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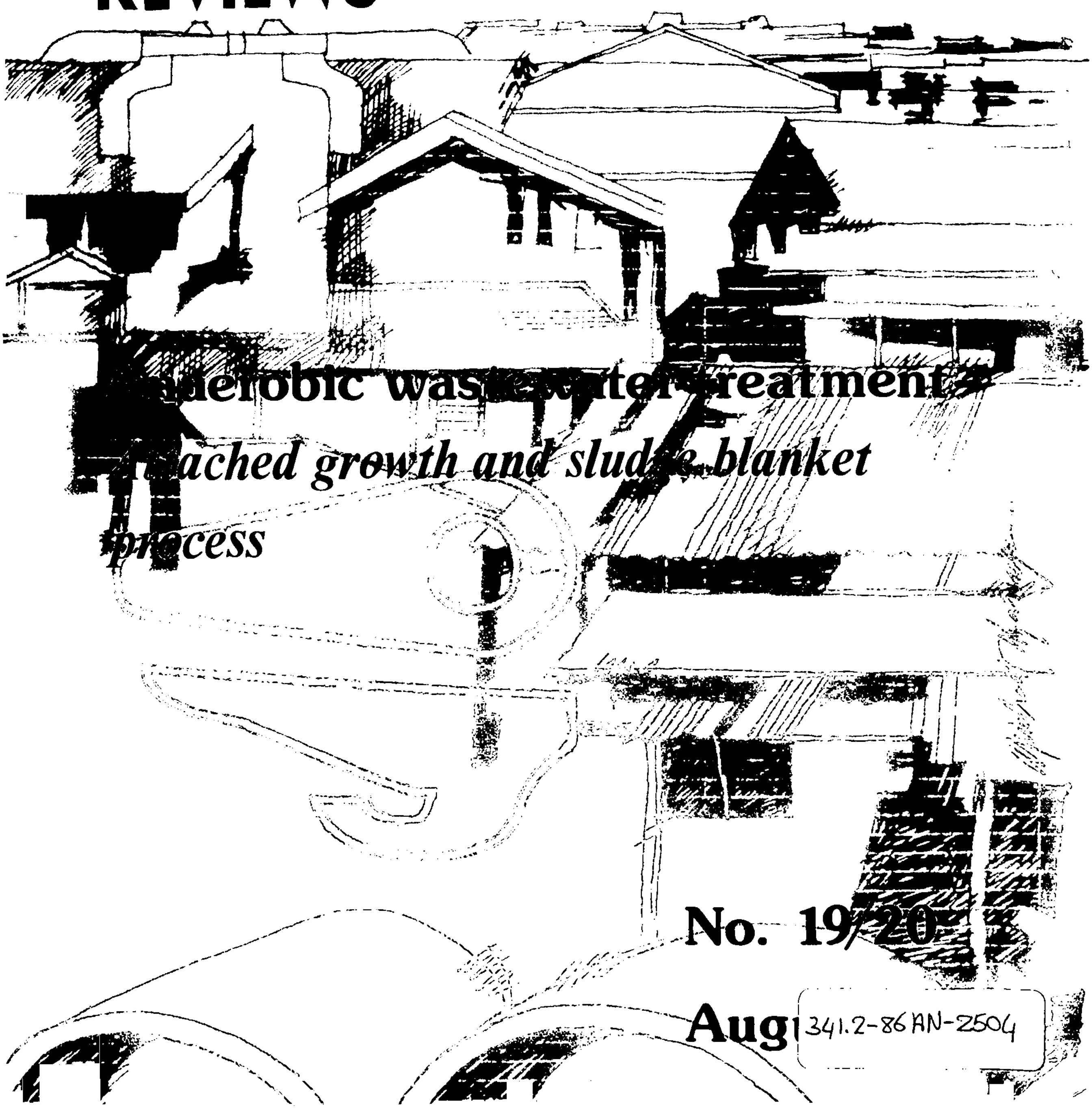

Anaerobic wastewater treatment: Attached growth and sludge blanket process

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ANAEROBIC WASTEWATER TREATMENT –
ATTACHED GROWTH AND SLUDGE BLANKET PROCESS

by

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I. ANAEROBIC TREATMENT

1.1 Introduction

In the fifties and sixties aerobic processes were very popular in biological treatment, which generally require energy equipment for transferring oxygen into the waste. The picture changed significantly as a result of environmental debate and the increase in energy prices both culminating in the seventies. Reuse and energy conservation became the current topics of research interest and anaerobic processes quickly emerged with a new potential.

Further, increasingly stringent pollution control regulations coupled with the rising energy costs of aerobic treatment systems in the early seventies greatly stimulated interest in anaerobic treatment as an energy saving waste treatment technology. This interest led to the development of range of reactor designs suitable for the treatment of high and low strength soluble wastewaters.

1.1.1 Mechanism of Anaerobic Treatment

The anaerobic fermentation process converts waste organic materials to methane and carbon dioxide in the absence of molecular oxygen. It is long recognized as a useful process for wastewater treatment. The mechanism of the anaerobic process is much more complex than the aerobic ones. This is a result of many pathways available for an anaerobic community. The pathways and microorganisms responsible for the reaction are not known in great detail, but during the last two decades a broad outline of the process has been established as described by a number of investigators. The reader is referred to a review article prepared by Henz and Harremoes (1982) for more details.

Basically anaerobic fermentation is performed by two groups of bacteria, the acid producing and the methane producing bacteria. Acid producing bacteria are subdivided into acid forming bacteria (butyric and propionic acid) and acetogenic bacteria (acetic acid and hydrogen) while methane producing bacteria are subdivided into acetoclastic methane bacteria (acetophilic) and methane bacteria (hydrogenophilic).

The present state of knowledge is summarized in Fig. 1.1 (Mosey, 1982), which shows the four groups of bacteria now believed to be involved in the conversion of glucose into carbon dioxide and methane.

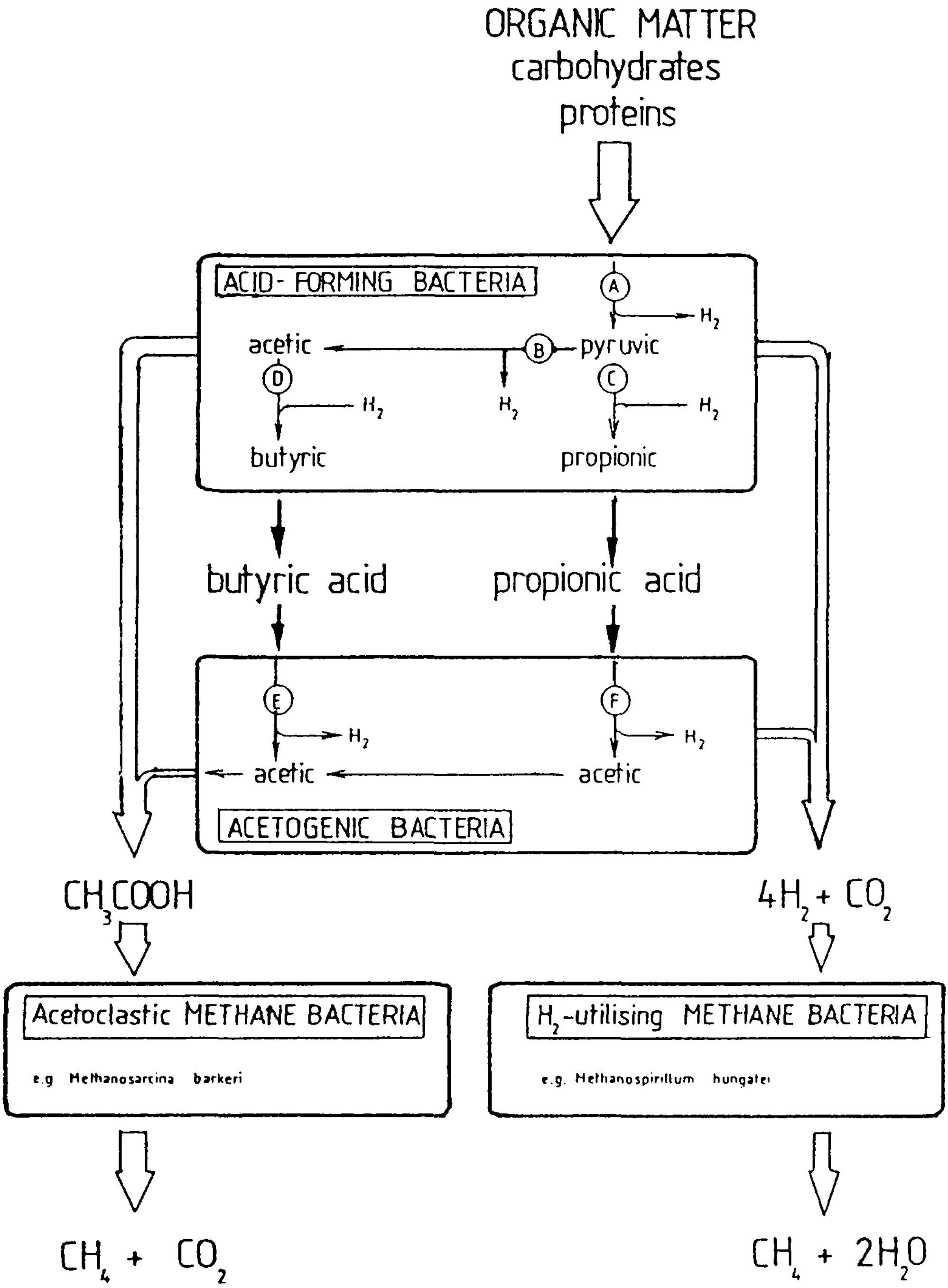


Fig. 1.1 The microbial ecology of the anaerobic digestion process (Mosey, 1982)

Steps of Reaction

The anaerobic metabolism of a complex substrate, including suspended organic matter, is regarded as a three stage process.

- In the initial stage, hydrolysis of suspended organics and soluble organics of higher molecular weight takes place. Hydrolysis of organic matter is a rather slow process brought about by extracellular enzymes. Lipids are hydrolized very slowly, with the result that overall, the hydrolysis might be (including methane production) a rate limiting step for a waste containing considerable amount of lipids and other slowly hydrolyzing compounds, like piggery waste (Kennedy & van den Berg, 1982a).

- In the second stage, these hydrolysis products are fermented to simple organic compounds, predominantly volatile fatty acids by the acid producing bacteria. This results in the formation of acetic acid or in case of instability, the higher fatty acids such as propionic and butyric acids. The acid production rate is high compared to the methane production rate, which means that a sudden increase in easily degradable (soluble) organics will result in increased acid production with subsequent accumulation of the acids.

- In the third state, the real waste stabilization occurs. During this stage methane is produced by methane producing bacteria, primarily from acetic acid but also from hydrogen and carbon dioxide. Methane production is a very slow process in anaerobic digestion.

The main features of the four groups of bacteria mentioned above are believed to be as follows (Mosey, 1982).

The Acid-Forming Bacteria

These are fast-growing bacteria (minimum doubling times around 30 minutes) which ferment glucose to produce a mixture of acetic, propionic and butyric acids according to the reactions:

$$C_6H_{12}O_6 + 2H_2O \longrightarrow 2CH_3COOH \text{ (acetic)} + 2CO_2 + 4H_2$$

$$C_6H_{12}O_6 \longrightarrow CH_3CH_2CH_2COOH \text{ (butyric)} + 2CO_2 + 2H_2$$

$$C_6H_{12}O_6 + 2H_2 \longrightarrow 2CH_3CH_2COOH \text{ (propionic)} + 2H_2O$$

Their preferred reaction is the first one, the conversion of glucose into acetic acid. It provides the acid-forming bacteria with the biggest energy yeild for growth and it provides the acetoclastic methane bacteria with their prime substrate for methane production. The other reactions, the formation of butyric and propionic acids are the bacteria's response to accumulations of hydrogen during surge loads. Diversion of glucose metabolism towards butyric acid reduces both the acid-forming bacteria's output of hydrogen and the acid load on the system. Formation of propionic acid actually puts hydrogen production into reverse.

The Acetogenic Bacteria

As their name implies, these are the bacteria that convert propionic and butyric acids into acetic acid according to the equations:

$$CH_3CH_2COOH \text{ (propionic)} + 2H_2O \longrightarrow CH_3COOH + CO_2 + 3H_2$$
$$CH_3CH_2CH_2COOH \text{ (butyric)} + 2H_2O \longrightarrow 2CH_3COOH + 2H_2$$

Their existence has not yet been demonstrated. As quoted by Henz & Harremoes (1982) it has only recently been deduced by McInerney and others (1971) from the inability of any known methane bacteria to metabolise propionate and butyrate directly.

The Acetoclastic Methane Bacteria

These are the bacteria that convert acetic acid into a mixture of carbon dioxide and methane according to the reaction :

$$CH_3COOH \longrightarrow CH_4 + CO_2$$

They also grow very slowly with minimum doubling times of 2-3 days, but are believed to be unaffected by the concentration of dissolved hydrogen gas in the growth media.

