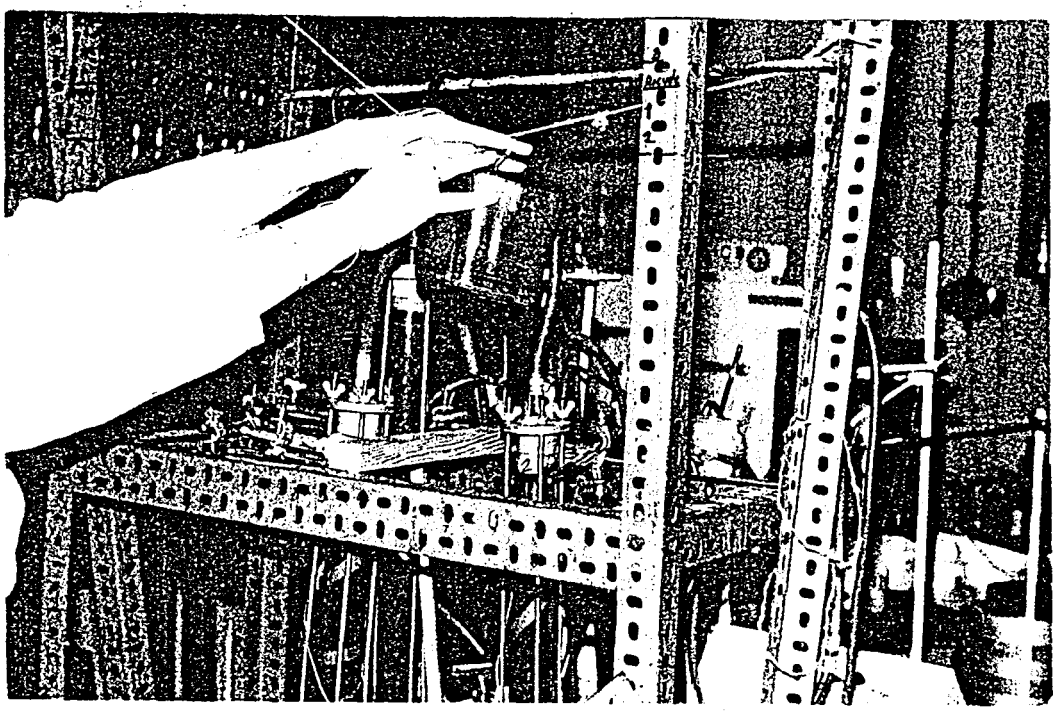


Figure 23. The sampling operation

b) The necessary tests explained in section 4.5 were performed.

c) Volume of gas produced and percentage of methane concentrations were determined.

4.4 Procedure followed in batch reactors:

In the batch treatment, the reactors were fed once and the effluent samples were taken at certain intervals without addition of new samples. The procedure given in sections 4.3-b and 4.3-c is the same for the batch treatment.

4.5 Methods of Analysis

Determination of the following parameters was done according to the procedures described in Standard Methods for the Examination of

Water and Wastewater (1981).

a) Total and Volatile Solids

b) Chemical Oxygen Demand (Sample was filtered through a "Whatman-Glass Microfibre Filters" and consequently the COD determination was performed).

c) Volatile Acids (Steam Distillation Method)

d) Solids (Total and Volatile)

e) Total Kjheldal Nitrogen

f) pH

g) Methane Analysis

The gas produced during the decomposition was mainly composed of Carbon Dioxide and Methane. The gas analyses were performed by an Orsat Analyzer.

The flowrate measurements were made with a one liter volumetric flask and a stopwatch. The necessary checks in flowrates were performed periodically (approximately once every 10 days)

CHAPTER 5

RESULTS AND DISCUSSIONS

The data and results obtained in this study can be separated into two classes.

a) Data obtained during semi-continuous feeding after reaching steady state conditions.

b) Data obtained during batch treatment.

5.1 Semi-Continuously Fed Reactors:

The data obtained for this part of the study is summarized in Table 1 in Appendix 1. The detailed discussion for some important parameters is given in the following sections.

5.1.1 Chemical Oxygen Demand

The variations in influent and effluent COD values with time of operation for all the reactors are given in Figure 24.Reactor 1 and reactor 2 give the data of the anaerobic attached film expanded bed reactors and Reactor 3, the data of the control reactor. Substrate I was used in Reactor 1 and III while Substrate II in reactor II.

During the initial period of the study there was no gas production in the reactors. This was an indication of inefficient removal of organic matter in the system. In reality COD measurements have shown that removal of organic matter was extremely low at the beginning. The low COD removal percentages may be due to the adaptation of microorganisms to new environmental conditions.

The COD reduction versus time and the COD reduction percent versus time are given in Figure 25. As can be observed in this figure, the general trend of COD removal percentage versus time for the expanded bed reactors (Reactor 1 and 2)are similar. COD removal efficiency

