KINETICS OF AN ANAEROBIC FLUIDIZED BED REACTOR USING BIOLITE INERT CARRIER PARTICLES

by

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Submitted in partial fulfilment
of the requirements for the degree of
Master of Applied Science

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University of Ottawa.

September 1994.

ISBN 0-612-04914-0

BIOGRAPHICAL SKETCH

Rajiv Prakash was born in Richmond, Virginia on August 2, 1968. Prior to entering Graduate School as a Master's student, he obtained a Bachelor of Applied Science degree from the Department of Chemical Engineering of the University of Ottawa in April, 1990.

ACKNOWLEDGEMENTS

The author wishes to express his sincere appreciation and thanks to Dr. Kevin Kennedy for his guidance, support and patience during the investigation and writing of this thesis.

A special feeling of appreciation is extended to Joe, Lou and Frank of the chemical engineering machine shop for their help in constructing the experimental apparatus. The author would also like to thank Kathleen and Frank for their general advice on the operation of the reactors and other equipment.

Finally, the author wishes to thank his father and mother who offered encouragement and support throughout his graduate education.

ABSTRACT

This work investigates the startup and steady state operation of anaerobic fluidized bed reactors (AFBRs) with "Biolite"as the inert carrier material. Biolite is a new product of Degremont S.A., France, and is claimed to be highly suitable for expanded and fluidized bed reactors due to its size, density, and surface characteristics.

Startup of AFBRs was performed using the maximum efficiency profile (MEP) and the maximum load profile (MLP) using sucrose based wastewater as feed. Although both loading profiles led to equally fast startup times of about 2 months, MEP resulted in more stable reactor operation during the startup period and thus should be preferred over MLP. The quick startup of this study confirms the high compatibility of Biolite particle's surface for bio-adhesion.

Steady state experiments were performed using three concentrations of sucrose substrate and five hydraulic retention times (HRT). Organic removal efficiencies over 80% were obtained for volumetric loadings as high as 20 g/l/d. Substrate removal efficiency decreased with increasing loading rate and decreasing HRT. It was found that the dependence of removal efficiency on HRT is influenced by substrate concentration.

Biomass accumulated in the AFBR in the form of biofilms on Biolite particles. Biomass increased exponentially with organic loading over the conditions of this study. For a organic loading of 25 g/l/d, biomass reached a maximum of 230 mg volatile film solids (VFS)/g Biolite corresponding to 69 g VFS/l of expanded bed volume. This high

concentration of biomass in AFBRs allows treatment of wastewaters at high loading rates and removal efficiencies.

Total yield was determined to be 0.1 g VFS/g COD removed demonstrating the low net synthesis of solids in the AFBR which is highly desirable for wastewater treatment systems. Under the experimental conditions tested, the AFBRs had an average solids retention time (SRT) of 150 days and a washout factor (f) of 0.01. Thus even at very low HRTs, reactor operation is stable with no washout of microorganisms. The extrinsic kinetics of the AFBRs was determined to be zero order with a maximum specific utilization rate (k) of 0.48 d^{-1}. This value of k is conservative since all of the biomass was considered active.

Municipal landfill leachate was treated at steady state conditions at two influent concentrations and four different loading rates. Removal efficiencies as high as 87% were obtained in the AFBR which corresponds to all the biodegradable waste present in the leachate (BOD_5:COD of 0.86). Compared to sucrose based waste, COD removal efficiencies for leachate were lower and not as dependent on loading rate. During the course of treating leachate, the biomass gradually became "mineralized" due to the precipitation of metal sulphides and carbonates. This resulted in loss of sludge activity and the need for higher pumping rates to maintain the same degree of bed expansion.

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NOMENCLATURE

b = decay coefficient (T^{-1})

C_I = inertial drag coefficient

d_s = diameter of sphere of equal volume (L)

D_p = particle diameter (L)

f = washout factor

F_k = drag force acting on a particle in a fluidized bed (MLT^{-2})

F_{ks} = drag force on a single sphere (MLT^{-2})

g = acceleration due to gravity (MLS^{-2})

g_c = factor to convert natural unit of force to derived unit

Ga = Galileo Number

k = maximum specific utilization rate (T^{-1})

K_S = half velocity or saturation constant (ML^{-3})

L = height of fluidized bed (L)

L_O = height of bed if porosity were zero (L)

M = total mass of particles in bed

Re = Reynolds Number

Re_{mf} = Reynolds Number at minimum fluidization velocity

S = substrate concentration (ML^{-3})

S_o = influent substrate concentration (ML^{-3})

u = actual velocity in the fluidized bed (LT^{-2})

U = specific utilization rate (T^{-1})

u_i = terminal free fall velocity of a single sphere (LT^{-2})

V_0 = superficial velocity (LT^{-2})

X = biomass concentration ($M\ L^{-3}$)

X_L = biomass concentration is reactor mixed liquor ($M\ L^{-3}$)

Y = yield coefficient

μ = specific growth rate (T^{-1})

μ' = net specific growth rate (T^{-1})

μ_m = maximum specific growth rate (T^{-1})

ϵ = porosity

Δp = pressure drop in fluidized bed ($ML^{-1}T^{-2}$)

ρ = fluid density (ML^{-3})

ρ_p = particle density (ML^{-3})

ϵ_M = minimum porosity for fluidization

Φ_S = particle sphericity

AFBR = anaerobic fluidized bed reactor

BOD_5 = five day biochemical oxygen demand

COD = chemical oxygen demand

DSFF = downflow stationary fixed film

FBR = fluidized bed reactor

FID = flame ionized detector

GC = gas chromatograph

HRT = hydraulic retention time

MEP = maximum efficiency profile

MLP = maximum load profile

OLR = organic loading rate

SBF = sludge bed filter

SRT = solids retention time

STP = standard temperature and pressure

TOC = total organic carbon

UASB = upflow anaerobic sludge blanket

VFA = volatile fatty acids

VFS = volatile film solids

VSS = volatile suspended solids

Chapter 1: INTRODUCTION

1.1 General Background

Anaerobic digestion is one of the key routes for the decomposition of organic matter in nature. The final products of this decomposition are methane, carbon dioxide, and in some cases water, along with the production of more cells. By controlling the habitat of the microbial population, this natural process may be accelerated, thus making it useful for wastewater treatment.

Stricter environmental regulations combined with the rising cost of energy has made the use of anaerobic processes for wastewater treatment economically attractive. Anaerobic processes have the capability of removing a major fraction of the organics present in most wastewaters. They also allow recovery of the energy value of the effluent in the form of methane gas, which for certain industries constitutes a substantial fraction of their total energy consumption. Adding an anaerobic pretreatment stage to existing aerobic treatment facilities helps in reducing aeration costs.

Methane fermentation has been used for municipal and industrial sludge stabilization for over a century. Over the last few decades, tremendous research efforts have yielded significant gains in the knowledge of the microbiology and biochemistry of anaerobic wastewater treatment processes. Despite these advances, anaerobic processes have been largely dismissed for their application in wastewater treatment due to the high sensitivity and low growth rates of anaerobic bacteria.

To minimize these limitations, researchers have designed several novel reactor configurations. These second generation reactors have the common ability to retain active biomass or microbial cells in the reactor beyond the hydraulic retention time instead of allowing it to leave with the treated effluent. One of these types of reactors is the anaerobic fluidized bed reactor (AFBR) in which anaerobic biomass is present in the form of a biofilm supported on inert carrier particles. These particles together with their attached biofilm are fluidized by the upwardly directed flow of wastewater. The large surface area to volume ratio available for bacterial growth combined with high sludge activity lead to high removal rates and good process stability.

The type of carrier particles used in an AFBR affects the concentration of biomass which can be maintained inside the reactor and degree of its contact with the incoming wastewater. Thus, the carrier particle material and size are important design considerations for fluidized bed reactors. In order to improve treatment efficiencies, there is a need to test new materials for the applicability to AFBRs.

Biolite, of Degremont S.A., France, is one such novel prototype material which from preliminary investigations appears to hold a lot of promise. It is light weight, strong, of appropriate size and is considered to have excellent surface adhesion properties. All these characteristics make it highly suitable for use in fluidized bed reactors. Although presently Biolite is relatively expensive (100-300 $ Cdn. per cubic metre), its cost will presumably drop once it is produced on a large scale.

Due to the recent development of Biolite, few studies have been done on the performance of AFBRs using this material. The availability of results that relate process efficiency to parameters such as biomass concentration, microbial activity and solids retention time is crucial before any full scale applications. Also, the need to evaluate start-up procedures using this material is of utmost importance since properties of microbial films which dictate final steady state reactor performance can be established early in development (Bryers and Charaklis, 1979).

1.2 Objectives

The general objectives of this study were to investigate startup and steady state operation of anaerobic fluidized bed reactors using Biolite carrier particles of Degremont Inc., France. The specific objectives were to:

1. Compare different startup procedures and their effect on early biofilm development.
2. Establish the suitability of Biolite as an inert carrier material for AFBRs.
3. Evaluate the effect of substrate concentration on steady state reactor performance and biofilm growth.
4. Determine biofilm microbial activity and show importance of the biofilm in the process.
5. Determine steady state kinetic constants for the removal of sucrose based wastewater.
6. Access the treatability of municipal landfill leachate in an AFBR using Biolite particles.

Chapter 2: LITERATURE REVIEW

2.1 Overview

Section 2 discusses the biochemical steps of anaerobic degradation and compares it to aerobic processes for wastewater treatment, highlighting some of the distinct advantages of anaerobic systems.

Section 3 outlines the various types of immobilized biomass anaerobic reactors.

Section 4 describes the AFBR, its specific design and operating parameters, and laboratory studies performed on organic carbon removal from various wastewaters.

Section 5 gives some background information on landfill leachate, its characterization, and its treatment in anaerobic reactors.

2.2 Anaerobic Wastewater Treatment

The net result of anaerobic digestion is the conversion of water soluble organic compounds to methane, carbon dioxide and small amounts of hydrogen, nitrogen, sulphide and bacterial cells. This is possible due to the complex interactions between many types of metabolically different bacteria, each utilizing a product excreted by another. The population diversity and inter-species interactions give the process stability and permits the degradation of a large variety of substrates.

2.2.1 Biochemistry and Microbiology

Due to the many pathways available to anaerobes, the biochemistry of anaerobic processes is much more complicated than that of aerobic processes. Although specific details of the pathways and microbes responsible for the reactions is not fully known, a broad outline of the processes has been established. As shown in Figure 1, the anaerobic metabolism of a complex substrate can be regarded as occurring in three steps which are: hydrolysis, fatty acids production, and methane generation (Henze and Harremoes, 1983). Each of these three processes is described below.

In the first step of anaerobic digestion, complex organic substrates are hydrolysed into simple compounds such as sugar monomers, amino acids and alcohols. This is achieved through the excretion of extracellular enzymes produced by the acidogenic bacteria. In the case of influent containing large amounts of lipids, the hydrolysis reaction may become the overall rate limiting step.

The second stage of anaerobic digestion is acid formation or "acetogenesis". Here, the products from the hydrolytic stage are transported into the bacterial cells and are fermented to the C_1 to C_4 monocarboxylic or short chain volatile fatty acids (VFAs), lactic acid, ethanol, methanol, ammonia, hydrogen and carbon dioxide. Among the fatty acids, acetic acid is preferentially formed although in the case of process instability, significant quantities of propionic, butyric, and valeric acids are also produced. According to Henze and Harremoes (1983), the level of propionic acid may be used as a stress indicator in anaerobic systems. Acid producing bacteria are chemoheterotrophic and are comprised of a mixed microbial population of obligate and facultative anaerobes.

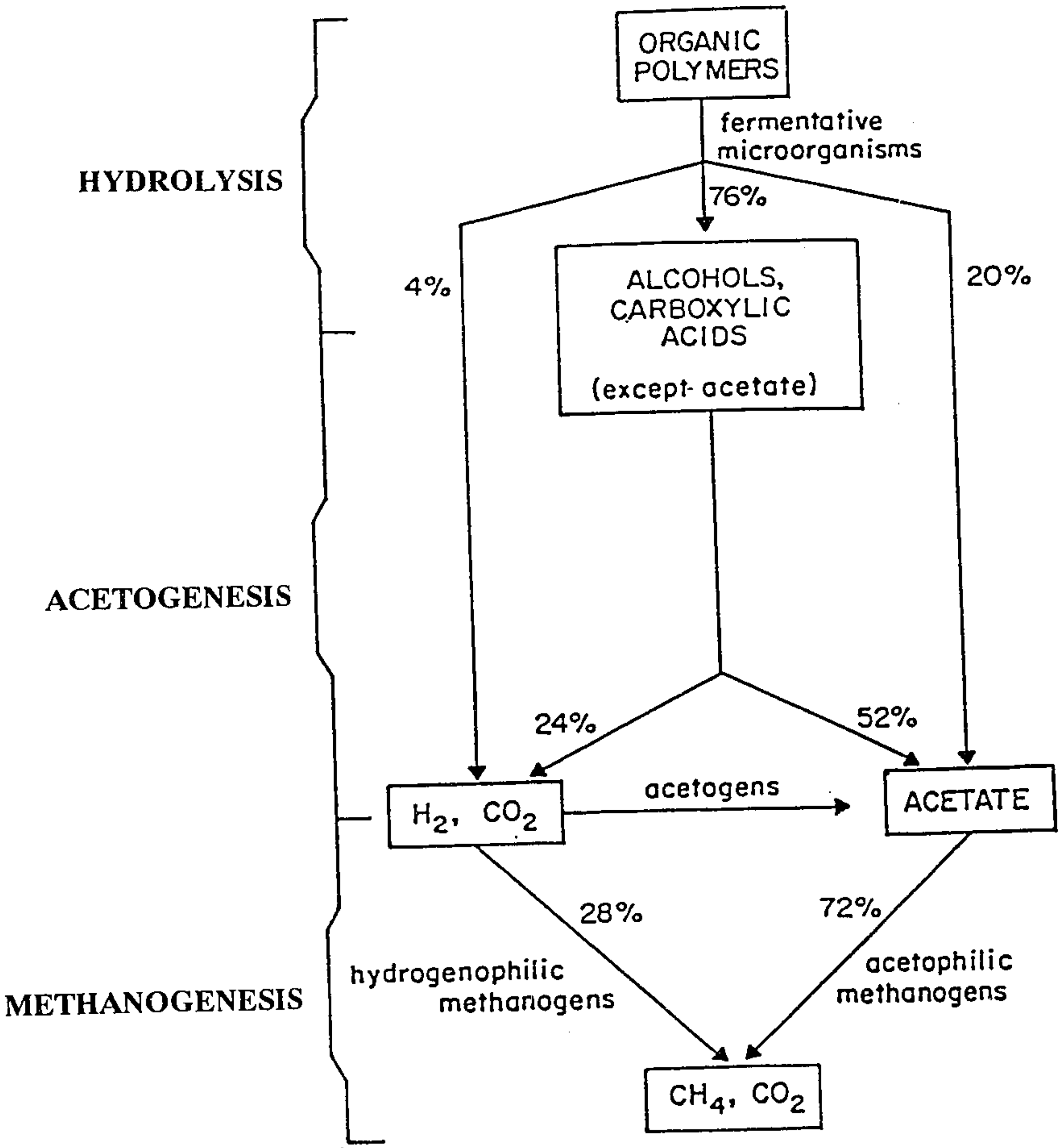

Figure 1. Anaerobic Decomposition of Organic Matter

Methanogenesis, the final stage of the anaerobic degradation of carbon, occurs through two distinct pathways. One is via the decarboxylation of acetate to form methane and carbon dioxide. The other pathway involves the oxidation of hydrogen and the reduction of carbon dioxide to form methane and water. The former usually accounts for 70% of total methane production, however this figure varies depending on the substrate and microbial consortia present.

Few categories of microorganisms are as diverse as methane producing bacteria. They may be in the form of sarcina (packets of irregular size cells), rods, cocci (spheres) and spirals. They also often appear as pairs, chains or long filaments. These cells demonstrate a tendency to adhere to solid surfaces and to aggregate into clumps, flocs, or granules. All methane forming bacteria are strict anaerobes and can utilize only a few substrates. Although acetate is the major methanogenic precursor, only the Methanosarcina species - Methanothrix soehngenii and Methanococcus mazei can form methane from acetate (Mah, 1982).

2.2.2 Anaerobic vs. Aerobic Processes

Anaerobic processes have a number of distinct advantages over aerobic treatment alternatives (Kennedy et al.,1990).

1) Anaerobic treatment does not require power intensive aeration. This leads to cost savings especially when treating concentrated or toxic effluent.

2) Due to the high biomass densities in advanced anaerobic reactors and the absence of oxygen transfer limitations, these reactors can be operated at high loading rates and are

able to withstand shock loadings. This also means that reactor space requirements are very small.

3) Anaerobic processes are characterized as having low nutrient and sludge disposal requirements. This is because of the low cell yield of approximately 5% of chemical oxygen demand (COD) removed compared to 50% for aerobic systems.

4) Rapid start-up is possible even after long dormant periods of up to a year.

5) Since thermophilic anaerobes can function in the temperature range of 50 to 65 °C, cooling requirement for hot effluents may be reduced or unnecessary.

6) The production of methane rich gas (theoretical maximum of 0.35 m^3/kg of COD removed) can potentially offset operating costs or provide additional revenue.

Some disadvantages associated with anaerobic processes are:

1) The specific substrate removal rate of anaerobic biomass is approximately 70% lower than for aerobic systems (Lee et al., 1989). This inherent deficiency can be partly overcome by maintaining higher densities of biomass and operating at pH and temperature optima.

2) The low cell yield of anaerobic systems which is advantageous during steady state operation is a significant drawback during reactor start-up and recovery from upset.

3) The pH of the reactor must be controlled near neutral for the methanogenic population to function optimally. This usually requires addition of costly buffering chemicals.

4) The methane forming bacteria are sensitive to oxidants such as oxygen, peroxide and chlorine, to sulphide and sulphite, and to heavy metals.

5) Because most anaerobic processes cannot effectively treat very dilute wastewater,

effluent from these reactors may be unsuitable for direct discharge to the receiving waters. Therefore, rather than entirely replacing aerobic treatment systems, anaerobic systems can remove substantial amounts of biological oxygen demand (BOD) prior to aerobic polishing, reducing the capital and operating costs of the operation as whole.

2.3 Types of Anaerobic Reactors

Anaerobic treatment of wastewater is carried out in several different reactor types. As shown in Figure 2, each of these reactors simply represents a different method of contacting wastewater with biomass. Listed below are four important immobilized reactor types of commercial significance. The anaerobic fluidized bed reactor, which is the topic of this study, is covered separately following this discussion.

2.3.1 Anaerobic Fixed Bed Reactors (Filter, DSFF)

In a fixed bed reactor, microorganisms are attached to inert media with a substantial amount of biomass in the form of suspended flocs trapped in the voids between the inert media. A suitable fixed film support should be chosen so that the requirement for a high specific surface area for biofilm growth and high void fraction to avoid plugging are balanced. Support media packing can have a random or regular orientation. Although random packing increases the separation of sludge from rising gas bubbles, it is more susceptible to clogging than packing in regular rigid vertically oriented channels. An optimum surface area to volume ratio in an anaerobic filter reactor with vertical media is

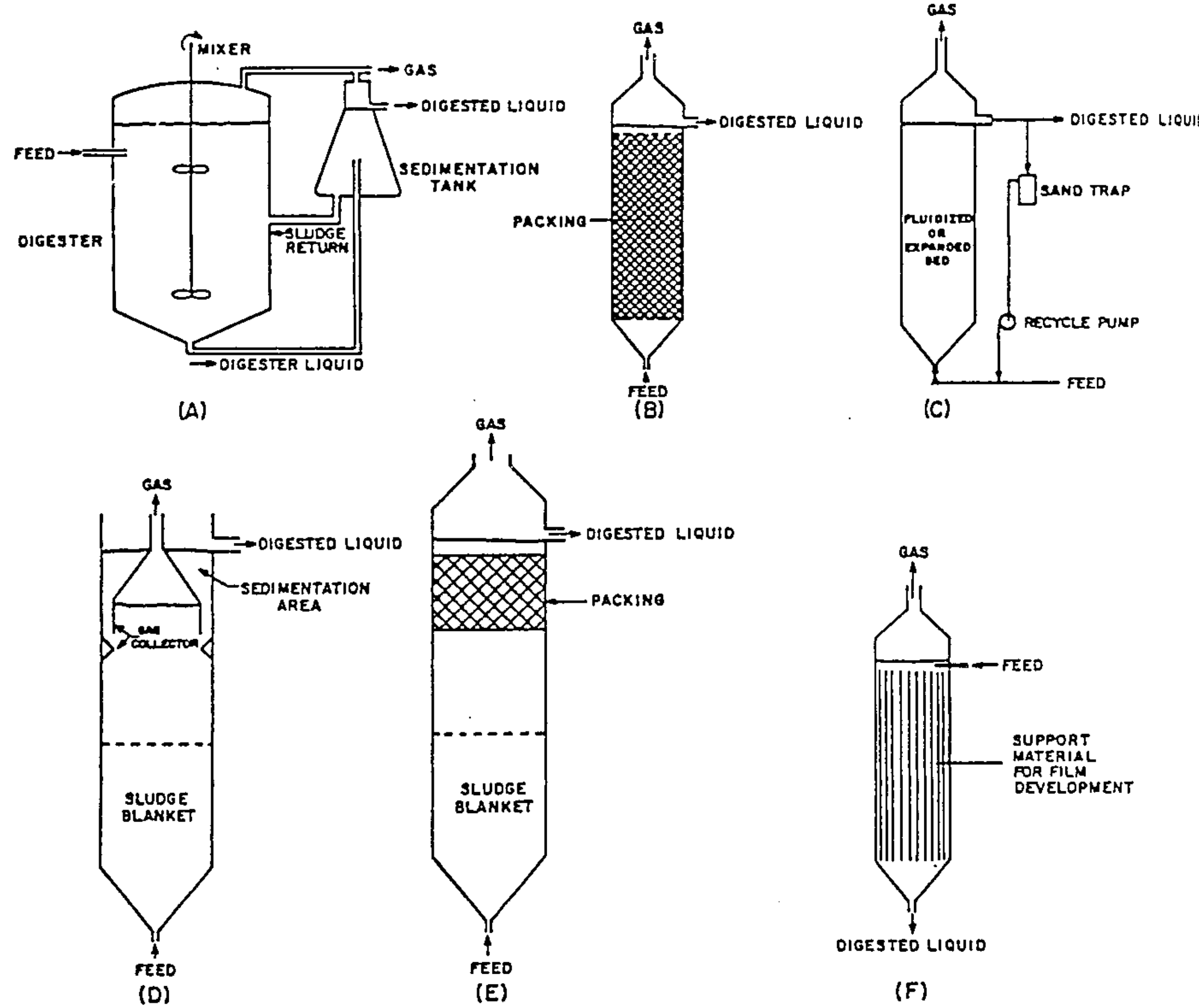

Figure 2. Types of Anaerobic Reactors: A) Anaerobic Contact, B) Upflow Anaerobic Filter, C) Fluidized Bed, D) UASB, E) UBF, F) DSFF

100-150 m^2/m^3 (van den Berg et al., 1985).

Wastewater flows upwards (or downwards) in the reactor, generally without recycle, resulting in plug-flow. When gas production is high, rising bubbles tend to mix the reactor fluid. Also, recycle may be employed to overcome problems related to wastewater toxicity or pH. In these situations reactors may deviate from ideal plug-flow towards a more mixed flow regime.

In the anaerobic filter good treatment efficiency at high or variable loading capacity is achieved without the need for specialized sludge separation (Rosen and Gunnarsson, 1986). However, high suspended solids in the feed or excessive biomass accumulations pose severe operating difficulties, leading to plugging and channelling and loss of active reactor volume. This problem can be minimized by restricting solids entry into the reactor, using vertically oriented media, designing for downflow operation so that suspended solids are flushed out, and by incorporating a means for backwashing to remove excess sludge.

2.3.2 Upflow Anaerobic Sludge Blanket

The upflow anaerobic sludge blanket (UASB) process was developed in The Netherlands, where it was first applied commercially for treatment of a beet-sugar waste. This type of reactor relies on the superior settling characteristics of biological flocs and granules instead of internal support structure to retain microbes within the reactor. Wastewater enters the reactor from inlet pipes at the bottom and flows upward through a bed of relatively dense, granular sludge, and a blanket of flocculated sludge particles.

Within these zones, much of the organic matter is converted to carbon dioxide and methane.

The amount of biomass retained within a reactor varies with the loading rate and with the form of the biomass. For UASB systems, OLRs are typically between 3-30 kg $BOD_5/m^3/d$ at 35 °C (Lee et al., 1989). Additionally, UASB reactors are generally best for treating low suspended solids wastewater since suspended solids tend to break up microbial granules resulting in sludge washout. Usually, no external agitation is employed for UASB reactors. When mixing due to rising gas bubbles and the flow of wastewater is insufficient, the reactor is operated with recycle. This enhances reactor control. In many commercial designs the upper part of the reactor is equipped with a gas-solids separator device which allows separation of entrapped or attached gas bubbles from the sludge and the return of the sludge particles to the bed below.

One of the critical elements in the proper operation of the process is the development and maintenance of granular sludge particles (0.5 - 2.5 mm diameter) that exhibit excellent settling properties (SVI values of less than 20 mL/g). Depending on the characteristics of the wastewater and the material used to "seed" the reactor, it may take several months to accumulate a sufficient mass of "pelletized" sludge. Supply of nutrients such as Ni and Fe, OLR, and calcium concentration are some factors which have been cited as affecting the development and properties of granules (de Zeeuw and Lettinga, 1981).

2.3.3 Hybrid Reactor

The hybrid reactor combines a sludge blanket and filter reactor to better accommodate elevated loading rates with high removal efficiencies while minimizing the limitations of each individual design. The lower section of this reactor consists of a granular sludge bed like the UASB reactor. Here the greatest fraction of COD reduction takes place due to the high densities and activities of the biomass (Rosen and Gunnarsson, 1986). However, suspended biomass systems require long acclimation periods and exhibit poor tolerance to shock loading (Geller and Gottsching, 1985).

By adding an inert film support structure to the upper portion of the reactor a fixed and active microbial film may be developed improving the total treatment capacity. The packed bed raised above the very dense sludge bed exhibits fewer of the hydraulic inefficiencies observed with a conventional filter reactor. The support matrix also acts as a gas-solid separator which aids the retention of poorly settling granular sludge. If the fixed bed is to be used primarily as a gas solid separator, it may be only 4 - 5% of the total reactor height (Kennedy et al., 1989). Commercially, there has been limited design and operating experience with hybrid reactors. Some key considerations are the optimal use of reactor volume and the cost of inert support media.

2.3.4 Expanded Bed Reactor

In an expanded bed reactor, microorganisms are attached to a large number of inert support particles ranging in diameter from 0.3 to 1.0 mm. These particles may be ones which are used in fluidized bed reactors such as sand, gravel, anthracite, or plastic although slightly heavier and/or larger particles may also be acceptable. Through the use of a recycle rate, sufficiently high vertical velocities are attained to expand the bed of biofilm covered particles. The bed expansion is maintained at a level at which the particles still remain in contact with each other and are not truly fluidized. Typical bed expansions of expanded bed reactors is 10 to 20% of the static bed height.

Unlike the filter reactor, in an expanded bed reactor the small particles are in constant motion. The attrition, caused by motion and physical contact of neighbouring particles, is responsible for much of the biofilm thickness control. Excess sludge may be removed from any place in the bed. Many designs incorporate some means of solids separation in order to prevent particles with thick films from being washed out of reactor. Like fluidized bed reactors, expanded bed reactors present a delicate hydraulic problem due to the need for uniform fluid distribution (Henze and Harremoes, 1983).

Due to the large surface area to volume ratio exceeding 1000 m^2/m^3 available for microbial growth, this reactor offers the advantage of high treatment capacity and low space requirements. It is also capable of treating dilute waste (200 mg/L) at removal efficiencies greater than 80% (Switzenbaum, 1978).

2.3.5 Reactor Comparisons

Several studies have been performed comparing the behaviour and treatment efficiencies of various anaerobic reactors under identical conditions. In particular, Kida et al. (1991) compared the performances of sludge recycling, sludge blanket, fluidized bed, and upflow filter reactors treating synthetic low strength wastewater (100-2000 mg/L

TOC) at a volumetric loading rate of 0.8 g/L/d. They found that the conventional sludge recycling process was not suitable for treating the low strength waste. The other three reactors successfully treated the wastewater with the fluidized bed and sludge blanket reactors removing as much as 80% of the organics for the influent TOC of 100 mg/L. The fluidized bed reactor was able to maintain high removal efficiencies even when the temperature was lowered to 15 °C and HRT was reduced to three hours.

Switzenbaum (1983) compared filter reactors to expanded bed reactors and concluded that the latter type of reactor design is more efficient for anaerobic fermentation reactions. His main reasons were the lower external mass transfer resistance and higher surface area to volume ratio in expanded bed reactors.