They normally control the pH value of the fermentation by removal of acetic acid and formation of carbon dioxide. They are responsible for most of the methane produced by the anaerobic digestion process.

The Hydrogen-Utilizing Methane Bacteria

These bacteria are hydrogen-scavengers. They obtain energy for growth from the reaction:

$$4H_2 + CO_2 \longrightarrow CH_4 + 2H_2O$$

and in doing so, they remove almost all of the hydrogen from the system. They grow quite quickly with minimum doubling times around 6 hours.

The traces of hydrogen left behind, regulate both the total rate of acid production and the mixture of acids that is produced by the acid-forming bacteria. Hydrogen also controls the rates at which propionic and butyric acids are subsequently converted back into acetic acid. These H_2-utilizing methane bacteria regulate the formation of volatile acids.

1.1.2 Advantages of Anaerobic Treatment

The anaerobic digestion process as a waste treatment process offers several advantages over conventional aerobic systems. These are as follows:

- a higher degree of waste stabilization;
- only a small amount of the organic material is converted into biomass, limiting the problem of further disposal of sludge;
- there is no need for energy and equipment for transferring oxygen into the waste, moreover the end product (methane) separates;

- methane is often a valuable source of energy, both for operating the anaerobic digestion process and for other operations such as producing electricity and for heating and cooling;
- a lower nutrient requirement.

1.1.3 Applications

In the past, broad scale application of the anaerobic process has been largely focused on the treatment of municipal sewage sludges and animal residues to achieve waste stablization and solids destruction.

The reactor configuration generally used contains a large holding tank into which wastes are fed continuously or intermittently (Fig. 1.2).

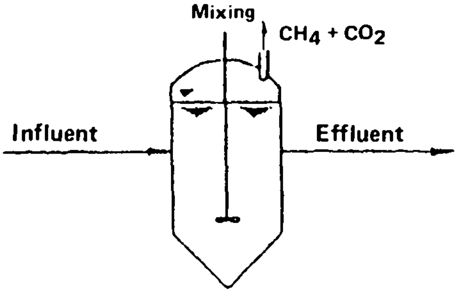


Fig. 1.2 Conventional anaerobic digester

The wastes are usually maintained in the reactor for 30 to 60 days. When first introduced several decades ago, these units were unmixed and unheated; hence the process was very slow and inefficient (the optimum temperature for mesophilic methonogenesis is 35°C).

More recent designs of the conventional process have led to the development of heated, completely mixed, high rate systems. These innovations have lowered the retention time to 15 days, or less, and greatly increased the allowable loading rates. Often anaerobic digestion systems are built as two stage systems, the first stage being the active digestion zone, and the second stage a quiescent-settling zone.

These anaerobic digesters are suitable for solids processing than for liquid waste treatment. Because of the high temperature requirement, high strength wastes are more suitable. Large quantities of methane produced by high strength wastes can be used to heat the reactors. In addition, the high solids retention time, which dictate large reactor volumes, preclude the processing of large waste flows.

1.1.4 Disadvantages of Anaerobic Treatment

There are however, several limitations on the fermentation process which are caused by the process and equipment used, the properties of many bacteria involved, and by lack of understanding of the sensitive nature of the bacteria. These are:

- the slow growth rate of the methane producing bacteria (methanogens);
- the long solids retention time, requiring a larger volume of reactor;
- the need for auxiliary heating to maintain the digester at the optimum

temperature (35°C) for the growth of essential bacteria;
- the sensitive nature of the methanogens;
- the general feeling of unreliability associated with the process.

1.2 New Developments

More recently, significant advances in both the fundamental understanding of the anaerobic process and the engineering application of this process have taken place. These new developments show a great deal of promise in overcoming many of the limitations associated with treatment for the processing of both municipal and industrial wastewater.

With the new interest, new approaches in the field of wastewater treatment developed resulting to the novel application of fixed films in the anaerobic treatment of wastewater. This field is in a rapid state of development and this review is intended primarily for researchers actively involved in this development.

1.2.1 Advanced Technologies Available at Present

The principal objective of any advanced biological reactor configuration should be to bring the substrates and enzymes into intimate contact for a sufficient time to allow the reactions to occur. For anaerobic methane fermentation processes, long microbial residence times are necessary due to the slow growth rate of the methane producing bacteria.

High rates of conversion of waste organic materials into methane have mostly been achieved by getting around the problem of slow growth of essential microorganisms (methanogenic bacteria). Until now, attempts to increase the growth rate of these bacteria have essentially been unsuccessful except by changing digestion temperature into thermophilic range (50-60°C) from the mesophilic range (35°C optimum) (Schraa & Jewell, 1984). By preventing bacteria from escaping in the effluent, the digestion process becomes eventually independent of growth rate. This way it is possible to reach high concentration of bacteria and hence high rates of reaction in spite of very slow growth rates. This is the principle on which advanced technologies are based.

1.2.2 The Anaerobic Contact Process

While a high solids retention time (SRT) is necessary for efficient methane fermentation, a low hydraulic retention time (HRT) is desirable for system economy. The conventional system is not able to separate SRT and HRT and thus larger reactor volumes are required. The anaerobic contact process was developed from the concept of recycling biological solids to obtain a larger retention time. It is an anaerobic activated sludge process (Fig. 1.3).

In this process, effluent from the bioreactor is pumped to a settling unit where a portion of settled sludge is returned to the reactor, enabling the contact unit to maintain a high concentration of active biomass. The solids concentration can be maintained independently of wasteflow using the method of biomass recycle.

The performance of this process depends on:

- degree of the mixing of digester contents;

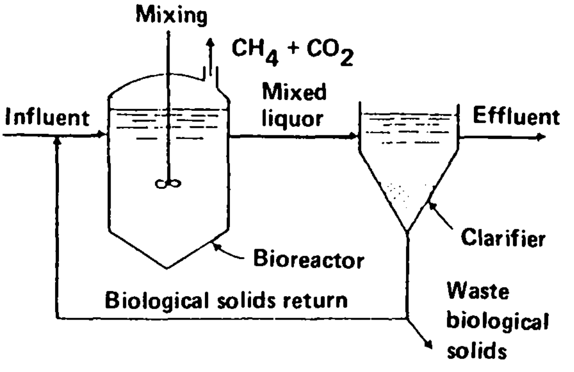


Fig. 1.3 Anaerobic contact process

- extent to which the bacteria settle out in the sedimentation tank for return to the reactor.

Problems and limitations of the process are caused by difficulties of control due to:

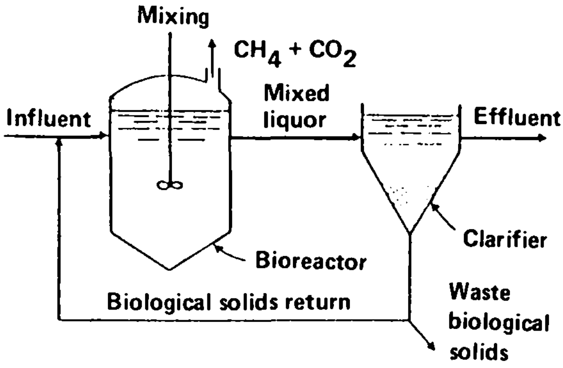
- difficulty to provide adequate mixing facility for large digesters;
- slow settling of microorganisms from the digester liquid;
- difficulty in predicting the amount of settling microorganisms in the clarifier;
- unstability of the process at high methane production rates, causing long term set backs.

Experience with the anaerobic contact process has shown that this process is best suited for the treatment of either concentrated or naturally warm wastes. In general, the process is not satisfactory for a waste containing less than 2,000 mg/L BOD_5 at temperatures less than 30°C.

1.3. Fixed Film Systems

Another means of providing an anaerobic process with a high solids retention time for the methane producing bacteria with a short hydraulic retention time for system economy is by using fixed film reactors. In these heterogeneous systems, the microorganisms grow in a film on a solid support while organic matter is removed from the liquid flowing past them. These systems may have very high loading capacities compared with aerobic process because they are not limited by oxygen transfer.

Various types of fixed film reactors have been developed for anaerobic treatment. These reactors have in common a retention of microbial biomass within the reactor by mechanisms which avoid the costly operational problems associated with the solids recycle system of the anaerobic contact process. In contrast to the earlier designs the new reactors are "retained biomass reactors" and their mode of operation relies on the propensity of bacteria, especially the methanogens, for attachment to solid surface (Colleran et al., 1982). The reader is referred to an excellent review on anaerobic fixed film reactors by Henz and Harremoes (1982). Upflow anaerobic filter, expanded/fluidized bed reactor, and downflow anaerobic stationary fixed film reactor are the attached growth reactor types considered in this review. In addition, the review includes information on upflow anaerobic sludge blanket reactors.



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1.3.1 Upflow Anaerobic Filter

The anaerobic filter concept was first developed by Young & McCarty (1969) and has found numerous applications in both high and low strength industrial wastewaters. The anaerobic filter (or packed bed or submerged filter) is composed of one or more vertical filter beds containing some inert material, such as rocks or plastic media, which acts as a stationary support surface for microbial film attachment. Wastewaters are pumped upwards through the support media, allowing contact between the attached microorganisms and wastewater. Microbial growth also takes place in the voids between the support media. This system permits adequate mean cell residence time for the methane producing bacteria and still allows a short hydraulic retention time for system economy. Washout of the suspended biomass from the filter is not considered to pose any operational problem since the support matrix physically hinder washout during hydraulic shock loads. A schematic diagram of an anaerobic filter in shown in Fig. 1.4.

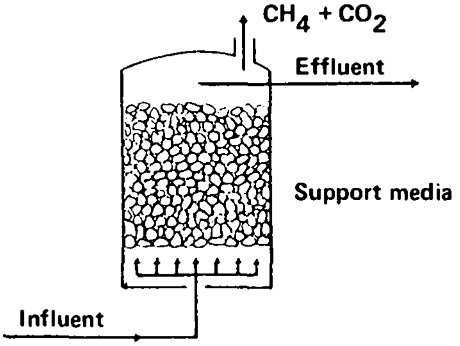

Fig. 1.4 Anaerobic filter

1.3.2 Anaerobic Expanded Bed/Fluidized Bed System

Anaerobic filter treatment process may be hampered by clogging and inefficient contact of microorganisms and the wastewater, therefore fluidized bed systems were introduced to overcome these problems. Anaerobic expanded or fludized bed, as applied to wastewater treatment, consists of inert-sand-sized particles (in a reactor) which expands (or remains in the fluidized state) by the upward flow of waste through the reactor. As in the case of anaerobic filter the inert particles act as support surfaces for the growth of attached organisms and the process relies on the retention of the biomass within the reactor. A schematic diagram of the process is given in Fig. 1.5.

The exact difference between expanded and fludized bed is somewhat ambiguous. In many cases, fluidization has been used to refer to more than doubling of the reactor volume as caused by the high flow rate of fluid through the filter composed of small particles (Switzenbaum, 1983). The term 'expanded bed' has been used to designate reactors that have a smaller degree of expansion of the static volume.

The first application of fluidized bed technology to anaerobic treatment was developed in Jewell's laboratory at Cornell University (Jewell, 1981) for the treatment of domestic wastewater.

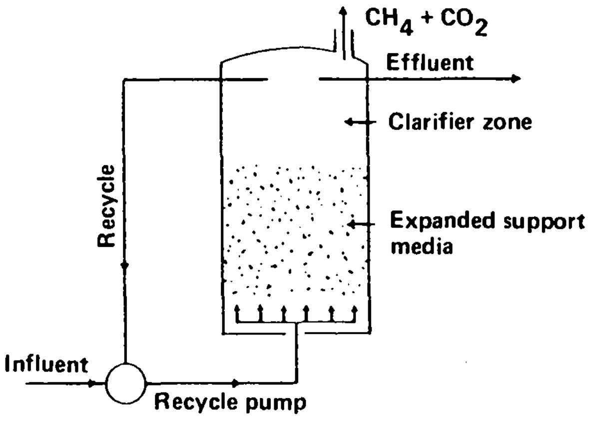

Fig. 1.5 Fluidized-bed reactor

1.3.3 Downflow Anaerobic Stationary Fixed Film Reactor

This reactor was developed by the National Research Council in Ottawa (van den Berg et al., 1980). In these reactors the active biomass is grown on fixed surfaces and retained as an attached film. Inert support materials such as glass, fired clay or plastic are used. In fixed film reactors relatively little suspended growth occurs or is retained. The latter is achieved by removing the effluent from the bottom rather than from top as in the case of most anaerobic filters. The schematic diagram is shown in Fig. 1.6.