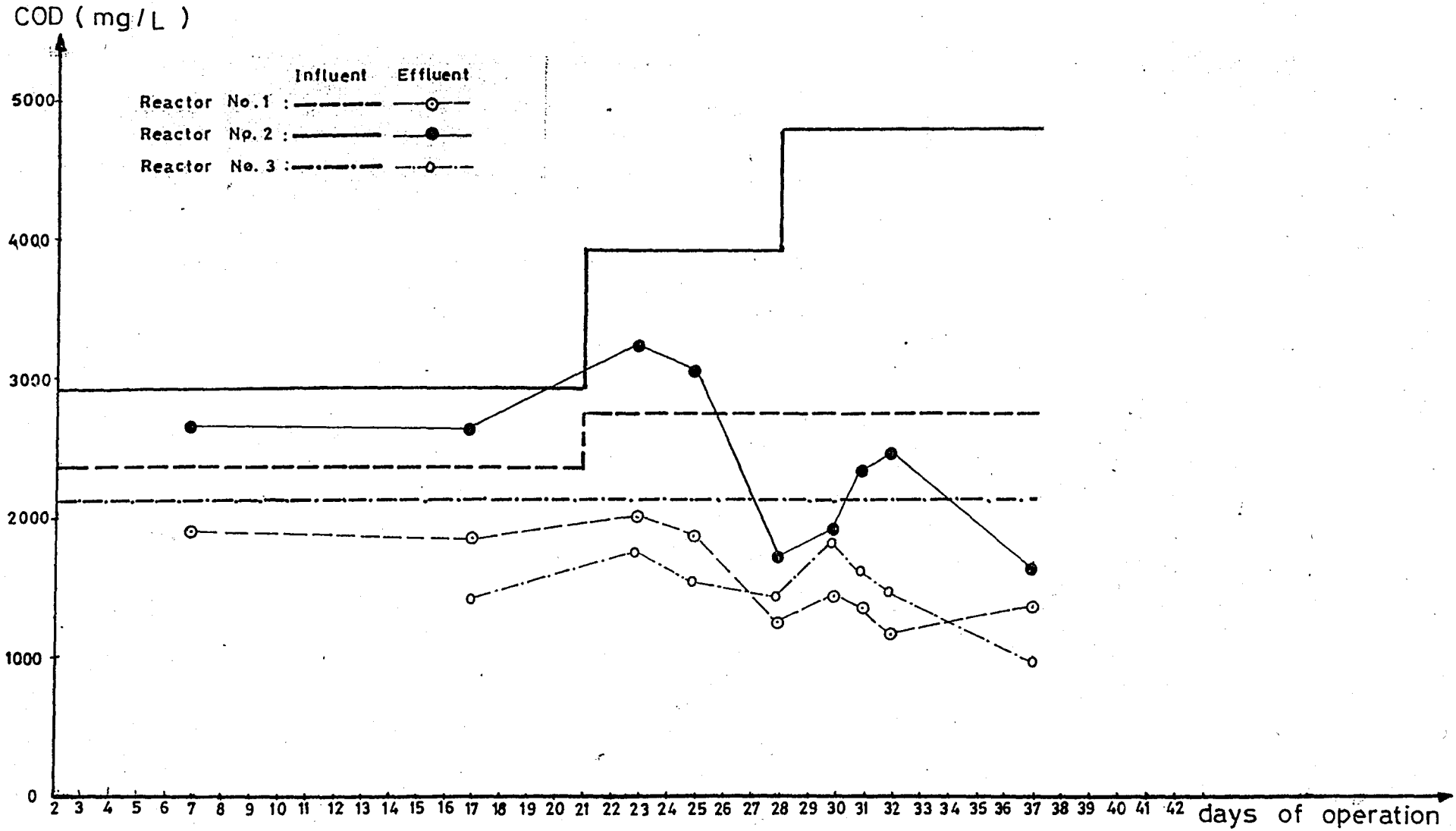
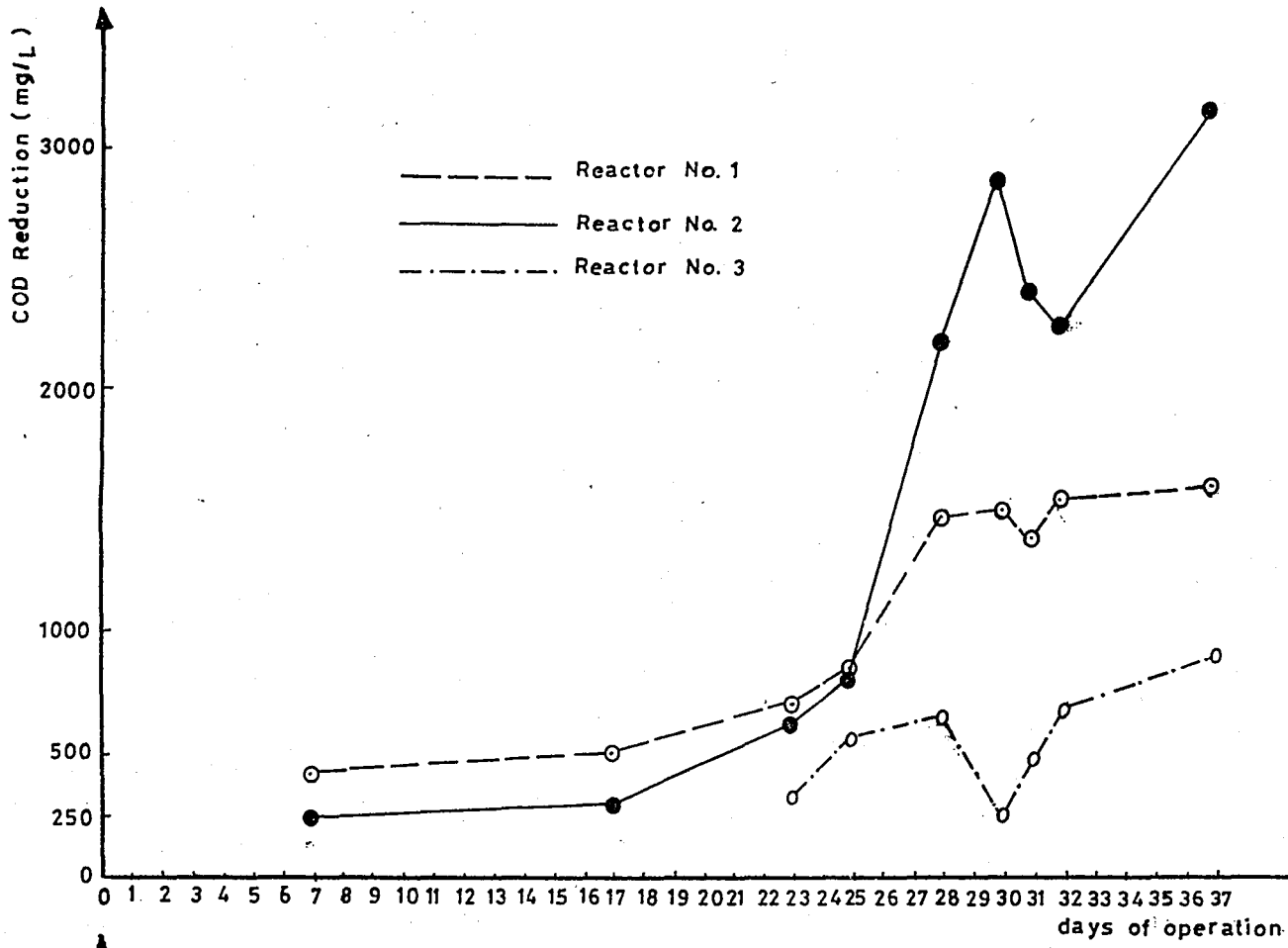


Figure:24. Variations of influent and effluent Chemical Oxygen Demand (COD) values



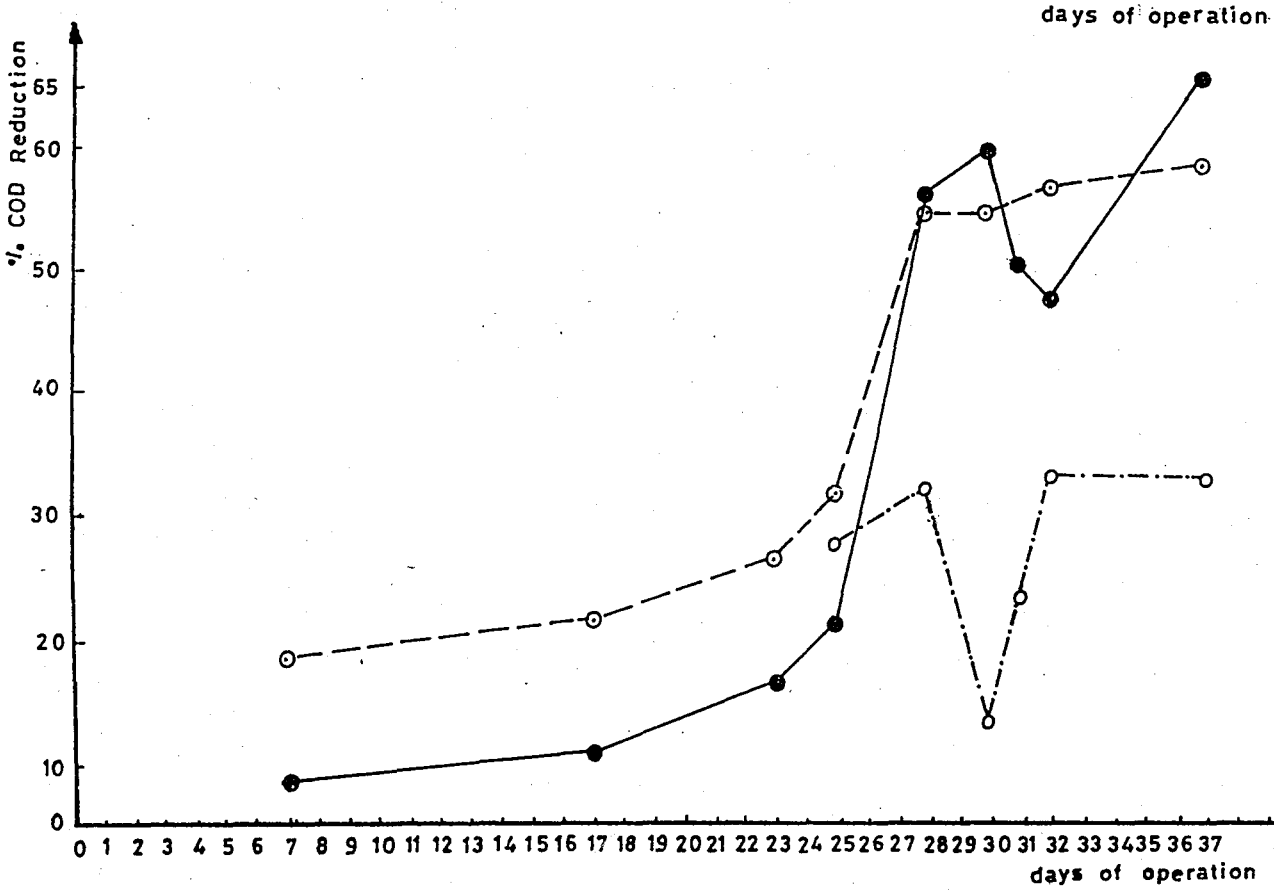


Figure:25. Variation of Chemical Oxygen Demand (COD) Reduction and COD percen reduction with time

increases with increasing influent substrate values. Although the COD removal percentages of Reactor 1 and 2 have similar values after the 28th day, the COD removal (mg/L) of Reactor 2 has a greater order than Reactor 1. This may be due to the fact that Substrate II which was fed in the 2nd reactor contained a certain amount of yeast.