Studies on anaerobic treatment also show a fundamental difference between biofilms developed in expanded and fluidized bed reactors compared to filter reactors. In the former, biofilm are usually highly dense and active. Because the biofilm is relatively thin, it offers little diffusional resistance. Switzenbaum (1978) in his study of expanded bed reactors reported dense active biofilms with thicknesses not exceeding 15 microns. This is in contrast to biofilms of 140 to 1700 microns in thickness found in filter reactors, only a portion of which is utilized to degrade waste.

2.4 Anaerobic Fluidized Bed Reactor (AFBR)

The general advantages of biological fluidized bed reactors (FBR) were first realized for aerobic oxidation of organics and for denitrification. It was only in the mid 70s that the suitability of FBRs to anaerobic processes was realized. After several years of research and development, by the early 80s a few full scale AFBRs were built. The following sections discuss the basic principles of AFBRs followed by specific design and operating factors which must be evaluated prior to building full scale treatment systems.

2.4.1 Principles of Operation

In a fluidized bed reactor the biomass grows in the form of a biofilm around small inert support particles. These particles are maintained in the fluidized state by the upwards-directed flow of water. Typically, beds are expanded to 100% or more of the static bed height. The AFBR possesses many key advantages over other reactor types. Due to the high settling velocity of the carrier particles of around 50 m/h, high liquid velocities may be utilized. This prevents accumulation of inert sediments and the sludge activity remains high. Since the support particles used are small (0.1-1 mm diameter) and are present in large numbers, the surface area to volume ratio is very high (around 2000 m^2/m^3). Thus, because of the high concentration and activity of biomass in the reactor, high treatment capacity is obtained.

2.4.2 Design and Operating Considerations

Carrier Type:

Most of the work in the area of fluidized bed wastewater treatment has been performed using sand as the carrier particle. Few studies have been conducted comparing the performance of AFBRs using different support particles. Some of the characteristics of the particles which affect treatment include the shape, size, density, porosity, surface roughness and charge.

The density and size of a particle determine the superficial velocity required for fluidization and the associated pumping costs. Using light, small particles means lower operating cost during start-up. However, as the biofilm grows, some of the particles may become too light and be easily entrained in the upward moving fluid. Thus, operating conditions such as loading rate, temperature, waste strength should be carefully considered when selecting particle type.

Petrozzi et al. (1991) used various support particles including limestone, porous glass, quartz sand, activated carbon, and treated coal for anaerobic degradation of an acidic wastewater from a cellulose process. They found that porosity is not important but surface roughness is. They also recommended small, high density particles for high loading rates since lighter particles tend to develop thick biofilms due to low shear and are carried out of the reactor. Narayanan et al. (1991) using granular activated carbon in an expanded bed reactor to treat municipal wastewater achieved greater than 94% removal of all volatile organic compound except chloroform.

Development and testing of new carrier materials of low cost and with suitable surface characteristics and density is an ongoing research area with immediate commercial application. Preliminary investigations on Biolite material, supplied by Degremont, France, have demonstrated its potential usefulness in fluidized / expanded bed reactors, thus warranting this study of its application in treating wastewater.

One or two stage AFBR configuration:

Single stage anaerobic reactors provide one environment for all microbes used for the degradation of incoming substrate. However, acid forming bacteria are quite different from methanogens with respect to growth rates, nutritional requirements, physiology, pH optima, and their ability to withstand environmental stresses. Thus, separating these two bacterial groups and placing each in a different reactor which optimizes environmental conditions would seem to improve the efficiency of the process.

In suspended systems, stage separation is achieved by using low retention times in the acid reactor, thus ensuring that the slow growing methane forming bacteria are not retained to any significant extent. However in immobilized biomass systems such as AFBRs, the population of methanogens is reduced in the acid phase reactor by controlling the reactor pH at a level considerably lower than the 6.8 to 7.4 optima of the methanogens.

The biochemical reactions in a conventional single stage anaerobic reactor cannot be equated to the sum of the acidogenic and methanogenic stages of a two stage configuration. The different bacteria populations, environmental conditions and accumulations of intermediate products may lead to newly favoured metabolic pathways (Wiegant et al., 1986).

The overall performance of the two stage system is affected by the relative sizes of the two reactors. Factors such as wastewater composition, reactor pH and temperature, and variability of hydraulic and organic loading rates influence the selection of the acid stage to methane stage volume ratio. Some studies have concluded that the required sizes of acidification and methanation fluidized bed reactors are the same (Li et al., 1984; Heijnen et al., 1985).

Some of the advantages of two stages are higher activity of methanogenic sludge, easier pH control and less alkalinity consumption, better process stability with respect to shock loadings, easier biofilm thickness control and higher purification capacity and efficiency (Li et al., 1984, Heijnen et al., 1985, Sutton and Li, 1983, Li et al, 1982). For the treatment of wastewater containing de-proteinized whey, Sutton and Li (1983) reported that by using a two stage AFBR configuration instead of a single reactor, a 22% reduction in total reactor volume and 34% reduction in system power requirements are realized.

Some disadvantages of separating the acidification and methanation steps are the higher initial capital costs due to the need for two reactors along with the corresponding increase in instrumentation and process control required. Also, for wastewaters containing high levels of readily hydrolysable substrates, hydrogen production without its removal will lead to its accumulation in the acid reactor liquor. Elevated hydrogen concentrations have been shown to influence the VFA distribution, shifting it away from acetic acid towards the longer chained VFAs as a direct result of feedback inhibition of acetogenesis (McInerney and Bryant, 1981).

Biolayer thickness and control:

Biofilm thickness measurements have been made by some investigators for different loading conditions and bed expansions. The biofilm thickness varies from 60-200 μm for carrier particles of 0.3-0.6 mm and reactor conversions of about 40 kg COD m^{-3} d^{-1} (Heijnen et al., 1989). Switzenbaum (1978) found that higher organic loading produces thicker films.

Biofilm thickness increases due to the growth of existing film plus the attachment of suspended biomass. Loss of attached biomass occurs mechanically due to collisions with neighbouring particles and liquid shear. Toxic loadings or large sudden changes in temperature are known to result in the sloughing off of biofilms. The turbulence created by gas bubbles rising through the bed increase detachment rate. The two opposing processes of biomass attachment / growth and detachment eventually balance each other to reach a dynamic equilibrium. However, the biofilm thickness thus obtained may be sub-optimal and active biofilm control through the separation of biomass from carrier particles outside the reactor is practiced. Usually, particles are removed from the top portion of the bed where the biofilm is thickest. Control of biofilm thickness in this way leads to better process analysis and control.

In contrast to AFBRs, biofilm thickness control in other fixed film reactors, such as the filter and downflow stationary fixed film reactors, is only partially achieved by means of adjusting recycle rate and backwashing. Thus the ability of the operator to effectively control biofilm thickness in an AFBR represents a significant advantage over other fixed film reactors.

Temperature:

Temperature is a critical environmental condition affecting the metabolic rate of microorganisms. The substrate composition, concentration and the predominant population of microbes are the important factors which affect the temperature response of biological processes (Lettinga, 1979). Anaerobic digestion can be conducted at the mesophilic range (20 to 40 °C) or the thermophilic temperature range (40 to 75 °C).

Although the mesophilic optima of approximately 35 °C is generally agreed upon, the corresponding temperature for thermophilic treatment is not well defined.

For the treatment of whey in a AFBR, Hickey (1982) reported that temperature does not have a major effect on process performance when operating in the mesophilic range. He found that lowering the temperature from 35 °C to 24 °C resulted in only a 10% reduction in COD removal rates over the entire loading range of 15 to 36.8 kg COD/m^3/day examined. Studies by Switzenbaum (1978) at 10, 20, and 30 °C using an expanded bed reactor showed that efficiency decreases with decreasing temperature but the Arrhenius relationship is not followed because of the low activation energy dependence on temperature.

Thermophilic operation can be advantageous if the effluent is hot since cooling is then not required. Also, as cited above, reaction rates are higher at thermophilic conditions albeit not as high as chemical reaction theory would predict. Disadvantages of operating at higher temperatures include the requirement of a higher degree of process control (Lin et al., 1986), possibly higher heating costs, dearth of thermophilic seed sludge, and limited experience with full scale systems. Also, reactors operated at thermophilic conditions are not as stable as their mesophilic counterparts because the consortia of thermophilic bacterial species is not as large and thus cannot tolerate large sudden changes in process conditions such as shock loadings.

pH:

Reactor pH is a powerful selective agent in determining which bacterial species predominate from a mixed culture inoculum. However, for many industrial effluents, the use of buffering chemicals for pH control is the main cost for anaerobic treatment (Prong and Chmelauskas, 1988). Thus, in order to establish a viable treatment system, the reactor pH must be selected to maximize biochemical conversions at minimum cost.

Non-methanogenic bacteria are not as sensitive as the methane formers and can function over a wide pH range of 4.0 to 8.5. Consequently, for single phase reactors, in which acidification and methanation reactions are carried out in one reactor, pH is usually maintained in the 6.4 to 7.6 range optimum for methanogens (Price, 1985).

For a two phase configuration, the acidification reactor pH is best controlled between 5.7 and 6.0 as this leads to high and stable butyric acid production rates (Kissalita et al., 1987). Butyric acid formation is regarded as highly favourable since its conversion to acetate, which is the primary substrate of methane forming bacteria, is very rapid. Running the reactor at much lower pH to save buffering chemicals results in the depression of acidogenesis. Undissociated acids, which predominate at lower pHs are presumed to easily penetrate the bacterial cell membranes inhibiting their metabolism (Neal et al., 1965).

Nutrients:

Anaerobic wastewater treatment processes require that a number of nutrients be biologically available. The deficiency of any single nutrient can be rate limiting. The macro-nutrients for bacterial growth and metabolism are a source of organic carbon, nitrogen and phosphorus. To meet this nutritional requirement, the commonly followed practice in AFBRs is to adjust the COD:N:P ratio of the feed to 300:5:1.

Micro-nutrients such as sodium, potassium, ammonia, calcium, magnesium and other metals are required to compliment macro-nutrients. These cations, at levels of one tenth to one half of their inhibiting concentrations help stimulate microbial metabolism. Since anaerobic treatment is not growth associated, determining nutritional requirements from elemental analysis of bacteria is unreliable. Also since nutrients may be adsorbed onto cells or accumulate within them, the difference in added nutrient levels in the influent and effluent is not very useful. Thus nutrient levels which stimulate biochemical conversions but are not present in unacceptable concentrations in the effluent are deemed important.

Start-up:

Little information on start-up procedure for AFBRs is found in the available literature despite of its being mentioned as problematic by some researchers. The problem lies in developing well attached biofilms on the carrier particles. Due to the difficulty in comparing studies using different methods of start-up because of differences in feed, loading rate, and type of carrier particle only some general trends may be observed.

Carrier conditioning has been investigated by some researchers as a method of hastening start-up. Stronach et al. (1987) found that synthetic polymer addition (millfloc C_{30}, C_{50}) enhanced biomass retention but did not improve process performance. It appears that carrier conditioning is neither practical nor particularly successful (Heijnen et al., 1989).

There are several different approaches that may be used to inoculate the AFBR with microorganisms. The inoculant can consist of reactor supernatant or sludge. It may be added only once initially, several times in batch mode, or continuously with the feed.

Batch or continuous inoculation lead to faster start-up since the rate of attachment of biomass increases with higher suspended biomass concentrations. However, single inoculation may be more practical for full scale reactors.

Two types of loading rate profiles emerge from the data: the maximum efficiency profile (MEP) and the maximum load profile (MLP). In the former, a low COD load (0.5 to 1 g/L/d) is applied in the beginning and is increased stepwise when treatment efficiency is maximum. In contrast, the maximum load profile calls for a high COD load from the start which is stepwise increased while the VFAs are still high and conversion efficiency still low. Reactor pH level may be used to control the rate of loading increase or extra alkali may be added in order to prevent the pH from falling below 6.7. Heijnen et al. (1989) after reviewing several studies conclude that the maximum load profile results in faster start-up. This conclusion is supported by Yen and Shiek (1992) who used the above start-up procedures with acetic acid substrate and Manville R-633 support beads.

Using an easily degradable and non-toxic substrate (such as methanol) during startup will result in quicker biofilm formation but the actual waste should be introduced gradually to prevent sloughing off of biofilm due to shock. Heijnen et al. (1989) have stated that for complete biolayer formation a short liquid residence time is better. At short HRTs, the suspended biomass is washed out and biomass in the reactor is predominantly in the form of a biofilm. In the case where the reactor is only inoculated once, the reactor should be run on recycle alone for several days to allow enough time for attachment.

For a new carrier material such as Biolite, it is important to determine the optimum startup procedure which will ensure a short startup period with stable reactor operation.

This requires comparing the rate of attached biomass accumulation, COD removal and stability using loading profiles such as MEP and MLP. Such studies also serve to add to the limited body of information on startup of anaerobic reactors.

Reactor Biomass:

In the AFBR, biomass is found either in attached or suspended form. With short hydraulic retention times (HRTs), biomass grows in a completely attached mode since the dispersed growth is washed out (Heijnen, 1984). From Table 1 it appears that biomass concentrations of up to 40 kg/m^3 can be obtained in AFBRs due to the large surface area for biofilm attachment and high biomass concentration in the biolayer of up to 150 kg VSS/m^3 (Switzenbaum and Jewell, 1980). The biomass activity is about 0.7 g COD/ g VSS/d (35 °C) for the first stage sludge and about 2 g COD/g VSS/d (35 °C) for second stage sludge fed with fatty acids (Shieh et al., 1985). From these data it is clear that very high volumetric stabilization capacities are possible in fluidized bed reactors.

During start-up using the maximum load profile, several interesting observations have been made regarding the development of microbial populations. In acidification reactors at 1 to 3 h HRT its has been observed that within days biofilms form containing mainly acidogens. Within a week, sulphur reducing bacteria begin to colonize the biolayer and after about one month of operation methanogens begin to appear in significant numbers (Heijnen, 1983a). In methanogenic reactors, fed with VFAs, initially only acetic acid and butyric acid are degraded and sludge activity is mainly due to Methanosarcina type microbes. However, at the end of the start-up period acetate levels drop below 200-500 mg/l and Methanosarcina are replaced by microbe resembling Methanotrix (Heijnen et al., 1985).

2.4.3 COD Removal in AFBRs

Table 1 lists some of the studies conducted over the last 15 years on the anaerobic treatment of wastewaters using fluidized bed reactors. The wastes vary from easily degradable glucose to fairly recalcitrant cellulosic waste to toxic phenolic waste. Wastewater concentration and reactor HRT used were in the ranges of 0.5-12 g COD/L and 0.3-37 h respectively. Many researchers have obtained high removal rates of over 90%.

It can be seen from Table 1 that the vast majority of work in the area of AFBR treatment has focused on sand as the carrier particle. This is due to the availability and low cost of this material. However, new materials such as Biolite have the advantage of being less dense thus requiring lower superficial velocities to fluidize. This translates to lower pumping costs. Also Biolite is presumed to have better surface properties which may lead to faster startup, higher removal rates and more reliable and stable operation.

Table 1. Summary of Anaerobic Fluidized Bed Treatment

Reference	Substrate Type	Carrier			Operating Conditions			Bio-mass (g/L)	%COD Rem.
		Type	d (mm)	V_{sup} (m/h)	T (°C)	HRT^{FB} (h)	COD (g/L)		
Cheng & Chian (1982)	phenol	act. carbon	_	13.3	35	24	4	_	90
Hall (1982)	brewery waste	sand	0.4	_	35	15	12	5	95
Sutton et al. (1982)	soy waste	sand	0.5	31	35	20	12	_	60
Heijnen (1983b)	yeast water	sand	0.1-0.3	7-12	37	1	2.5	40	>90
Barnes et al. (1983)	molasses	sand	_	_	36	8	4	_	90
Elmaleh et al. (1984)	acetate/ isobutanol	sand	0.4	34	35	0.3-8	1-8	_	>90
Biver (1984)	glycerin	sand	0.4	13	20	4	1.4	40	95
Bull et al. (1984)	acidified glucose	sand	0.2	4.6-7.6	35	20	12	5-20	80
Biver (1984)	whey permeate	sand	0.2	5.4	35	2.4	3	27	70
Chen et al. (1985)	glucose	carbon	0.6	_	35	0.5-8	0.5-9	30	90
Petrozzi et al. (1991)	cellulose waste	6 types	0.2-0.6	_	37	2-12	4.5-7	_	26-69
Barascud et al. (1992)	papermill waste (synthetic)	Biolite	_	_	35	7-37	8.4	_	65-95

2.5 Landfill Leachate

Landfilling is the most common form of disposal for a wide variety of municipal, commercial and industrial wastes, both hazardous and non-hazardous. For example about 90% of all domestic and commercial solid wastes generated in the United Kingdom are disposed of by landfill methods (Harrington and Maris, 1986). In the United States, about 80% of the municipal solid waste produced is landfilled, while 10% is recycled and 10% is incinerated. A major environmental problem which occurs at landfills is the generation of leachate. Leachate is the combined organic and inorganic wastewater which is reproduced when snow, rain and other liquids seep through solid and liquid wastes deposited in landfills. In order to prevent groundwater and surface water pollution, leachate must be contained, collected and treated before discharge.

2.5.1 Characterization of Leachate

Due to the complex and variable nature of leachate, the treatment system installed at a landfill site is generally tailored to that particular site and specific leachate characteristics. Therefore, to design an efficient and cost-effective leachate treatment plant in compliance with all prevailing environmental regulations, it is important to characterize leachate composition as thoroughly as possible, to establish accurate leachate flows, and identify the most appropriate treatment options through carefully conducted bench-scale and/or pilot plant treatability studies.

Leachate characteristics of particular interest include five-day biochemical oxygen demand (BOD_5), COD, the BOD_5 / COD ratio, molecular weight and/or size of the organic molecules, VFAs, ammonia and metals. Forgie (1988) conclude that biological leachate treatment is most appropriate when the BOD_5 /COD ratio is high (ie. >0.4) and the molecular weight of the majority of the organics is less than 500 g/mol.

Generally, raw sanitary landfill leachate contains high levels of BOD, COD, suspended and dissolved solids, is coloured and odoriferous. The distinct odour of the leachate is due to the presence of free volatile acids and reduced sulphur compounds (Harrington and Mavis, 1986; Venkataramani et al., 1984). The metals mostly present are iron, manganese, zinc and copper. The organic constituents in sanitary landfill leachate are largely a function of landfill age. For relatively unstable or young landfills, up to 90% of the soluble organic carbon can be accounted for as short-chain volatile fatty acids. For stabilized landfills VFAs tend to decrease over time and such landfills also contain relatively high concentrations of ammonia-nitrogen (Lu et al., 1981). It is common to speak of young, acid-phase leachate and old, methanogenic leachate as well as high, medium, and low strength leachate. Henry et al. (1987) suggested a BOD_5 / COD ratio of 0.7 for a "young" leachate, 0.5 for "mature", 0.3 for "aging" and 0.1 for "old". A typical time period for the transition from young to old leachate-types after the start-up of the landfill is from 6 to 10 years (Chain and DeWalle, 1976) but can be shorter, e.g. 2 years, in specific cases (Harmsen, 1983).

Leachate quality also depends on factors such as landfill depth, whether or not leachate is recycled, and the yearly season (Ehrig, 1984). In general, deeper landfills produce leachate with higher concentrations of organics and other contaminants since the water must travel through a longer path, picking up contaminants, before it is collected. Recycling of leachate, or the redistribution of leachate back to the top of the landfill, leads to a generation of a more uniform, lower strength leachate (Leckie et al. 1979). Considerable seasonal variations of leachate strength occur; in places with snow cover, leachate collected during spring is fairly dilute due to thawing of snow.

Industrial landfill leachate resemble very high strength industrial wastewaters and are much more complex than sanitary landfill leachate. Moreover, unlike sanitary landfill leachate, industrial landfill leachate is usually not composed of easily metabolizable substrate (Venkataramani et al., 1984).

As mentioned earlier, the leachate treatment process is selected based on the age of the landfill and characteristics of leachate. Bull et al. (1983), Ehrig (1984), Maris et al. (1984) and Harrington and Maris (1986) have reviewed the various biological treatment options. In general, it was concluded both anaerobic and aerobic biological processes can be used to treat all sanitary leachate. By using anaerobic treatment followed by aerobic polishing, high strength leachate may be treated to levels premitting direct discharge into receiving waters.

2.5.2 Anaerobic Treatment of Landfill Leachate

Pretreatment:

They are three main reasons for pretreating leachate. Firstly, many leachates are deficient in important nutrients such as nitrogen and phosphorus. Chain and DeWalle (1977) reported that the COD:N:P ratio should be adjusted to 500:13.5:1 prior to anaerobic treatment to ensure that nitrogen and phosphorus are not limiting. More recently, many researchers have successfully treated leachate with much lower additions of nutrients (COD:P=1000:1) (Thirumurthi et al., 1990) or with no additions at all (Wu et al., 1988; Gourdon et al., 1989). Thirumurthi (1990), in a study using six filter reactors, found that the minimum concentration of total P required to support anaerobic decomposition is 0.7 mg/l and the maximum ratio of COD/P that can be applied was 30,300:1. Although the addition of nutrients represents an additional cost associated with biological treatment; compared to aerobic systems (COD:N:P ratio of 100:5:1), the nutritional requirements of anaerobic systems are very low.

A second reason for pretreatment is to eliminate possible toxicity due to heavy metals. Iza et al. (1992) strongly recommend removal of heavy metals prior to leachate treatment. This has been achieved in laboratory scale studies by addition of lime slurry followed by flocculation and settling. Finally, pretreatment is used to increase pH of leachate from acidic to near neutral values. Chang (1989) treated municipal landfill leachate in a sludge bed filter reactor. After comparing removal efficiencies of runs with leachate pH adjusted to 6.4-7.0 using sodium bicarbonate to runs treating raw leachate (pH of 5.5), he concluded that no buffer addition is necessary. However, most researchers do add base to their leachate to ensure optimal conditions for methanogenic bacterial population.

Biomass:

Some researchers using different anaerobic reactors have shown that biomass concentration increases with organic loading rate (Chang, 1988; Lin, 1991). This behaviour is expected from reactors treating ideal waste and indicates that under the experimental conditions of these investigations, the effect of inhibitory substances present in the leachate was small. Chang (1988) reports a net solids production of 0.041g VSS/g COD removed for leachate treated in a sludge bed filter. Surprisingly, this value is high compared to previously reported values of 0.027 g VSS/g COD (Guiot and van den Berg, 1985) for a sugar waste using a similar reactor and 0.012 g VSS/g COD (Chain and DeWalle, 1977) for high strength acidic wastewater using a completely mixed filter. Using a conventional anaerobic digester fed semi-continuously with municipal landfill leachate, Lin (1991) found a yield of 0.08 g VSS/g COD and an endogenous decay coefficient of 0.15 L/d. The above biomass yields indicate that solids production in an anaerobic reactor treating leachate is comparable to similar feed types.

Toxicity:

Most investigators have found that at concentration levels present in leachate, iron and zinc do not inhibit anaerobic degradation of organic material (Kirsh and Sykes, 1971; Iza et al., 1992; Wu et al., 1988). This is probably due to the low redox potentials in anaerobic reactors which cause precipitation of heavy metals as metal sulphides and insoluble carbonate species. Iza et al. (1992) noted dense gritty sludge at the bottom and hollow granules at the top of their UASB reactor. They quantify the "mineralization of sludge" by the ratio of inert solids to total solids. Similar built-up of precipitates leading to reactor clogging of fixed film reactors has been reported by Kennedy et al. (1988) and Mennerich (1987).

Gas Production:

Gas production from anaerobic reactors treating leachate is generally seen to rise linearly with organic loading. Chang (1988) reports an average methane production rate of 0.31 L CH_4(STP)/g COD removed which represents 94.7% of the maximum theoretically possible. However, he found that methane content and energy recovery is affected by sulphate loading. At a level of 683 mg/L/d SO_4^{2-}, energy recovery drops down to 52% with 66.9% of the gas being methane (Chang, 1988). This observation may be attributed to the competition between sulphate reducing bacteria and methanogenic bacteria for electrons.

COD Removal:

Boyle and Ham (1974) were among the first to investigate the anaerobic treatment of leachate. Since then, several other researchers have conducted their own studies, most with a high degree of success. Table 2 is a summary of the more recent work done.

Table 2. Summary of Reported Data on Anaerobic Leachate Treatment

Reference	System Type	BOD_5/COD	Loading Criteria	Detention Time	Removal	Additional Comments
Carter et al. (1984)	Anaerobic Filter	0.58	8.02 kg $BOD/m^3/d$	5.5 d	80 to 90% Organics	
Henry et al. (1985)	Anaerobic Filter	0.3 to 0.7	<2 kg $COD/m^3/d$	0.5 to 4 d	>80% COD	longer times are better
Maris et al. (1985)	UASB	0.67	3.6 to 19.7 kg $COD/m^3/d$	1.0 to 4.3 d	>85% BOD_5 >82% COD	heated to 29 °C
Muthukrishanan & Atwater (1985)	Anaerobic Filter	0.48 to 0.85	2.0 to 4.2 kg $COD/m^3/d$	1 d	50 to 85% COD	
Wright et al. (1985)	Anaerobic Filter	0.7	4 kg $COD/m^3/d$	5 d	92% COD 97% VFA	heated to 35 °C
Schafer et al. (1986)	Anaerobic Filter	0.76	0.72 kg $BOD/m^3/d$	>4.9 d	95% BOD_5	
Thirumurthi et al. (1986)	Anaerobic Filter	0.76	1.6 to 2.0 kg $COD/m^3/d$	12.5 to 14.9 d	96 to 99%	plastic media
Henry et al. (1987)	Anaerobic Filter	0.3 to 0.7	1 to 2 kg $COD/m^3/d$	24 to 96 h	90% COD	21 to 25 °C
Thirumurthi & Groskopf (1988)	Anaerobic Filter	0.53	1.39 kg $COD/m^3/d$	13.9 d	>97% COD	
Chang (1988)	SBF	0.68	21.8 kg $COD/m^3/d$	2.7 d	68% COD	no pH adjustment
Kennedy et al. (1988)	SBF	0.6	14.7 to 44.0 kg $COD/m^3/d$	36 to 10 h	88 to 97% COD	pretreated with lime
Kennedy et al. (1988)	DSFF	0.6	14.7 kg $COD/m^3/d$	36 h	94% COD	pretreated with lime
Lin (1991)	Activated Sludge	0.69	1.1 to 2.8 $COD/m^3/d$	8 to 20 d	93 to 95 % COD	semi-continuous
Iza et al. (1992)	SBF	0.56	<10 kg $COD/m^3/d$	18 h to 1 d	65 to 90% COD	pilot plant study
Schroeder et al. (	Expanded Bed	0.2 to 0.6	4 to 15 kg COD/ m^3/d	6 h	26 to 82% COD	spiked with synthetic organics

CHAPTER 3. THEORY

3.1 Overview

Section 2 describes principles of fluidization

Section 3 presents the Monod model for microbial processes

Section 4 very briefly covers the theory behind the adhesion and subsequent growth of bacteria on solid surfaces.

3.2 Fluidization

If a liquid or a gas is passed at low velocities through a porous bed of solid particles, the particles do not move. The fluid passes through the small, tortuous paths losing pressure energy. If the velocity of the fluid is steadily increased, the pressure drop of passing fluid increases as shown in Figure 3 (McCabe and Smith, 1976) until the pressure drop equals the force of gravity on the particles. At this point (point A on the graph) the particles begin to move, the bed porosity increases and the pressure drop rises more slowly than before. At point B the bed is in its loosest possible state with the particles still in contact. At fluid velocities higher than at point B, the bed is truly fluidized with the particles separating from each other. Beyond point F, the particles move more and more vigorously in random directions and the bed is referred to as a boiling bed. When point P is reached, entrainment becomes appreciable. At point Q, the porosity approaches unity and entrainment is complete. This process is called disperse-phase or

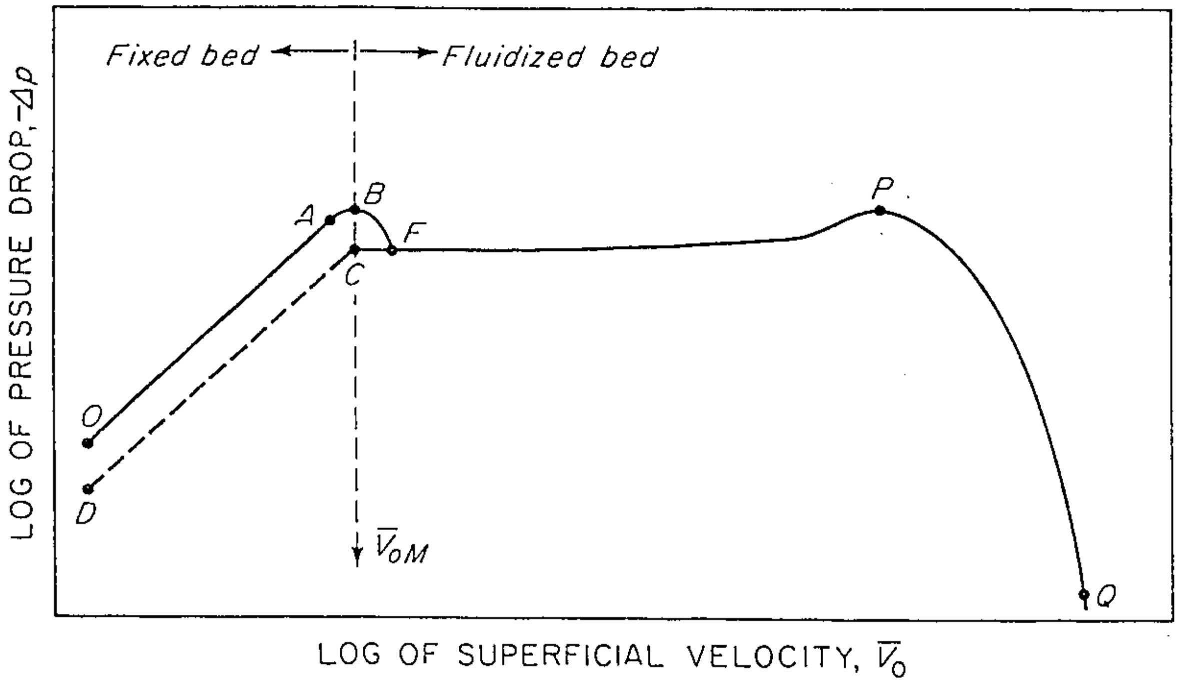

Figure 3. Pressure Drops in Fluidized Beds

lean-phase fluidization with pneumatic transport. Fluidization without solid entrainment is known as dense-phase fluidization.