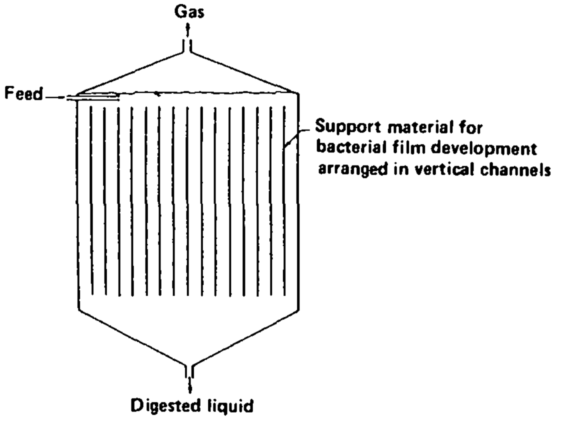

Fig. 1.6 Sketch of anaerobic fixed film reactor (van den Berg et al., 1980)

The fixed film reactor has the advantage of greater resistance to wash out of biomass attached to the stationary support where as the suspended biomass is susceptible to wash out by hydraulic shock loading. In these reactors the amount of retained biomass depends on the surface to volume ratio, and therefore limited by the support matrix area. This fixed film reactor with effluent removal from the bottom has been found to produce methane from wastes with high suspended solids contents where as in the other three reactor design the suspended particles which are undigestible and/or inorganic in nature may seriously decrease the reactor efficiency by plugging or clogging of the sludge blanket or the filter matrix (van den Berg, 1982). Stationary fixed film reactor configuration is simple compared to other advanced designs.

1.3.4 Upflow Anaerobic Sludge Blanket (UASB)

This reactor was developed in the Netherlands in the seventies by Lettinga and his coworkers (1980). The essential feature of this reactor is the presence of a very active sludge blanket in the bottom of the reactor. In the UASB reactor, the microbes attach themselves to each other or to small particles of suspended matter to form conglomerate or granules. Other important feature relates to gas removal without interference with the settling of microorganisms and their return to the sludge blanket. In this process the waste is introduced at the bottom of the reactor into the sludge bed where most of it is converted into methane and carbon dioxide. The gas formed causes sufficient agitation to keep the sludge bed particles moving around to keep the bed fully mixed. Some particles are lifted up above the sludge blanket, but, as they lose the entrapped gas, they settle back. The UASB reactor is equipped with a gas-solids separator in the upper part of the reactor as shown in Fig. 1.7.

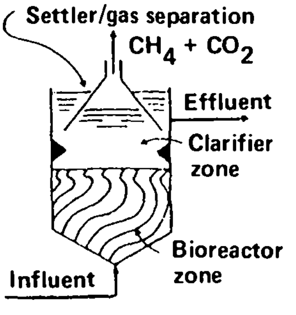

Fig. 1.7 Upflow anaerobic sludge blanket reactor

The separator acts to separate the gas produced by the methane reaction, from the dispersed sludge particles. This is very important for the retention of sludge in the reactor. The amount of retained biomass is generally larger per unit reactor volume than in downflow fixed film or upflow filter reactor (Colleran et al., 1982). The system has been applied to both low and high strength waste waters.

The purpose of this review is to describe the above four configuration processes, compare their efficiencies, summarize the design criteria and the effect of operating parameters and their application in the field of municipal and industrial wastewater treatment.

II. ANAEROBIC FILTER TREATMENT (UPFLOW FILTERS)

2.1 General

Most of the earlier works on anaerobic treatment was related to anaerobic digestion of municipal waste sludges including uncontrolled digestion in a septic tank. It was followed by anaerobic treatment of high strength industrial wastewaters. Recently several attempts have been made to treat dilute wastewater (raw and settled) and septic tank effluent by anaerobic processes, especially by the fixed film processes such as anaerobic filters.

2.2 Anaerobic Upflow Filtration

2.2.1 Introduction

Anaerobic filters are typically operated in an upflow mode to ensure the medium is in submerged condition in order to maintain the anaerobic conditions. The filter can also be operated in the downflow (van den Berg et al., 1980) or in the horizontal mode (Landine et al., 1981; Landine et al., 1982).

The anaerobic filters operating under submerged conditions is a useful and practical method for secondary treatment of septic tank effluent in areas where soil adsorption or leaching systems for effluent disposal are precluded by dense soil conditions, high water table and limited availability of land.

Septic tank systems followed by anaerobic filter or composite anaerobic filter are some of the simple treatment devices where localized collection and treatment of sewage could be expedient (Raman & Khan, 1982).

The use of upflow anaerobic filters for industrial wastewater treatment is well documented. Mueller and Mancini (1977) listed information on anaerobic filters treating various industrial wastes. A state-of-the-art review on the treatment of high strength organic wastes by submerged media anaerobic filters has been recently prepared by Wu et al. (1982). The reader is referred to an excellent review article of Henz and Harremoes (1982) for a more complete discussion of anaerobic filter application. Switzenbaum (1983) and Anderson et al. (1984) have also summarized the full scale treatment experiences with this system.

Recently attempts have also been made to evaluate the performance of the anaerobic filters as a treatment device for sanitary wastewater and septic tank effluent. Pilot-scale experimental studies on anaerobic filter treatment of septic tank effluent treatment have been carried out in India (National Environmental Engineering Research Institute), in U.S.A. (Washington State University) and in Thailand (Asian Institute of Technology).

2.2.2 Principles

The principle of operation of an upflow anaerobic filter is that the wastewater is passed through at low velocities through a column filled with packing material. The packing material acts as a surface for the attachment of micro-organisms and as an entrapment mechanism for unattached flows of organisms. The attached and entrapped anaerobic biomass will convert both soluble and particulate organic matter in the influent wastewater to methane and carbon dioxide as wastewater flows upwards through the column.

The removal mechanisms are primarily adsorption, filtration, and oxidation (Kennedy, 1981). Polprasert and Hoang (1983) found similar mechanisms for the removals of bacterial and bacteriophages and also included natural die-off under anaerobic conditions as an additional mechanism.

2.2.3 System Design

The essential features of the upflow anaerobic filter design are: (i) a distributor in the bottom of the column; (ii) a media support structure; (iii) inert packing material; (iv) a free board above the packing material; (v) effluent draw off and (vi) operational features such as recycling facilities, backwashing facilities or a sedimentation zone below the packing material. The distributor is designed for the even distribution of the incoming waste stream, over the whole cross-sectional area of the anaerobic filter column to avoid short-circuiting.

The free board above the packing material is designed as a head space to allow for the accumulation and capture of methane gas. Various provisions are made for the draw-off and utilization of the gas, which consists mainly of methane and carbon dioxide, with smaller amounts of hydrogen sulphide, hydrogen and other trace gases.

2.2.4 Recycling

Recycling of effluent back to the influent is not usually practised in the operation of the anaerobic filter. Anaerobic filters are generally designed as one-pass-plug flow reactors, taking advantage of the high driving force of the undiluted incoming waste stream. However, recycling might be necessary with certain wastes to dilute organic materials present in high concentration, to dilute toxic materials, or for pH control. Thus, provisions may be made for effluent recycling.

2.3 Design Criteria

2.3.1 General

Two of the major design parameters for the design of anaerobic filter are hydraulic retention time (HRT) and organic volumetric loading rate (OVLR). The value of the parameters will depend upon the influent waste concentration, degradability, temperature, and the desired efficiency of removal or rate of gas production. As far as possible these design parameters should be based on experience obtained with full scale experiments (Switzenbaum, 1983).

Laboratory and pilot scale filters have been operated at HRT values of three to several hundred hours and at organic volumetric loading rates of 0.4 to 27 kg COD/m^3-d (Henz & Harremoes, 1982).

Full scale industrial applications employ HRT values in the range of one to ten days, COD loadings from 4 to 16 kg/m^3-d (Switzembaum, 1983).

2.3.2 Filter Media, Size and Shape

A wide variety of materials and sizes have been used in the anaerobic filter studies. In general the material selected should have a high specific surface (surface-area-to-volume ratio) to provide a large surface for attached biofilms, while maintaining a sufficient void volume to prevent the reactor from plugging either from particulate solids entering with the influent waste stream or bacterial floc growth within the reactor. Since the anaerobic filter consists of both attached and entrapped biomass, the choice of packing material is quite important (Switzenbaum, 1983).

The proper selection of the material, however, will depend to a larger extent on the waste characteristics. Two key elements are the concentration of particulates in the incoming waste and organic composition. Higher strength waste will produce a larger yeild of biomass than proteinaceous wastes. Because of these factors some designs give provisions for media backwashing or solid wasting to avoid plugging.

All the laboratory scale and full scale anaerobic filters used so far to treat septic tank effluent have used only rock medium, although different types of media have been tried or used in anaerobic filters treating other wastewaters. Viraraghavan & Kent (1983) reported on the studies conducted by Frostell (1979) on the influence of media on the loading capacity of anaerobic filters. Three different types of filter media as given below were examined by him.

(i) Approximately spherical hard rock particles (porosity 0.42)
(ii) Ceramic raschig rings (porosity 0.68)
(iii) Plastic berl saddles (porosity 0.91)

A soluble synthetic wastewater with a COD of 1200 mg/L was used and responses to different loading rates were quite similar in all filters. A COD reduction of slightly below 90% was achieved at 0.5 kg COD/m^3-d, decreasing to 69% in the plastic, 76% in the ceramic and 77% in the rock filters at 1.8 kg COD/m^2-d; no improvement in sludge retention was observed with the more porous ceramic and porous media.

However, studies by van den Berg & Lentz (1981), van den Berg & Kennedy (1981) and Young & Dahab (1982) showed quite different results when using different media. van den Berg & Lentz (1981) found that area loading rates were up to 20% higher with baked clay than with glass or PVC plastic as film support material. They also found that active films formed faster on clay than on glass or plastic and that clay support provided far greater process stability than glass or plastic. van den Berg & Kennedy (1981) tested needle punched polyester construction material and red draintile clay. They found that rates of film development were very fast on these two material in comparison with potters clay and especially polyvinyl chloride tested earlier by their group. Both the needle punched polyester and red draintile clay were very effective. Red draintile clay is easily available and cheap. Investigations conducted by Young & Dahab (1982) using four different media showed the importance of medium type, size and shape on wastewater treatment perfomance. Corrugated modular blocks were used as medium in two reactors, while the other two contained cylindrical pall rings and polypropylene spheres as medium respectively. Fig. 2.1 shows the effect of various media on COD removal at different hydraulic retention times.

In the case of anaerobic filter treating septic tank effluent, it is preferable to use a medium which is relatively inexpensive and which is sturdy in comparison (Viraraghavan, 1983).

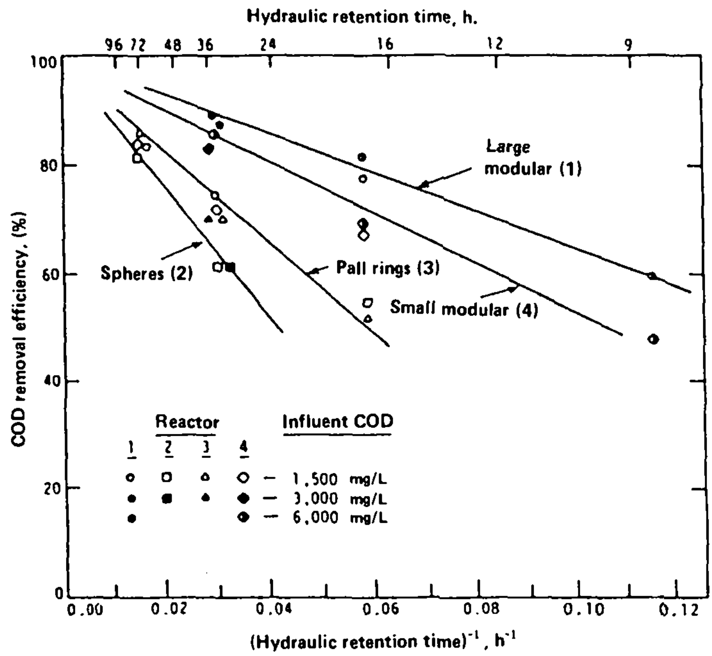

Fig. 2.1 COD removal efficiency versus the inverse of the theoretical hydraulic retention time when operating at a number of influent concentrations (Young & Dahab, 1982)

Raman & Khan (1982) in their study of anaerobic filter treatment for sewage used gravel and broken stone with a size of 2.0-2.5 cm as the media.

2.3.3 Filter Depth

An increase in filter depth to a certain extent would result in better removal efficiency. But beyond a certain depth, removal efficiency will not be significantly improved, but the headloss increment will be rapid. A study of anaerobic upflow filter treatment for septic tank effluent (Sutharkar, 1981) with 600-700 mg/L total COD showed that 60% reduction was obtained at a depth of 60 cm.

Askinin (1983) reported from his laboratory study for sewage treatment, that a depth of 120 cm is sufficient for required treatment.

Raman & Khan (1982) provided a depth of 115-125 cm in their anaerobic upflow filter treating either raw settled sewage or septic tank effluent.