The variation of COD removal rate in anaerobic attached bed reactors with organic load is given in Table 9 . and presented graphically in Figure 26. As can be seen in this figure, the COD removal rate increased with the increase in organic loading.

Table 9. Chemical Oxygen Demand (COD) Removal Rate at Different Organic Loads Applied at the Given Date.

Reactor No	Days of Operation	Organic Load Applied (kg COD/m day)	COD removal rate (kg COD/m day)
1	17	0.420	0.0907
1	28	0.490	0.2675
1	37	0.490	0.2880
1	42	0.580	0.3320
2	17	0.522	0.0566
2	28	0.699	0.3950
2	37	0.855	0.5660
2	42	1.050	0.7170

These results are in agreement with the findings of Frostel(1981). A comparison of the curves obtained in this study (Figure 26) with Frostel's curve (Figure 6) show many similarities.

The data obtained in relation to COD removal are given in Appendix I, Tables 1 and III.

5.1.2 Total Volatile Solids

The influent and effluent variations in Total Volatile Solids as a

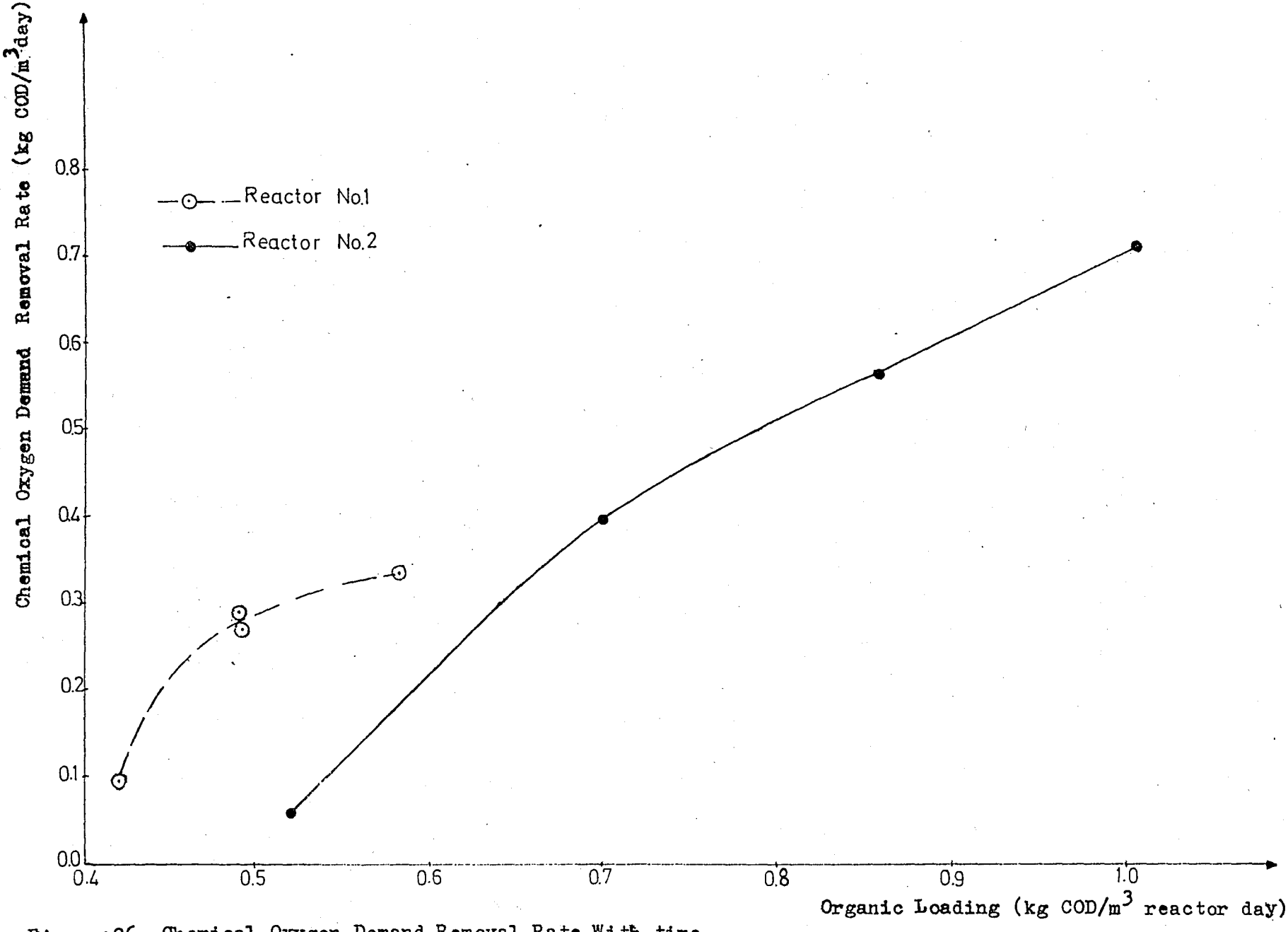


Figure:26. Chemical Oxygen Demand Removal Rate With time

function of time of operation in the Anaerobic Attached Film Expanded Bed Reactor are given in Figure 27. An increase in the influent substrate concentration resulted in an increase of the effluent substrate concentration at first; then a decrease was observed in effluent values. Similar results were observed in Figure 25.

The data of Table IV in Appendix 1 were derived by using the steady state data of the three reactors and are represented Figure 28. This Figure gives the Total Volatile Solids Reduction and the Total Volatile Solids Reduction Percentages at the given time. Although Total Volatile Solids reduction of reactor 2 is greater than that of reactor 1, a comparison of the percent reduction showed a similarity between reactor 1 and 2. The total volatile solid removal efficiency of the expanded bed reactors was considerably higher than the control. The percent decrease in volatile solids for Reactor 2 reached values as high as 48.7 per cent while for Reactor 1 49.2 per cent and for Reactor 3, 26.5 per cent. This is a clear indication that the attached biological film expanded bed has a positive effect on the volatile solids removal efficiency.

The rates of volatile solids destruction in Anaerobic Attached Film Expanded Bed Reactors at different organic loadings are given in Table 10 and Figure 29.

It is obvious from this figure that the volatile solids removal rate increased with the increase in organic loading. This results is in aggrement with the results obtained by Leuschner et.al.(1978).

A comparison of Figures 26 and 29 shows a similarity between the removal rates of COD and Volatile Solids with COD load applied. This expected result is an indication of the accuracy of the results obtained in this study.