Dense phase fluidization with a liquid occurs in the mode of action called particulate fluidization. Here particles move as individuals throughout the bed, although strong transient currents may cause many particles to move temporarily in one direction. This is contrast with aggregative fluidization, seen when the fluid is a gas.

3.2.1 Deviations from Ideal Behaviour - Channelling

Fluidization of solid particles by a liquid generally gives rise to a particulate behaviour which may be described as homogeneous. Channelling, which is characterized by locally organized movement of solids, represents an important deviation from ideal behavior. Upward and downward motions in different regions of the bed are connected by horizontal flows near the top and bottom regions of the bed. This pattern distinctly defers from the ordinary slow erratic motion of particles in liquid fluidization.

It has been clearly demonstrated that maldistribution of the liquid at the base of the bed gives rise to channelling. Liquid velocities are important at the base. Where the velocity is high, solids develop an upwards movement; where it is low, solids move downwards. Channelling can be avoided by using a homogenizing section and distributor which produces a uniform velocity profile. The homogenizing section generally consists of a fixed section right below the bed, filled with solid packing. Current full scale fluidized bed reactors treating wastewater utilize complex distributing systems in which liquid entering the bottom of the reactor is deflected off walls of appropriate geometries leading to a flat velocity profile (Dorr-Oliver, 1980). This type of design is more resistant to clogging by suspended matter in the wastewater.

3.3 Film Models

Based on fundamental theory, several attempts have been made to mathematically describe the microbial process in order to obtain a rational design procedure for film processes. Theoretically, a system is either described as homogenous or heterogeneous. A homogenous system is characterized by the absence of mass transfer resistances. Suspended systems in which individual and small clusters of microbes are completely dispersed throughout the reactor volume may be approximated as homogenous systems. In a heterogeneous system, the microbes are considered to be in a separate phase than the liquid containing the substrate. Between these two phases exists a definite interface. This is true for systems which use inert media to support biomass growing in the form of a film.

When analyzing the factors that affect the rate of substrate utilization by biological films, it is convenient to simplify the overall process into three major steps:

(1) Diffusion of substrate from the bulk of the liquid to the interface between the liquid and the biofilm.

(2) Diffusion of substrate within the biofilm.

(3) Biochemical reaction (substrate conversion) within the biofilm.

In order to adequately model the substrate uptake reaction by biological films, a proper understanding of each of these processes is required.

If an AFBR is operated with high fluid velocities and liquid substrate concentrations then external and internal diffusional resistances are negligible. Kennedy and Droste (1987) have shown that in downflow stationary fixed film reactors, biofilms up to 2.6 mm thickness are completely penetrated by substrate and do not exhibit any diffusional limitation. They attributed this to a lack of need for an electron acceptor, high saturation constant values of anaerobic bacteria, and micro-mixing in the biofilm caused by gas evolution. Such systems, where diffusion is not rate limiting may be considered as homogenous. For simplicity, this will be assumed in this study.

The following empirical steady state model development is based on the assumptions below.

1. The rate limiting substrate is the soluble organics as determined by COD analysis.
2. Substrate completely penetrates the biofilm and no substrate gradient exists across the biofilm depth.
3. Biofilm accumulation is small and can be neglected.
4. Concentration of nonbiodegradable organic matter in the reactors is small and can be neglected.
5. Reactor is completed mixed in terms of soluble substrate.
6. Influent to the reactor is totally soluble and contains no biomass.

Based on the above assumptions a mass balance for substrate in the reactor gives

$$\frac{dS}{dt} = \frac{Q}{V}(S_0-S) - \frac{\mu X}{Y} \qquad (1)$$

where Q is the influent volumetric flowrate, V is the reactor volume, S and S_0 are the mixed liquor and influent substrate concentrations respectively, Y is the biomass yield, X is the concentration of biomass in the reactor and μ is the specific growth rate of microbes. A mass balance for the microorganisms can be expressed as

$$\frac{dX}{dt} = (\mu - b)X - \frac{Q}{V} X_L \qquad \textbf{(2)}$$

where X_L is the suspended biomass present in mixed liquor and b is the decay coefficient. The decay coefficient includes the effects of endogenous metabolism and microbial death.

At steady state, equations 10 and 11 are written as

$$X = Y\,(S_0 - S)\,\frac{Q}{V\mu} \qquad \textbf{(12)}$$

$$\mu = \frac{QX_L}{VX} + b \qquad \textbf{(13)}$$

Defining the SRT or sludge age as

$$\theta_S = \frac{VX}{QX_L} \qquad \textbf{(5)}$$

equation 13 can be written as follows

$$\mu = \frac{1}{\theta_S} + b \qquad \textbf{(6)}$$

The term $1/\theta_S$ is the net specific growth rate (μ').

The specific growth rate of microorganisms μ in a homogenous system may be described by the empirical equation developed by Monod (1949).

$$\mu = \frac{\mu_m S}{K_s + S} \qquad (7)$$

Here K_s is the saturation or half velocity constant and μ_m is the maximum specific growth rate. K_s determines how rapidly the curve approaches μ_m and is defined as the substrate concentration at which μ is equal to half of μ_m. Although this equation was originally proposed for continuously fed, completely mixed, suspended pure cultures, it is widely accepted for describing the extrinsic kinetics of many types of biological reactors.

The specific substrate utilization rate U is related to the specific biomass growth rate by the yield Y.

$$U = \frac{\mu}{Y} = \frac{kS}{K_s + S} \qquad \textbf{(8)}$$

Here, k is the maximum substrate utilization rate.

One way to convert the Monod equation into a linear form is using the Hanes method (Grady and Lim, 1980).

$$\frac{S}{U} = \left(\frac{1}{k}\right)S + \frac{K_s}{k} \qquad \textbf{(18)}$$

Plotting S/U vs. S results in a straight line with slope 1/k and y-intercept of K_s/k. Thus the values of the kinetic constants K_s and k may be found.

The washout factor (f) relates the HRT to the SRT as given by the equation below.

$$\mu' = \frac{1}{\theta_S} = \frac{f}{\theta_H} \qquad \textbf{(10)}$$

Combining and rearranging equations 13, 15, 16, and 19, COD removal efficiency, E, and reactor biomass concentration, X, can be predicted as a function of HRT using the following equations.

$$E = [1 - \frac{K_S (f / \theta_H + b)}{S_0 (\mu_m - f / \theta_H - b)}] \, 100 \qquad \textbf{(20)}$$

$$X = \frac{Y (S_0 - S)}{(f / \theta_H + b) \, \theta_H} \qquad \textbf{(21)}$$

Equation 21 describes the total biomass concentration in the reactor. Biofilm biomass can be determined by subtracting biomass in the mixed liquor.

3.4 Biofilm Development and Adhesion

As illustrated in Figure 4, biofilm development is composed of the following processes (Bryers, 1980): (1) surface adsorption of organic macromolecules, (2) transport of microorganisms to the surface, (3) microorganism attachment, (4) growth of attached microorganisms and surrounding polymer matrix and (5) detachment of biofilm.

Several investigators (Loeb and Neihof, 1973; Baier et al., 1968) have observed the formation of an organic layer prior to cell adhesion, which they believe shifts the critical surface tension toward a range more compatible for bio-adhesion. Zobell (1943) suggested that bacterial attachment occurs in two steps: (1) cellular surface adhesion and (2) permanent attachment through the aid of extracellular binding materials. Biofilm photomicrographs show that films consist of microorganisms entrapped within a gelatinous matrix or glycocalyx; the volume of microbial cells being a small fraction of the total biofilm volume. Cellular reproduction and exopolysaccharide production lead to biofilm growth. Most of the work on biofilm growth rates has focused upon kinetics of limiting nutrient removal. Increasing the carbon to nitrogen ratio causes the microbes to produce more "slime" (Characklis, 1970). Bryers and Characklis (1979) suggest that biofilm detachment rate is indirectly proportional to biofilm thickness as well as directly proportional to shear stress. Their predictions have correlated well with experimental results. Biofilm development is the net result of the above processes occurring simultaneously, some predominating others at certain stages.

Attached microorganisms have several advantages compared to suspended microorganisms. Due to their attached state, they are more persistent and are not readily

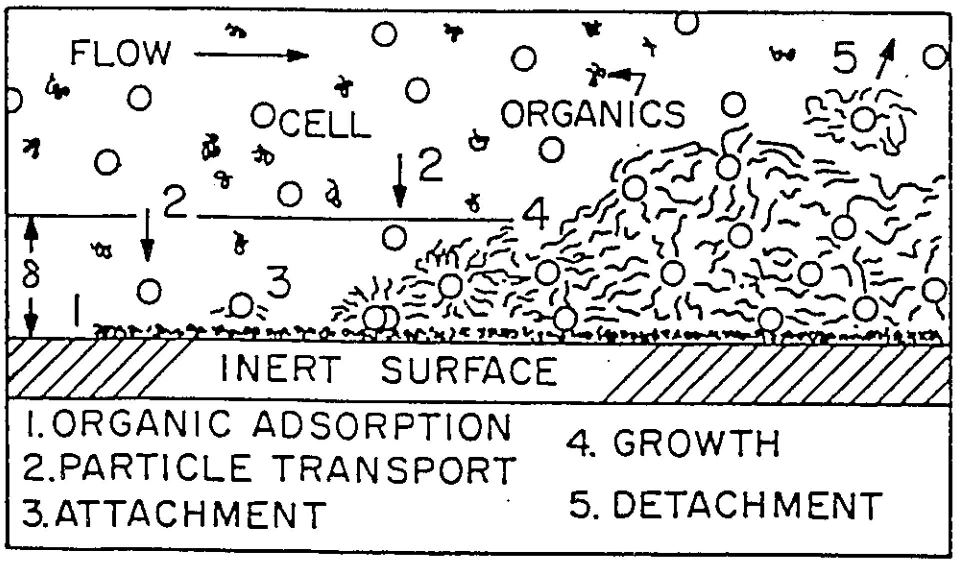

Figure 4. Biofilm Development Process

washed out. This allows for high biomass accumulation in fixed film reactors, leading to high efficiency treatment of wastewater. Addition of solid surfaces also facilitates bacterial growth. Additionally, higher metabolic activity of attached bacteria has been attributed to more favourable microenvironment near the solid surface. Studies have shown that bacterial films are more resistant to toxins than their floating counterparts. It takes 500 times more antibiotics to kill biofilm bacterium as it does planktonic bacterium (Costerton et al., 1985). Kennedy et al. (1987) showed that anaerobic DSFF reactors were not inhibited by a 24 h dose of 60 mg/L oxytetracycline. The anionic property of the polysaccharide matrix is believed to inhibit transport of the toxic compounds.

CHAPTER 4. MATERIALS AND METHODS

4.1 Overview

Section 2 describes in detail the materials of construction and design of the AFBR used in this study and outlines properties of Biolite carrier particles.

Section 3 presents the procedures used to collect data during the startup, steady state, and batch operation of the AFBRs including the composition of synthetic and leachate feeds used.

Section 4 discusses the various analytical techniques employed during the course of this investigation.

4.2 Apparatus

As shown in Figure 5, three laboratory scale anaerobic fluidized bed reactors were constructed for use in the experimental study. A schematic diagram of the AFBR system is shown in Figure 6. Each glass reactor had an outside diameter of 6.0 cm and was 75.0 cm tall. The inside diameter of the reactor column was 5.6 cm and the total empty bed reactor volume to the level of the liquid recycle line was 1.3 l.

As shown in Figure 7, the reactors were made completely air tight by means of a glass lid in conjunction with a ring clamp, Teflon "O" ring and silicon grease. A slight positive pressure was maintained inside the reactors which could be checked visually from the U-tube at the effluent exit.

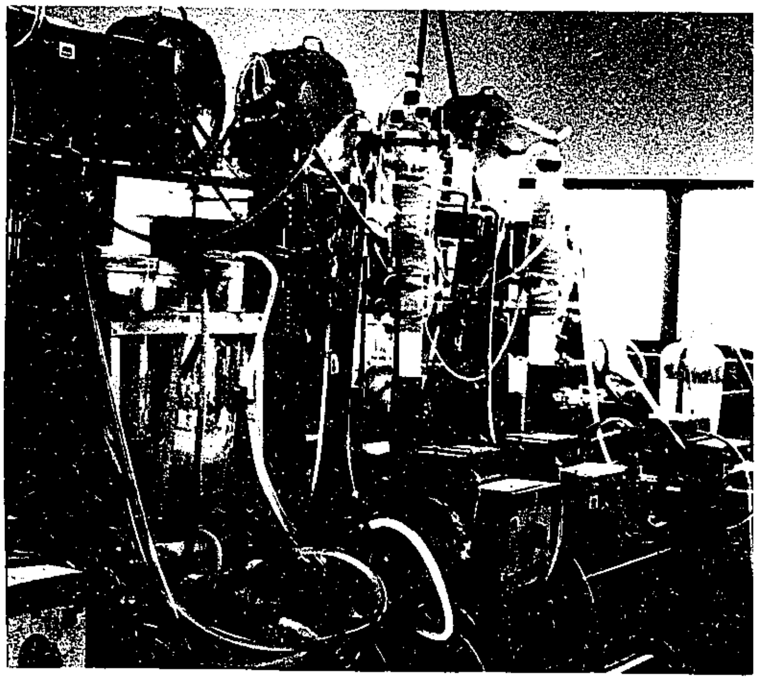

Figure 5. Photograph of three AFBR Systems used in this study

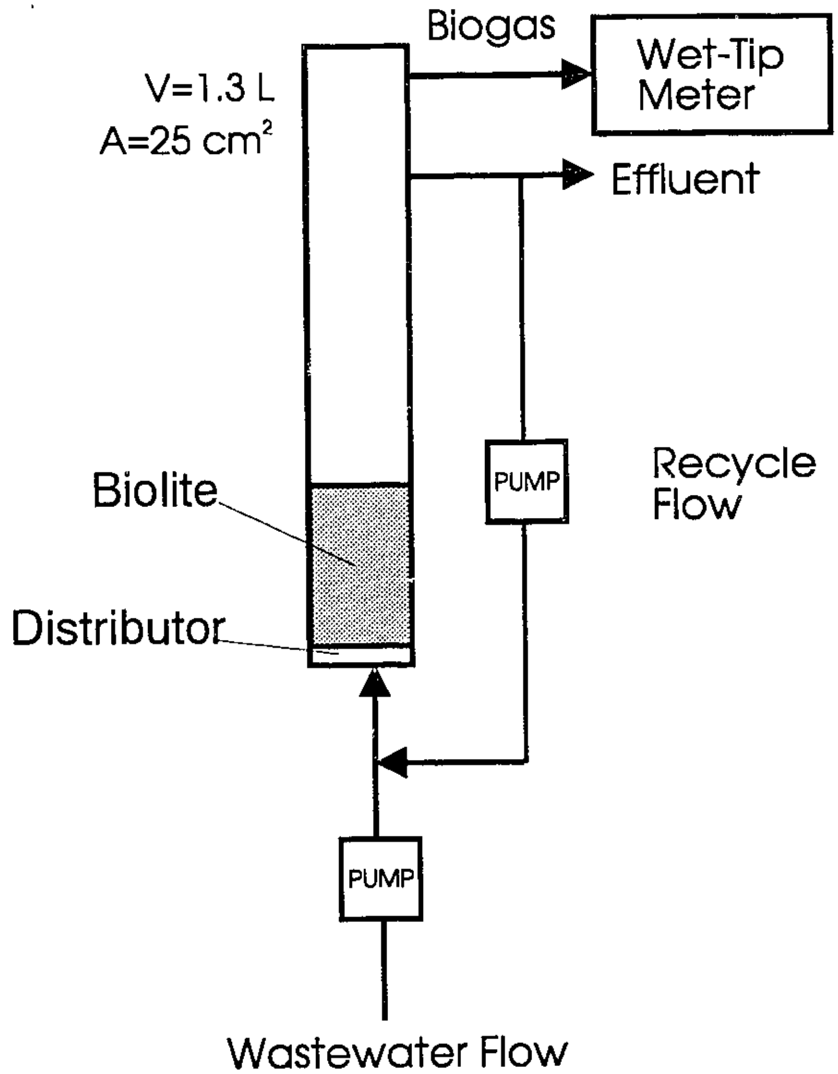

Figure 6. Schematic diagram of the AFBR system used in lab study

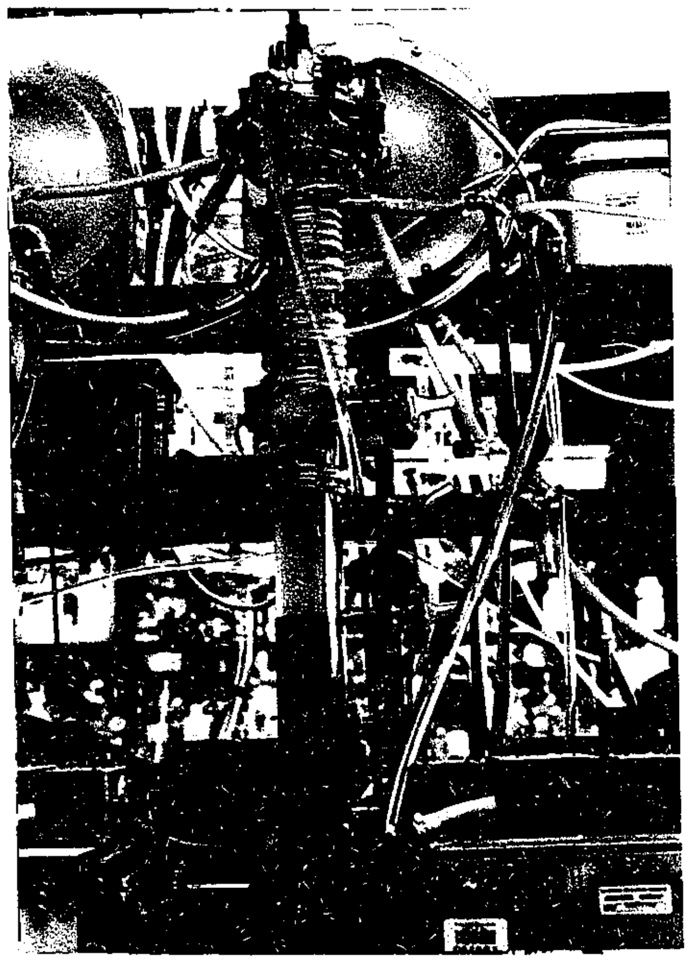

Figure 7. Photograph of an AFBR

This prevented air from entering the reactor which would inhibit growth of obligate anaerobes. The reactor fluid was maintained at a temperature of 35 °C which is optimum for mesophiles. This was accomplished by circulating water at 50 °C through tygon tubing wrapped around part of the outside of the reactors.

Figure 8 shows a fluidized bed of biofilm covered particles along with the distributor. Fresh influent and recycled mixed liquor was uniformly distributed at the base of the reactor by means of a multi-orfice plate. This plate was 5.5 cm in diameter, 1.0 cm in thickness and had 1/8" holes arranged in a concentric fashion. The areas of the holes comprised about 20% of the total area of the plate. Over the multi-orifice plate lay a 1.5 cm high packed section of glass beads. The glass beads were smooth and 3.0 mm in diameter. Since the glass beads were smaller than the plate holes, a nylon mesh was placed between the two to prevent the holes from plugging. Inert support material, Biolite, with mass of 150 g and surface area of 2.7 m^2 was placed directly over the glass beads. The glass beads not only aided with the uniform distribution of the fluid entering the base of the reactor but also prevented the small carrier particles from falling through the distributor when the recycle pump was temporarily shut off. Due to their size and density, the glass beads remained stationary throughout the experiment and were not fluidized like the carrier particles. The unexpanded bed was 250 mL in volume and 10 cm in height..

The inert support medium used in this study is called Biolite and is shown in Figure 9. Biolite, a clay based material, is a new product of Degremont S.A., France. This

Figure 8. Photograph of a Fluidized Bed of Bioparticles

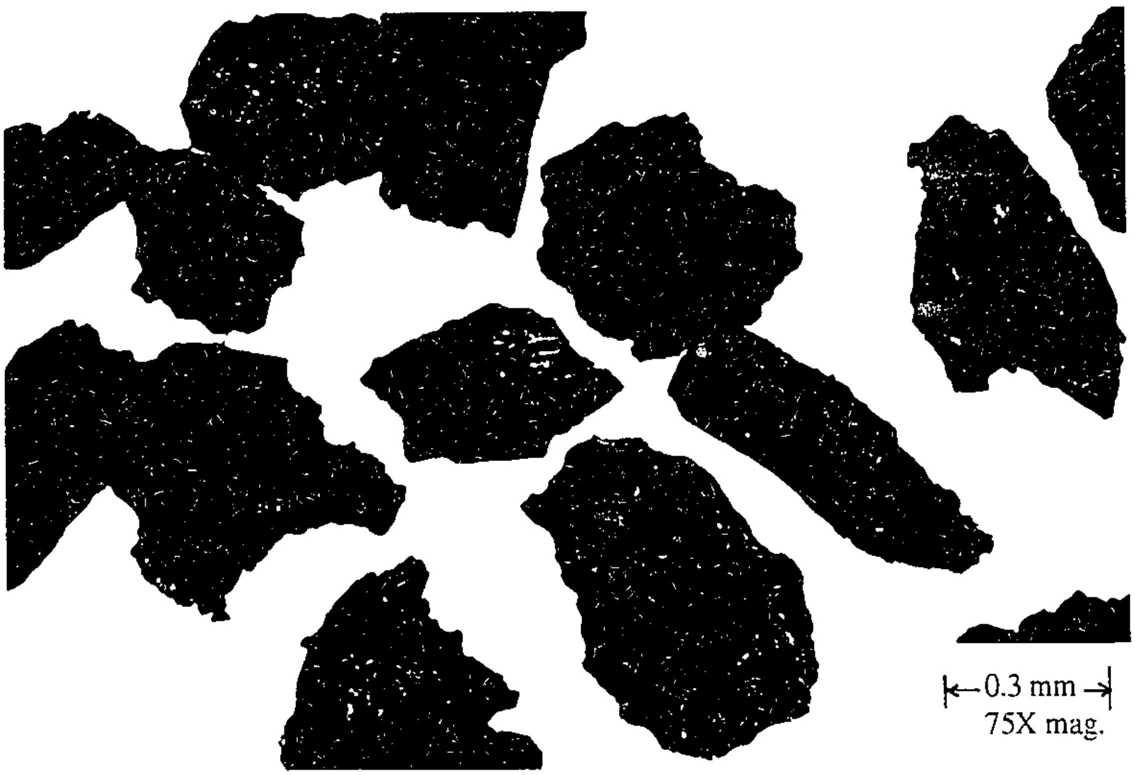

Figure 9. Photograph of Clean Biolite Particles

material is still at the research stage and has not yet been made commercially available by its manufacturers. Various properties of Biolite measured as part of this study are given in Table 3. This material was chosen for two primary reasons. Firstly, there is a lack of experimental studies in the literature using this material due to its recent development. Secondly, and perhaps more importantly, Biolite is a strong, light weight material and is claimed to have excellent surface properties for bio-adhesion. These properties make this material highly suitable for use in biological fluidized bed reactors.

Table 3. Physical Properties of Biolite

Material of Composition	Expanded Clay
Appearance	Light Grey (uncoated)
Average Diameter	0.3 mm
Density	1.6 g/mL
Bulk Density	0.6 g/mL
Sphericity	0.7
Specific Surface	18 m^2/kg

Figure 10 shows feed and recycle pumps used with the AFBR. A variable speed drive peristaltic pump (Masterflex, 1-600 rpm) was used to recycle the reactor fluid back to the bottom of the reactor. A high precision, low flowrate peristaltic pump (Harvard Apparatus) was used to introduce feed into the recycle line. Over the entire study, pumping rates for the recycle and feed pumps ranged from 75- 500 mL/min and 0.1-1.7

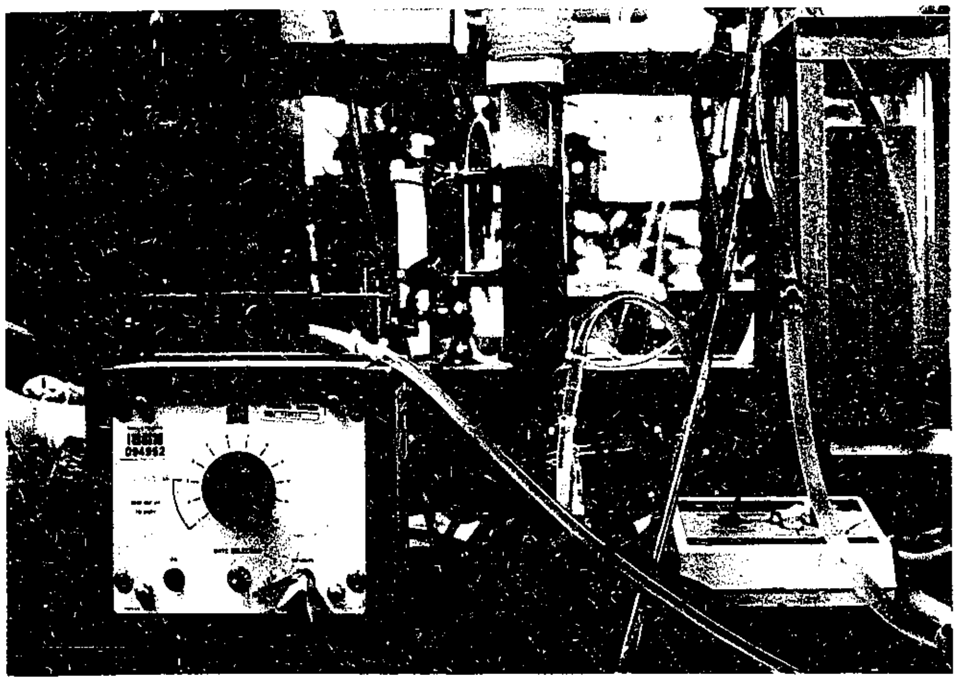

Figure 10. Photograph of Feed and Recycle Pumps used in AFBR System

mL/min respectively. Tygon tubing was used to connect the feed reservoirs, pumps and reactors in the AFBR system. Wear resistant Norprene tubing (by Cole Parmer) was used inside the peristaltic pumps. Feed was contained in jacketed glass vessels and was maintained at 4°C by circulating cold tap water to slow down its degradation.

Biogas generated was measured using a wet-tip gas meter (by Wettip Inc.) placed above the reactor. A glass tee fitted with a septum was placed in the gas line between the reactor and the gas meter to facilitate determination of biogas quality. Gas samples for chromatographic analysis were taken using a 1 mL syringe. All pumps and meters were calibrated before experiments were started.

A dip tube extended from the reactor lid to about 4 cm below the liquid level in the reactor. Sampling of reactor fluid and biomass was done using a long glass tube, which fit inside the dip tube, and a rubber suction bulb. This technique allowed removal of small amounts of biomass covered particles or bioparticles from any depth of the bed without interrupting fluidization or opening the reactor. An additional sampling port with a rubber septa was placed in the recycle line for convenience.

A common problem encountered in some bench scale reactors is the difficulty in maintaining the desired feed rate despite the use of an accurate pump. This is due to the production of gas in the feed line running between the feed tank and the feed pump. The problem was circumvented here by adding a high speed centrifugal pump in the feed line which maintained a high pressure in the line and prevented gas from accumulating.