2.3.4 Temperature

A laboratory study conducted at the University of Washington (Seabloom et al., 1981) indicated an average BOD removal of approximately 80% or more through an anaerobic filter treating septic tank effluent at temperatures of 7°C to 14°C, but studies of field anaerobic filter units showed average BOD and COD removals of only 30% at approximately 12°C, (Seabloom et al., 1981; Hamilton, 1976). The reasons for the wide difference in organic matter removal efficiency are not apparent (Viraraghavan, 1984).

Anaerobic filter studies with certain industrial wastewaters and blackwater (Prasad & Heinke, 1981) showed that although removal efficiencies may not be high, the anaerobic filter unit is capable of adapting to low temperatures without failure.

Raman & Khan (1982) conducted an anaerobic upflow filter study to treat either settled sewage or septic tank effluent at a temperature ranging from 23-33°C. 70-80% BOD removal efficiency was achieved with an influent BOD concentration of 110-300 mg/L and at an organic volumetric loading rate of 0.34 kg BOD/m^3-d.

2.3.5 Hydraulic Retention Time

Hydraulic Retention Time (HRT) is calculated on the basis of liquid volume of the filter (or void volume). HRT is the ratio between the void volume and the influent flow rate of the wastewater.

Data collected by Young (1968) and reported by Young & McCarty (1968) indicated that a linear relationship existed between COD removal efficiency and the inverse of the HRT in the voids within the rock-filled reactors. This can be represented by the following equation:

$$E = 100(1- \quad /HRT)$$

where E = COD removal efficiency, percent

HRT = hydraulic retention time in void volume, and

= a proportionality coefficient

The COD removal data for the investigations carried out by Young & Dahab (1982) followed a similar trend but had a different value for each media type and size (Fig. 2.1). The above equation provides only an empirical description of COD removal in the anaerobic filter. For the data of the four media studied by Young & Dahab (1982), plus the rock media by Young (1968), (when operating over a wide range of organic loadings, waste types and waste strengths), it was suggested that the relationship is fairly dependable. One objective of the pilot tests would be to determine values of for specific types of media proposed for use in full-scale anaerobic filters.

Raman & Khan (1982), in their study of upflow filter treating raw sewage obtained 79.5% BOD_5 and 88.5% suspended solids (SS) reduction at retention time of 6 to 8 hours. Suthakar (1981) reported that a retention time of 1 day is an optimal condition in treating septic tank effluent and it can reduce 62.1% TCOD (total COD) and 48.8% SS.

2.4 Different Design Configurations

2.4.1 Winneberger's Design

Winneberger's concept of an anaerobic filter for the treatment of septic tank effluent as quoted by Viraraghavan & Kent (1983) is presented in Fig. 2.2. The filter is relatively shallow and it is operated essentially in a horizontal flow mode.

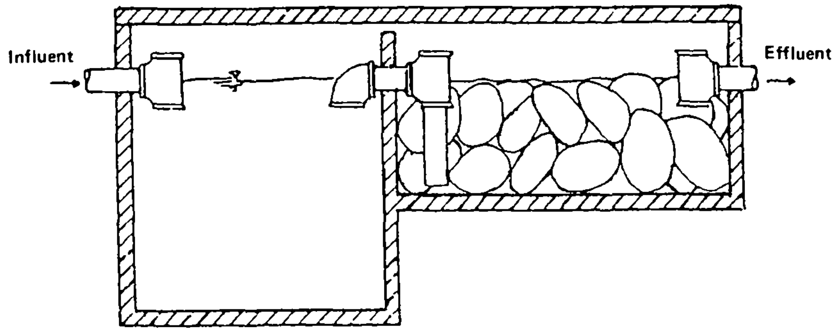

Fig. 2.2 Winneberger's design of anaerobic horizontal filter (Viraraghavan & Kent, 1983)

2.4.2 Hamilton's Design

Hamilton's system (Figs. 2.3, 2.4 and 2.5) as quoted by Viraraghavan & Kent (1983) consists of two rock filters connected in series with a septic tank. This system is reported to have operated for almost four years without clogging or solids removal.

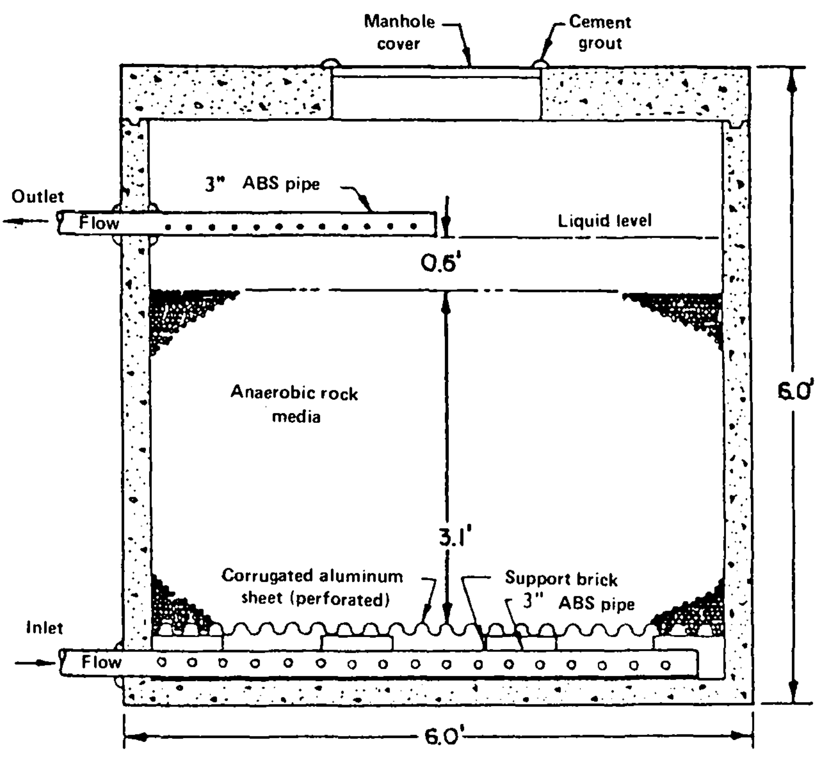

Fig. 2.3 Anaerobic filter detail

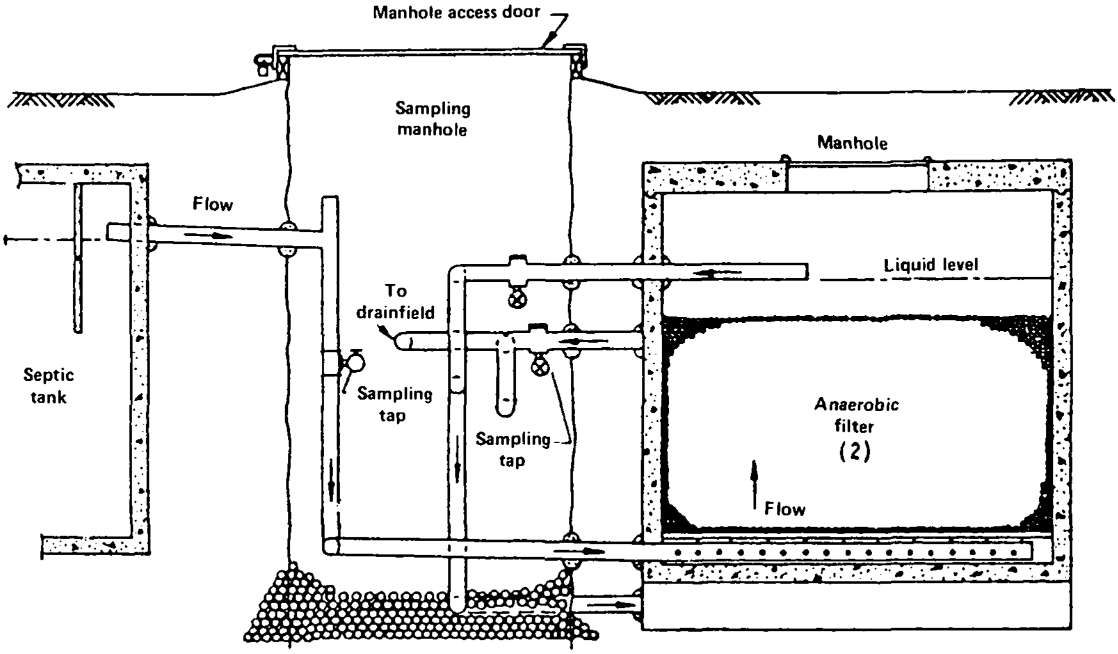

Fig. 2.4 Septic tank, manhole, and anaerobic filter (profile view)

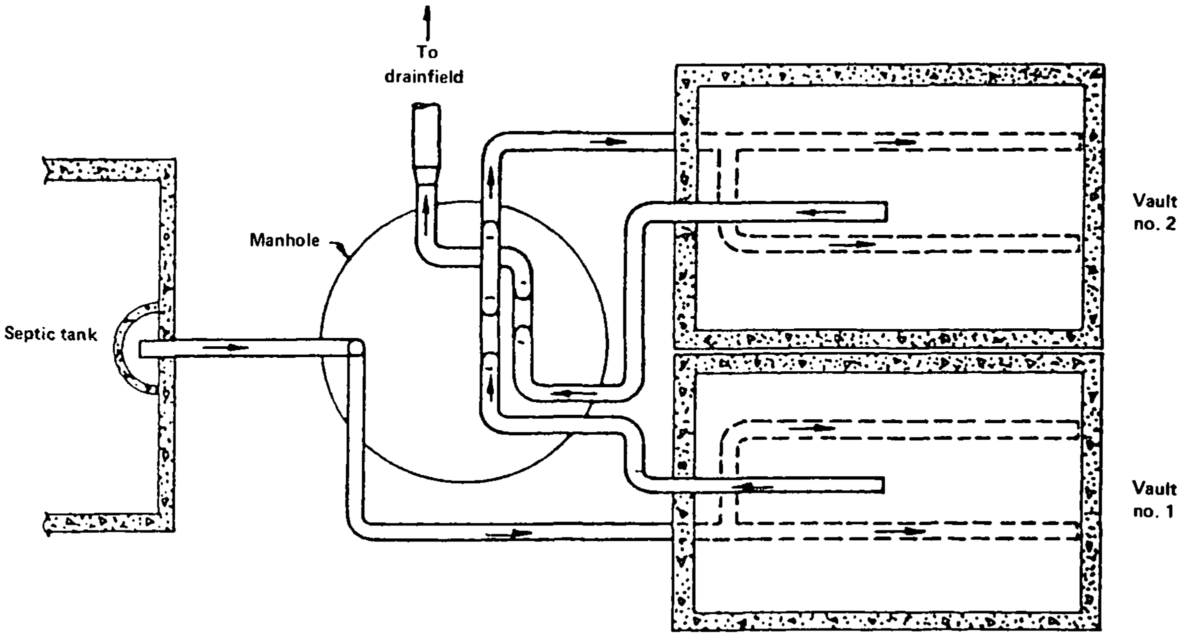

Fig. 2.5 Septic tank, manhole and anaerobic filter (plan view)

2.4.3 Raman and Chakladar's Design

Raman & Chakladar's (1972) devices are used to treat septic tank effluent.

Type 1: An upflow (anaerobic) filter with a rectangular chamber (Fig. 2.6), 1.2. m by 0.7 m in plan (4' by 2' 3"), was constructed to treat the effluent from a septic tank (of liquid capacity 2.8 m^3) of a house with 17 people in a village near Singur, West Bengal, India. The chamber was filled to a depth of 69 cm (2'3") with overburnt brick-bats of sizes ranging from 1.9 cm (0.75 inch) at the bottom, to 0.2 cm (0.0625 inch) at the top. The media rested on a perforated concrete false bottom of the filter chamber. The influent entered through a 15 cm (6 inches) diameter asbestos cement pipe and was distributed upward through the media, through the perforated bottom. After several months of operation, the filter media was changed to a uniform size of 1.27 cm (0.5 inches) at a depth of 45 cm (18 inches) at the top and 1.9 cm (0.75 inches) at the bottom 15 cm (6 inches). The volume of the filter was also reduced by reducing its length by 0.3 m.