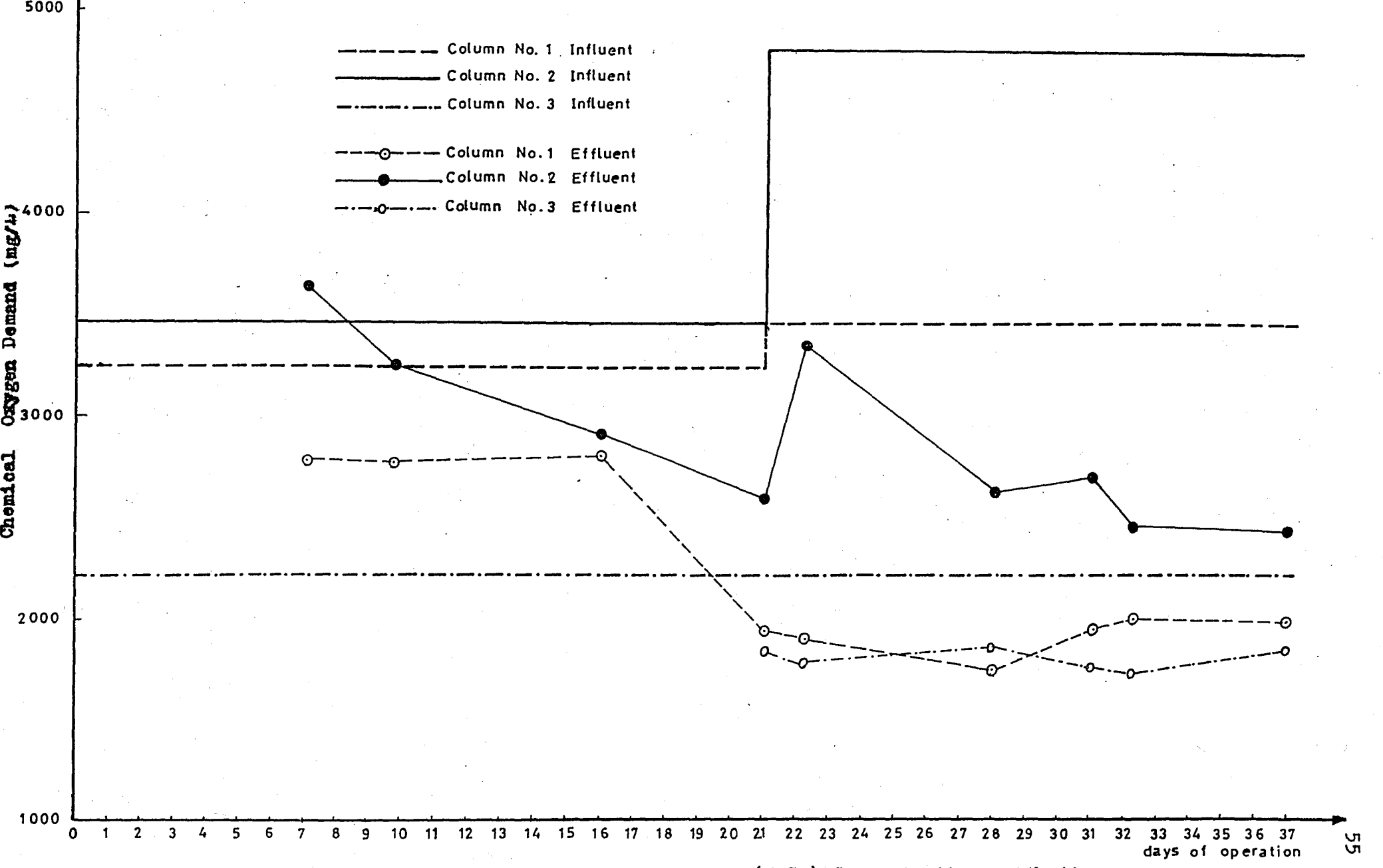


Figure:27. Variation of influent and effluent Volatile Solids (V.S.) Concentrations with time

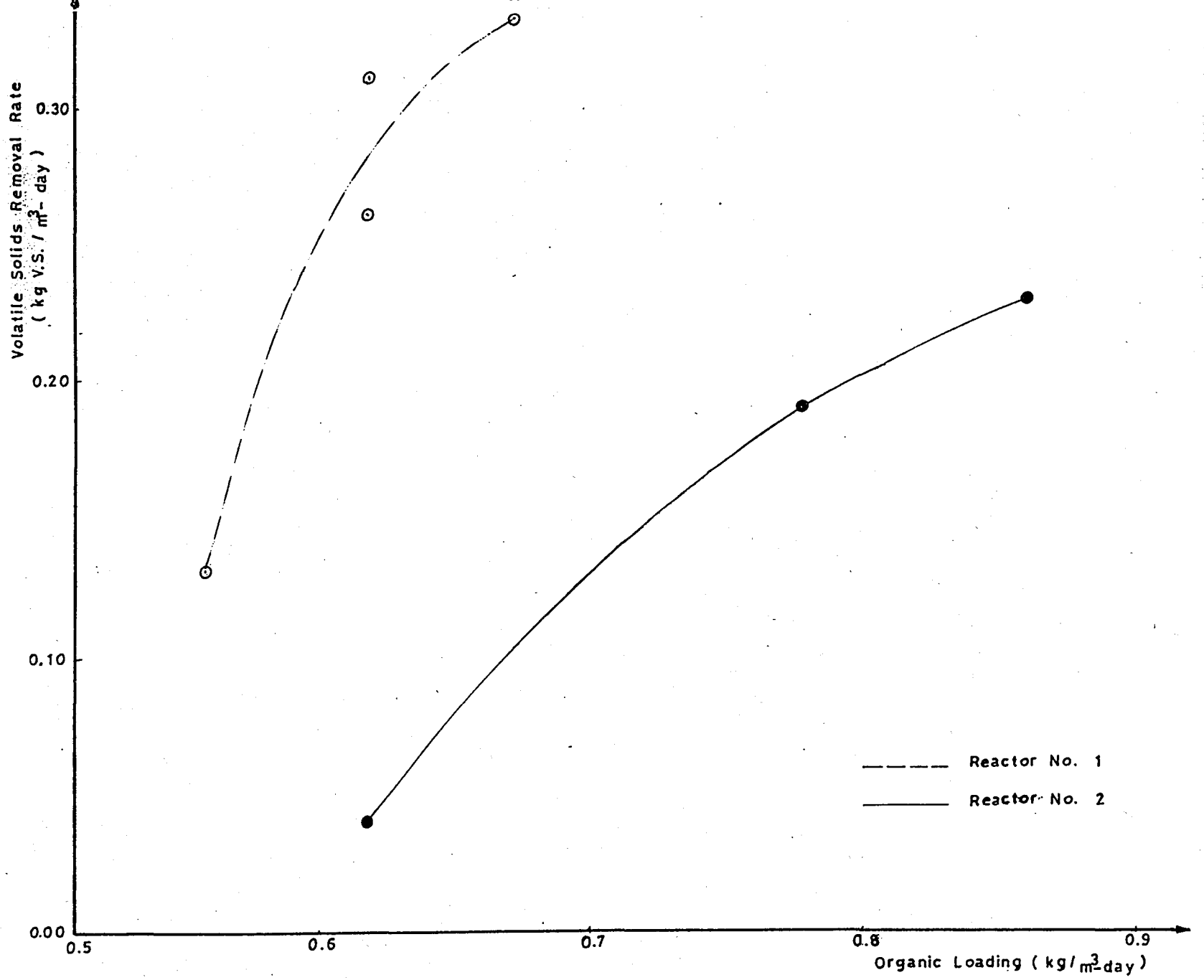


Figure:29: Rates of Volatile Solids destruction with Organic Loading.

Table 10. Volatile Solids Removal with Applied Organic Loads.

Reactor No	Number of days of operation	Organic load applied(kg V.S/m-day)	V.S. removal rate (kg V.S./m -day)
1	17	0.580	0.13
1	28	0.617	0.31
1	37	0.617	0.26
1	42	0.669	0.33
2	17	0.618	0.04
2	28	0.855	0.39
2	37	0.859	0.21
2	42	0.774	0.19

5.1.3 Gas Production

The cumulative amount of gas and methane produced is drawn in Figure 30. Gas Production in Reactor 1 was 0.052 L CH_4/kg. COD removed and 0.048 L CH_4 /kg. Volatile Solids removed. Gas Production in Reactor 2 was 0.081 L CH_4 /kg. COD removed and 0.099 L CH_4 /kg. Volatile Solids removed. These results are in aggrement with the results of Oleszkiewicz et.al.(1982) who obtained gas production rates of 0.024 to 0.172 L CH_4/ kg. COD removed and 0.03 to 0.222 L CH_4/kg.Volatile Solids removed. The values obtained in this experiment fall in this range.

5.1.4 pH

pH values measured during the study were in the range of 6.6 to 7.4 for Reactor 1, 6.7 to 7.4 for Reactor 2 and 6.6 to 7.8 for Reactor 3. These pH values are within the optimum range given in the literature for Anaerobic Treatment.