4.3 Program of Experimentation

The AFBRs were operated for a period of about 8 months. The first 2 months were for startup of the reactors during which period a healthy biofilm on the Biolite carrier particles was developed. During the next 4 months, steady state experiments were run using sucrose as carbon source in the feed. Here a total of 15 runs were performed at different feed concentrations and HRTs. The last 2 months of this study were utilized for carrying out the treatment of landfill leachate and batch kinetic runs using acetate as feed. Film thickness measurements were also made during the last week of the experimental program to get estimates on biofilm densities.

4.3.1 Reactor Feed

Sucrose Feed:

The synthetic sucrose feed was composed of various components other than sucrose to provide the anaerobic bacteria with sources of nitrogen, phosphorus, miscellaneous essential nutrients and alkalinity. These nutrients with their respective concentrations for each synthetic feed used in this study are listed in Table 4. As this table indicates, for all three feed concentrations the COD:N:P ratio was maintained at 110:5:1. Salt solution containing all components of the feed except sucrose and yeast extract was made in 10 L quantities as required and stored at 4 °C. This hastened the feed making process and reduced the chances of errors which would lead to variations in feed.

Leachate Feed:

Landfill leachate was obtained from the Trail Road Landfill in Nepean in three 50 L containers. Chemical and physical characteristics of the leachate as measured in the laboratory are listed in Table 5. The leachate was kept frozen by storing the containers at -5 °C and were transferred to cold room at 4 °C only when required. Before use, leachate

Table 4. Composition of Sucrose Wastes in g/L

	5 g COD/L	10 g COD/L	15 g COD/L
Sucrose	5.0	10.0	15.0
$(NH_4)HCO_3$	1.0	2.0	3.0
$Na(HCO_3)$	2.5	5	7.5
KH_2CO_3	3.1	6.2	9.3
$(NH_4)_2SO_4$	0.25	0.5	0.75
K_2HPO_4	0.13	0.26	0.39
KH_2PO_4	0.1	0.2	0.3
Yeast Extract	0.05	0.1	0.15

in each container was completely thawed and mixed well. These procedures ensured minimal degradation of wastewater prior to treatment and consistent feed composition. Prior to treatment, the total nitrogen and phosphorus levels of the leachate were determined to check for any possible nutritional deficiency. Due to phosphorus deficiency in the original leachate, total phosphorus of the leachate was increased to a COD:P ratio of 300:1 by the addition of K_2HPO_4. The total nitrogen level of 240 mg/L was more than adequate for anaerobic treatment. Thus the leachate treated had a COD:N:P ratio of 300:14:1. High leachate alkalinity of 6250 mg/L made addition of pH buffers unnecessary.

Table 5. Characteristics of Trail Road Landfill Leachate

Component	Concentration in mg/L (except pH)
Total COD	5600
Soluble COD	5300
Total BOD_5	5390*
BOD_5:COD	0.86*
Acetic acid	1180
Propionic acid	680
Butyric acid	450
Total VFA (as acetate)	2900
Total VFA (as COD)	3110
pH	7.25
Alkalinity	6250 (as $CaCO_3$)
Total Phosphorus	0.5
Total Nitrogen	240*
Total Solids	10600
Total Volatile Solids	3450
Total Fixed Solids	7150
Suspended Solids	125
Volatile Suspended Solids	99
Fixed Suspended Solids	26

* Taken from Trail Road Landfill Data

4.3.2 Startup

For a carrier material to be considered suitable for use in fluidized bed reactors, it must allow fast and reliable startup of the reactor. This requires the particle to have surface properties which are conducive to bacterial attachment. Startup studies on AFBRs using Biolite were performed to evaluate the biofilm formation on this newly developed material, and to compare its performance to other carrier particles like sand. More specifically, two different startup protocols were tested to establish the superior startup method based on length of time required and stability of the process during startup period.

After careful consideration of the AFBR start-up literature, start-up protocols for the maximum efficiency profile (MEP) and the maximum load profile (MLP) were investigated. Following MEP, the first reactor received an initial loading of 0.25 g/L/d at a HRT of 10 d. This loading rate was incremented by 0.25 g/L/d when VFA levels dropped below 500 mg/L and remained there for three consecutive days. The second reactor was started-up using MLP by applying a organic load of 2.5 g/L/d initially (HRT of 1 d) and holding it at that level till reactor performance stabilized and start-up was deemed complete. HRTs were based on empty bed reactor volume.

Initially, beds were expanded to only 30% of static bed height to provide low shear environment which would facilitate microbial attachment. As the biofilms developed, bioparticles got lighter and expansion level gradually increased. Reactors were inoculated twice a week with inoculant (2% of reactor volume) containing about 40 g COD/L of biomass for the entire startup period. The inoculant was prepared by finely grinding seed sewage sludge in a blender and adding enough water to make a slurry. During start-up, total and soluble COD, attached biomass and pH were measured twice a week. Reactor mixed liquor VFAs, biogas and temperature were monitored daily.

4.3.3 Steady State Experiments

Steady state experiments were conducted to provide valuable information on the kinetics and stability of AFBRs treating sucrose based substrate. They also were used to evaluate the suitability of Biolite as a carrier material in AFBRs and the treatability of municipal landfill leachate.

Steady state run conditions performed for synthetic sucrose and leachate feeds are presented in Tables 6 and 7. As Table 6 shows, for the study using sucrose feed, the reactors were loaded at five different HRTs, each at three different influent concentrations giving a total of 15 runs. Since landfill leachate was used only for a treatability study, fewer runs were performed. Leachates with organic concentrations of 5000 mg COD/L and 2000 mg COD/L were each treated at four different HRTs. Volumetric loading rates for both sucrose and leachate feed were chosen after carefully reviewing the literature on similar reactors and wastewater. Chosen loading rates for leachate are lower than those for synthetic waste due to possible presence of inhibiting compounds such as heavy metal ions and chlorinated organics.

For each steady state run, three sets of data were taken over three consecutive days. The average of these values was used for calculations and graphs. Ten HRTs (up to a maximum of 5 d) were allowed after each change in loading conditions and prior to data collection. This ensured that the system had reached a pseudo steady state and that any changes in biomass or effluent quality after this period would occur very slowly.

Table 6. HRTs corresponding to various Sucrose Feed COD Concentrations and Loading Rates

Feed Conc.	Loading (g/L/d)				
(g COD/L)	5.0	10.0	15.0	20.0	25.0
	HRT (h)				
5.0	24.0	12.0	8.0	6.0	4.8
10.0	48.0	24.0	16.0	12.0	9.6
15.0	72.0	36.0	24.0	18.0	14.4

Table 7. HRTs corresponding to various Leachate COD Concentrations and Loading Rates

Feed Conc.	Loading (g/L/d)			
(g COD/L)	2.1	4.3	6.4	10.7
	HRT (h)			
2	22.4	11.2	7.5	4.5
	Loading (g/L/d)			
	2.7	5.4	8.0	13.3
	HRT (h)			
5	44.8	22.4	15	9

4.3.4 Batch Studies

Batch tests were performed as a means of cross-checking results of steady state substrate utilization kinetics. Batch studies have the advantage of taking little time to conduct since there is no need to wait for the system to reach steady state following changes in operating conditions.

After steady state experiments using sucrose as feed, batch studies were performed. Only one reactor was utilized for these studies. Three trials were performed using initial concentrations of 534, 899 and 1160 mg/L acetate. The biomass in the reactor was 200 mg COD/g Biolite and assumed to be constant.

After bringing the reactor mixed liquor to the desired concentration by adding acetic acid, samples of mixed liquor were taken periodically for VFA analysis and determination of maximum specific activity. The temperature of the reactor was maintained at 35 °C throughout all trials.

4.3.5 Determination of Biofilm Thickness

Biofilm covered particles were removed from a reactor treating 15 g sucrose/L synthetic wastes at a loading rate of 25 g/L/d. Samples of particles were taken from the top portion of the bed, where biofilms were thick and less dense, and from the bottom portion of the bed where biofilm were thinner and denser. Thickness determination was done by measurement under a light microscope according to the procedure given in the next section.

4.4 Analytical Techniques

4.4.1 Biogas Quality

Biogas composition (methane and carbon dioxide) was determined by the gas chromatography method of van Huyssteen (1967). This technique involved using a Porapak T column (6.35 mm x 304.3 cm) on a Hewlett-Packard 5710a gas chromatograph equipped with a thermal conductivity detector and model 3380a integrator. The column was held at 70 °C and helium flowing at a rate of 40 mL/min was used as the carrier gas.

4.4.2 Volatile Fatty Acids

VFAs were determined by the method of Ackman (1972) using a Hewlett-Packard 5721A gas chromatograph equipped with an automatic sampler, flame ionization detector (FID), model 3380A integrator, and Chromosorb 101 glass column (6.35 mm x 365.76 cm). The column was kept at 180 °C and the temperature of the FID was 350 °C. Helium passing over formic acid at a flow rate of 15 mL/min was used as the carrier gas. Samples were prepared by adding 0.5 mL of sample to 0.5 mL of an internal standard that contained 1000 mg/L of isobutyric acid. The chromatograph was calibrated with a standard that contained 1000 mg/L each of acetic, propionic and butyric acids. The volume of sample injected (on the column packing) was 0.8 uL.

4.4.3 Chemical Oxygen Demand (COD)

COD determinations were made using the closed reflux colorimetric method as outlined in 5220 D, APHA (1989) . Absorbance was measured using a Perkin-Elmer spectrophotometer (Coleman 295) set at 600 nm. Soluble COD was determined by centrifuging liquid (IEC, Model B-20, Rotor #870) at 5000 rpm for 5 min and taking supernatant as COD sample.

4.4.4 Attached Biomass

The COD of biomass was determined by removing 0.5 g of the bed from the top of the reactor using a long glass tube and a suction bulb. The bioparticles were then washed twice using 2x10 mL of distilled water by gently swirling in a 50 mL beaker followed by careful decantation. Depending on the concentration of biomass expected, small amounts of bioparticles were transferred into two Kimax glass culture tubes. After the COD test was completed, the reagents were carefully decanted and the remaining Biolite particles were vigorously washed with tap and distilled water using a vortex. The clean Biolite particles were then dried at 105 °C for 1 hr and weighed in a sensitive analytical balance. The COD of the biomass was expressed as mg COD per g of clean Biolite.

4.4.5 Biofilm Thickness

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

Small amounts of biofilm covered particles and mixed liquor were removed from the reactor and transferred to glass vials. Bioparticles were then gently rinsed in distilled water to remove loosely attached and suspended material which would otherwise interfere with size measurement. Size measurements on the coated particles were made while the particles were wet, since moisture loss would have caused film shrinkage. Biofilm covered particles ashed at 550 °C for 1 hr were used as the uncoated sample. The size of fresh Biolite particles was also measured to determine the amount of attrition.

The second longest diameter of particles was measured using a light microscope under a total magnification of 50X and field of view of 4.0 mm. The smallest division of the calibrated scale was 20 μm. Measurement of the diameter in all cases was done using a sample size of 100.

4.4.6 Analytical Errors

A summary of the variations which could be expected for the analyses used in this study is given in Table 8 . The systematic and analytical errors associated with the laboratory measurements were considered to be within the range of values published in APHA (1989). Other systematic and analytical errors were calculated as necessary for the study.

Table 8. Summary of Analytical Errors

Quantity Measured	Equipment / Method used	Precision of Analytical Technique	Reference
biogas quality	HP 5710 A GC (TC detector)	± 1 %	van Huyssteen (1967)
biogas quantity	Wet-Tip meter	± 5 %	this study
volatile fatty acids	HP 5721 A GC (FI detector)	± 2 %	Ackman (1972)
pH	Fisher glass electrode and pH meter	± 0.1 unit	APHA (1989)
alkalinity	titration method (APHA 2320 B)	± 2 %	APHA (1989)
COD	closed reflux colorimetric method (APHA 5220 D)	± 8 %	APHA (1989)
biomass COD	see section 4.4.3	± 10 %	this study
solids	Gooch crucible, glass fibre filter (APHA 2540 B, D, E)	± 10 %	APHA (1989)
total phosphorus	stannous chloride method (APHA)	± 10 %	APHA (1989)

CHAPTER 5. RESULTS AND DISCUSSION

5.1 Overview

Section 2 presents results on reactor startup and early biofilm development

Section 3 describes steady state AFBR operation and presents data for organic removal efficiency, biomass concentration, biogas production, effluent suspended solids and refractory organics.

Section 4 develops relationships between important process variables which can be used for design purposes.

Section 5 shows the data from batch kinetic studies and how they compare to those determined from steady state data.

Section 6 presents a brief comparison of kinetic constants determined from this study to literature values.

Section 7 presents both predicted and experimental values of removal efficiencies and biomass concentrations as a function of of reactor HRT.

Section 8 presents results on the treatability study of landfill leachate and the practical problem encountered when treating a real waste.

5.2 Reactor Startup

Reactors 1 and 2 were started up using the maximum efficiency profile (MEP) and the maximum load profile (MLP) respectively. Startup duration was about two months with both reactors achieving >90% COD removal at an organic loading rate of 2.5 g/L/d with feed concentration of 2.5 g/L and HRT of 1 d at the end of this period. It should be noted that HRT was based on empty bed reactor volume. Table 9 lists organic loading rates, total and soluble effluent COD, and attached solids COD for each reactor. The loading rate profile of each reactor is shown in Figure 11. As it can be seen, under MEP, reactor 1 is gradually brought to the desired end loading rate of 2.5 g/L/d over a period of five weeks. The average time between increments was about 4 days. In contrast, the loading profile under the MLP used for reactor 2 is constant. Here an initial loading rate of 2.5 g/L/d was held constant during the entire startup period.

Figure 12 is a plot of volatile film solids expressed in both mg COD/g of Biolite and g VFS/L of fluidized bed volume against time in days. Both reactors 1 and 2 reach about 30 mg COD/g Biolite (or 6.4 g VFS/L of expanded bed) by the end of startup. Therefore it can be concluded that both loading rate profiles lead to equally fast startups.

The level of propionic acid in the mixed liquor represents the amount of stress being placed on the reactor. From Table 10, which shows levels of acetic, propionic and total volatile fatty acids, it can be seen that the propionic acid levels of reactor 2 (subjected to MLP) are on the average higher than those of reactor one. Subjecting a biological reactor to high stress levels usually results in unstable operation. Figure 13, which is a plot of effluent soluble COD and COD removal vs. time, supports this conclusion. Here large peaks in the effluent COD are seen to approach 900 mg COD/L for reactor 2. Corresponding peaks for COD of reactor 1 mixed liquor are substantially lower indicating more stable operation. Since the reactors were inoculated with biomass

Table 9. Summary of Startup Data

	Reactor1 (MEP)				Reactor 2 (MLP)			
Day	Loading Rate (g/l/d)	Total Effluent COD (mg/l)	Soluble Effluent COD (mg/l)	Attached Solids COD (mg/g)	Loading Rate (g/l/d)	Total Effluent COD (mg/l)	Soluble Effluent COD (mg/l)	Attached Solids COD (mg/g)
1	0.25	900	400	0.0	2.50	1160	320	0.0
4	0.50	660	390	5.0	2.50	810	630	7.5
9	0.75	830	440	9.8	2.50	1270	870	9.8
11	0.75	790	390	12.5	2.50	640	410	11.1
15	1.25	450	280	17.0	2.50	550	410	14.2
18	1.50	930	470	15.9	2.50	960	470	12.1
25	1.75	520	290	22.0	2.50	520	380	20.0
30	2.00	740	590	29.6	2.50	1010	880	28.6
33	2.25	960	460	23.5	2.50	1660	830	31.4
37	2.50	560	270	38.7	2.50	660	350	23.8
52	2.50	480	200	27.0	2.50	400	260	32.0

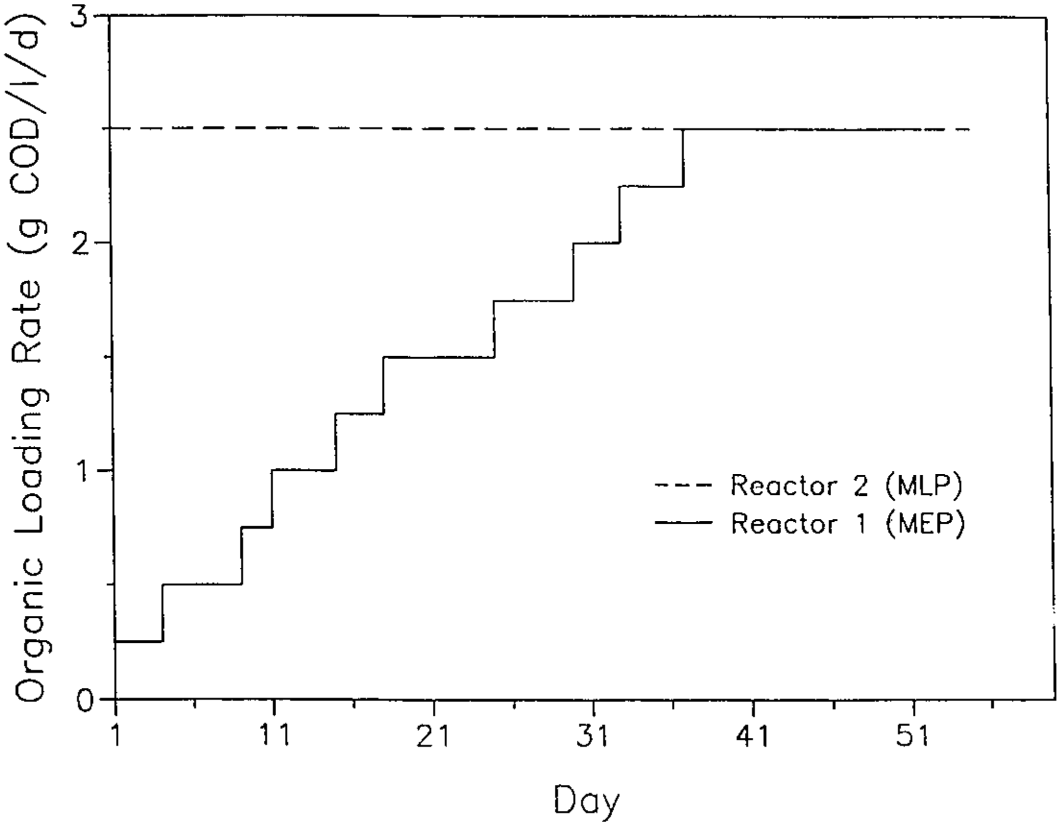

Figure 11. Startup Loading Rate Profiles of AFBRs

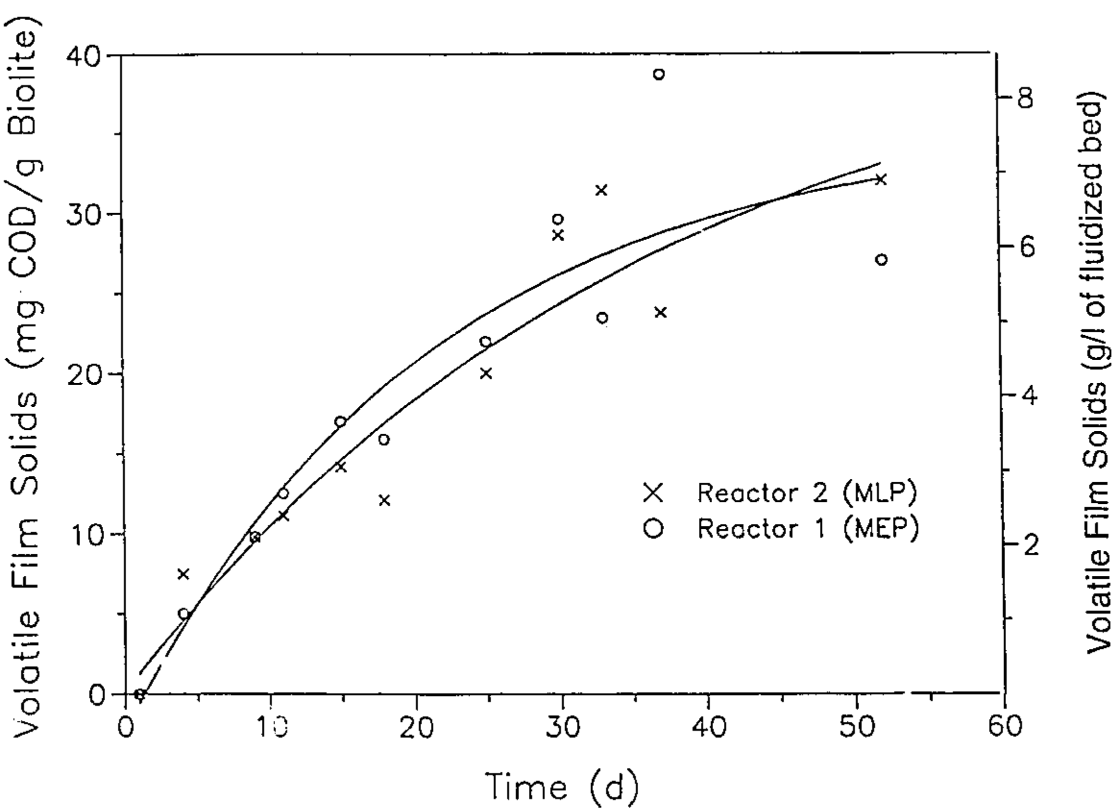

Figure 12. Volatile Film Solids Accumulated during Reactor Startup

Table 10. Startup VFA Levels for Reactors 1 and 2

	Reactor 1 (MEP)			Reactor 2 (MLP)		
Day	Acetic Acid (mg/l)	Propionic Acid (mg/l)	Total VFA as Acetate (mg/l)	Acetic Acid (mg/l)	Propionic Acid (mg/l)	Total VFA as Acetate (mg/l)
2	18	8	29.2	26	9	38.6
4	6	0	6	18	7	27.8
5	35	0	35	51	15	72
6	37	20	65	9	0	9
7	108	16	130.4	212	31	255.4
8	196	31	239.4	490	92	618.8
12	202	0	202	620	76	726.4
13	121	19	147.6	251	15	272
14	105	0	105	117	9	129.6
15	74	12	90.8	144	0	144
16	19.5	0	19.5	75	0	75
18	38	0	38	86	0	86
19	352	43	412.2	525	100	665
20	495	44	556.6	359	10	373
21	167	0	167	237	0	237
22	370	101	511.4	429	39	483.6
23	205	23	237.2	231	10	245
25	19.3	0	19.3	90.5	0	90.5
26	268	99	406.6	468	135	657
28	72	9	84.6	164	17	187.8
29	306	123	478.2	445	161	670.4
32	0	0	0	9.7	0	9.7
33	314	127	491.8	481	167	714.8
34	170	30	212	210	10	224
35	0	0	0	0	0	0
36	135	115	296	365	205	652
37	0	0	0	0	0	0
39	0	0	0	0	0	0
40	60	85	179	145	95	278
41	0	0	0	0	0	0
48	90	0	90	0	0	0
49	305	100	445	0	0	0
53	20	95	153	140	0	140
55	35	15	56	35	20	63
56	80	125	255	85	185	344
57	60	75	165	55	165	286
60	5	0	5	0	0	0
Avg.	121	36	171	178	43	237

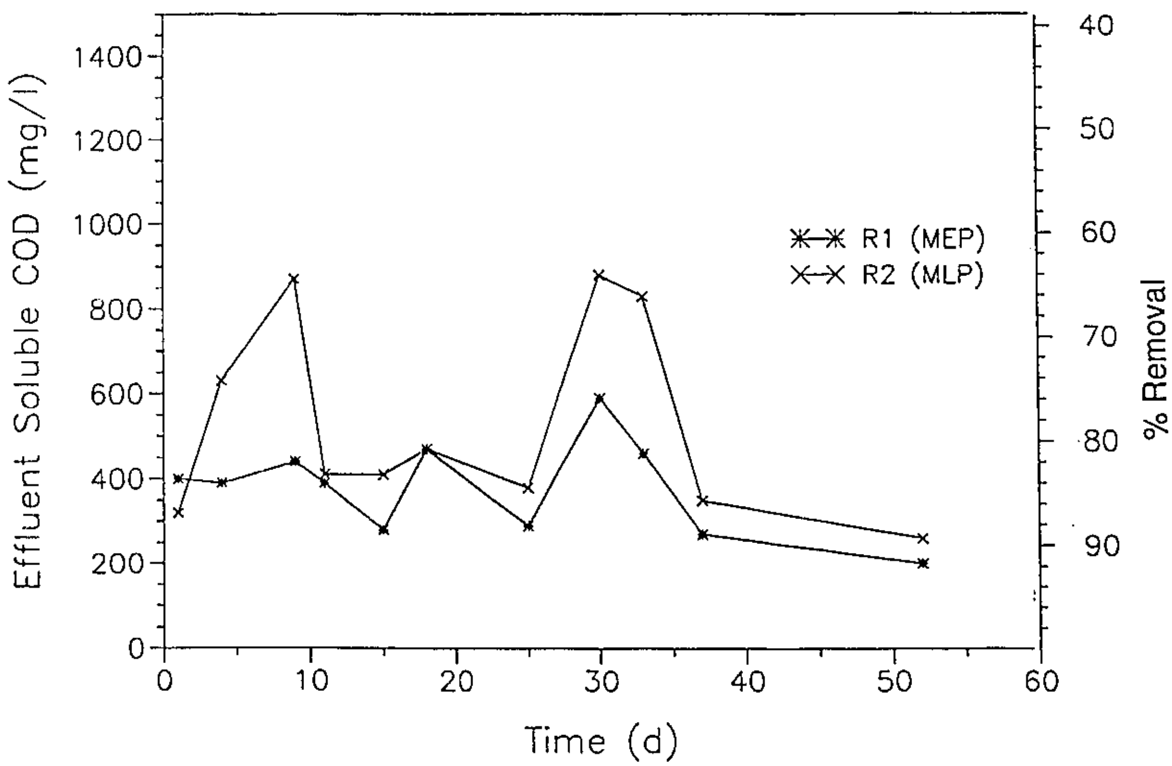

Figure 13. Effluent Soluble COD levels during Startup

twice a week, soluble COD levels of the reactors rise and fall due to removal of organics by suspended biomass of inocula.

Figures 14 and 15 plot the specific loading and removal rates vs. time respectively for MEP and MLP. From these figures it can be seen that the reactor 2 subjected to MLP received an initial specific loading rate which was four times higher than reactor 1. The corresponding specific removal rates are also much higher. However, at the end of the startup period, both reactors reach similar loading and removal rates.

Startup conditions of reactor 2 were more severe and closer to the point at which danger of reactor failure exists. Thus, the MEP (used for reactor 1) should be preferred over the MLP because of safer and more stable reactor operation.

Summary

Compared to other startups reported in the literature, the startup of the two AFBRs in this study was remarkably fast. For instance, Hsu and Shiek (1993) required a minimum of 140 days using the MLP to bring their reactor up to a loading of 6 g TOC/L/d at 98% COD removal. They used Manville R-633 beads (flux calcined diatomic earth clay) as growth support media. Switzenbaum (1978) spent 9 months to startup his anaerobic expanded bed reactors with glass beads as carrier particles. Reported long startup periods are not isolated to AFBRs and expanded bed reactors. Van den Berg and Kennedy (1981) found that startup of DSFF reactors using glass support takes more than 200 days, however they report that startup using clay based material as support only took 30 days. The fast startup of this study is believed to be due the excellent surface adhesion properties of Biolite carrier particles, thus supporting Degremont Inc. claims of Biolite.