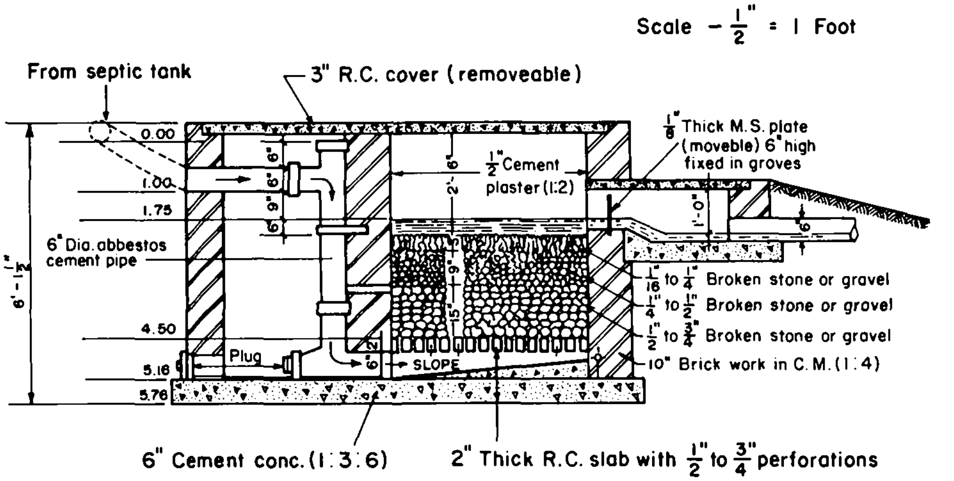

Fig. 2.6 Type 1, upflow filter (Raman & Chakladar, 1972)

A vertical pipe was fitted with a tee at the bottom, one branch of which led to the filter and the other branch was kept plugged while the filter functioned. The plug could be removed to facilitate emptying into an adjoining chamber (Fig. 2.6) and cleaning the filter when required. The effluent from the top of the filter bed escaped over a V-notch to a nearby ditch, the sill level being kept 15 cm (6 inches) above the top of the media.

Type 2: A down and upflow filter consisted of two 0.69 m x 0.60 m (2'3" x 2'0") interconnected compartments filled with stone media for treating effluent from a septic tank of 3.9 m^3 (140 cu.ft.) capacity serving 10 people. The tank effluent was discharged over a perforated plate in the chamber, passed down, entered the second laterally at bottom and flowed upward through the filter discharging over a V-notch at the top (Fig. 2.7). The design was conceived to reduce the total depth, while increasing the length of travel of the waste in the filter.

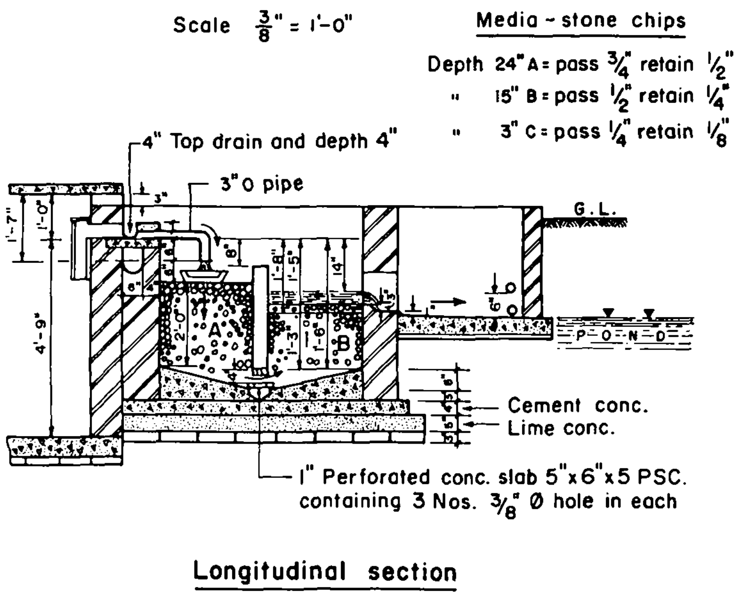

Fig. 2.7 Type 2, down and upflow filter (Raman & Chakladar, 1972)

2.4.4 Polprasert & Hoang's Design

Polprasert & Hoang (1983) design units are similar to the units used by Raman & Chaklader (1972) except that the septic tank is contiguous to the filter unit in the Polprasert & Hoang's (1983) design. Each unit built consisted of three rectangular chambers; the first was a single chamber septic tank, the second relatively small in configuration, housed a 12 cm diameter polyvinyl chloride pipe to convey flow of septic tank effluent to the anaerobic filter, and the third is the anaerobic filter chamber (Fig. 2.8).

2.4.5 Green's Design (U.S. Patent October 6, 1981, 4, 293, 421)

A very practical design requiring only slight modification of the current standard septic tank installation practices (Fig. 2.9) was developed by Green and Associates of USA (1981). Gravel media is filled around the excavation space to a depth equal to that of the septic tank and effluent is then piped around the tank to the influent side of the excavation. Effluent filters through the media towards a perforated riser which is the influent pipe to the drainfield.

2.4.6 Proposed World Bank Design

A configuration shown in Fig. 2.10 is proposed in a recent publication of the World Bank as an alternative design (World Bank, 1980). It shows two-compartment septic tank with an anaerobic upflow filter. Mara (1976) proposed a similar design and stated that the filter effluent may be discharged into a stream or disposed of in drainfield trenches or evapotranspiration beds.

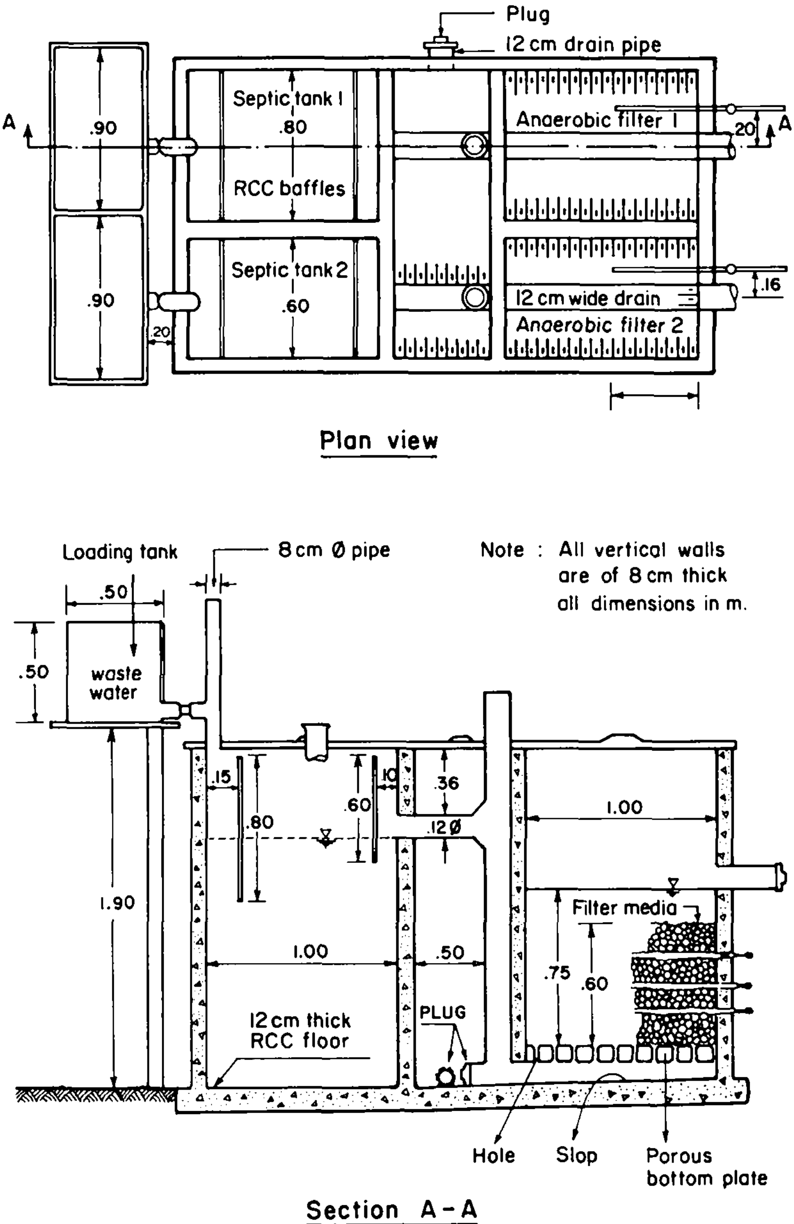

Fig. 2.8 Septic tank anaerobic filter units
(Polprasert & Hoang, 1983)

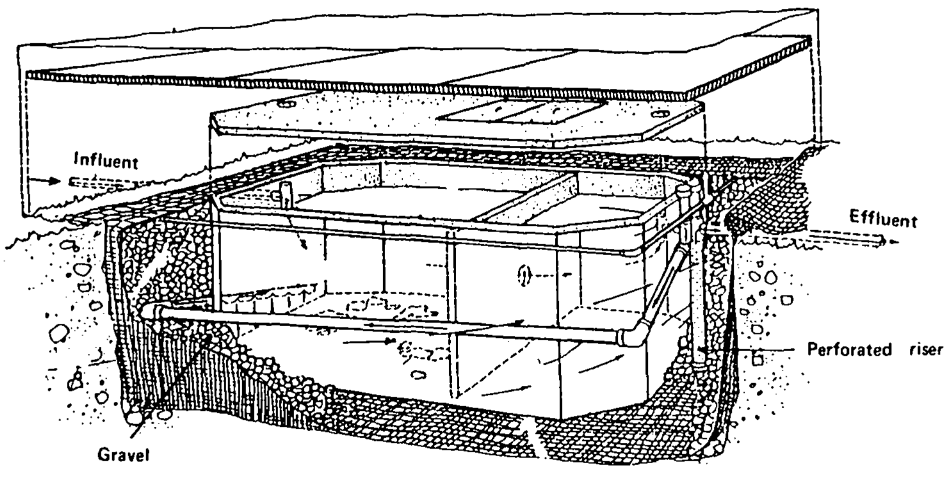

Fig. 2.9 Wrap-around filter (Green & Associates, 1981, U.S. Patent No. 4, 293, 421)

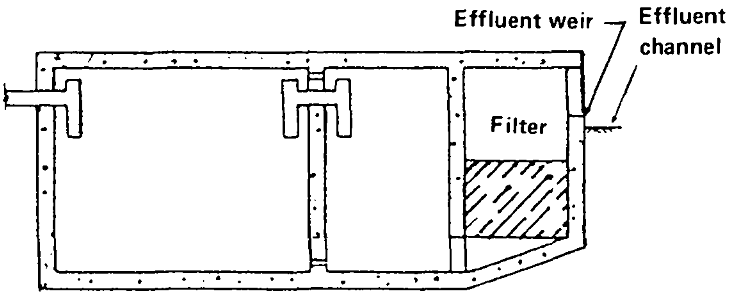

Fig. 2.10 Two-compartment septic tank with upflow filter (World Bank, 1980)

2.5 Application Status

Both full-scale and laboratory, or pilot-scale applications are categorized as municipal (low strength) applications and industrial (high strength) applications.

2.5.1 Municipal Applications

A summary of anaerobic filter investigations on dilute wastewaters including sanitary wastewaters is summarized by Viraraghavan & Kent (1983) and is presented in Table 2.1.

DeWalle et al. (1981) reported on studies at the University of Washington which evaluated the applicability of the anaerobic filter to treat sanitary wastewater. Viraraghavan & Kent (1983) reported on the studies conducted by Caudill (1968) to treat settled domestic wastewater where a 60% COD removal was achieved at a loading rate of 0.7 kg COD/m^3-d. According to Viraraghavan & Kent (1983), experiments conducted by Thulow (1974) achieved a 75% COD removal and 90% SS removal at a maximum loading rate of 0.18 kg COD/m^3-d for similar wastewater.

Table 2.1. Anaerobic treatment of sanitary wastewater and dilute wastewater (Viraraghavan & Kent, 1983)

Nature of wastewater	Characteristics of wasterwater	Process	Type of study	Loading rate	HRT (h)	Removal (%)	Temperature (°C)	References
Settled wastewater (Washington State)	COD - 226 mg/L BOD - 90 mg/L	Anaerobic filter	Laboratory study	0.1 kg COD/m^3-d	25	60 (COD)		DeWalle et al. (1981) (Quoted by Viraraghavan & Kent, 1983)
Settled wastewater (Washington State)	BOD - 124 mg/L	Anaerobic filter	Laboratory study	0.18 kg COD/m^3-d	19	75 (COD) 90 (SS)		DeWalle et al. (1981)
Diluted raw wastewater (Pretoria, South Africa)	COD - 500 mg/L	Contact digester	Laboratory and pilot	0.48 kg COD/m^3-d	24	90		Pretorius (1971)
Dilute synthetic wastewater	COD - 900 mg/L	Anaerobic filter	Laboratory study	0.8 kg COD/m^3-d 1.2 kg COD/m^3-d 1.6 kg COD/m^3-d 2.2 kg COD/m^3-d	12.8 9.1 6.4 4.8	96 95 91 90	24-33 24-33 24-33 24-33	Khan & Siddiqui (1976)
Sanitary wastewater (India)	BOD - 210 mg/L SS - 272 mg/L	Upflow filter	Fullscale	0.01 kg BOD/m^3-d	6 to 8	79.5 88.5	30°	Raman & Khan (1982)
Sanitary wastewater	COD - 288 mg/L	Upflow	Laboratory study	0.32 kg COD/m^3-d	24	73	25-35	Kobayashi et al. (1983)
Diluted city wastewater	BOD-300-400 mg/L COD-1100-1300 mg/L	Contact digester	Pilot	0.64 kg BOD/m^3-d		80 (BOD)		Simpson (1971) (Quoted by Viraraghavan & Kent, 1983)
Synthetic wastewater	COD - 780 mg/L	Two stage anaerobic filter	Laboratory study	0.5 kg COD/m^3-d 1.0 kg COD/m^3-d 2.0 kg COD/m^3-d	15 8 4	83 79 71	21-26 21-26 21-26	Frostell (1979) (Quoted by Viraraghavan & Kent, 1983)
Dilute synthetic wastewater	COD-50-600 mg/L	Anaerobic attached film expanded bed	Laboratory	Up to 8 kg COD/m^3-d	6-0.33	80	10-30	Switzembaum & Sewell (1980) (Quoted by Viraraghavan & Kent, 1983)
Raw wastewater (U.S.A)	BOD - 220 mg/L (mean)	Upflow filter	Pilot scale	0.24 kg BOD/m^3-d	10	55	10-25	Genung et al. (1981) (Quoted by Viraraghavan & Kent, 1983)
Raw wastewater (India)	BOD - 175 mg/L - 210 mg/L	Upflow filter	Pilot scale	0.26 kg BOD/m^3-d 0.23 kg BOD/m^3-d	5	73 80	28 31	Raman & Khan (1978); Kennedy (1980) (Quoted by Viraraghavan & Kent, 1983)
Raw wastewater (Washington state)	BOD - 120 mg/L SS - 140 mg/L	Upflow filter	Bench			61 (BOD) 64 (SS)		Kennedy (1980) (Quoted by Viraraghavan & Kent, 1983)

Pretorius (1971) investigated the applicability of the anaerobic process for raw wastewater pretreatment. He observed that about 90% of the raw wastewater COD was reduced after a detention time of 24 hours at a temperature of 20°C. He further stated that the hydraulic loading rate could be a better parameter for design purposes than waste concentration.