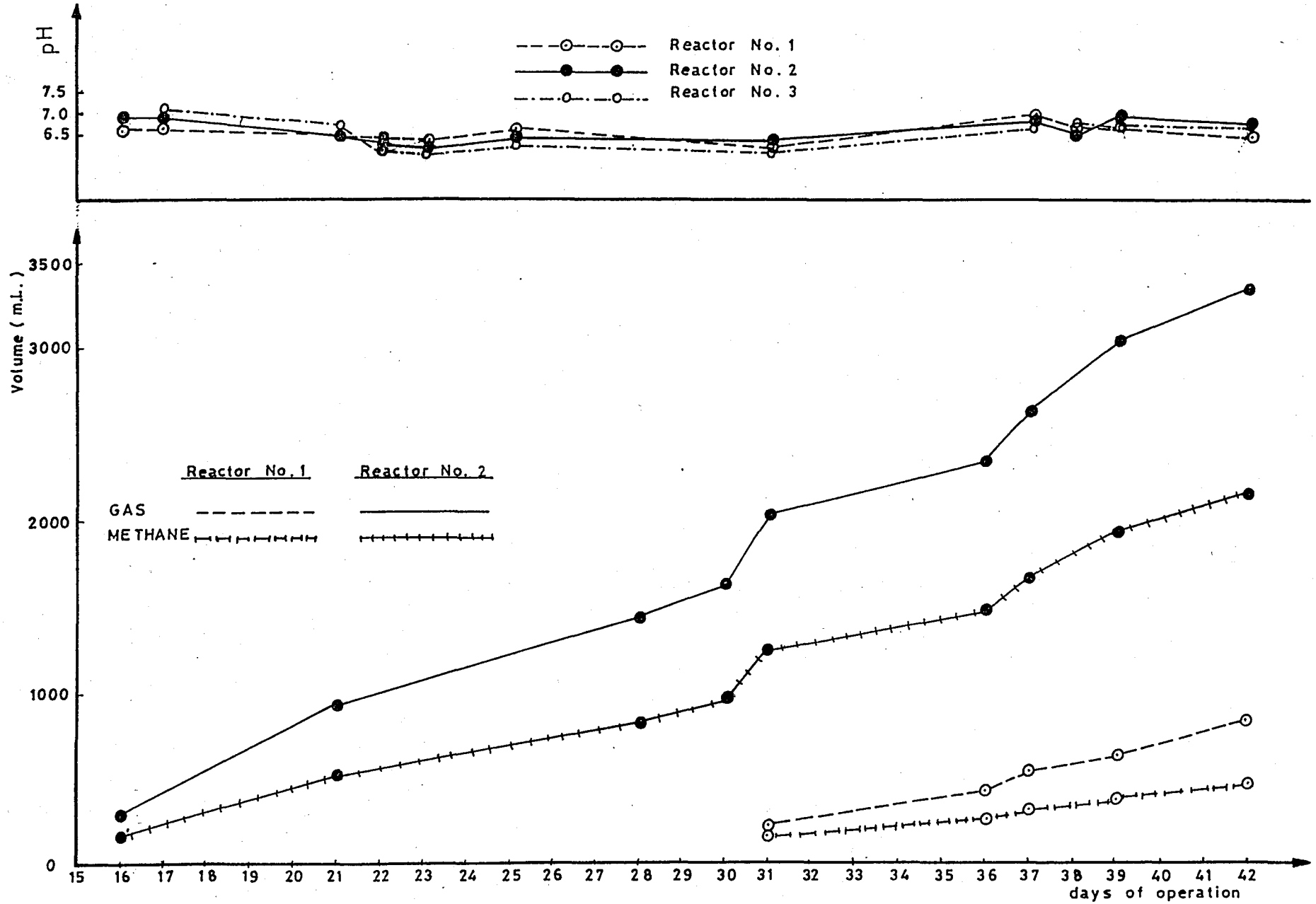


Figure: 30. Cumulative Gas Production and Methane Production

5.2 Batch Reactors

In this part of the study the performence of the Anaerobic Attached Film Expanded Bed Reactor (Reactor 1) and the control reactor (Reactor 3). under batch conditions was examined. The reactors were both fed with Substrate I and worked with for 20 days. During this period the necessary tests were performed. The results can be given as follows:

5.2.1 Chemical Oxygen Demand (COD)

The variations in effluent COD concentrations with time for the Anaerobic Attached Film Expanded Bed Reactor (Reactor 1) and for the control reactor (Reactor 3) are drawn in Figure 31 (a).As can be seen in this figure the COD reduction rate is considerably higher in Reactor 1 than in Reactor 3. This becomes more obvious in Figures 31 (b) and (c) where the reduction rate can be observed readily. Although the initial COD was 29520 mg/L in both reactors,after 20 days the COD in reactor 1 was 5638 mg/L while in Reactor 3 it was 12910 mg/L.

Another interesting observation which can be made from these Figures is that the control reactor reaches steady state faster than the Anaerobic Attached Film Expanded Bed Reactor. Although after the 13^{th} day COD reduction of the control reactor remains constant, the COD reduction of the Anaerobic Attached Film Expanded Bed Reactor continues. Both observations are clear indications that the efficiency of COD removal in AAFEB is greater. This may be the result of the fact that the bacterial population in the Anaerobic Attached Film Expanded Bed Reactor is larger than the control reactor.

If a parameter N is defined as the COD per cent reduction difference between Reactor 1 and Reactor 3,divided by COD reduction per cent of the Reactor 1 at a given time,

$$N: \frac{\text{COD reduction per cent of (Reactor 1-Reactor 3)}}{\text{COD reduction per cent of Reactor 1}} \times 100$$

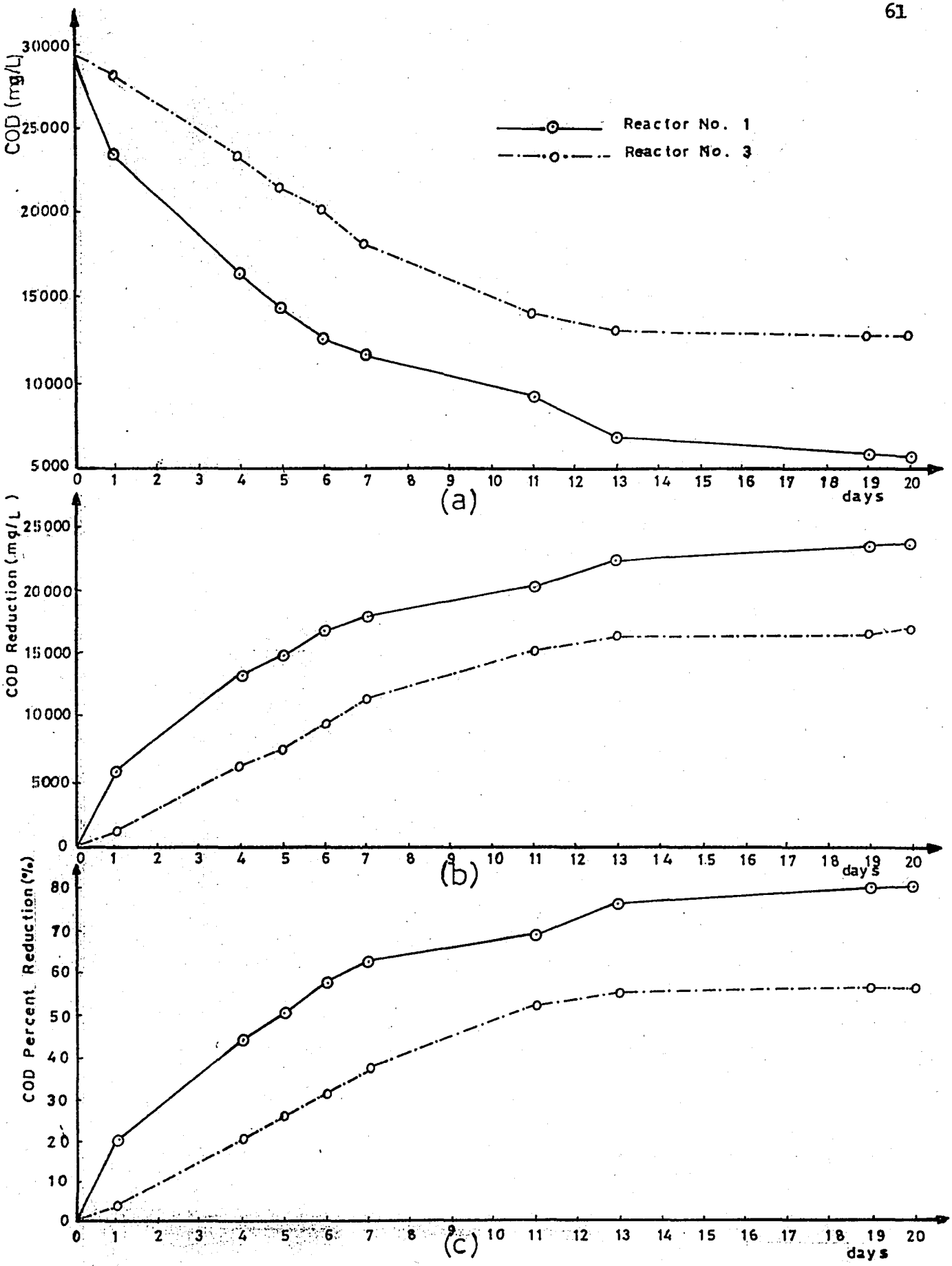


Figure: 3I(a) Variations in effluent Chemical Oxygen Demand (COD) concentrations in Batch study

(b) Chemical Oxygen Demand (COD) reduction versus time

(c) Chemical Oxygen Demand (COD) reduction per cent versus time

the N value for the first day is 79 per cent and for 20 days is 30 per cent. The decrease in the N value gives the decrease in the performance of reactor 1 with time. The increase in hydraulic detention time decreases the performance of the Anaerobic Attached Film Expanded Bed with respect to the control reactor. The raw data related to COD reduction are given in Appendix 1, Table 5.