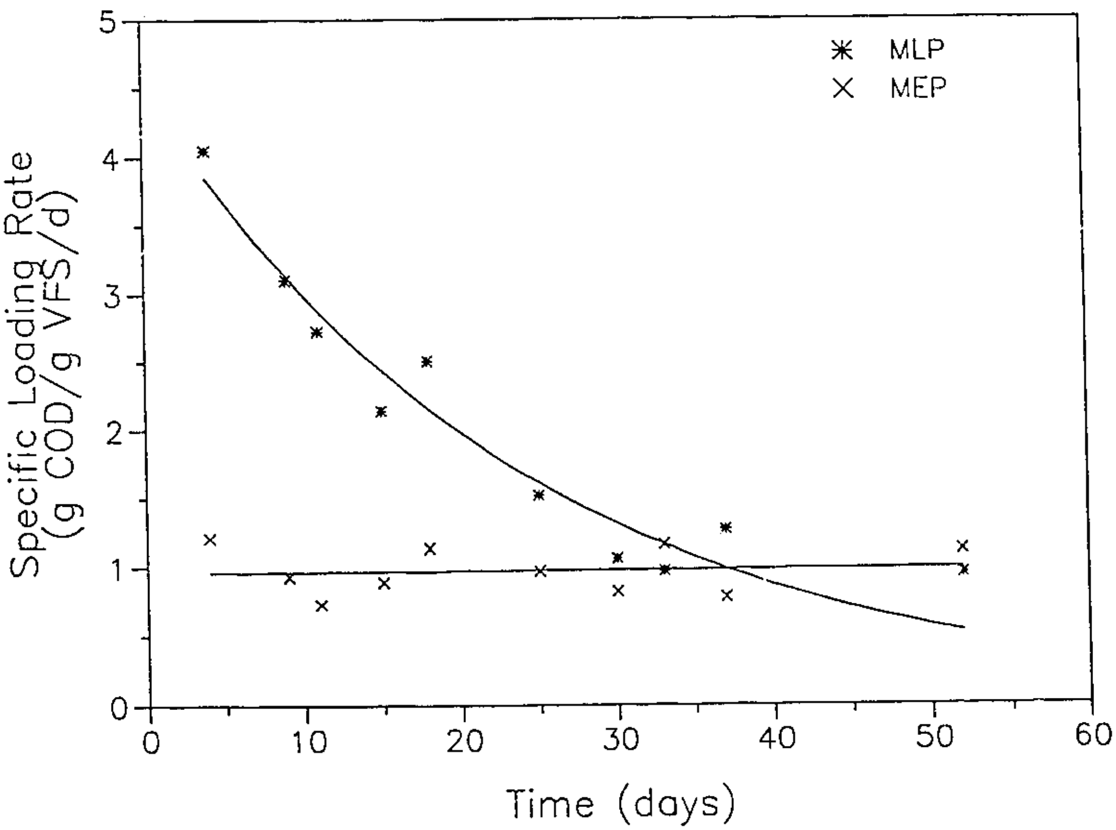

Figure 14. Specific Loading Rate vs. Time (during startup period)

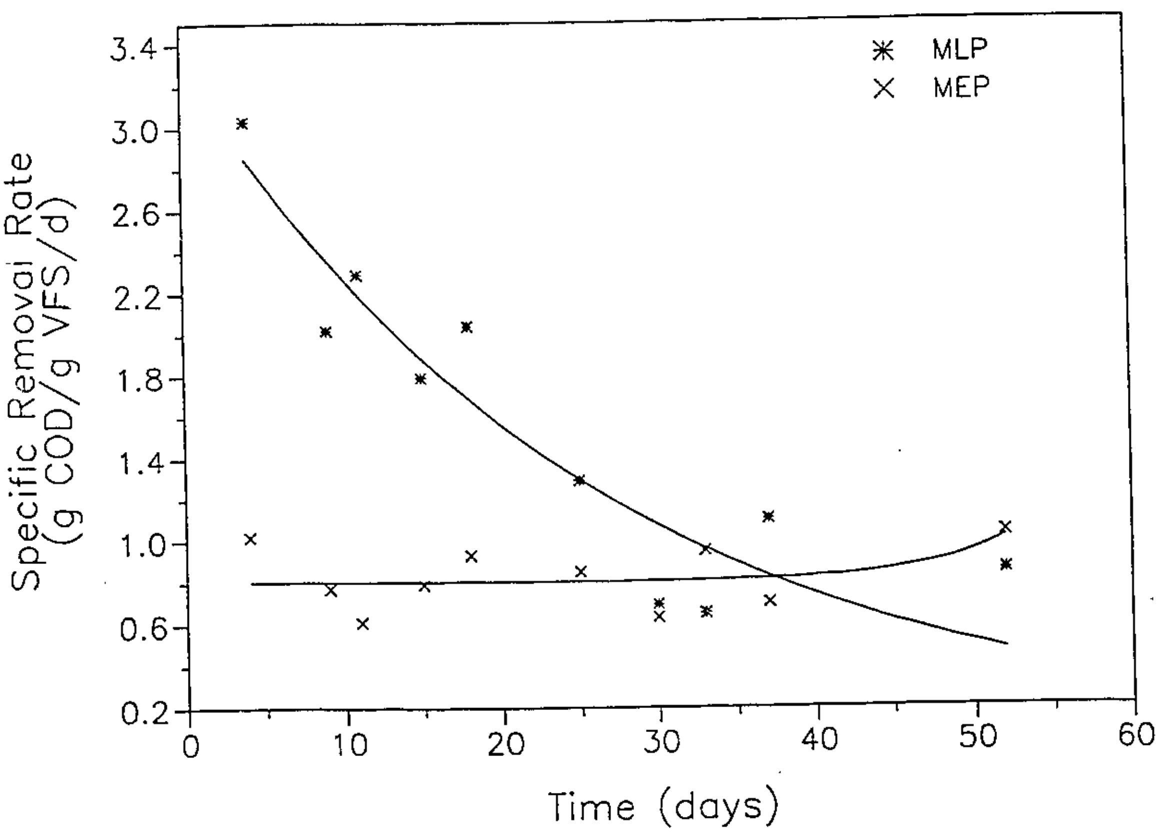

Figure 15. Specific Removal Rate vs. Time (during startup period)

5.3 Steady State AFBR Operation

5.3.1 Summary of Steady State AFBR performance

Summary data for each of the 3 substrate concentrations evaluated are presented in Table 11. Results are presented for effluent soluble and insoluble COD, VFA, biogas production and attached biomass. Each of these results is an average of three sets of measurements taken over three consecutive days at steady state conditions. The COD removal based on total COD takes into account the suspended biomass leaving the reactor in the effluent while the removal based on soluble COD only considers soluble organic substrate remaining which was not utilized completely by the microbes. Thus numbers for the %COD removal based on soluble COD are slightly higher as expected and are >96% for low loading rate of 5 g/L/d.

Since the organic substrate (sucrose based synthetic feed) to the reactor was completely soluble, the suspended effluent COD represented the biomass leaving the reactor in the effluent. This was calculated by taking the difference between the total and soluble mixed liquor CODs. Methane production is based on the methane content of the biogas. In this study approximately 70% of the biogas generated is methane, the remaining 30% being carbon dioxide. Volatile fatty acid levels are expressed as mg/L of acetate by using the conversion that 1 g of propionic acid is equivalent to 1.4 g of acetate. The levels of butyric acid in the effluent were very low and consequently not included in

Table 11. Summary of Steady State Data for Sucrose Feed

Loading (g/l/d)	Feed COD (g/l)	HRT (h)	Total Effluent COD (mg/l)	Soluble Effluent COD (mg/l)	Attached Solids COD (mg/g)	Methane Production (l/d)	Suspended Effluent COD (mg/l)	% Removal (based on total effluent COD)	% Removal (based on soluble effluent COD)	VFA (as acetate) (mg/l)
5	5	24	560	390	69	0.7	170	88.8	92.2	300
10	5	12	400	180	66	1.4	220	92.0	96.4	144
15	5	8	800	240	56	1.9	560	84.0	95.2	180
20	5	6	1440	1160	210	2.7	280	71.2	76.8	1084
25	5	4.8	1390	1260	220	3.3	130	72.2	74.8	1071
5	10	48	600	430	70	0.9	170	94.0	95.7	176
10	10	24	710	420	53	1.8	290	92.9	95.8	85
15	10	16	1220	660	110	3.0	560	87.8	93.4	515
20	10	12	3330	3100	260	2.1	230	66.7	69.0	2450
25	10	9.6	3200	2700	220	3.5	500	68.0	73.0	2500
5	15	72	800	550	100	1.1	250	94.7	96.3	515
10	15	36	1370	960	72	2.4	410	90.9	93.6	940
15	15	24	2390	1840	120	2.1	550	84.1	87.7	790
20	15	18	3570	3510	210	2.1	60	76.2	76.6	2126
25	15	14.4	5000	4380	320	3.8	620	66.7	70.8	4570

the calculations.

5.3.2 Organic Carbon Balance

An organic carbon balance can be evaluated for the AFBRs over the period of steady state operations. These balances are useful because they give additional support to the experimental results obtained by acting as a cross checking mechanism. The mass balance on organic carbon was evaluated in the following manner:

Carbon in = Soluble carbon in effluent + Gaseous methane production +

Methane dissolved in effluent + Effluent volatile suspended solids.

The right hand side of this equation represents the organic carbon out of the reactor. The ratio of carbon in to carbon out is calculated in the last column of Table 12. All numbers in Table 12 are expressed as g COD/d. Methane was converted to a COD equivalent using the relationship that at 35 °C and one atmosphere, one gram of COD is equivalent to 400 ml methane (Switzenbaum, 1978).

The "methane in effluent" column expresses the rate of loss of dissolved methane in the effluent stream. The rate of loss of dissolved methane was assumed to be the product of the solubility of methane at 25 °C and one atmosphere pressure (0.0328 l CH_4/L of solution) and the waste flow in litres per day. This assumes that the effluent is saturated with methane in solution. The volume of dissolved gas is converted to its COD equivalent.

The "effluent VSS" accounts for organic carbon lost as solids in the effluent. The effluent solids quantity was converted to a COD equivalent by using the conversion factor

Table 12. Organic Carbon Balance

HRT (h)	Carbon In (g COD/d)	Soluble Effluent (g COD/d)	Methane Production (g COD/d)	Methane in Effluent (g COD/d)	Effluent VSS (g COD/d)	Total Carbon Out (g COD/d)	Rate In/ Rate Out
24	2.50	0.20	1.85	0.04	0.09	2.17	1.15
12	5.00	0.18	3.52	0.08	0.22	4.01	1.25
8	7.50	0.36	4.73	0.12	0.84	6.05	1.24
6	10.00	2.32	6.68	0.16	0.56	9.72	1.03
4.8	12.50	3.15	8.21	0.21	0.33	11.89	1.05
48	2.50	0.11	2.25	0.02	0.04	2.42	1.03
24	5.00	0.21	4.50	0.04	0.15	4.90	1.02
16	7.50	0.50	7.58	0.06	0.42	8.55	0.88
12	10.00	3.10	5.25	0.08	0.23	8.66	1.15
9.6	12.50	3.38	8.74	0.10	0.63	12.84	0.97
72	2.50	0.09	2.74	0.01	0.04	2.88	0.87
36	5.00	0.32	5.93	0.03	0.14	6.41	0.78
24	7.50	0.92	5.33	0.04	0.28	6.56	1.14
18	10.00	2.34	5.33	0.05	0.04	7.76	1.29
14.4	12.50	3.65	9.45	0.07	0.52	13.69	0.91
						Average:	1.05

of 1.4 mg COD/mg VSS determined in this study. This value agrees well with the value of 1.42 mg COD/mg VSS found by Kennedy (1985).

The mean ratio of carbon in : carbon out was found to be 1.05. This value which is close to 1.0 indicates a good account of organic carbon coming in and out of the fluidized bed reactors and verifies the reliability of the experimental results obtained in this study. A sample organic carbon balance calculation is shown in detail in Appendix B.

In a similar type of calculation, Table 13 shows a comparison of COD removed to methane produced. In this table the total methane produced (methane captured as gas and methane dissolved in effluent) is compared to the methane equivalent of the COD removed by the AFBRs as determined by the difference between influent and effluent soluble COD converted to its methane equivalent. The same conversion factors were used as in the previous organic carbon balances.

The results indicate that an average of almost 90% of soluble COD was recovered as methane. These results suggest that the net synthesis of biological solids in AFBRs is low.

5.3.3 Organic Removal Efficiency vs. Hydraulic and Organic Loading Rates

The variation of organic removal efficiency with organic loading rate and HRT are shown in Figures 16 to 19. As expected, the substrate removal efficiency decreased with increasing organic loading and decreasing HRT. In Figures 18 and 19 where removal efficiency is plotted against HRT, three separate curves emerge for each influent

Table 13. Comparison of COD Removed to Methane Produced

HRT (h)	Influent Waste (g COD/d)	COD Removed (g COD/d)	Total Methane Produced (l/d)	Methane Equiv. of COD Removed (l/d)	Conversion of COD to Methane
24	2.50	2.31	0.7	0.92	0.80
12	5.00	4.82	1.4	1.93	0.73
8	7.50	7.14	1.9	2.86	0.66
6	10.00	7.68	2.7	3.07	0.87
4.8	12.50	9.35	3.3	3.74	0.88
48	2.50	2.39	0.9	0.96	0.94
24	5.00	4.79	1.8	1.92	0.94
16	7.50	7.01	3.0	2.80	1.08
12	10.00	6.90	2.1	2.76	0.76
9.6	12.50	9.13	3.5	3.65	0.96
72	2.50	2.41	1.1	0.96	1.14
36	5.00	4.68	2.4	1.87	1.27
24	7.50	6.58	2.1	2.63	0.81
18	10.00	7.66	2.1	3.06	0.70
14.4	12.50	8.85	3.8	3.54	1.07
		Totals:	32.8	36.7	0.89

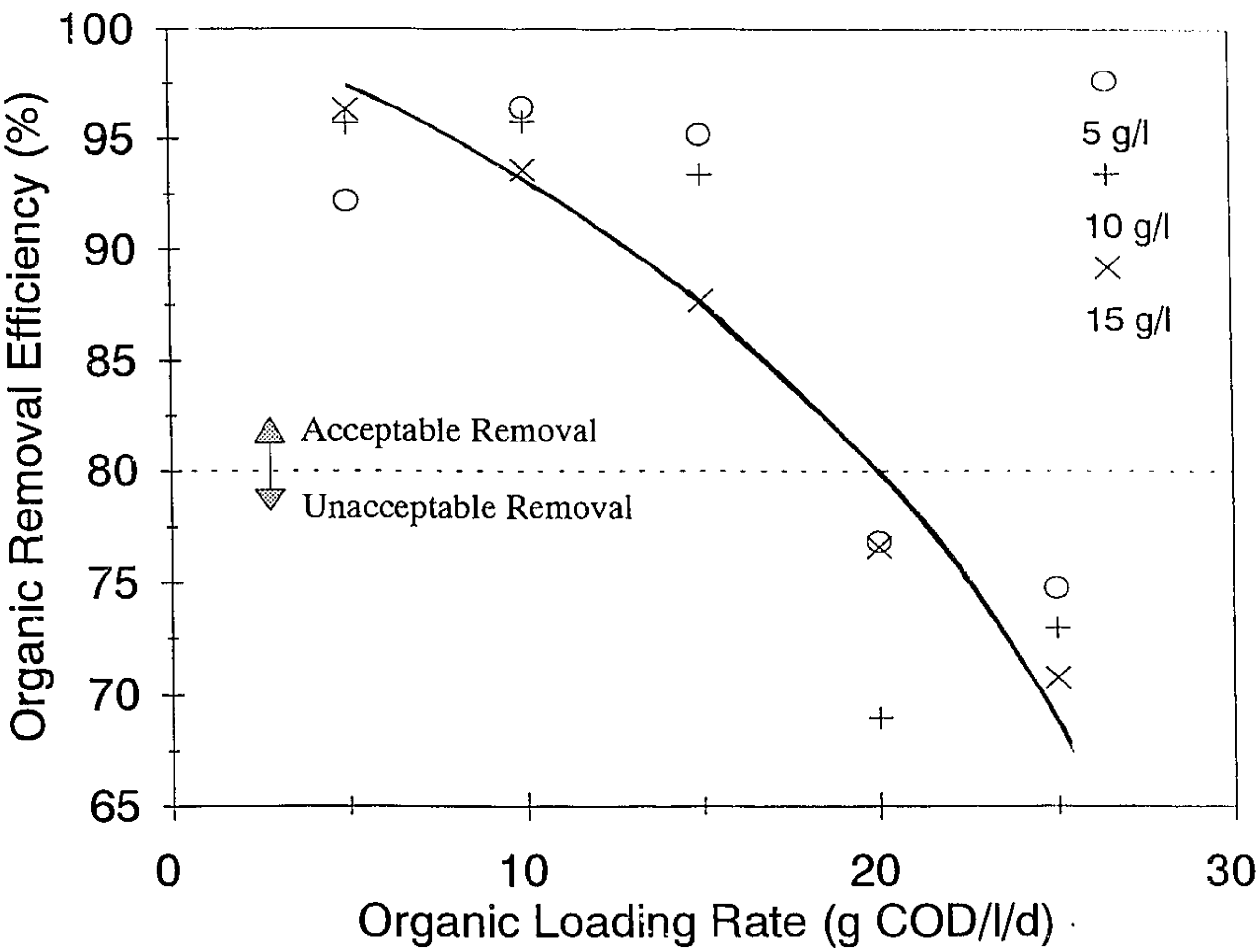

Figure 16. Organic Removal Efficiency (based on soluble COD) vs. Organic Loading Rate

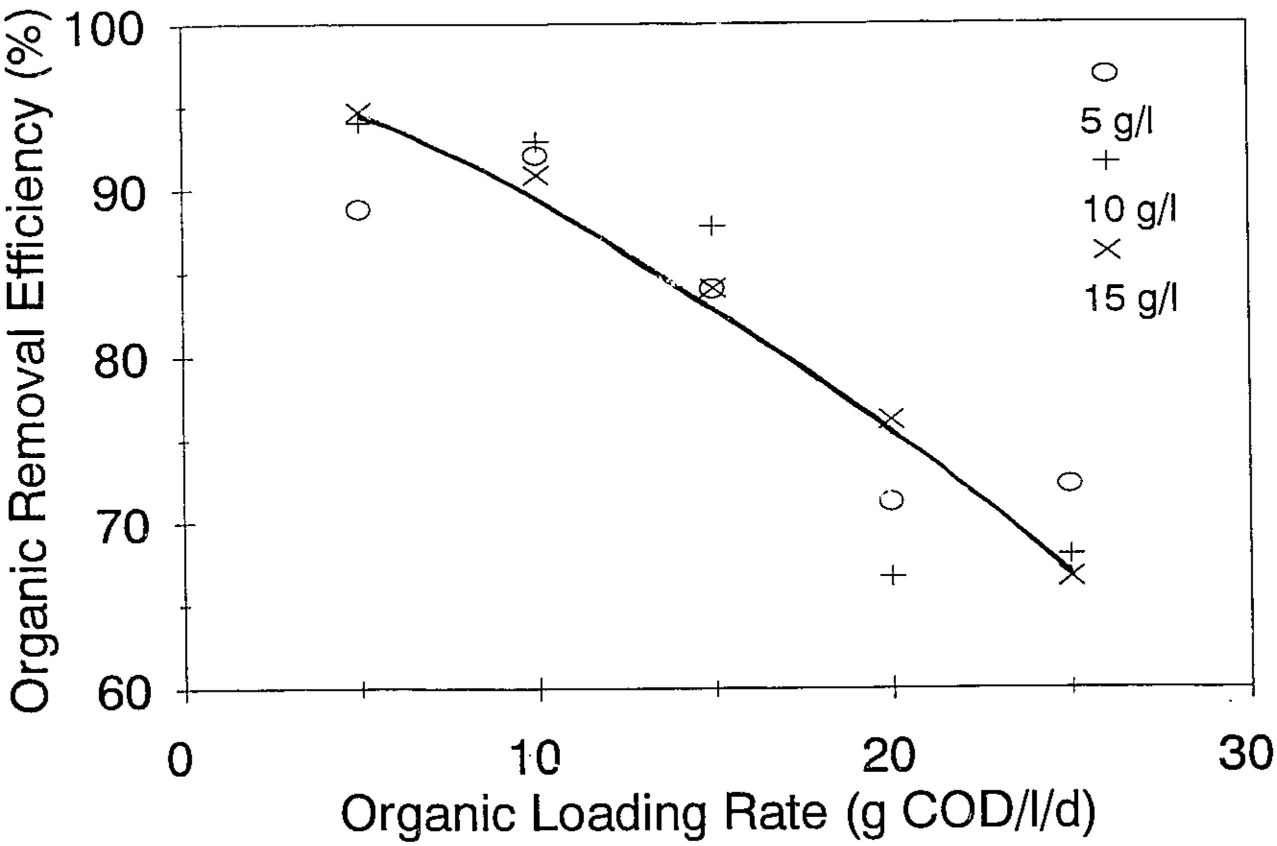

Figure 17. Organic Removal Efficiency (based on total COD) vs. Organic Loading Rate

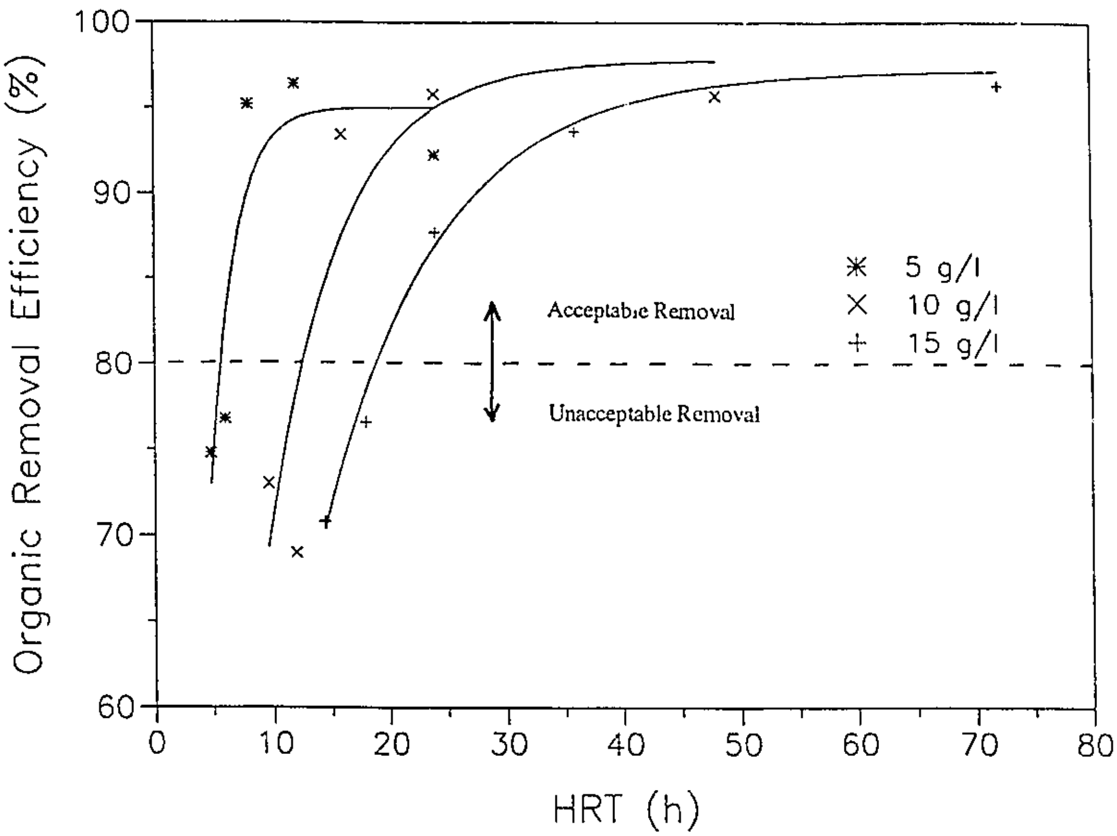

Figure 18. Organic Removal Efficiency (based on soluble COD) vs. HRT - Influence of influent substrate concentration

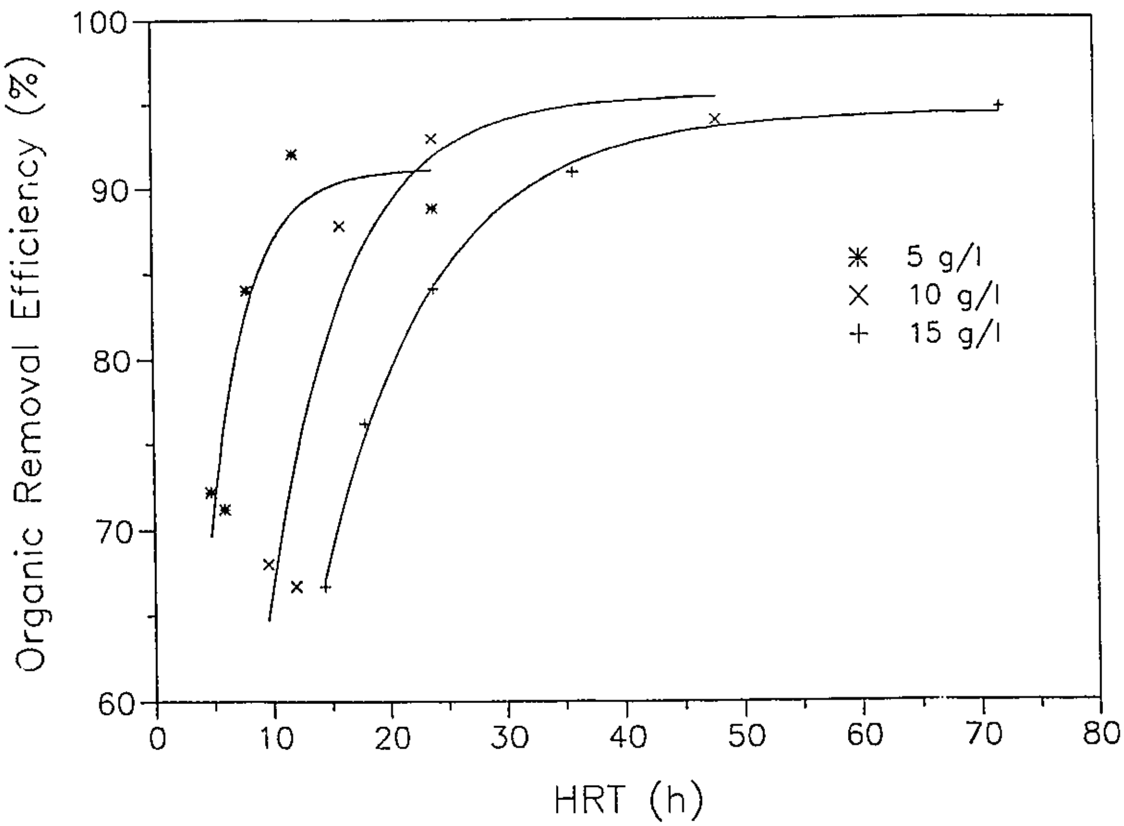

Figure 19. Organic Removal Efficiency (based on total COD) vs. HRT - Influence of influent substrate concentration

concentration. This suggests that the relationship between removal efficiency and HRT is dependent on influent concentration. This is not the case for Figures 16 and 17 where only a single curve is plotted to represent the functional relationship between removal efficiency and organic loading rate.

Below organic load of 20 g/L/d the percent removal based on soluble effluent COD is >80%. Thus, for sucrose feed, loadings of over 20 g/L/d may not be practical due to unacceptably low removal rates of a completely biodegradable feed. As seen in Figures 18 and 19, at substrate concentration of 5 g/L, the drop in removal efficiency occurs rapidly below 10 h HRT. As the feed concentration is increased however, this lowering of removal efficiency occurs more gradually and begins at higher HRTs.

5.3.4 Biomass Concentration

In an AFBR, biomass is present in the form of a biofilm which coats support carrier particles. Figure 20 is a plot of reactor attached biomass vs. organic loading rate. Here a single curve is shown since no dependence on influent concentration was apparent. As can be seen, attached volatile solids increase with higher organic loading rates. For a high loading of 25 g/L/d, biomass reached a maximum of 320 mg COD/g Biolite (230 mg VFS/g Biolite). Since each reactor contained 150 g of Biolite, this translates to 69 g VFS/L of expanded bed. This high concentration of biomass is possible because of the large surface area available for biofilm growth in AFBRs and results in high treatment efficiency at low HRTs.