Khan and Siddiqui (1976) using both low and high strength raw wastewater found that the filter depth was optimized at 1.2 m. Witherow et al. (1958) were unable to improve the efficiency of the filter by doubling the depth of media from 1.22 to 2.45 m sufficiently enough to recommend this modification.

Raman and Khan (1982) recommended that the raw wastewater should be free from grit and preferably homogenized prior to anaerobic upflow treatment. They also reported the effluent to be relatively odor free and that overland grass filtration was a useful adjunct for further polishing.

Kobayashi et al. (1983) used a laboratory scale anaerobic filter packed with synthetic high surface area media to treat domestic wastewater. He found that effluent quality declined as the temperature decreased from 25°C to 20°C but that performance at 25°C and 35°C degrees was not significantly different. It was further noted that the filter was relatively insensitive to daily fluctuations of influent wastewater quality.

As quoted by Switzenbaum et al. (1984), Genung et al. (1983) reported on a 19 m^3/d pilot plant anaerobic upflow (ANFLOW) fixed-film reactor at Oak Ridge National Laboratory treating raw municipal wastewaters at ambient temperatures with a HRT of 9.43 - 62.4 hours. A summary of the 18 month period data showed a SS and BOD_5 removal efficiencies of 69.4% and 52.9% respectively. Average influent SS, and BOD_5 concentrations were 140.6 mg/L and 135.7 mg/L respectively.

The performance of an anaerobic upflow (ANFLOW) fixed-film bioreactor was studied on a near-commercial scale in Knoxville, Tennessee, with a 190 m^3/d facility from August 1981 to October 1983 (Harris et al., 1985). During treatment of low-strength municipal wastewater before primary sedimentation, the effluent met the U.S. EPA secondary treatment discharge limits of 30 mg/L for TSS and BOD the vast majority of the time, with only an occasional increase to the 39 to 40 mg/L range for perhaps 1 to 2 days. Loading rates were approximately 0.25 kg/m^3-d of TSS and BOD each, and the hydraulic retention time was 9 to 10 hours.

A recent study prepared by ADI Limited (1984) for the National Research Council of Canada investigated the possibility of incorporating the static bed submerged media anaerobic reactor (SMAR) technology into municipal wastewater treatment in Canada. The study concluded that there may be some potential for SMAR application, especially for several plants handling high-strength wastewater where average flow currently exceeded the design capacity; the study indicated that greater opportunity existed in warm-climate developing countries for utilizing SMAR technology in municipal wastewater treatment.

2.5.2 Septic Tank Effluent Treatment

A summary of anaerobic filter performance while treating septic tank effluent is presented in Table 2.2 (Viraraghavan & Kent, 1983).

Table 2.2. Performance summary for anaerobic filters treating septic tank effluent (Viraraghavan & Kent, 1983)

Location of study	Characteristics of wasterwater	Process	Type of study	Loading rate	HRT (h)	Removal (%)	Temperature (°C)	References
India	BOD - 240 mg/L	Upflow	Full-scale	0.02 kg BOD/m^3-d	6 d	71		Raman & Khan (1978)
India	BOD - 210 mg/L	Downflow and upflow	Full-scale	Low loading	High	75		Raman & Khan (1978)
Washington State	BOD - 241 mg/L	Upflow	Full-scale			26		Seabloom et al. (1981)
	COD - 486 mg/L					28	12.4	
India	BOD - 290 mg/L	Upflow	Laboratory	0.34 kg BOD/m^3-d		76	23-33	Raman & Khan (1978)
Thailand	COD - 310 mg/L	Upflow	Full-scale	0.5 kg COD/m^3-d	24	60		Suthakar (1981)
	SS - 170 mg/L					70		
Washington State	BOD - 103 mg/L	Upflow	Full-scale	0.04 kg BOD/m^3-d	38.4	28		Hamilton (1976) (Quoted by Viraraghavan & Kent, 1983)
	SS - 67 mg/L					42		
	COD - 305 mg/L					23		
Washington State	BOD - 217 mg/L	Upflow (trench)	Full-scale	0.027 kg BOD/m^3-d		39		Hamilton (1976) (Quoted by Viraraghavan & Kent, 1983)
	SS - 33 mg/L					33		
	COD - 542 mg/L					21		
Washington State		Upflow	Bench scale			85 (BOD) 90 (SS) 80 (COD)	7 and 14	Hamilton (1976) (Quoted by Viraraghavan & Kent, 1983)
Washington State	BOD - 120 mg/L	Upflow	Full-scale	0.19 kg COD/m^3-d	19	28 (COD)	11.8	Hamilton (1975) (Quoted by Viraraghavan & Kent, 1983)
Washington State		Downflow sand filter				60 (COD) 28 (SS)		DeWalle et al. (1981)

Raman and Chakladar (1972, 1978) found that the high efficiencies of anaerobic filter can be maintained even at low influent concentrations of 125 to 120 mg/L BOD, with either continuous or intermittent flow. They noted that although temperatures varied from 12.5°C to 26°C during the winter and 25°C to 36°C during the summer, the treatment efficiencies did not vary appreciably. It was further noted that it may not be practical to reduce the effluent BOD below 30 mg/L with the reported sizes of filter and media, and recommended that further effluent polishing could be accomplished by aeration.

Polprasert and Hoang (1983) reported that after a four day retention period in the anaerobic filter, the effluent was almost free of bacteria. However, five of eight samples still contained 10^2 to 10^3 MPN/100 ml of bacteriophages and may not be safe for disposal to areas where viral infection is prevalent.

Raman and Khan (1977) found that the removal of helminthic ova (Ascaris eggs) was almost 100 percent in most of the field and laboratory units. After two years of field observations, BOD removal was 70 to 78 percent based on an influent concentration of 170 to 250 mg/L at an average loading of 0.9 kg BOD/m^3-d.

Hoang (1981) found that, although there was no direct relationship between organic loading and viral removal in an anaerobic upflow filter, at the same retention time, bacteriophage reduction increased with lower level organic loading to the filter. Greatest reductions were found at a retention time of four days.

Suthakar (1981) reported that a one day retention time, correspoding to a 0.43 m^3/m^2/day hydraulic loading, was optimal for the reduction of organic components.

As quoted by Viraraghavan & Kent (1983), Hamilton (1976) reported findings from two full scale studies treating septic tank effluent from single family dwelling in Washington State. This system is reported to have operated for almost four years without clogging or solids removal. At the first site, after a 1.6 day retention time, 42% SS, 23% COD, 28% BOD_5 and 28% filtered COD removal rates were observed. At the second site 39% BOD_5, 33% SS and 21% COD removal rates were produced after acclimatization.

2.5.3 Industrial Applications

During the last two decades many investigations have been carried out with a wide variety of organic industrial wastes. The tendency has been to try to treat increasingly more difficult wastes. A summary of several anaerobic filter studies for variety of industrial wastes is summarized by Switzenbaum (1982) and is given in Table 2.3. In general 70% and greater substrate organic removal efficiencies are reported for these systems. Switzenbaum (1983) reported on full-scale experiences with anaerobic filter systems and detailes are presented in Table 2.4.

Anderson et al. (1984) reported on several full-scale units now in operation, with a particular reference to shock loading. A soluble carbohydrate waste water with maximum and average COD concentrations of 90,000 mg/L, and 58,000 mg/L respectively was treated successfully in an anaerobic filter (COD removal = 99.4%) at an organic loading rate of 14 kg COD/m^3-d; the treated effluent SS was 350 mg/L. Further, the response to shock load was investigated. In the case of high strength carbohydrate wastewater, the organic loading rate has been as high as 28.5 kg COD/m^3-d for eleven days, at which point the treated effluent COD increased to 3,100 mg/L. On reduction of the

Table 2.3. Anaerobic filter studies for industrial wastes (Switzenbaum, 1982)

Waste	Influent (mg/L)	Effluent (mg/L)	Removal (%)	Loading rate (kg/m^3-d)	HRT (h)	Temperature (°C)	References	Remarks
Glucose-nutrient broth	COD-1500 COD-3000	112-950 204-1105	36.7-92.1 63.0-93.4	0.42-3.4 0.42-3.4	4.5-36 9-72	25	Young & McCarty (1969)	-
Volatile acids	COD-1500	24-476	68.4-99.4	0.42-3.4	4.5-36	25	Young & McCarty (1969)	
	COD-3000 COD-6000	42-240 139-794	92-98.6 86.9-97.7	0.42-0.85 1.7-3.4	36-72 18-36			
Metrecal	COD-10,000	-	75-90	-	18	-	El-Shafie & Bloodgood (1973)	6 beds in series
Pharmaceutical waste	BOD_5-2000	-	94% BOD_5 70-80% COD	-	36	35	Jennett & Rand (1981) (Quoted by Switzenbaum, 1982)	-
Protein-carbohydrate waste	-	-	50-90% COD	3.2-27.2	3-24	35	Mueller & Mancini (1975)	
Whey	COD-8100	-	98%	1.9	-	22-25	Hakannson (1972) (Quoted by Switzenbaum, 1982)	
Whey permeate	COD-5000	-	95%	0.7	-	22-25	Hakannson (1977) (Quoted by Switzenbaum, 1982)	

Table 2.3. Anaerobic filter studies for industrial wastes (Switzenbaum, 1982) (cont'd)

Waste	Influent (mg/L)	Effluent (mg/L)	Removal (%)	Loading rate (kg/m^3-d)	HRT (h)	Temperature (°C)	References	Remarks
Food processing carbohydrate waste	COD-8475 COD-5200	546-5000 975-3890	41.0-93.6 25.2-81.3	-	0.54-3.45 days 0.54-3.45 days	35	Plummer *et al.* (1968)	
Waste sulfite liquor	BOD-1300-BOD-5300	-	27-58	-	3.7-4.0 days		Wilson & Timpany (Quoted by Switzenbaum, 1982)	
Potato processing	COD-3000	-	41-79	-	0.5-2.5	19-22	Paitrop *et al.* (1971)	
Fish processing	COD-466 COD-407 BOD-310	90 102 58	80.7 74.9	-	3.1 days 1.6 days	26 23.5	Hudson *et al.* (1978)	
Sludge heat treatment liquor	COD-10,000	-	75-85% BOD 55-66% COD	6.5	1.5 days		Donavan (1981) (Quoted by Switzenbaum, 1982)	Design values from pilot study
Chemical plant high strength industrial waste			65% COD	16.330 kg/day			Ragan (1981) (Quoted by Switzenbaum, 1982)	Design value for the Celanese system
Domestic wastewater				0.95	1.5 days		Koon, *et al.* (1980) (Quoted by Switzenbaum, 1982)	Design values, the Anflow system

Table 2.4. Full-scale upflow anaerobic filter installations (Switzenbaum, 1983)

Location	Type	COD (mg/L)	Design flow (m^3/d)	HRT (day)	Media	COD load (kg/m^3-d)	Removal (%)
Spokane, WA[a]	Starch gluten	8800	490	0.9	Graded rock, 2.5-7.6 cm	3.8	64
Vernon, TX	Guar gum	9140	823	1	9 cm pall rings	16	60
San Juan, PR[b]	Rum distil.	95000	1325	7-8	Synthetic ("vinyl core")	8.9	75
Bishop, TX	Chemical	12000	3785	1.5	9 cm pall rings	9.6	80
Pampa, TX	Chemical	14400	3785	1.5	9 cm pall rings	10.4	90

[a] No longer in operation
[b] Downflow anaerobic filter

organic loading rate to 14 kg COD/m^3-d, the treated effluent COD started to fall immediately and had returned to approximately 500 mg/L within five days.