5.2.2 Volatile Solids

Variations in effluent Volatile Solids concentrations for the Anaerobic Attached Film Expanded Bed Reactor (Reactor 1) and for the control reactor (Reactor 3) are drawn in Figure 32 (a).As can be seen in this figure the amount of volatile solids remaining in the AAFEB reactor was smaller than that in the control reactor. Although initially the volatile solids influent values were the same (10184 mg/L) for both reactors,in one day these values became 8480 mg/L for the AAFEB reactor and 9465 mg/L for the control reactor,in seven days 4074 mg/L for The AAFEB reactor and 6470 mg/L for the control reactor and in 20 days 1700 mg/L for the AAFEB reaçtor and 5010 mg/L for the control reactor. Figure 32 (b) shows the total volatile solids reduction versus time and Figure 32 (c) shows the total percentage reduction of volatile solids versus time.

When Total Volatile Solids reduction percentages of the AAFEB reactor and the control reactor are considered,the AAFEB reached the higher Total Volatile Solids reduction percentage in 6 days, while the control reactor in 20 days. The total Volatile Solids reduction per cent difference between the AAFEB reactor and the control reactor was obtained as 32.5 per cent in 20 days hydraulic detention time.

If a parameter N_1 is defined as the per cent difference between the AAFEB reactor and the control reactor in the Total Volatile Solids, divided by the per cent total volatile solids reduction of the AAFEB

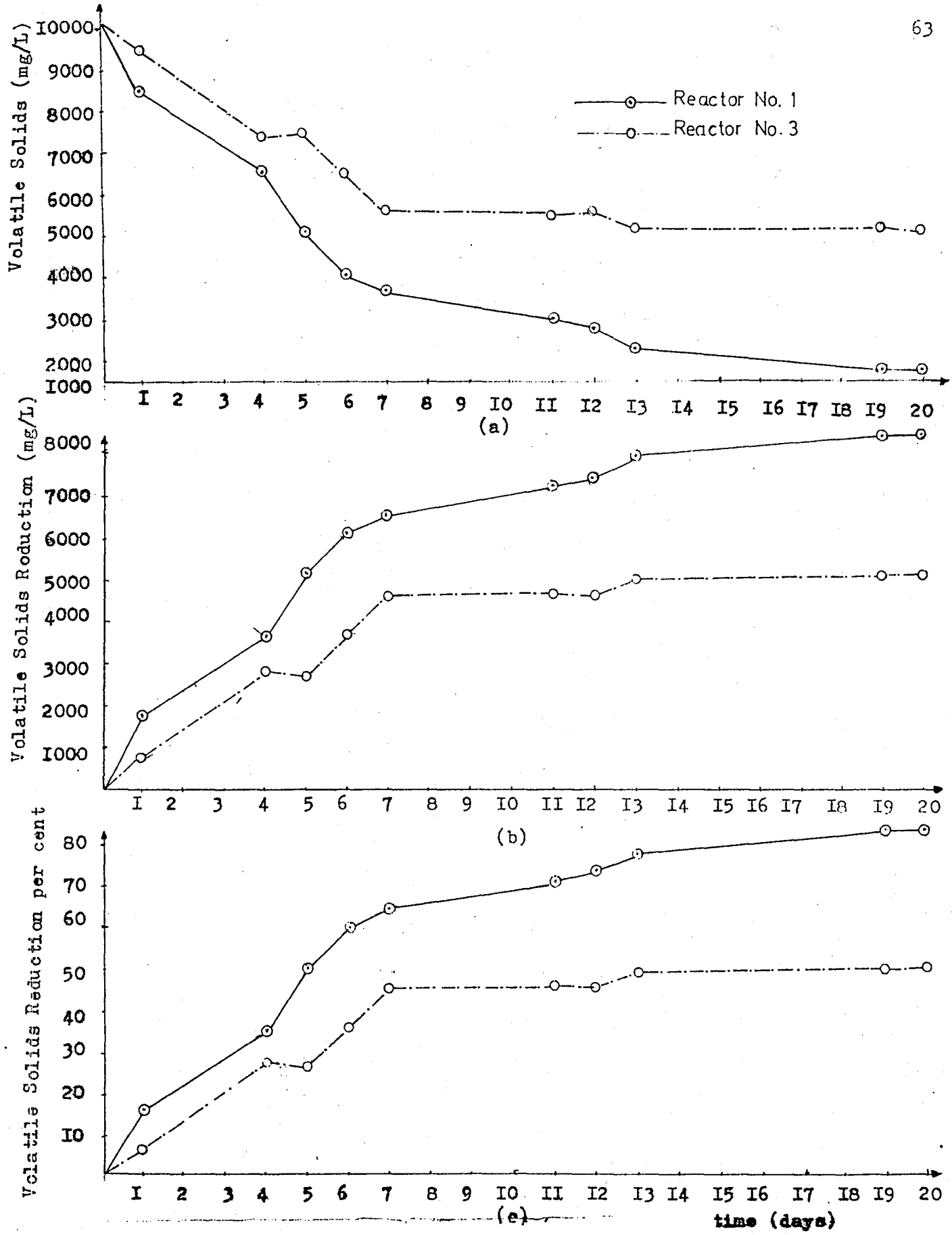


Figure: 32(a) Variations in effluent Volatile Solids concentrations with time

(b) Volatile Solids Reduction versus time

(c) Per cent Volatile Solids Reduction versus time

reactor at a given time,

$$N_1 = \frac{\text{Total Volatile Solids Reduction per cent of (Reactor 1-Reactor 3)}}{\text{Total Volatile Solids Reduction per cent of Reactor 1}} \times 100$$

the N_1 value for the first day is 58.1 per cent while on the 20th day it is 39.0 per cent. The only difference in the definitions of N and N_1 is that N is for COD and N_1 is for Total Volatile Solids. The N and N_1 values attain greater values in shorter times and smaller values over longer periods. Therefore, it can be concluded that, the increase in hydraulic detention time also decreases the performance of the AAFEB for volatile solids when compared with the control reactor.

When the COD Reduction percentage versus time (Figure 31-c)and the Total Volatile Solids Reduction percentage versus time (Figure 32-c) curves are compared for both reactors, the AAFEB reactor (Reactor 1)and the Control Reactor (Reactor 3)show a similarity. The per cent difference between COD reduction and Volatile Solids Reduction in 20 days is only 2.4 per cent in Reactor 1 while 5.5 per cent in Reactor 3. The small treatment differences between Chemical Oxygen Demand and Volatile Solids for a given time gives an idea about the reliability of the experiments.

5.2.3 Gas Production

It is well known that one of the end products of anaerobic decomposition is methane. The results obtained in relation to this parameter are represented graphically in Figure 33.