Although the shape of the curve depicted in Figure 20 indicates an exponential

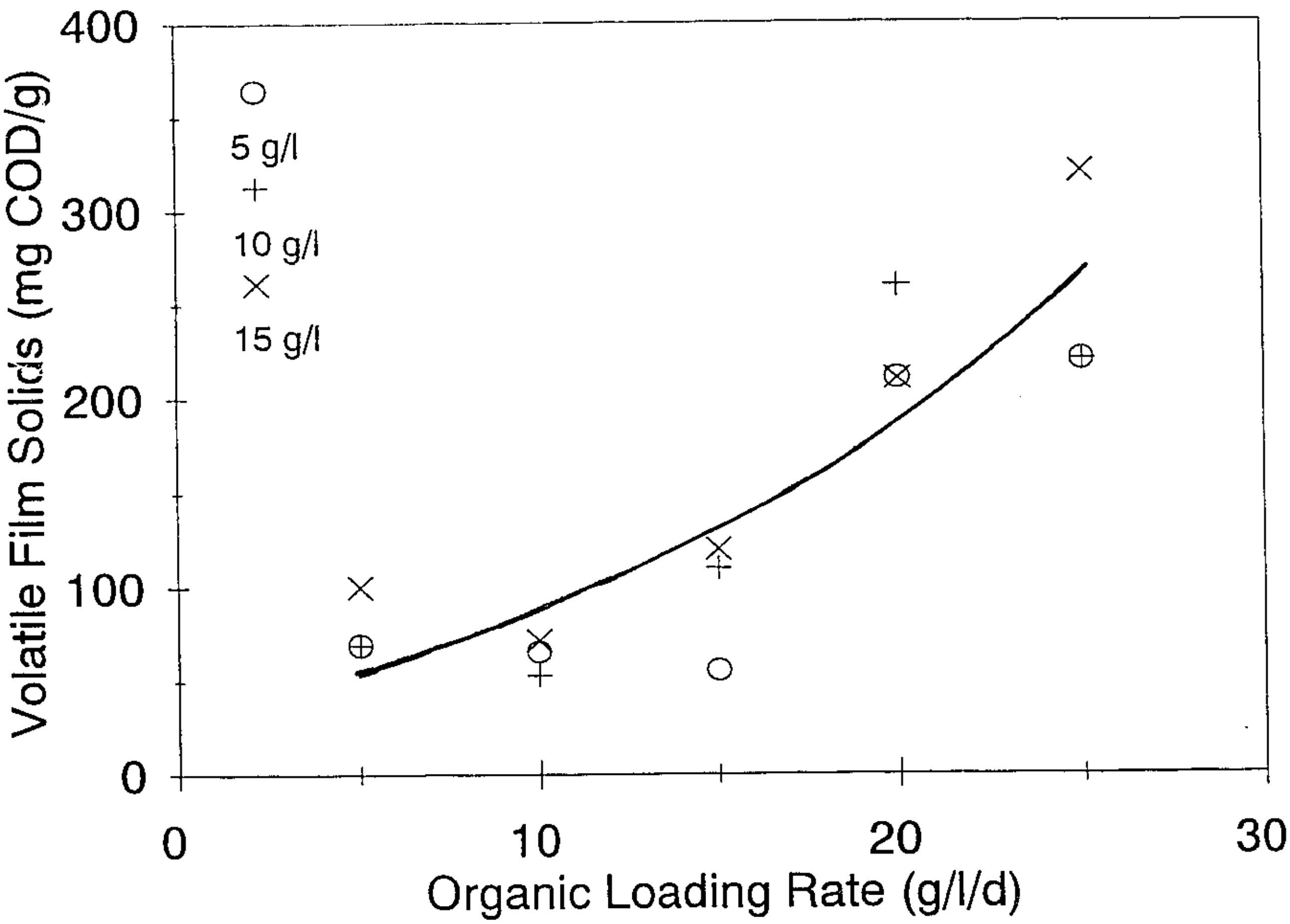

Figure 20. Volatile Film Solids vs. Organic Loading Rate

increase of biomass with organic loading, it is expected to plateau when the ability of carrier particles to support the film has reached its maximum. Also, as the bioparticle gets larger, shear forces acting on the surface of the film become increasingly important resulting in sloughing off of portions of the biofilm. This limits the amount of biomass that can be accumulated in the reactor. The highest biomass concentration of 69 g VFS/L of fluidized bed is higher than the value of 40 g VFS/L reported by Heijnen (1983) and Biver (1984) using sand as carrier particles.

Figure 21 is a plot of volatile film solids against HRT in days. For the same HRT, higher influent substrate concentration results in higher reactor biomass levels. VFS levels increase with decreasing HRT due to the greater availability of growth limiting substrate.

Biomass measurements were for well attached films since bioparticles were washed prior to determining biofilm COD. Biomass in the fluidized bed is also present as loosely attached biomass and biomass entrapped in the interstitial spaces between particles. However this type of biomass has been found to make up only a small fraction (4-6%) of the total reactor biomass for expanded beds (Switzenbaum, 1978). In the case of fluidized bed reactors, this fraction can be expected to be even smaller due to the higher superficial velocity and bed expansion. This may be easily verified using steady state data given in Table 11. For example, at a feed concentration of 5 g COD/L and HRT of 1 d, the attached biomass and effluent suspended solids are 69 mg COD/g Biolite and 170 mg COD/L respectively. Since there are 150 g of Biolite in the reactor, the total attached biomass is 10.4 g COD. Noting that the empty bed reactor volume is 1.3 l, the fraction of suspended solids of the total biomass present is only 2%. For this reason, entrapped

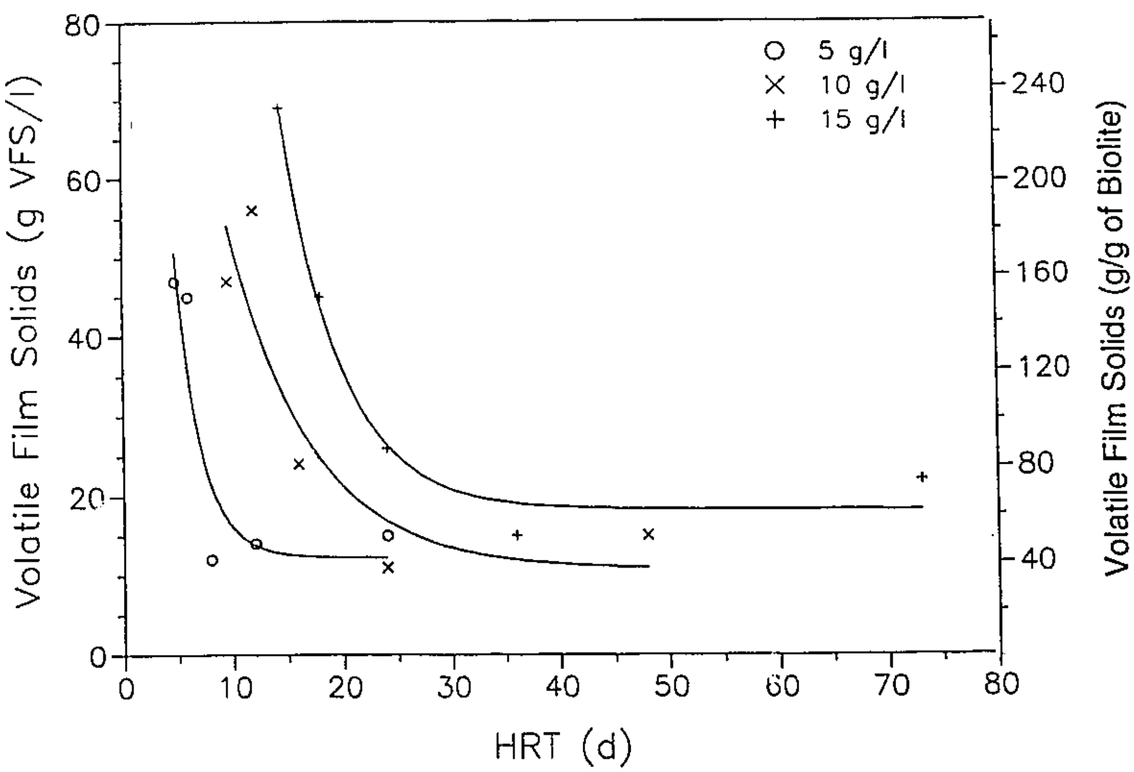

Figure 21. Volatile Film Solids vs. HRT - Influence of influent substrate concentration

biomass was not measured and assumed to be negligible.

Figures 22 and 23 show photographs of bioparticles removed from the upper and lower portions of the bed of an AFBR, respectively. These particular photographs were taken when this reactor was loaded at 25.0 g/L/d using sucrose feed of 15 g COD/L. Due to lower shear forces and fewer collisions between neighbouring particles, attrition of biofilm at the top of the bed is significantly lower than at the bottom of the bed. This explains the difference in biofilm thicknesses of 250 μm and 70 μm of particles removed from the top and bottom portions of the fluidized bed respectively. These numbers compare well with biofilm thickness of 60-200 μm sited by Heijnen et al. (1989) but are much higher than those reported by Switzenbaum (1978), for expanded bed reactors treating low strength waste, which never exceeded 15 μm.. The fraction of volatile solids in the sludge was determined to be 0.73. The calculated densities of the top and bottom portions of the bed were 0.13 g/cm^3 and 0.27 g/cm^3 respectively (see Appendix C for more details) which is comparable to the value of 0.15 g VSS/cm^3 given in a study by Switzenbaum and Jewell (1980).

5.3.5 Effluent Suspended Solids

An important variable for effluent quality is the suspended solids concentration. Values for this parameter are listed in Table 11. The effluent suspended solids concentration as a function of HRT is shown is Figure 24. Suspended solids in the effluent increase as HRT is lowered. This happens because any biomass which gets detached from carrier particles is more quickly washed out of the reactor at low HRT

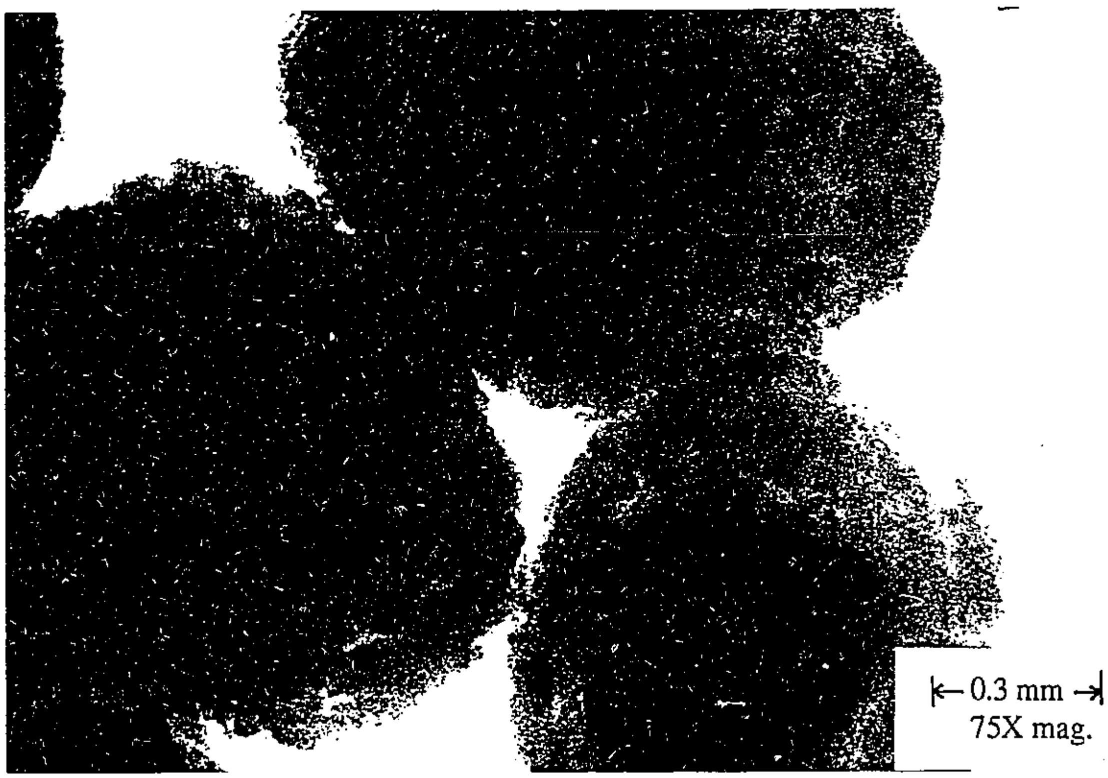

Figure 22. Photograph of Biolite Particles with Thick Biofilms (from upper portion of bed)

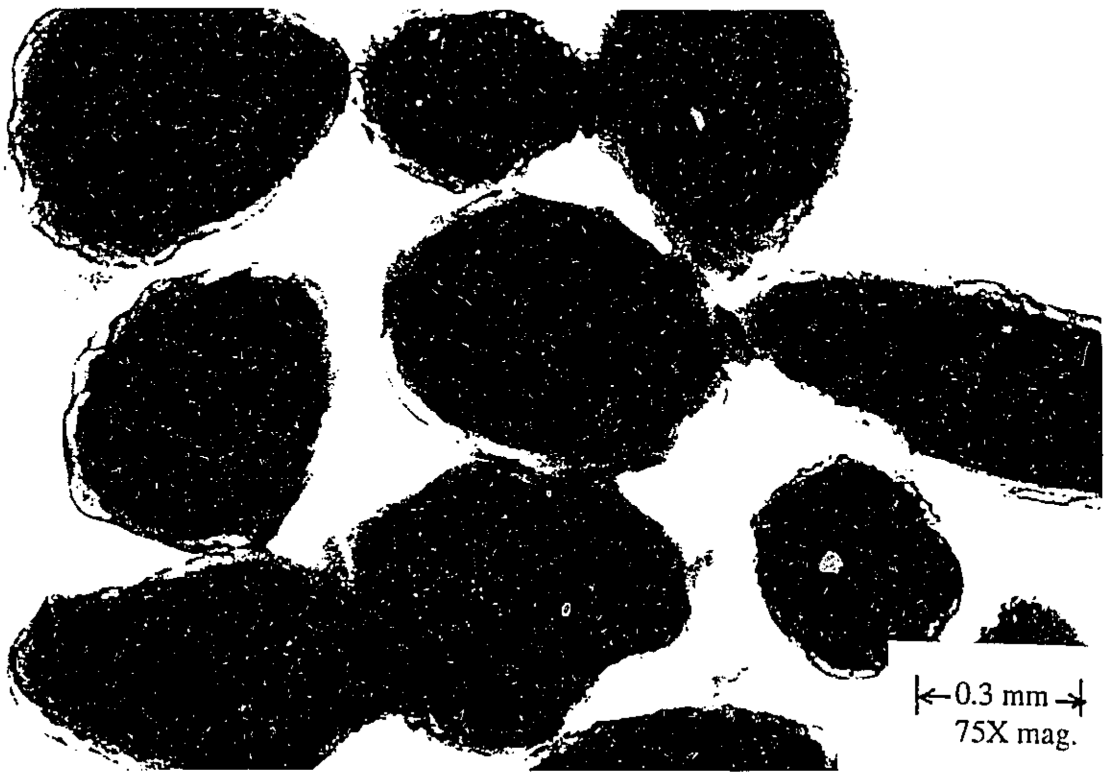

Figure 23. Photograph of Biolite Particles covered with Biofilms (from lower portion of bed)

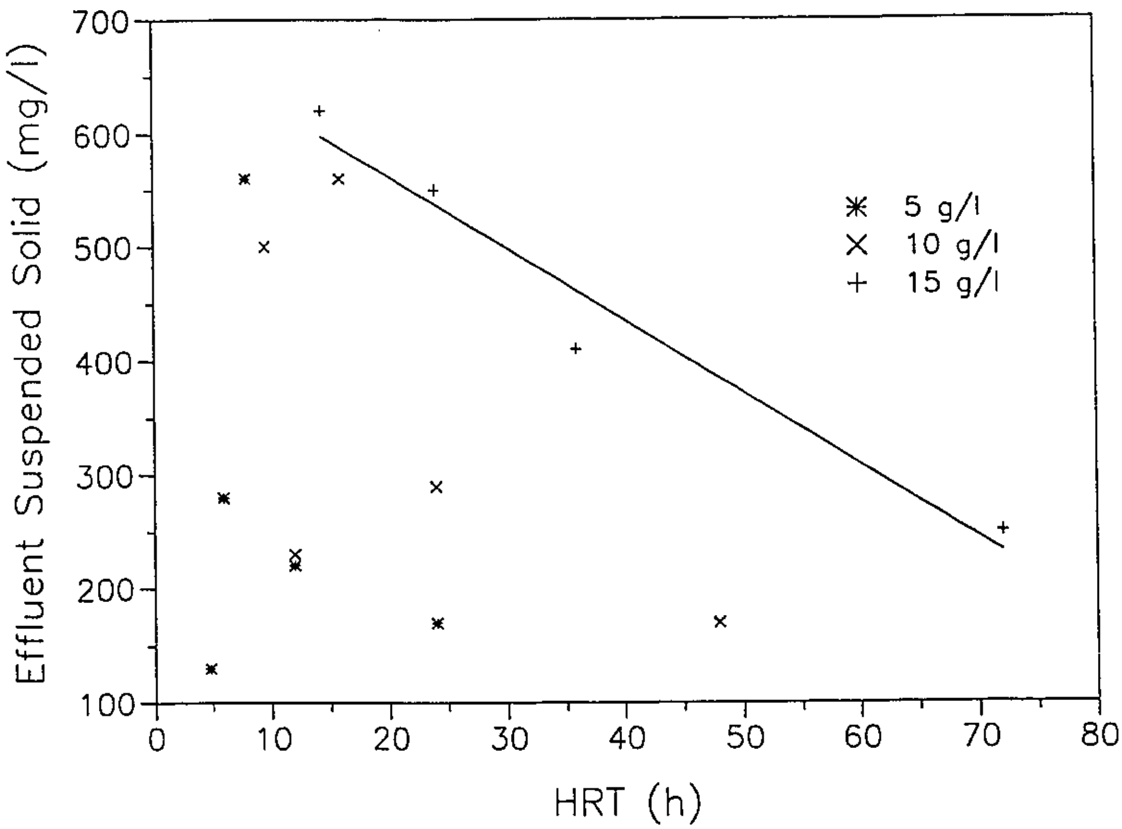

Figure 24. Concentration of Effluent Suspended Solids vs. HRT - Influence of influent substrate concentration

conditions. Suspended solid levels increase with an increase in influent concentration because higher concentrations provide an organically enriched environment for microorganisms to grow in. Due to the lack of a strong linear relationship, straight lines have not been shown for influent concentrations of 5 and 10 g/L sucrose.

5.3.6 Refractory Organics

As mentioned earlier, the anaerobic degradation of complex substrate can be regarded as occurring in three steps: hydrolysis, fatty acid production, and methane generation. The first two steps take place relatively quickly for a substrate like sucrose, the production of methane from VFAs being the rate limiting step. This is confirmed by data presented in Table 14 in which VFA levels are shown along with the fraction of VFA (as acetate) of the effluent. The average fraction of effluent COD which is VFA for the 15 steady state runs performed is 0.80. Of these VFAs 70% is in the form of acetic acid while the remaining 30% is propionic acid. Thus the majority of effluent COD is biodegradable with only a small fraction possibly being of refractory nature. Since the feed is 100% biodegradable, the refractory organics are probably of microbial origin.

5.3.7 Biogas Production

As mentioned earlier, one of the advantages of anaerobic processes over aerobic ones is the production of methane which can be used for fuel. The relationship between methane production and organic loading rate is shown in Figure 25. As this figure indicates, the amount of methane produced by the methanogenic bacteria increases linearly

Table 14. Effluent VFA levels

Feed COD (g/l)	HRT (h)	Acetic Acid (mg/l)	Propionic Acid (mg/l)	Total VFA as Acetate (mg/l)	Fraction of VFA in Effl.	Propionic Acid fraction of VFAs
5	24	198	23	300	0.82	0.11
5	12	86	41	144	0.86	0.43
5	8	108	51	180	0.80	0.43
5	6	660	160	1084	1.00	0.22
5	4.8	643	306	1071	0.91	0.43
10	48	120	8	176	0.44	0.07
10	24	47	19	85	0.22	0.34
10	16	309	147	515	0.83	0.43
10	12	1600	210	2450	0.85	0.13
10	9.6	1500	400	2500	0.99	0.24
15	72	340	39	515	1.00	0.11
15	36	450	310	940	1.05	0.49
15	24	311	355	790	0.46	0.67
15	18	1424	132	2126	0.65	0.09
15	14.4	2700	790	4570	1.12	0.26
				Average	0.80	0.30

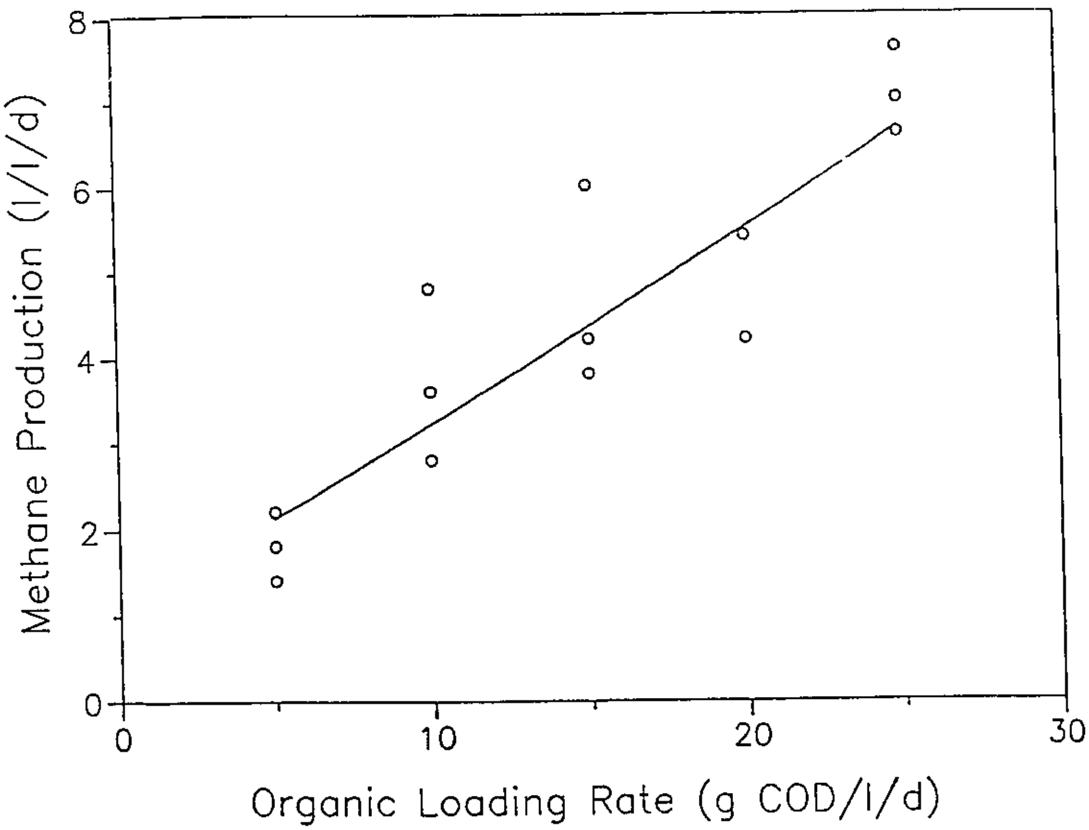

Figure 25. Methane Production vs. Organic Loading Rate

with loading rate. Also, from the previous comparison of COD removed to methane produced (Table 13), 89% of COD removed is used towards the production of methane gas.

5.4 Determination of Steady State Kinetic Constants

5.4.1 Computation of Specific Film Utilization Rate and Net Specific Film Growth Rate

The values of net specific growth rate (μ) and specific film utilization rate (U) are listed in Table 15. Values of μ' were obtained by dividing the mass leaving the system (mg VSS per day) by the mass in the system (mg of total volatile solids). Since the feed to the units was totally soluble, that is with no particulates, it can be assumed that no biomass is being introduced with the feed. Also implicit in the calculation of the kinetic parameters is the assumption that the system is at steady state. For biomass concentration this condition is satisfied since it remains almost constant during the length of the evaluation testing. For example, if a biomass concentration of 30 g VFS/L and net yield of 0.05 are assumed and 80% of influent substrate is removed at a loading rate of 15 g COD/L/d, then the biomass would increase by less than 2% over a one day period. However if true steady state conditions have not been reached then the biomass leaving the system as VSS is not equal to the net growth rate due to biomass accumulation.

The solids retention time (SRT) is given by the inverse of the net specific growth rate and is a widely used design and operational parameter for biological waste treatment

Table 15. **Calculated Microbial Growth and Substrate Utilization Rates for various HRTs and Feed Concentrations**

Loading (g/l/d)	HRT (h)	Feed COD (g/l)	Net Specific Growth Rate (1/d)	Specific Utilization Rate (1/d)	SRT (d)
5	24	5	0.00817	0.3103	122
10	12	5	0.02234	0.6851	45
15	8	5	0.10060	1.1971	10
20	6	5	0.01778	0.3413	56
25	4.8	5	0.00985	0.3967	102
5	48	10	0.00407	0.3205	246
10	24	10	0.01813	0.8383	55
15	16	10	0.02545	0.5944	39
20	12	10	0.00590	0.2477	170
25	9.6	10	0.01894	0.3871	53
5	72	15	0.00277	0.2240	361
10	36	15	0.01265	0.6067	79
15	24	15	0.01528	0.5118	65
20	18	15	0.00127	0.3404	788
25	14.4	15	0.01076	0.2581	93

systems. As Table 15 shows, high SRT values were maintained in the AFBRs averaging 150 days. This means that even at low HRTs washout was avoided as the organisms were still growing at a rate greater than they were being removed. This is a very important factor in maintaining a stable and reliable system and represents a significant advantage of fixed film reactors over suspended systems.

The value of U was obtained by dividing the amount of substrate removed per day by the amount of biofilm biomass in the system. Here it is assumed that all of the biomass present in the reactor is active. Although this assumption is highly unlikely, it was made because the fraction of total biomass which is inactive is unknown. Thus, calculated values of specific substrate utilization rates are conservative since they are based on total VFS and are necessarily lower that those based on active biomass only.

In Table 15, the values of μ' and U calculated for a influent concentration of 5 g/L and HRT of 8 h are an order of magnitude higher than the rest. From Table 11 it is seen that at these conditions the attached biomass reading was lower than expected while the effluent suspended solids were higher. This may have been due to loss in biomass from the reactor due to a sudden adverse environmental condition. Thus the above mentioned values are probably outliers and have been left out in the determination of the yield and decay coefficients in the next section.

5.4.2 Comparison of Observed Yield to Measured Yield

Equation 12 shows the relationship between net growth rate and specific utilization rate. Since the values of U and μ' have been calculated, they can be used to find the biomass yield (Y) and decay coefficients (b). Figure 26 shows a plot of μ' vs. U. Using linear regression, the slope (Y) and y-intercept (-b) were determined to be 0.028 (g VFS/g COD removed) and $3x10^{-4}$ 1/d. Since the decay coefficient must be a positive number its value is assumed to be negligible. The insensitivity of the analysis can easily account for a small difference between an expected small positive number and the actual very small negative one obtained.

Using the comparison of COD removed to methane produced table (Table 13), a measured yield can be formulated by the difference between the COD removed and the methane produced. This difference is assumed to be converted to biomass and hence is a measured yield. The calculated average measured yield is 0.079 (g VFS/g COD removed) which much higher than the observed yield of 0.028 (g VFS/g COD removed). Futhermore, during startup about 2.5 g VSS/L were produced during the first 30 d at an average loading of 1.0 g/L/d (reactor 1 in Table 9). From this a biomass yield of 0.08 may be calculated

Thus most of the evidence points towards a yield of about 0.1 g VSS/g COD. Due to the very long SRTs of the AFBR system used in this study, true steady state was probably not acheived. This means that biomass was still accumulating in the reactor resulting in low measured yields. Low yields of 0.1 demonstrate the low net synthesis of solids in the AFBR, which in the context of wastewater treatment is very attractive.