A plant containing effluent from a molasses distillery with a COD in the range 40,000-85,000 mg/L, BOD_5 20,000-36,000 mg/L, SO_4^{-2} 4,000-8,000 mg/L and suspended solids 3,000-8,000 mg/L and with a pH 4.2-4.8 has been treated to achieve 70-45% COD removal efficiencies associated with 85% to 50% BOD_5 removal efficiencies. With a reactor pH of 7.2 and 75% BOD removal efficiency the methane yield of 0.35 m^3/kg COD removal has been achieved. The plant was operated continuously, on a five days per week basis, and was tested for its response at start-up following a winter shut down. Following a two-day cessation in feeding, the plant produced gas immediately when the feed was reinstated. During the two week shut down in the winter the temperature within the plant decreased to 9°C. At restart, the plant was initially heated to 30°C before feeding commenced at the maximum rate. The rate of increase in gas production rate under these conditions was identical to that observed following a two-day cessation in feeding, with a maximum production rate being observed within 8 hours of feed reinstatement.

An upflow anaerobic filter treating an effluent from whey processing which has a COD of 8,000-10,000 mg/L with an associated BOD of 5,000-6,500 mg/L was reported. To date, (Anderson et al., 1984) approximately 85% BOD removal was observed at organic loading rates in excess of 15 kg COD/m^3-d. The plant has demonstrated good stability with SS concentrations in the effluent not exceeding 350 mg/L.

Experimental work by Anderson et al. (1984) showed the feasibility of the application of anaerobic filters for denitrification for a methanol and potassium nitrate waste. Results are summarized in Table 2.5.

Table 2.5. Anaerobic filter (denitrification) (Anderson & Donnelly 1984)

Waste	Temp. (°C)	Hydraulic retention (days)	Feed COD:NO_3-N	NO_3-N loading (kg/m^3-d)	Organic loading (kg COD/m^3-d)	Percent removal NO_3-N	Percent removal COD
Methanol and potassium nitrate	Room Temp. (20-24)	0.10	4.42	9.840	43.50	94.5	64.0
		0.14	3.36	6.860	23.10	93.0	97.5
		0.24	3.09	2.840	8.79	97.5	97.0
		0.49	5.43	0.984	5.34	99.2	59.0
		1.00	2.76	0.645	1.775	86.2	98.7
		1.40	3.12	3.570	1.775	91.5	99.4

2.6 Applicability

In terms of biofilm thickness control, the anaerobic filter is frequently stated to need a special biofilm control device in the form of backwash (Switzenbaum, 1982). To some extent, anaerobic filters are suitable for wastes with suspended organics but may encounter filter clogging. It tolerates severe hydraulic overloading and is able to withstand high loading variations and to operate at a once per day or twice per day loading pattern.

Concerning the treatment of liquid wastes with a high fractional content of suspended solids, the anaerobic filter presumably is less suitable, particularly if the biodegradability of the suspended matter is poor. Due to lack of mixing, these solids will accumulate in the lower part of the filter and ultimately it may drive the viable sludge out of the reactor (Lettinga et al., 1982). Moreover, except in removing SS, the anaerobic filter process performs quite well in removing dissolved biodegradable organic matter. As far as medium strength dissolved wastes are concerned, quite satisfactory results have been obtained with small pilot plant anaerobic filter processes, although so far this only seems to be the case at organic loading approximately below 10 kg COD/m^3-d (Lettinga et al., 1982).

Anaerobic filters are capable of minimizing the severity of the response to toxic slugs due to its quasiplug flow regime, which passes the toxic slug quickly without washing the biomass (Parkin & Speece, 1984). In the case of chronic toxicity, optional recycle facilities facilitate the dilution and maximum acclimation potential for micro-organisms which also maximize the potential for degradation of biodegradable toxicants.

Retention times of the order of several days are needed for successful operation (Switzenbaum, 1983). Based on the studies conducted at University of Washington (Seabloom et al., 1981), and studies on anaerobic filter treatment of septic tank effluent carried out at AIT, Bangkok (Polprasert & Hoang, 1983), the anaerobic filter should be operated at a hydraulic retention time of at least four days to reduce the chances of microbial pollution especially that of enteric bacteria. However, it is observed that even with a hydraulic retention time of two days, approximately 75% fecal coliform removal and 60% bacteriophage removal occured, thus providing a better effluent for further treatment through soil (Viraraghavan & Kent 1983).

2.7 Problems Associated with Anaerobic Filter

2.7.1 Channeling and Short Circuiting

In anaerobic filters, gas bubbles may adhere to flocs/bed particles and cause these to rise in the reactor, and may result in washout of biomass or deterioration of the effluent quality. Gas bubbles entrapped in filters may cause channeling and short-circuiting in the reactor (Henz & Harremoes, 1982).

2.7.2 Filter Clogging

Due to clogging only part of the retained sludge will effectively contact the wastewater, as a result the contact time between the sludge and the wastewater will be relatively short.

In full-scale studies (Raman & Chakladar, 1983; Khan & Siddique, 1976; Raman & Khan, 1977; Raman & Khan, 1978), filter clogging has been reported after 18 months of continuous operation. Wasting of sludge from the filter can be accomplished by flushing water from the top through an idle filter and removing the solids (Raman & Chakladar, 1972).

2.7.3 Start-up of an Anaerobic Filter

The start-up time of an anaerobic filter is directly proportional to the concentration of the microbial population. Lag times may be reduced by seeding. Young & McCarty (1969) noted micro-organisms in an unseeded filter remained dispersed and a significant fraction washed out with the filter effluent, whereas in a highly seeded filter, rapid flocculation was observed, causing the biomass to remain in the filter.

Raman & Chakladar (1972) and Raman & Khan (1977) reported that without seeding, four to six weeks of continuous operation at temperatures between 25 and 32°C were required for start-up, and three months were required before the filter became mature. Additional information from the literature concerning the start-up of anaerobic filter processes as summarized by Letting et al. (1982) is presented in Table 2.6.

2.7.4 Removal of Nutrients

The anaerobic upflow filter generally does not remove nutrients such as nitrogen and phosphates. Investigators (Raman & Khan, 1982; Kobayashi et al, 1983; Hamilton, 1975) report both ammonia-nitrogen and phosphate levels increasing in the effluent. Their results are summarized in Table 2.7.

Kennedy (1981) states that the anaerobic filter can be used as a denitrifying chamber preceeded by aerated pretreatment, rather than primary settling. An additional source of carbon is normally required to complete the denitrification process. The carbon source could either be the traditional methanol, or greywater (Laak et al. 1981).

2.8 Advantages and Disadvantages

Advantages (Viraraghavan & Kent, 1983)

The main advantages are as follows:

- simplicity in construction, operation and maintenance;
- more efficient waste stabilization since organics are removed from the wastewater as methane and carbon dioxide gases, rather than fixed cells, resulting in a low production of waste biological sludge;
- low head loss - less than 15 cm in normal operation;
- low nutrient requirements;
- clear, odor and nuisance-free effluent;
- efficiency is not affected by intermittent or transient nature of flows;

Table 2.6. Period of time required for the first start-up of anaerobic filters (literature data) (Lettinga *et al.*, 1982)

A.F. Process investigated	Wastewater used in the investigation	Initial space load (kg COD/m^3-d)	Inoculum used in the start-up	Results obtained (days for 90% red)	References
28.5 L quarz-stone 28.5 L quarz-stone	3.0 g/L VFA-mixture 3.0 g/L VFA-mixture	0.42 (30°C) 0.84 (30°C)	2 x 30 g VSS Dig. Sew. Sl. light seed Dig. Sew. Sl.	40 days 180 days	Young & McCarty (1969)
14.2 L quarz gravel	(glucose) pharmar-ceutical, 1-16 g COD/L	0.37 (30°C) 1.76 (30°C)	30 g VSS Dig. Sew. Sl. 30 g VSS Dig. Sew. Sl.	38 - 40 days 100 days	Jennett & Dennis (1975)
33.4 limestone	Brewery press liquor waste (6 g/L)	0.8	Supernatant anaerobic digester	25 days	Lovan & Foree (1976)
24 L coke packed	Potato starch waste 4-6 g/L	1.0 (30°C) 0.5 (30°C)	14 g VSS slightly adapted Dig. Sew. Sl.	36 days 24 days	Lettinga & Janssen (Quoted by Lettinga *et al.*, 1982)
6.7 L plastic porous media packed	Soluble starch COD: 8.7 g/L	0.7 first week 1.4 second week 2.5 week 3-5 5 week 5-8	63 g VSS Dig. Sew. Sl.	first week second week week 4-5 week 6-8	Frostell (1981)

Table 2.7. Ammonia-nitrogen and phosphates in anaerobic filter effluent

Parameter	Influent (mg/L)	Effluent (mg/L)	Increase (%)	References
NH_3-N	25.4	34.8	37	Hamilton (1975) (Quoted by Viraraghavan & Kent, 1983)
NH_3-N	41	52	27	Raman & Khan (1982)
NH_3-N	33	44	33	Kobayashi _et al._ (1983)
Ortho-PO_4^{-3}	6.4	7.5	17	Hamilton (1975) (Quoted by Viraraghavan & Kent, 1983)
Total PO_4^{-3}	3	5	67	Kobayashi _et al._ (1983)

- due to the presence of packing material, wash-out of the suspended matter is reduced;
- is more efficient than a septic tank alone;
- the filter can be used alone or in conjunction with a septic tank;
- when used in conjunction with a septic tank, the filter prevents most of the solids discharged from the septic tank from entering the tile field;
- requires only occasional cleaning;
- provides excellent removal of pathogens;
- the filter requires no energy to operate;
- the anaerobic filter may be used for denitrification;
- long solids retention times afford a high level of treatment without long hydraulic retention times; and
- extreme actions such as passing air through the unit or permitting the pH to drop as low as 5.5 or letting the filter stands for weeks or months with zero loading, do not affect the ability of the filter for rapid recovery.

<u>Disadvantages</u>:

- channeling (short-circuiting) will be promoted due to the fact that gas bubbles select a limited number of channels for escaping upwards through the filter;
- mixing of the sludge is seriously hampered;

- depending on the type of packing a smaller or greater fraction of the reacter volume is lost for retaining sludge;

- rapid clogging of filter may occur in treating wastes containing a relatively high fraction of suspended solids;

- longer period are required for starting the process than with an aerobic process;

- the process is not likely to achieve effluent quality better than 30 mg/L BOD; and

- long solids retention times are required; thereby, allowing the filter to adjust less readily to environmental changes.

2.9 Conclusion

At present, the use of anaerobic filters for municipal wastewater treatment, is limited to pilot scale systems, but the trend indicates that such systems may become operational in the future, especially in the developing countries.

A review of the published studies showed that septic tank effluent treatment using an anaerobic filter is practical and generally cost-effective for achieving a much better effluent quality (Viraraghavan & Kent, 1983).

A variety of industrial wastes are amenable to treatment by anaerobic filter process. In general, 70% and greater organic removal efficiencies are reported for these systems (Switzenbaum, 1982). In many cases final effluent quality is not suitable for direct discharge, but significant portion of the organic load are removed. In addition, the benefits of methane production and lower residuals products are achieved.

III. ANAEROBIC STATIONARY FIXED FILM TREATMENT

3.1 General

Downflow stationary fixed film (DSFF) reactors are a relatively recent addition to the family of advanced high-rate anaerobic reactors, all of which are based on retention of the active biomass. The DSFF reactor distinguishes itself from other type of advanced reactors by the downflow mode of operation, the architecture of its packing (fixed biofilm support), and the absence or near-absence of suspended growth.

3.2 Principles

The method of operation of DSFF reactors is shown in Fig. 3.1. Waste is pumped in at the top, together with recycle effluent when desired, and effluent is withdrawn from the bottom. When started, inoculum from an active digester is recirculated and bacteria attach themselves to the channel walls of the support material to form a biofilm. The downflow mode allows any settleable material which might otherwise accumulate in the system to be removed with the effluent and reduce the risk of column plugging. Mixing in the reactor is produced entirely by the action of rising gas bubbles. Hence the high concentration of substrates are immediately dispersed and there is little need for an elaborate agitation system. With counter-current two-phase flow, the system exhibits good mixing characteristics and reduces potential for high localized concentrations of inhibitors or volatile acids (Hall & Melcer, 1984).