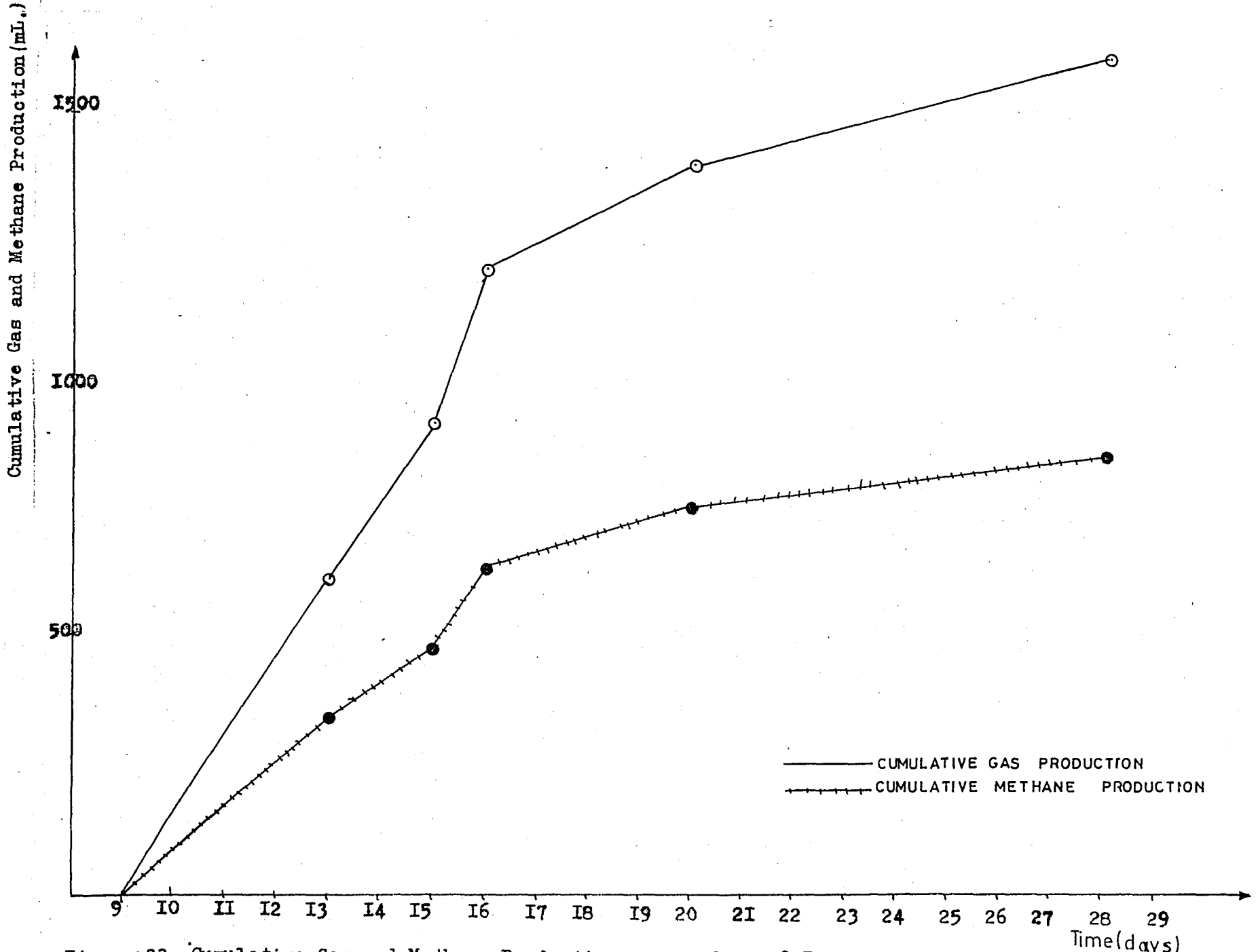


Figure:33. Cumulative Gas and Methane Production versus days of Expanded Bed Reactor

CHAPTER 6

CONCLUSIONS

The Anaerobic Attached Film Expanded Bed system is a feasible organic load removal alternative. The high efficiency obtained in this process is due to the large amount of active microorganisms accumulated on the surface of the filter media as well as in the voids among them.

The most important conclusions of this study are:

a) The Chemical Oxygen Demand removal efficiency increased both for batch and semi-continuous reactors with the increase in organic loading.

b) In a similar way the removal rate of total volatile solids was directly proportional to the load of total volatile solids applied.

c) The performance of the Anaerobic Attached Film Expanded Bed System was better than the performance of the simple anaerobic reactor.

d) The substrate containing yeast showed a better treatability than the substrate without yeast. Thus the results obtained from previous studies in Boğaziçi University (Alpaslan 1979, Baban 1982, Kocasoy 1982) were verified. Definitely, addition of yeast into wastes to be treated anaerobically will increase the efficiency. The detention periods observed in this study were larger than expected. It is believed however that with further investigation the system may be improved, and thus the Anaerobically Attached Film Expanded Bed Reactor can become a feasible process.

Due to the fact that the load applied per unit area of the Anaerobic Attached Film Expanded Bed Reactor is comparatively higher than the load applied in conventional anaerobic digesters, the system may become financially comparable with other anaerobic techinques. The extra energy required in the Anaerobic Attached Film Expanded Bed Reactor for

expanding the bed may be obtained from methane produced during the decomposition.

CHAPTER 7

RECOMMENDATIONS FOR FURTHER RESEARCH

7.1 Recommendations for Improvement of the Experimental Set-up

The following alterations should be made on the system if further studies are conducted.

1. Precautions should be taken to prevent heat losses during the study. This will help to make the system more economical.
2. The feeding system should be improved.
3. Intermediate sampling points should be introduced.
4. A more effective sludge removal unit should be incorparated into the system.

7.2 Topics for Further Research

Related to the Anaerobic Attached Film Expanded Bed Reactor system the following subjects will be interesting for further research.

1. Determination of the best size, type, shape and depth of filter media to be used in the Anaerobic Attached Film Expanded Bed Reactor.
2. Investigation of the performance of the system when Anaerobic Attached Film Expanded Bed Reactor units are connected in series.
3. Investigation of the performance of a continuous system.
4. Use of the system for the treatment of different industrial wastes like brewery wastes and olive oil wastes.

APPENDIX

TABLE 1. Semi-Continuously Fed Reactors Data (At 35+ 1 C)

DATE	@ DAYS	DAYS FROM START	COLON No.	COD INFL. mg/L	COD EFFL. mg/L	TOTAL SOLID INFL. mg/L	TOTAL SOLID EFFL. mg/L	VOLATILE S. INFL. mg/L	VOLATILE S. EFFL. mg/L	N_2 (KJ-HELDAL mg/L	VOLATILE ACID mg/L	GAS QUANTITY mL	CH_4 %	pH
May, 2	5.6	7	1	2356	1919	6864	3976	3250	2843					
"	5.6	7	2	2928	2682	5062	4559	3464	3658					
May, 5	5.6	10	1			6864	3878	3250	2770					7.2
"	5.6	10	2			5062	4476	3464	3268					7.4
May, 11	5.6	16	1			6864	4104	3250	2836					7.2
"	5.6	16	2			5062	4200	3464	2906			300	60.0	7.4
May, 12	5.6	17	1	2356	1848						960			7.2
"	5;6	17	2	2928	2611						1022			7.4
"	5.6	17	3	2125	1408									7.6
May, 16	5.6	21	1			6864	3200	3250	1946					7.0
"	5.6	21	2			5062	5130	3464	2595			650	56.0	7.0
"	5.6	21	3			4988	2988	2224	1843					7.2
May, 17	5.6	22	1			5665	3120	3456	1910					6.9
"	5.6	22	2			6556	4786	4788	3350					6.8
"	5.6	22	3			4998	2754	2224	1788					6.6
May, 18	5?6	23	1	2746	2015					93.06	1160			6.8
"	5.6	23	2	3916	3270					344.86	1201			6.7
"	5.6	23	3	2125	1768					165.20				6.6
May, 20	5.6	25	1	2746	1872									7.1
"	5.6	25	2	3916	3078									6.9
"	5.6	25	3	2125	1535									6.8
May, 23	5.6	28	1	2746	1248	5665	3212	3456	1754					
"	5.6	28	2	3916	1706	7061	4224	4788	2616			500	68.0	
"	5.6	28	3	2125	1442	4998	3359	2224	1856					

TABLE I. (Continued)