5.4.3 Biofilm Performance

Figure 27 shows a plot of specific utilization rate vs. specific loading. The straight

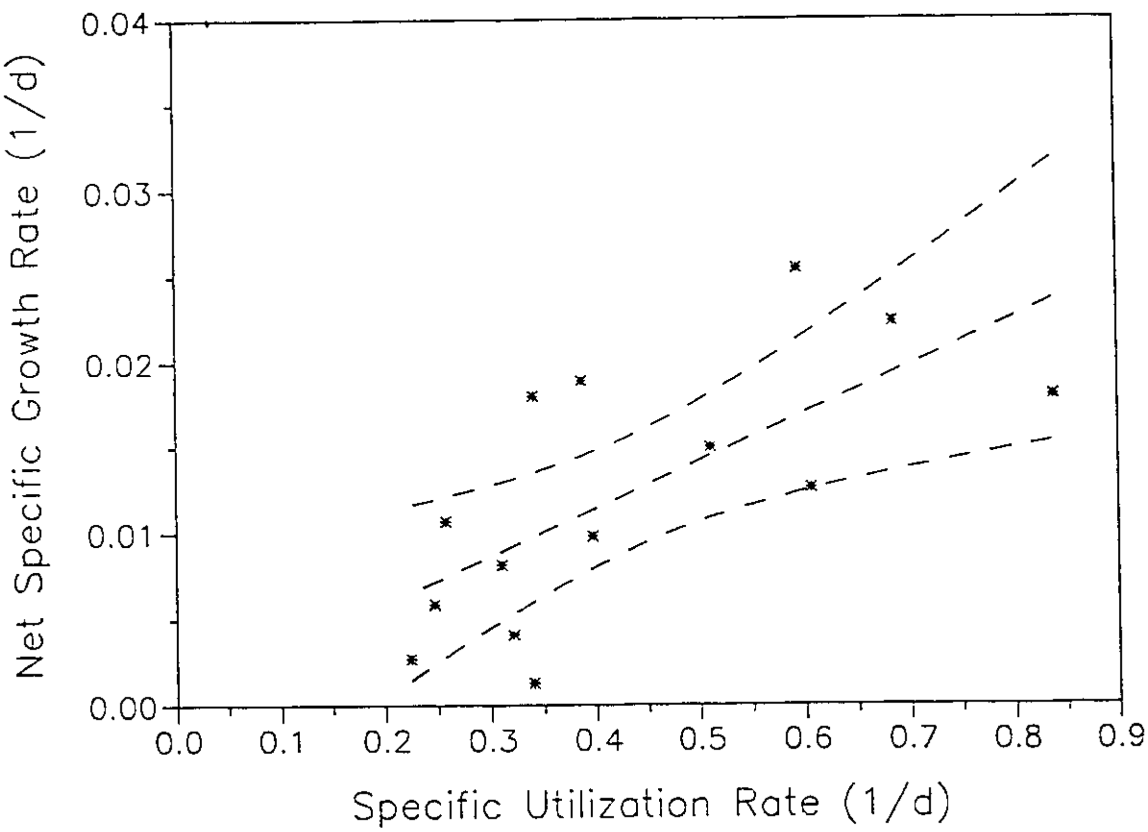

Figure 26. Net Specific Growth Rate vs. Specific Utilization Rate - showing linear regression line and 95% confidence interval

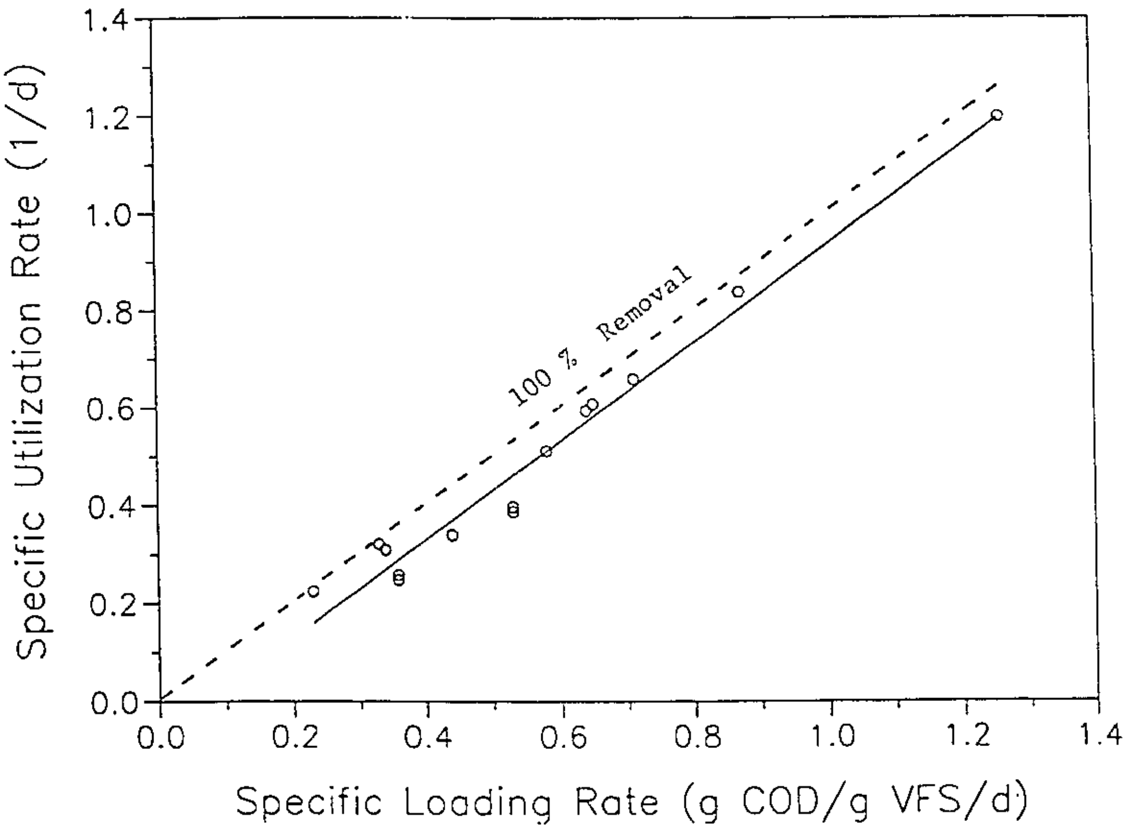

Figure 27. Specific Utilization Rate vs. Specific Loading Rate

solid line represents the least squares fit of the data and shows the high positive correlation between the plotted variables (r^2=0.97). Since the slope of the line is 1.01, each unit increase in specific loading results in one unit increase of specific utilization rate. As can be seen from the plot, this line lies below the 100% removal line, as expected.

To check possible correlation between values of SRT and HRT Figure 28 was plotted. Assuming a linear relationship, a slope of 100 and y-intercept of 13.8 d was obtained (r^2=0.69). Kennedy (1985) developed a washout factor (f) for fixed film reactors by taking the inverse of the slope of this plot. Using a DSFF reactor treating 10 g COD/L sucrose waste, he reported a f value of 0.115 which is about ten times higher than the value of 0.01 obtained from this study. This is the result of better adhesion of biomass in an AFBR compared to a DSFF reactor leading to much longer SRTs. Unlike Kennedy's (1985) findings, no apparent relationship between f and influent substrate concentration was observed in this study.

Thus solids in the reactor are retained longer when the reactor is operated at longer HRT. It is interesting to note even when HRT approaches 0.0 d, the SRT is still 13.8 d. This signifies again that even under high hydraulic loading conditions, there is minimal threat of biomass washout in an AFBR.

In Figure 29, reactor volatile film solids are plotted against substrate utilization rate. This plot is similar to Figure 18 where attached biomass is plotted against organic loading. Although the data points are scattered, the overall trend was an exponential increase in biomass with substrate utilization. This relationship was relatively unaffected by influent substrate concentration.

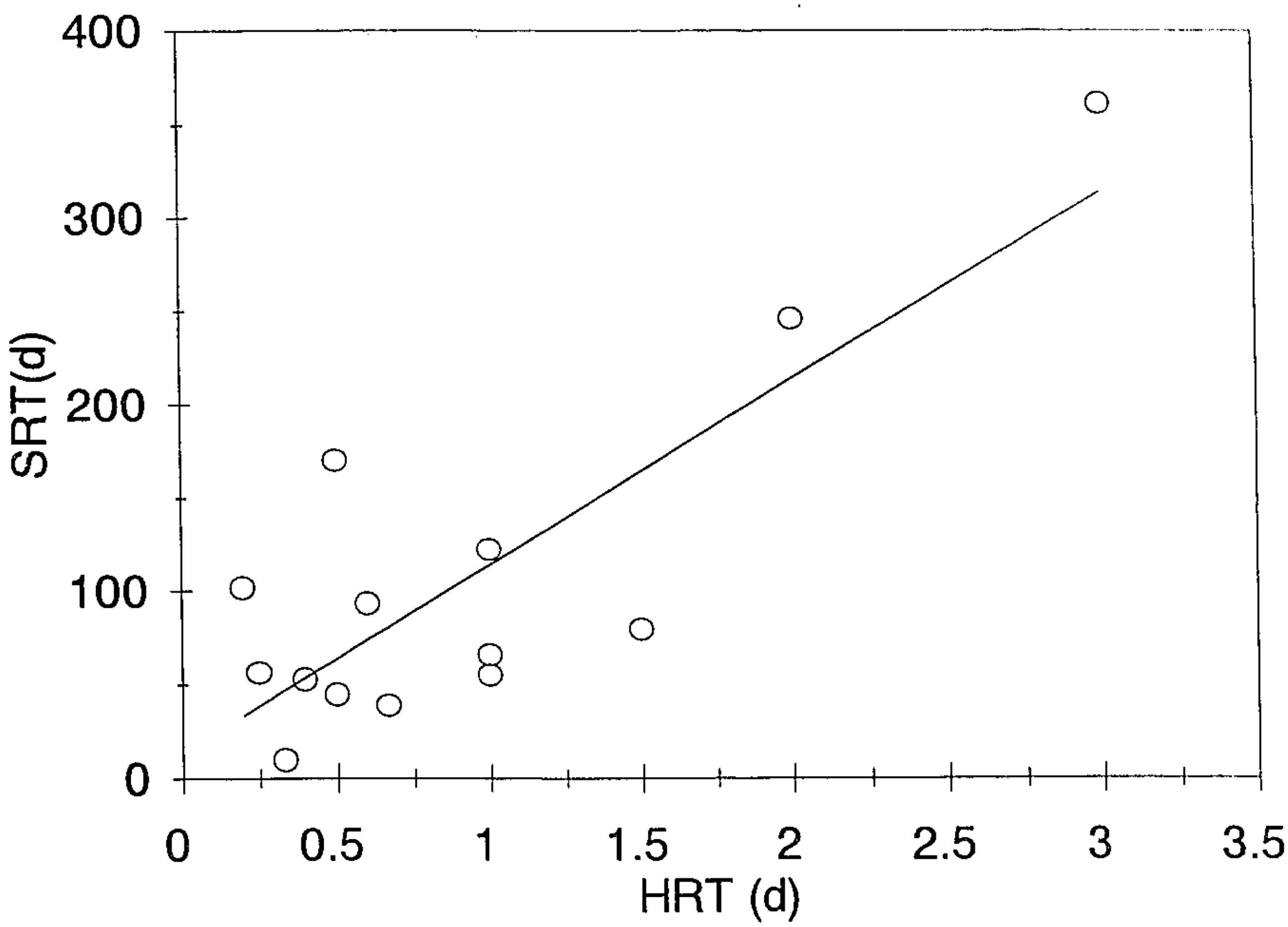

Figure 28. SRT vs. HRT - showing linear regression line

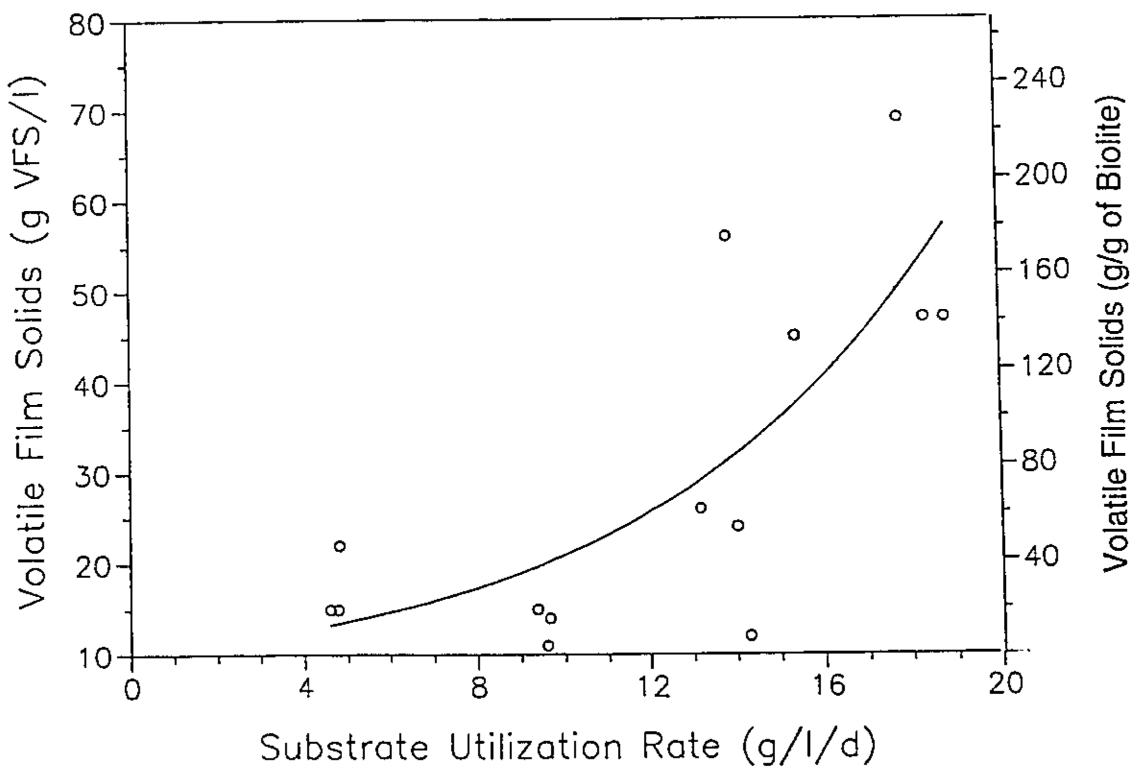

Figure 29. Volatile Film Solids vs. Substrate Utilization Rate

5.4.4 Determination of Maximum Specific Utilization Rate and Saturation Constant

Equation 17 was derived from the Monod equation and expresses the relationship between specific substrate utilization and effluent substrate concentration. This equation may be linearized according to the Hanes method and the value of the maximum specific substrate utilization (k) and saturation constant (K_s) determined using Equation 18. However, doing a linear regression on S/U vs. S data from this study leads to a negative y-intercept and consequently a negative K_s value (Figure 30). In the Monod model, the saturation constant is the substrate concentration of the effluent when the U is half its maximum value of k, and thus a negative K_s value has no theoretical meaning. Therefore, K_s was assumed to be approximately equal to zero. This gave a k value of 0.27 d^{-1}.

As indicated in Table 15, the mean value of the specific utilization rate is 0.48 d^{-1} which is higher than the value obtained using Hanes Method. Thus the value of k obtained using the Hanes method was not truly the maximum substrate utilization rate. In order to see the influence of influent substrate concentration on U, mean values for each concentration were calculated. For feed concentrations of 5, 10 and 15 g COD/L these values are 0.59, 0.48, 0.39 d^{-1} respectively. Using the yield value of 0.1, the corresponding maximum growth rates are 0.059, 0.048 and 0.039 d^{-1}.

5.5 Batch Kinetic Studies

Following steady state sucrose experiments, three batch runs were performed

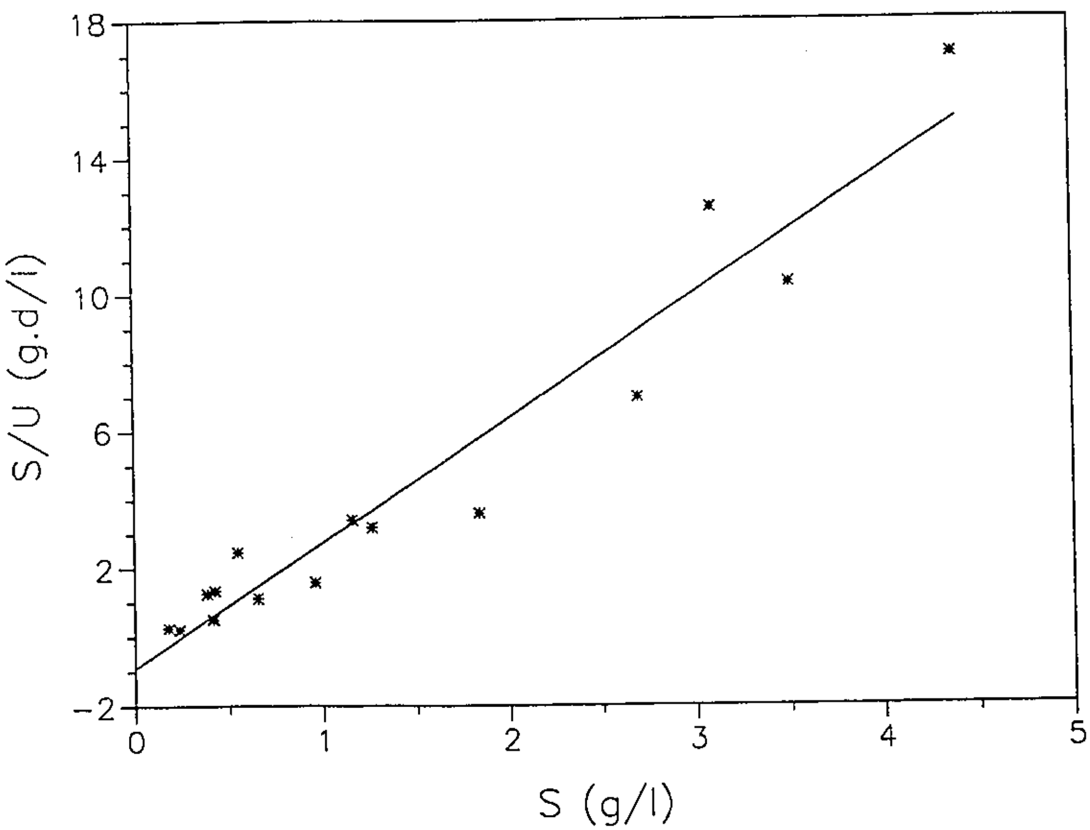

Figure 30. Hanes Plot for the determination of constants k and K_S

using three different initial concentrations of acetate. The acetate level of the reactor mixed liquor at several time intervals are presented in Table 16. Acetate concentration vs. time data is plotted in Figure 31. A linear relationship exists between these two variables meaning that the reaction is zero order and independent of acetate concentration. The average value of the reaction rate constant or slope for the three trials is 2.39 g/L/d. Since the biomass in the reactor is 200 mg COD/g of Biolite and is assumed to be constant for the duration of all 3 trials, the net specific utilization is 0.14 d^{-1} (See Appendix B for details on calculations).

From Figure 31, it can be seen that specific substrate utilization rate remains constant even at substrate mixed liquor concentrations of 200 mg/L. This supports the conclusions reached from the analysis of the steady state data that the value of K_S is very small. From the data of trial 3 shown on Figure 30, K_S may be estimated at 100 mg/L.

In anaerobic reactors degrading sucrose based wastewater, the population of acidogenic bacteria is about three times higher than that of methanogenic bacteria since the activity of the former group of bacteria is three times lower than the later group (Shieh et al., 1985). If such a reactor is suddenly switched to acetate based feed for which only methanogenic bacteria are required, then only a quarter of the reactor biomass is active. Using the mean value of U determined from steady state experiments (0.48 d^{-1}) the expected U of the total biomass in the batch studies is 0.24 d^{-1}. When determined from batch tests, a further lowering of this value occurred due to the recalcitrant organic metabolic products produced from the endogenous decay of acetogenic biomass. This explains the discrepancy between the biomass activities determined from steady state and

Table 16. **Acetate Concentration of Reactor Mixed Liquor vs Time Data for Batch Kinetics Study**

	Acetate Concentration (mg/l)	Time (h)
Trial 1	899	0.00
	875	0.50
	833	1.00
	738	1.50
	645	2.00
	575	3.00
	468	4.00
	363	5.00
	264	6.00
	167	7.00
Trial 2	534	0.00
	425	1.00
	307	2.00
	204	3.00
	153	4.00
	92	5.00
Trial 3	1160	0.00
	1095	1.00
	1002	2.00
	907	3.00
	790	4.00
	606	5.83
	514	6.83
	380	7.83
	288	8.83

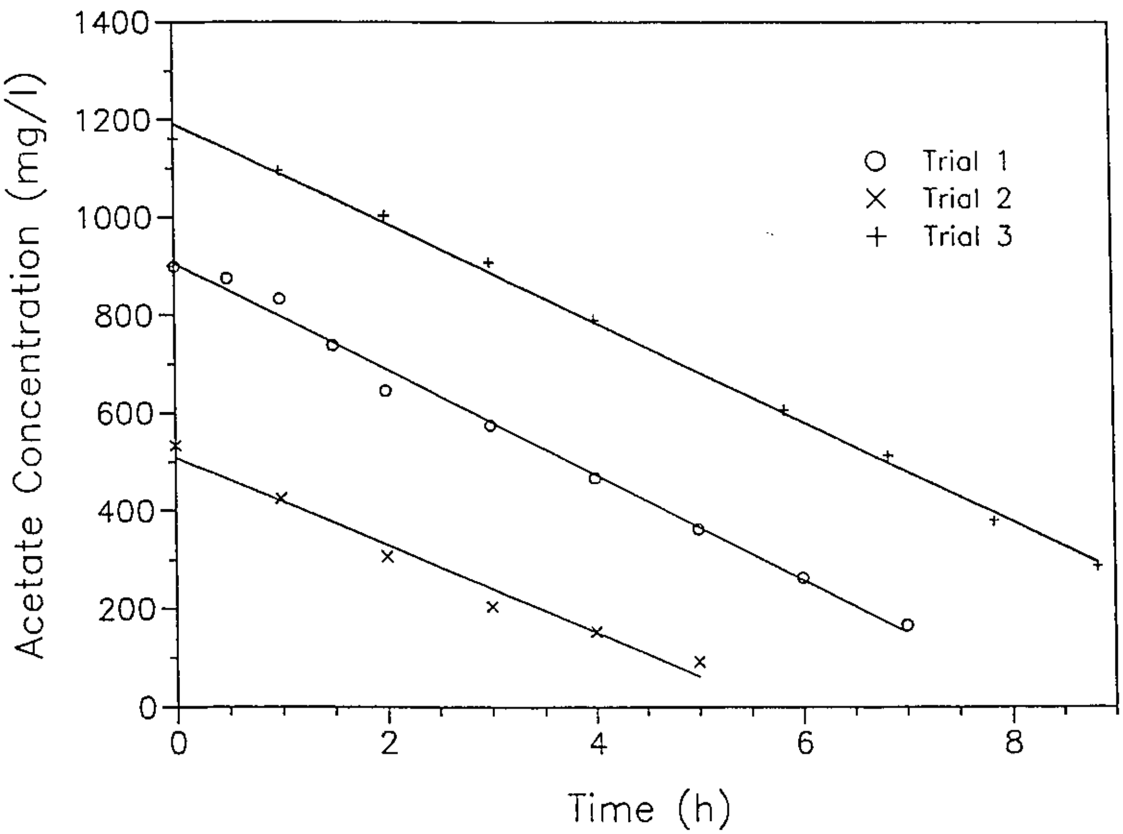

Figure 31. Concentration of Acetate in Reactor Mixed Liquor vs. Time - For 3 different initial acetate concentrations

batch tests.

5.6 Comparison of Kinetic Constants with Literature Values

Table 17 below presents values of the kinetic constants determined in this study along with those found in the literature for anaerobic treatment systems. The mean value of the maximum specific growth rate found in this study is lower than those found by Anderson and Duarte (1980) and Kennedy (1985). This is due to the calculated k value of this study which is conservative since it is the mean of specific utilization rates found. Values of Y and b are comparable to literature values listed.

Table 17. Kinetic Constants of Anaerobic Processes

μ_m (d^{-1})	Y (g VSS/ g COD)	k (g COD/g VSS/d)	b (d^{-1})	K_S (mg/L)	Temp-erature (°C)	Type of Waste	Reference
0.16	0.21	-	0.03	200	35	Glucose	Anderson and Duarte (1980)
-	0.14	0.9	0.04	4000	35	Palm Oil	Chin (1981)
-	0.046-0.076	-	-	-	30	Synthetic Stillage	Dahab and Young (1981)
-	0.12	-	0	-	30	Sucrose	Switzenbaum (1978)
0.22	0.18	0.93	0.06	300	35	Sucrose	Kennedy (1985)
0.05	0.1	0.48	≈0.0	100	35	Sucrose	This Study

5.7 Determination of Rational Design Relationship Equations

Using the values of the kinetic constants determined in this study, the system's steady state removal efficiency and biomass concentration may be predicted using Equations 20 and 21. Since, due to lack of sensitivity, no value of the decay coefficient was determined, a b value of 0.02 will be assumed. Figure 32 shows the predicted and measured values of organic removal efficiency plotted against HRT. Predicted curves for influent concentrations of 5 and 10 g/L fit the data reasonably well. However for feed COD of 15 g/L the model predicts removal efficiencies over 95% for HRTs down to 0.7 d which is not observed experimentally. Figure 33 shows predicted and measured values of attached biomass over the range of HRTs investigated. Although the predicted curves do not fit the data well, the general trend of decreasing VFS with increasing HRTs is followed.

5.8 Treatment of Landfill Leachate using AFBR

Landfill leachate is the wastewater formed when water percolates down into a landfill picking up organic carbon and various other contaminants. The leachate treated in this study came from a municipal landfill located in the city of Nepean, Ontario.

Original leachate characteristics are listed in Table 5. Since the BOD_5:COD ratio is high (0.86), this landfill may be classified as "young". About 58% of BOD_5 is in the form of soluble and readily degradable VFAs. The leachate contains very little VSS which make up only 5% of total COD. Prior to treatment, the leachate was enriched in

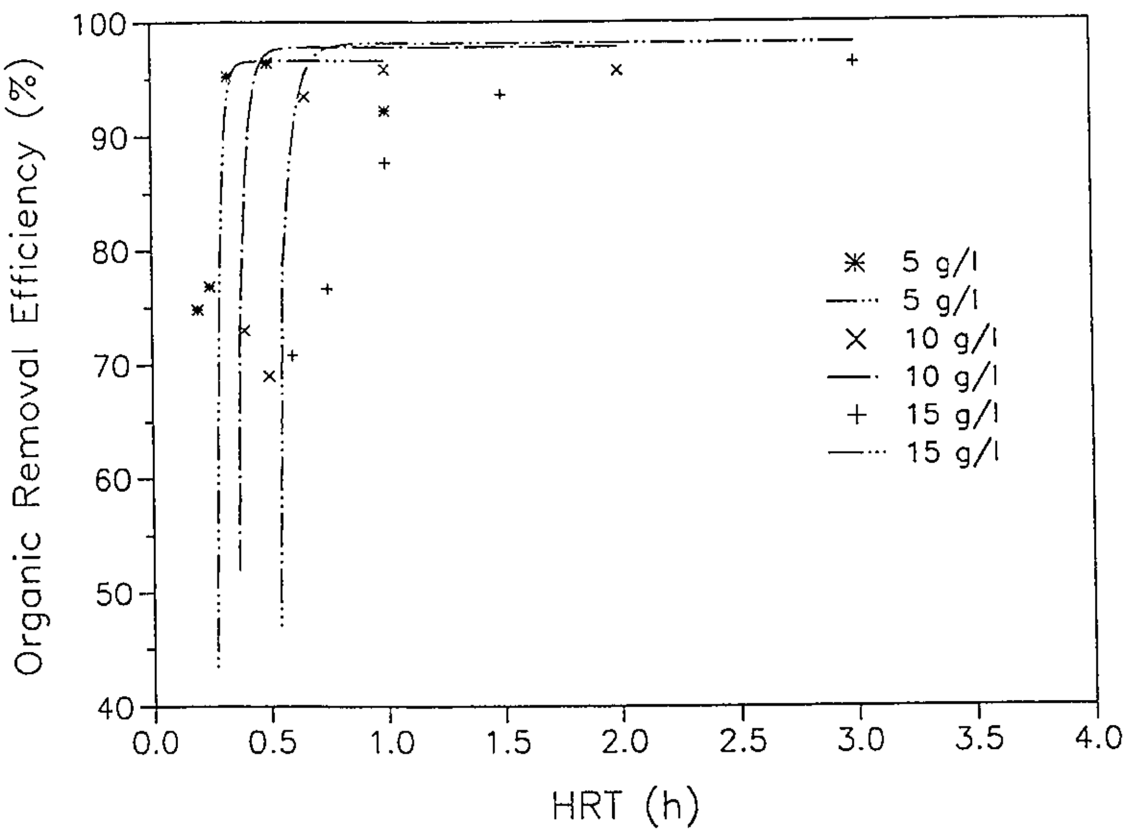

Figure 32. Organic Removal Efficiency vs. HRT - Experimental and Predicted values

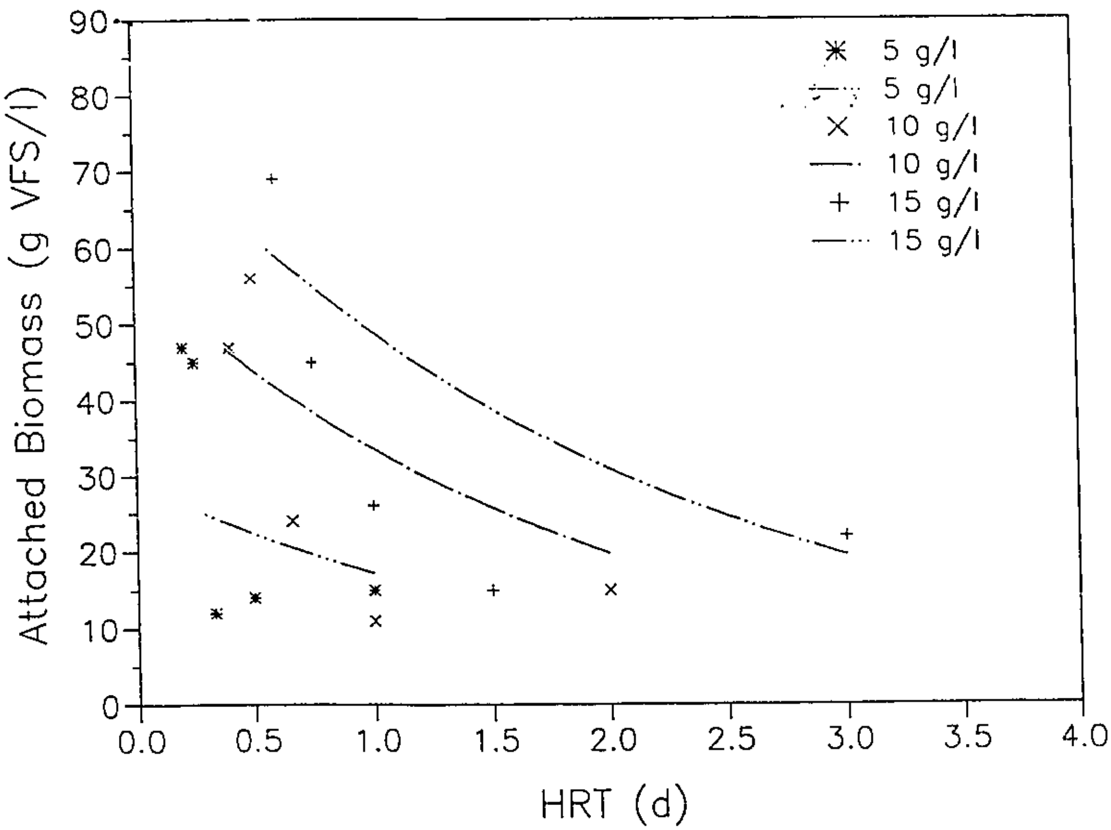

Figure 33. Attached Biomass Concentration vs. HRT - Experimental and Predicted values

phosphorus by the addition of di-potassium hydrogen phosphate to bring the COD:N:P ratio to 300:14:1. Leachate containing 5 g COD/L and 2 g COD/L were each treated at 4 different organic loading rates. These concentrations were made by suitably diluting the original leachate.