DATE	0 DAYS	DAYS FROM START	COLON No;	COD INFL. mg/L	COD EFFL. mg/L	TOTAL SOLID INFL. mg/L	TOTAL SOLID EFFL. mg/L	VOLATILE S. INFL. mg/L	VOLATILE S. EFFL. mg/L	N_2 (KJ-HELDAL) mg/L	VOLATILE ACID mg/L	GAS QUANTITY mL	CH_4 %	pH
May, 25	5.6	30	1	2746	1246									
"	5.6	30	2	4790	1914							200	68.0	
"	5.6	30	3	2125	1843									
May, 26	5.6	31	1	2746	1352	5665	3276	3456	1952			250	71.0	6.8
"	5.6	31	2	4790	2371	7061	4020	4810	2704			400	74.6	6.9
"	5.6	31	3	2125	1625	4998	3158	2224	1762					6.7
May, 27	5.6	32	1	2746	1186	5665	3228	3456	2006					
"	5.6	32	2	4790	2496	7061	4138	4810	2466					
"	5.6	32	3	2125	1421	4998	3012	2224	1737					
May, 31	5.6	36	1									200	7100	7.4
"	5.6	36	2									300	74.6	7.3
"	5.6	36	3											7.1
June, 1	11.2	37	1	2746	1131	5665	3418	3456	1992			100	52.0	7.4
"	11.2	37	2	4790	1622	7061	4652	4810	2442			300	68.0	7.3
"	11.2	37	3	2125	1840	4998	3207	2224	1844					7.2
June, 2	5.6	38	1											7.2
"	5.6	38	2											7.1
"	5.6	38	3											7.2
June, 3	5.6	39	1							10.9		100	54.0	7.2
"	5.6	39	2							395.4		400	64.0	7.4
"	5.6	39	3							99.2				7.2
June, 6	5.6	42	1	3248	1388							200	52.0	7.0
"	5.6	42	2	5880	1867							300	76.3	7.3
	5.6	42	3	2125	947									7.2

TABLE II. Batch Reactors Data (At 35 1C)

DATE	COLON No.	O Days	COD INFL. mg/L	COD EFFL. mg/L	T.S. INFL. mg/L	T.S. EFFL. mg/L	V.S. INFL. mg/L	V.S. EFFL. mg/L	N (KJ-HELDAL) mg/L	VOLATILE ACIDS mg/L	PHOSPHATE mg/L	GAS QUANT. mL	CH %	pH
June, 9	1		29520		14962		10184		2658	2880	180			7.2
"	3		29520		14962		10184		2658	2880	180			7.2
June,10	1	1		23527		14520		8480				600	56	7.0
"	3	1		28260		13010		9465						7.1
June,13	1	4		16440		11709		6540						6.7
"	3	4		23360		11735		7330						7.0
June,14	1	5		14545		11650		5013						
"	3	5		21680		11240		7430						
June,15	1	6		12560		9800		4074				300	42	6.6
"	3	6		20160		10280		6470						6.9
June,16	1	7		11520		5160		3638				300	54	6.6
"	3	7		18240		10835		5510						6.9
June,20	1	11		10273		4630		2960				200	58	6.5
"	3	11		14080		9275		5467						6.8
June,21	1	12				4909		2718						6.5
"	3	12				9207		5520						6.7
June,22	1	13		6967		4220		2260	296	3225	140			6.5
"	3	13		13042		8880		5120	234	2522	165			6.7
June,28	1	19		5875		3980		1731				200	51	6.4
"	3	19		12880		7755		5010						6.7
June,29	1	20		5638		3990		1700						6.3
"	3	20		12910		7750		5010						6.6

TABLE III. Chemical Oxygen Demand (COD) Data for Semi-Continuously Fed Reactors

DATE	DAYS FROM START	REACTOR 1				REACTOR 2				REACTOR 3			
		COD Infl. mg/L	COD Effl. mg/L	COD Reduc. mg/L	COD Reduc. %	COD Infl. mg/L	COD Effl. mg/L	COD Reduc. mg/L	COD Reduc. %	COD Infl. mg/L	COD Effl. mg/L	COD Reduc. mg/L	COD Reduc. %
May , 2	7	2356	1919	437	18.5	2928	2682	246	8.4				
May, 12	17	2356	1848	508	21.6	2928	2611	317	10.8	2125	1408	717	33.7
May, 18	23	2746	2015	731	26.6	3916	3270	646	16.5	2125	1768	357	16.8
May, 20	25	2746	1872	874	31.8	3916	3078	838	21.4	2125	1535	590	27.8
May, 23	28	2746	1248	1498	54.6	3916	1706	2210	56.4	2125	1442	638	32.1
May, 25	30	2746	1246	1500	54.6	4790	1910	2876	60.0	2125	1843	262	13.3
May, 26	31	2746	1352	1394	50.8	4790	2371	2419	50.5	2125	1625	500	23.5
May, 27	32	2746	1186	1560	56.8	4790	2496	2294	47.9	2125	1421	704	33.1
June, 1	37	2746	1131	1615	58.8	4790	1622	3168	66.1	2125	1840	285	13.4
June, 6	42	3248	1470	1860	57.3	5880	2205	3675	62.5	2125	1723	402	18.9

TABLE IV. Volatile Solids in Semi-Continuously Fed Reactors

DATE	DAYS FROM START	REACTOR 1				REACTOR 2				REACTOR 3			
		V.S. Infl. mg/L	V.S. Effl. mg/L	V.S. Reduc. mg/L	V.S. Reduc. %	V.S. Infl. mg/L	V.S. Effl. mg/L	V.S. Reduc. mg/L	V.S. Reduc. %	V.S. Infl. mg/L	V.S. Effl. mg/L	V.S. Reduc. mg/L	V.S. Reduc. %
May, 2	7	3250	2843	407	12;5	3464	3658	-	-				
May, 5	10	3250	2770	480	14.8	3464	3268	196	5.7	-			
May,11	16	3250	2836	414	12.7	3464	2906	558	16.1				
May,16	21	3250	1946	1304	40.1	3464	2595	869	25.1	2224	1843	390	17.5
May,17	22	3456	1910	1546	44.7	4788	3350	1438	30.0	2224	1788	436	19.6
May,23	28	3456	1754	1702	49.2	4788	2616	2172	45.4	2224	1856	368	16.5
May,26	31	3456	1952	1504	43.5	4810	2704	2306	47.9	2224	1762	462	20.8
May,27	32	3456	2006	1449	41.9	4810	2466	2344	48.7	2224	1737	487	21.9
June,1	37	3456	1992	1464	42.4	4810	2442	2366	49.2	2224	1844	380	17.1

TABLE V. Chemical Oxygen Demand (COD) Reduction and COD Reduction Percentages at the given day in Batch Reactors

DATE	DAYS FROM START	REACTOR I				REACTOR 3			
		COD Influent mg/L	COD Effluent mg/L	COD Reduction mg/L	COD Reduction Per Ceht	COD Influent mg/L	COD Effluent mg/L	COD Reduction mg/L	COD Reduction Per Cent
June , 9		29520				29520			
June,10	1		23527	5993	20.3		28260	1260	4.27
June,13	4		16440	13080	44.3		23360	6160	20.80
June,14	5		14545	14975	50.73		2168D	7862	26.6
June,15	6		12560	16960	57.45		20160	9360	31.7
June,I6	7		11520	18000	62.1		18240	11280	38.2
June,20	11		10273	20247	66.5		14080	15440	52.3
June,22	13		6967	22553	76.4		13042	16480	55.8
June,28	19		5875	23645	80.1		12880	16640	56.4
June,29	20		5638	23882	80.9		12910	16610	56.3

TABLE VI. Volatile Solids (V.S) Reduction and V.S. Reduction Percentages at the Given Day for Batch Reactors

DATE	DAYS FROM START	REACTOR 1				REACTOR 3			
		V.S. Influent mg/L	V.S. Effluent mg/L	V.S. Reduction mg/L	V.S. Reduction Per cent	V.S. Influent mg/L	V.S. Effluent mg/L	V.S. Reduction mg/L	V.S. Reduction Per cent
June, 9		10184				10184			
June, 10	1		8480	1704	16.7		9465	714	7.0
June, 13	4		6540	3644	35.8		7330	2854	28.0
June, 14	5		5013	5171	50.8		7430	2754	27.0
June, 15	6		4074	6110	60.0		6470	3714	36.5
June, 16	7		3638	6546	64.3		5510	4674	45.9
June, 20	11		2960	7224	70.9		5467	4717	46.3
June, 21	12		2718	7466	73.3		5520	4664	45.8
June, 22	13		2260	7924	77.8		5120	5064	49.7
June, 28	19		1731	8453	83.3		5010	5174	50.8
June, 29	20		1700	8474	83.3		5010	5174	50.8

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