5.8.1 Summary of Steady State Performance of AFBR Treating Leachate

Summary data for two leachate organic concentrations used in this treatability study are presented in Table 18. Results are presented for total and soluble effluent, attached biomass and gas production. As in the case of steady state experiments using sucrose feed, each result represents the average of three sets of data taken on three consecutive days.

The COD removal based on total COD considers the small amount of volatile suspended material in the influent leachate and also the VSS in the mixed liquor which includes biomass. Removal based on soluble COD only considers the soluble COD in the influent leachate and reactor mixed liquor. As Table 18 indicates, the steady state attached biomass concentration and COD removal efficiencies for leachate treatment are significantly lower than those for synthetic sucrose feed. Comparing the highest removal efficiency of 87% (based on soluble COD) obtained here with the BOD_5:COD ratio of 0.86 of the leachate, it may be concluded that at low leachate loadings, inhibition due to toxic compounds present in leachate are negligible. At higher loadings or lower biodegradable organic concentrations, removal efficiencies decrease.

Table 18. Summary of Steady State Data for Leachate Feed

Loading (g/l/d)	Feed COD (mg/l)	HRT (h)	Total Effluent COD (mg/l)	Soluble Effluent COD (mg/l)	Suspended Effluent COD (mg/l)	Attached Solids COD (mg/g)	Gas Production (l/d)	% Removal (based on total COD)	% Removal (based on soluble COD)
2.1	2000	22	633	553	80	33	0.4	68	71
4.3	2000	11	730	573	157	30	0.4	64	70
6.4	2000	8	823	660	163	17	0.5	59	65
10.7	2000	5	843	597	247	36	1.6	58	69
2.7	5000	45	687	597	90	74	0.2	86	87
5.4	5000	22	1167	1077	90	44	0.8	77	77
8.0	5000	15	1250	1093	157	69	1.4	75	77
13.3	5000	9	1163	1033	130	46	1.6	77	78

5.8.2 Effect of HRT and Organic Loading Rate on COD Removal Efficiency

The influent COD concentration of 2.0 COD/L and the HRTs applied simulate dilute and higher volumes of leachate collected in the spring. During this time there is lot of run-off due to thawing snow and ice. The higher leachate concentration of 5.0 g COD/L and HRTs used are representative of leachate produced during the summer and fall periods.

Figures 34 and 35 show that AFBRs can treat leachate at high biodegradable removal efficiencies. The biodegradable organic removal efficiency is based on the BOD_5:COD ratio of 0.86 of the leachate. Figure 34 shows a plot of organic removal efficiency (based on soluble organics) of both COD and biodegradable organics vs. loading rate. The plots for each influent concentration fairly flat, indicating that removal efficiency does not vary significantly with organic load for the range of organic loading rates tested. Higher removal efficiencies were obtained for 5 g COD/L than for 2 g COD/L. leachate. Removal efficiency is plotted against HRT in Figure 35. Removal efficiency increases with longer HRTs however thet remain in a small range even when large changes in HRT are made.

5.8.3 Reactor Biomass

After the four months of steady state runs treating sucrose waste, AFBRs were

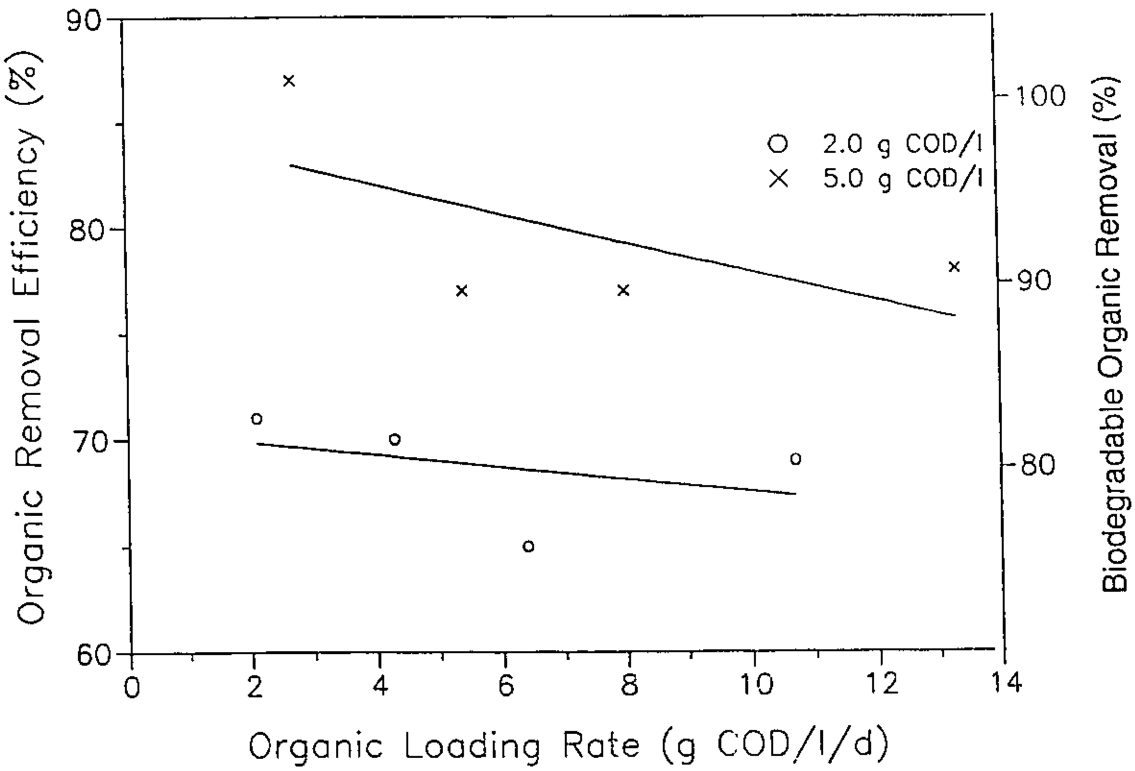

Figure 34. Organic Removal Efficiency (based on soluble COD) vs. Organic Loading Rate - Influence of influent leachate concentration

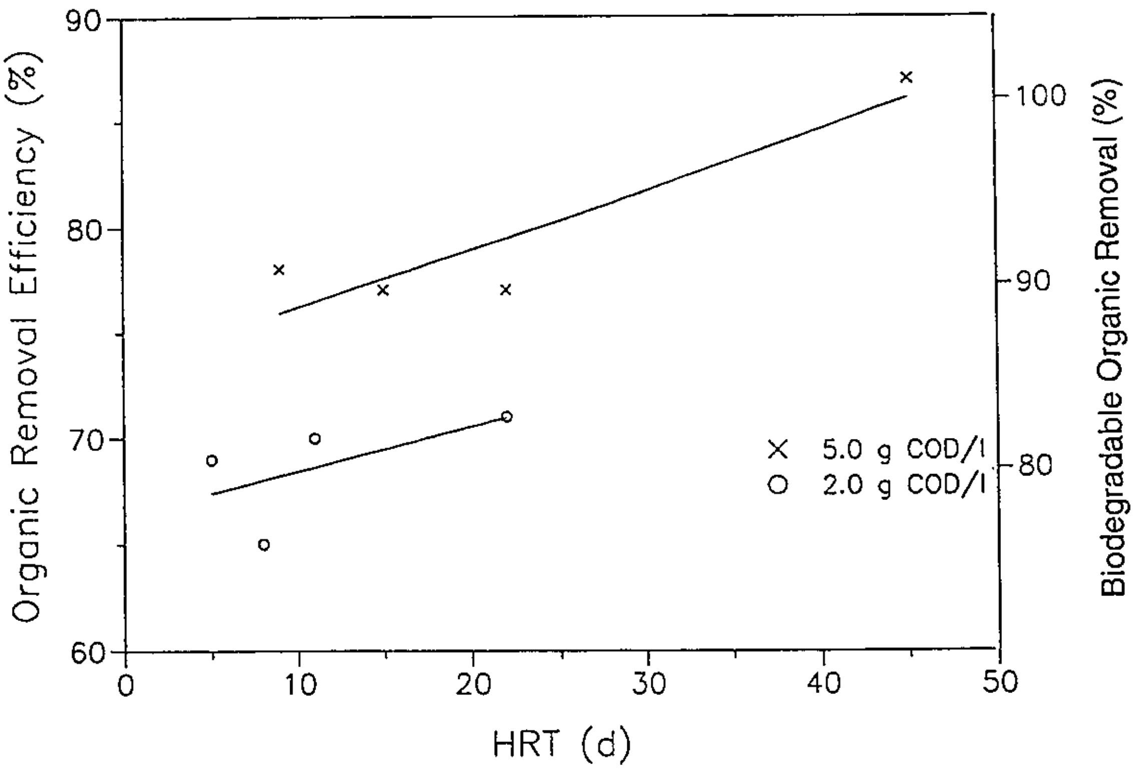

Figure 35. Organic Removal Efficiency (based on soluble COD) vs. HRT - Influence of influent leachate concentration

used for treating leachate . As leachate was fed to the system, loss of biomass was observable and the attached biomass color slowly changed from light grey to black. This dark appearance of the attached biomass was probably due to the precipitation of metal sulphides and carbonates over the carrier particles. This conclusion is supported by the fact that the recycle pump speed had to be steadily increased in order to maintain the same bed expansion level.

This "mineralization" of biomass has been reported by several other researchers (Iza et al., 1992; Kennedy et al., 1988; Mennerich, 1987) and poses a problem for leachate treatment using AFBR due to higher pumping costs required. One solution is to precipitate the heavy metals out of the leachate prior to treatment. This however represents an additional expense in equipment and chemicals.

For the reactor treating 5.0 g COD/L leachate at 13.3 g COD/L/d, film thickness measurements were made. Thickness values of 0.34 mm and 0.32 mm were measured for biofilms from the top and bottom portions of the fluidized bed respectively. Sample size used was 100.

For reactors treating sucrose at a loading of 25 g/L/d it was found that 73% of attached solids were volatile solids. Although efforts were made to determine a similar ratio for AFBR treating leachate, none was obtained due to the insensitivity of the method used. However assuming the same fraction of volatile solids for leachate as for sucrose feed, biofilm densities for the top and bottom portions of the bed are 0.52 g/cm^3 and 0.60 g/cm^3 respectively. These values are more than twice as high as those calculated for AFBR treating sucrose based feed. Actual biofilm densities are probably even higher than these calculated values due to mineralization of biomass which results in a higher fraction of fixed film solids.

CHAPTER 6. CONCLUSIONS AND RECOMMENDATIONS

Based on the results of this study, the following conclusions may be drawn:

1. Biolite is highly suitable material for use in AFBRs with favourable density, size and surface properties. Using Biolite as the inert carrier material in AFBRs results in superior startup and steady state performance.
2. Start-up using maximum efficiency profile is more stable than maximum load profile. Both loading profiles led to the development of healthy biofilm in a period of two months capable of treating >90% of the influent waste at a loading of 2.5 g COD/l/d.
3. When treating synthetic sucrose waste, organic removal efficiencies of over 80% are obtained at loading rates as high as 20 g/l/d. Removal efficiency decreased with increasing loading rate.
4. Organic removal efficiency of sucrose wastewater decreases with a lowering of reactor HRT. This relationship is influenced by influent substrate concentration.
5. For 5-15 g/l sucrose feed, AFBR follows zero order overall kinetics ($K_s \approx 0$) and has a maximum substrate utilization rate of 0.48 d^{-1}
6. An AFBR using Biolite media can accumulate large amounts of biomass (up to 220 mg VFS/g Biolite) resulting in high removal efficiencies at high organic loading rates.
7. Due to low yield (Y=0.1) and growth rate (μ_m=0.048 d^{-1}) solids production in an AFBR is low.
8. At low loading rates, the AFBR can effectively treat municipal landfill leachate with removal efficiencies up to 87% which represented all the biodegradable matter of the

particular leachate treated in this study.

9. When treating raw leachate, precipitation of metal sulphides and carbonates cause the reactor biomass to become mineralized. This leads to loss in sludge activity and need for higher superficial velocities to maintain bed expansion level. Thus application of AFBR to treat landfill leachate may be limited without appropriate pretreatment.

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

1. Conduct a detailed study, using several protocols, to determine ideal startup procedure for Biolite media.
2. Evaluate kinetics of AFBR using Biolite for different bed expansion levels and substrates to determine optimal process conditions.
3. Perform experiments with biofilm thickness as a controlled variable to determine optimum range of biofilm thicknesses. This can be done by positioning of the reactor recycle line at the proper height and incorporating a solid-liquid separator.
4. Conduct a more fundamental study on the effect of substrate concentration and loading rates on expansion behaviour of fluidized beds using Biolite and other novel carrier particles.
5. Conduct a longer term study on treating leachate to determine effect of toxins and metal precipitates.
6. Compare Biolite to other carrier media for use in AFBR using different wastes and operating conditions.

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APPENDICES

APPENDIX A: RAW DATA

Raw data for steady state experiments using sucrose based and leachate feeds is presented in Tables 19 and 20 below. The numbers used in the calculations and graphs are the average of the three sets of data shown here.

Table 19. **Steady State Data for Sucrose Feed - Raw Data**

Feed COD (mg/l)	HRT (h)	Total Effluent COD (mg/l)	Soluble Effluent COD (mg/l)	Solids COD (mg/g)	Gas Production (l/d)
5	8	1000	270	55	2.1
		820	280	40	2.3
		580	170	72	1.9
10	16	1150	620	130	3.4
		1080	590	97	3.7
		1420	780	100	3.0
15	24	2130	1490	130	2.4
		2260	1780	130	2.4
		2780	2260	97	2.3
15	18	3060	3250	170	2.1
		3400	3110	150	2.6
		4250	4180	300	2.4
5	24	590	410	70	0.6
		540	340	76	0.8
		540	430	62	0.9
10	48	520	310	59	1.0
		630	570	76	0.9
		660	410	74	1.1
10	12	3440	3260	220	2.4
		3380	2870	260	2.2
		3170	3170	300	2.4
10	24	700	390	50	2.0
		790	500	45	1.9
		630	370	65	2.1
15	72	860	590	91	1.1
		800	550	100	0.9
		740	500	110	1.1
5	6	1310	1080	190	2.6
		1540	1150	220	3.3
		1480	1240	220	3.0
15	36	1540	1170	60	2.4
		1420	880	80	2.8
		1150	820	76	2.7
5	4.8	1380	1200	220	3.6
		1410	1280	210	3.7
		1390	1300	220	3.6
5	12	420	190	63	1.4
		370	130	76	1.7
		410	230	58	1.6
10	9.6	3110	2800	200	3.9
		3690	2880	220	3.4
		2790	2430	230	4.5
15	14.4	4690	3840	290	4.5
		5230	4070	340	4.2
		5090	5230	320	3.9

Table 20. Steady State Data for Leachate Feed - Raw Data

Feed COD (mg/l)	HRT (h)	Total Effluent COD (mg/l)	Soluble Effluent COD (mg/l)	Solids COD (mg/g)	Gas Production (l/d)
5000	44.8	680	580	76	0.3
		680	560	75	0.2
		700	650	70	0.1
2000	11.2	720	590	26	0.5
		720	540	31	0.6
		750	590	34	0.4
5000	15.0	1090	1010	42	1.6
		1260	1050	74	1.6
		1400	1220	90	1.7
5000	9.0	1000	880	23	1.9
		1180	1090	40	1.9
		1310	1130	75	1.9
2000	4.5	890	680	35	1.7
		820	490	38	1.5
		820	620	36	2.3
5000	22.4	1190	1130	44	0.9
		1060	950	40	0.8
		1250	1150	48	1.1
2000	7.5	800	650	17	0.6
		920	700	19	0.7
		750	630	15	0.5
2000	22.4	610	540	31	0.4
		630	550	38	0.5
		660	570	29	0.4

APPENDIX B: SAMPLE CALCULATIONS

Organic Loading Rate (Table 9):

The organic loading rate is equal to the influent substrate concentration divided by the HRT.

Eg. 1. On day 1 of startup, the HRT was 10.0 d and the feed concentration was 2.5 g COD/l.

Organic Loading Rate = 2.5 g COD l^{-1}/10.0 d = 0.25 g l^{-1} d^{-1}

Total VFA Concentration as Acetate (Table 10):

The total VFA as acetate in mg/l is determined by using the conversion that 1 g of propionic acid is equivalent to 1.4 g of acetic acid.

Eg. 2. For day 2 of startup the concentrations of acetic and propionic acids are 18 and 8 mg/l respectively.

Total VFA as acetate = 18 mg l^{-1} + (8 mg l^{-1} x 1.5) = 29.2 mg l^{-1}

Specific Loading (Figure 14):

The specific loading rate was calculated by dividing the loading rate by the concentration of attached biomass in the reactor.

Eg. 3. On day 4 of startup, reactor 1 (MEP) was loaded at 0.5 g COD/l/d and had attached solids of 5.0 mg COD/g of Biolite. First the biomass is converted to g VSS/l of empty bed reactor volume by using the conversion of 1.4 g COD/g VSS and knowing that the empty bed reactor volume of 1.3 l contains a total of 150 g of Biolite.

g of VSS per l $= (5.0 \text{ mg COD g}^{-1}/1.4 \text{ g COD g}^{-1} \text{VSS}) \times (150 \text{ g}/1.3 \text{ l}) \times (1 \text{ g}/1000 \text{ mg}) =$ 0.412 g VSS/l

Specific Loading Rate $= 0.5 \text{ g COD l}^{-1} \text{d}^{-1}/0.412 \text{ g VSS l}^{-1} = 1.21 \text{ g COD g}^{-1} \text{VSS d}^{-1}$

Specific Removal Rate (Figure 15):

The specific removal rate was determined by multiplying the specific loading rate with the fractional soluble COD removed.

Eg. 4. On day 4 of startup, the soluble effluent COD was 400 mg/l. The feed COD for the entire startup period was fixed at 2.5 g COD/l. From eg. 3, the specific loading rate was 1.21 g COD/g VSS/d.

Specific Removal Rate $= ((2.5 \text{ g l}^{-1} - 0.4 \text{ g l}^{-1})/2.5 \text{ g l}^{-1}) \times 1.21 \text{ g COD g}^{-1} \text{VSS d}^{-1} = 1.02 \text{ g COD g}^{-1} \text{VSS d}^{-1}$

Effluent Suspended Solids (Table 11):

Suspended solids are calculated by substrating the soluble effluent COD from the total effluent COD.

Eg. 5. During steady state operations, at a feed COD of 5.0 g/l and HRT of 24 h, the total and soluble effluent (or mixed liquor) CODs are 560 mg/l and 390 mg/l respectively.

Suspended Solids $= 560 \text{ mg COD l}^{-1} - 390 \text{ mg COD l}^{-1} = 170 \text{ mg COD l}^{-1}$

% Removal of COD (Table 11):

This is the fractional COD of the feed which was removed expressed as a percentage. For the % removal based on total effluent COD, the total effluent COD was used which includes both the soluble and the suspended solids CODs.

Eg. 6. During steady state operations, at a feed COD of 5.0 g/l and HRT of 24 h, the total effluent (or mixed liquor) COD was 560 mg/l.

% Removal based on total effluent COD = ((5.0 g l^{-1} - 0.56 g l^{-1})/5.0 g l^{-1}) x 100 % = 88.8 %

The % removal based on the soluble effluent COD is similarly calculated using the soluble instead of the total effluent COD.

Organic Carbon Balance (Table 12):

All sample calculations below are done for a HRT of 24 h.

"Carbon In" was calculated by multiplying the organic loading rate by the expanded bed volume.

Eg. 7. At an HRT of 24 h the loading rate was 5 g/l/d.

Carbon In = 5 g COD l^{-1} d^{-1} x 0.5 l = 2.5 g COD d^{-1}

"Soluble Effluent" was calculated by multiplying the flowrate by the soluble effluent COD.

Eg. 8. The soluble effluent COD was 390 mg/l.

Soluble Effluent = (0.5 l / 1 d) x 0.39 g COD l^{-1} = 0.2 g COD d^{-1}

"Methane Production" was calculated by converting the methane produced in l/d into its COD equivalent.

Eg. 9. The rate of methane production was 0.7 l/d.

Methane Production = 0.7 l d^{-1} x (1/0.4 l g^{-1} COD) = 0.8 g COD d^{-1}

"Effluent VSS" was calculated multiplying the effluent suspended solids COD by the flowrate.

Eg. 10. The VSS in the effluent were 170 mg COD/l.

Effluent VSS = 0.17 g COD l^{-1} x (0.5 l / 1 d) = 0.09 g COD d^{-1}

"Total Carbon Out" was determined by adding the "Soluble Effluent", "Methane Production", "Methane in Effluent", and "Effluent VSS".

Eg. 11. Total Carbon Out = 0.2 + 1.85 + 0.04 + 0.09 = 2.17 (g COD d^{-1})

Eg. 12. Rate In / Rate Out = 2.5 g COD d^{-1}/ 2.17 g COD d^{-1} = 1.15

Net Specific Growth Rate (Table 15):

The net specific growth rate is found by dividing the suspended biomass leaving the system by the total attached biomass in the system..

Eg. 13. For HRT of 24 h and feed COD of 5 g/l the effluent suspended solids and attached biomass were 170 mg COD/l and 20.7 g COD/l of expanded bed (see eg. 3).

Net Specific Growth Rate = (0.17 g COD l^{-1} / 1 d) / 20.7 g COD l^{-1} = 0.0082 d^{-1}

Specific Utilization Rate (Table 15):

The specific utilization rate is found by dividing the removal rate by the total attached biomass in the system.

Eg. 14. For HRT of 24 h and feed COD of 5 g/l the effluent soluble COD was 390 mg/l and the total attached biomass was 14.8 g VFS/l (see eg.3).

Specific Utilization Rate = $((5.0\ g\ l^{-1} - 0.39\ g\ l^{-1})\ /\ 1\ d))\ /\ 14.8\ g\ VFS\ l^{-1} = 0.31\ d^{-1}$

Solids Retention Time (SRT) (Table 15):

The SRT is the defined as the inverse of the net specific growth rate.

Eg. 15. For HRT of 24 h and feed COD of 5 g/l the net specific growth rate was 0.0082 d^{-1}

$SRT = 1\ /\ 0.0082\ d^{-1} = 122\ d$

APPENDIX C. BIOFILM DENSITIES

Shown below is a sample calculation for biofilm density of particles removed from the top portion of an AFBR treating sucrose waste at a loading rate of 25 g/l/d. Similar calculations were done to find densities of biofilm from reactors treating leachate waste.

The biomass concentration of particles removed was 1400 mg COD/g Biolite or expressed as VFS was

$$1400/1.42 = 986 \text{ mg VFS/g Biolite}$$

The average bioparticle diameter using sample size of approximately 100 was determined to be 0.80 mm. From microscopic measurements and screen analysis, average diameter of clean Biolite was 0.30 mm. Assuming both Biolite covered particle and clean Biolite particle are perfect spheres, the average volume of a single biofilm is

$$4/3\, \Pi \left((0.8/2)^3 - (0.3/2)^3 \right) = 0.254 \text{ mm}^3$$

Since the average mass of a clean Biolite particle is 0.025 mg, there are 40,000 particles in 1.0 g of sample. Thus VFS of one biofilm covered particle was

$$986 \text{ mg VFS} / 40000 = 0.0246 \text{ mg VFS}$$

The fixed film solids in the biomass was found to be 27% with the remaining 73% volatile film solids. This gives a biofilm density of

$$0.0246 / (0.254 \times 0.73) = 0.13 \text{ g/cm}^3$$

APPENDIX D. SCREEN ANALYSIS

Screen analysis was conducted on Biolite particles which had been thoroughly dried in an oven at 105 ^{0}C. In this procedure, particles in a mixture are separated by their size using testing sieves. A set of standard screens is arranged serially in a stack, with the smallest mesh at the bottom and the largest at the top. An analysis is performed by placing the sample on the top screen and shaking the stack mechanically. The particles retained on each screen are removed and weighed. Table 18 shows cumulative fraction retained ϕ for each screen used.

Using a cumulative analysis, the area per unit mass of sample A_w also called the specific surface is given by

$$A_w = \frac{6}{\Phi_s\ \rho_p} \int_0^{1.0} \frac{d\phi}{D_p}$$

where Φ_s is the particle sphericity (assumed to be 0.7), ρ_p is the particle density, and D_p is the mesh size of screen n. The quantity ϕ is the mass fraction of the sample that consists of particles larger than D_p. The plot of screen opening vs. ϕ is shown in Figure 31. Integration of the above equation was done numerically using the trapezoidal rule. This gave a value of the specific surface of 174 cm^2/g.

The volume surface mean diameter D_s is related to the specific surface and defined by the equation

$$D_s = \frac{6}{\Phi_s A_w \rho_p}$$

which gives a mean diameter for Biolite particles of 0.31 mm.

Reference: McCabe and Smith, 1976.

Table 21. **Cumulative Fraction Retained for each Screen Opening Size**

Screen Opening (mm)	Cumulative Fraction Retained
0.417	0.00
0.351	0.18
0.295	0.76
0.246	0.92
0.208	0.98
0.175	1.00

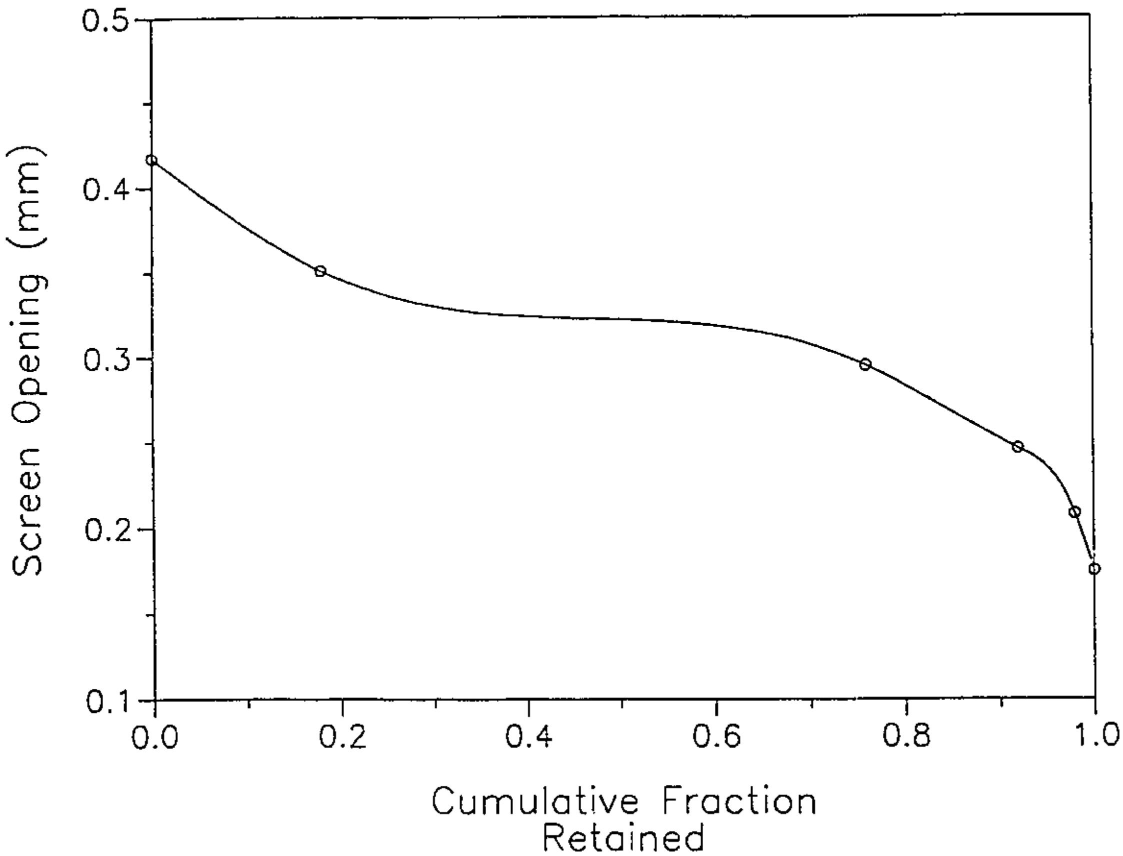

Figure 36. Screen Opening vs. Cumulative Fraction Retained