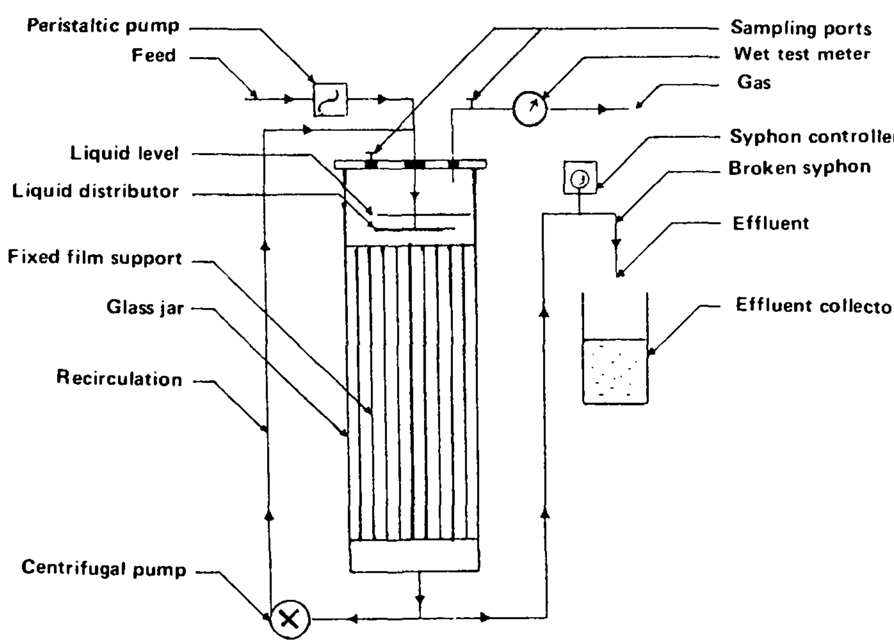

Fig. 3.1 Set-up multi-channel fixed film digester
(after Kennedy & van den Berg, 1982)

At present time, the anaerobic DSFF reactor is in an infant stage of development and its application for municipal waste is not being tested although studies with laboratory, pilot scale reactors and industrial scale reactors are documented.

3.2.1 System Design

The design of DSFF reactor consists of wastewater distribution system, a biofilm support structure, head space, effluent draw off and recycle facilities.

The downflow mode of operation dictated a stationary film support to maintain the film of micro-organisms in the reactor. Additionally, to prevent settling of suspended solids on parts of the film support surface, the stationary film support is arranged in more or less vertical channels and are made of potters clay, draintile clay, needle punched polyester or polyvinyl chloride. Reactor size and height have relatively little effect on the performance (when expressed in surface-to-volume ratio) but the reactor configuration and operation have marked effect on performance. Generally multi-channel reactors have not performed as well as reactors with only few channels (van den Berg et al., 1985).

3.2.2 Recirculation

As quoted by van den Berg et al. (1985), recirculation generally improves the performance of the DSFF reactor (Duff & Kennedy, 1982; Samson et al., 1984). With wastes containing large amounts of hard-to-digest suspended solids (e.g. piggery waste) recirculation helps to keep these in suspension and aid in their degradation (Kennedy & van den Berg, 1982b). Recirculation of effluent helps to maintain a uniform and relatively thin film (van den Berg et al., 1985).

3.3 Design Criteria

3.3.1 Support Material

Support material affects the rate of start-up markedly. The effect of support material on start-up and steady-state performance is shown in Fig. 3.2.

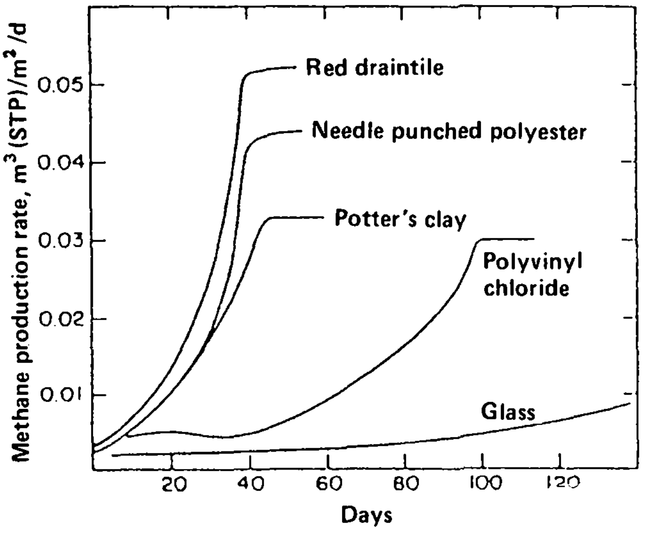

Fig. 3.2 Effect of support material on start-up and steady-state performance (35°C); bean blanching waste at 10 g COD/L; reactor volume; 0.8-1.2 L; area-to-volume ratio, 100-150 m^2/m^3) (van den Berg et al., 1985)

Reactors made from rigid foam polyvinyl chloride could not be started at all, while the glass reactors were slow to start up, presumably because bacteria had difficulty attaching themselves to the smooth inert surface. Solid polyvinyl chloride (PVC, used extensively in biological wastewater treatment) was substantially better than glass as a film support, but not as good as the fired clay and needle punched polyester. The earlier studies also indicated that inert support media with a rough surface enhanced biofilm accumulation and reactor performance. Physical roughening of smooth plastic surfaces and addition of sawdust to clay support media before firing (unpublished results) have been shown to enhance start-up and overall reactor performance (van den Berg et al., 1985).

3.3.2 Surface-to-Volume Ratio

The amount of retained biomass in a DSFF reactor depends on the surface to volume ratio and is therefore limited by the support matrix area. The importance of surface-to-volume ratio on start-up and ultimate loading rates for reactors of the same height is shown in Fig. 3.3. Reactors with larger surface-to-volume ratio achieved higher space loading rates and higher rates of methane gas production (Kennedy & Droste, 1984).

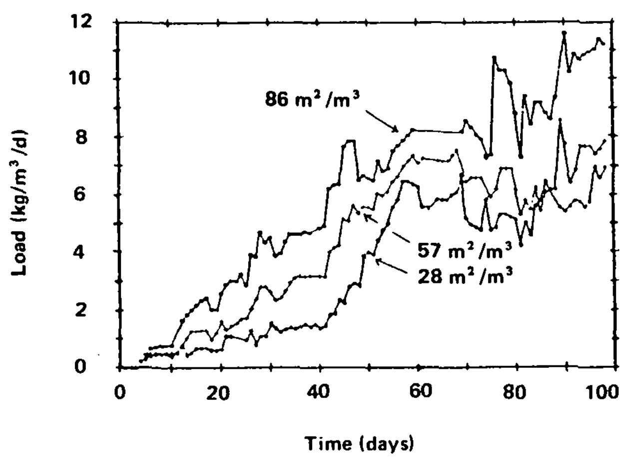

Fig. 3.3 Effect of surface to volume ratio in reactor of equal height on changes in loading rate during start-up (van den Berg et al., 1985)

3.3.3 Organic Volumetric Loading Rate

Loading rates and organic removal efficiencies depended on the total amount of active biomass retained as well as on type of waste. For a wide variety of industrial wastes, loading rates of 5 to 15 kg COD/m^3-d were readily obtained with 70-95% COD removal efficiency, depending on loading rate and type of waste (van den Berg et al., 1985).

van den Berg et al. (1980) with their experience with bean blanching waste, found that the small fixed film reactors could be loaded substantially higher than larger ones and this difference is not explained. The size of the reactor did not affect the COD removal efficiency.

Maximum COD loading rates, using bean blanching waste (9,500-10,500 COD mg/L, mainly soluble) were as high as 20 kg COD/m^3-d (depending on surface to volume ratio and size) to achieve 86% removal efficiency (Table 3.1).

Table 3.1. Performance data for anaerobic fixed film reactors of two different sizes, fed with bean blanching waste (van den Berg et al., 1980)

Parameters	Fixed film reactor size	
	0.7-liter[a]	35-liter[b]
Minimum hydraulic retention time, days	0.5	1
Maximum COD loading rate (M-COD-LR), kg/m^3-d	20[c]	10
COD removal efficiency, %, at M-COD-LR	86	86
Suspended COD of effluent		
1) %	0.09	0.09
2) % of total effluent COD	65	65

[a] Surface to volume ratio, 140 m^2/m^3.
[b] Surface to volume ratio, 120 m^2/m^3 - reactor may not have reached maximum loading rate in test run
[c] Independent of waste strength (0.5-2.0% COD)

For simulated sewage sludge (55,000 COD mg/L, over 45,000 which is suspended) the maximum COD loading rate of 12 kg COD/m^3-d with an efficiency of 70% removal is reported.

3.3.4 Temperature

Kennedy & van den Berg (1981) reported the effect of temperature on the performance of DSFF reactors treating bean blanching waste and chemical industry wastes. The maximum loading rates decreased linearly with the temperature. The fixed film reactor treating bean blanching waste, mainly containing soluble starch and protein (Total COD = 10 g/L) was reported to be capable of achieving high loading rates without a substantial change in COD removal efficiency at a temperature range of 10°C to 35°C. They observed that a decrease in temperature from 35°C to 25°C, decreased the maximum steady-state loading rate by 37%, while at 10°C the loading rate was reduced by 75% of the maximum loading rate at 35°C (i.e. 18.4 kg COD/m^3-d). The COD removal efficiency was independent of temperature and remained at 88 ± 3%.

Similar results were obtained for chemical industry waste (Total COD = 14 g/L). The maximum steady-state loading rate and the volumetric methane production rate decreased by less than 25% between 35°C and 25°C. As with the bean blanching waste there was no appreciable change in the COD removal efficiency or in digester gas composition with temperature. The effect of temperature on the performance of stationary fixed film reactors digesting chemical industry waste at maximum loading rate is presented in Table 3.2. The effect of temperature on loading rate and rate of methane production is given in Fig. 3.4.

Table 3.2. Effect of temperature on the performance of stationary fixed-film reactors digesting chemical industry waste at maximum loading rates (Kennedy & van den Berg, 1981)

Parameters	Temperature (°C)	
	25	35
Loading rate (kg COD/m^3-d)	14.0	17.9
Film surface loading rate (kg COD/m^2-d)	0.100	0.128
Hydraulic retention time (days)	1.0	0.78
COD removal (%)	81	84
Methane content of digester gas (%)	54	55
Volumetric rate methane production (m^3(STP)/m^3-d)	3.7	4.9
Methane production rate of film (m^3(STP)/m^2-d)	0.026	0.036
Volatile acids (mg/L)		
Acetic acid	280 ± 20	180 ± 20
Propionic acid	170 ± 20	180 ± 20

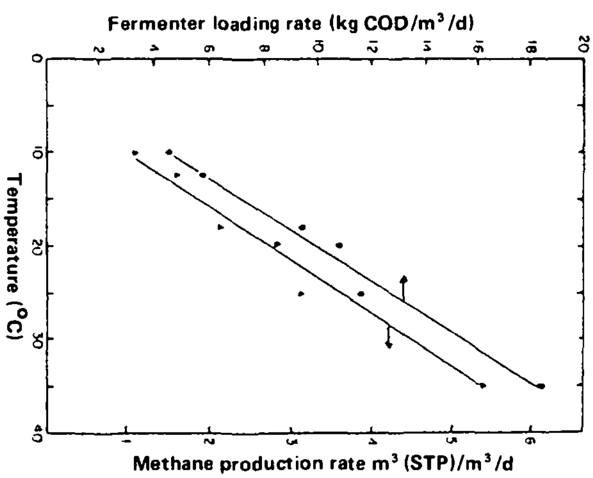

Fig. 3.4 Effect of temperature on loading rate and rate of methane production (Kennedy & van den Berg, 1981)

Kennedy & van den Berg (1981) proposed relationships between temperature and loading rate and temperature and methane production rate expressed by:

$$GP = 0.167T - 0.692$$

where: GP = methane production rate ($m^3(STP)/m^3$-d)
T = operating temperature (°C)

and

$$LR = 0.557T - 1.546$$

where: LR = COD loading rate (kg COD/m^3-d)

The coefficients in these relationships presumably depend on the nature of the substrate and the type of fixed-film support material as well as the support material configuration and surface-to-volume ratio.

van den Berg et al. (1985) reported the studies conducted by Kennedy & van den Berg in 1982 which stated that the DSFF reactors can be operated at thermophilic temperatures (55°C), but maximum loading rates and COD removals were similar to those at mesophilic temperatures.

3.3.5 Hydraulic Retention Time

Hydraulic retention time (HRT) is the ratio between the void volume of the reactor and the volumetric flow rate. An intermediate HRT is desirable for some high strength waste due to poor conversion of wastes to methane at very short HRT's. With laboratory experience, it was found that the HRT varies from few hours to number of days depending on the wastes characteristics and the strength.

3.4 Application Status

van den Berg et al. (1985) summarized the performance data of downflow stationary fixed film reactors (DSFF) at steady-state loading (Table 3.3) for a wide variety of wastes tested and for a wide range of reactor sizes. COD loading rates between 5 to 15 kg/m^3-d were readily maintained and COD removal efficiencies of 60 to 95% are reported. The composition of wastes treated in DSFF reactors is given in Table 3.4.

Samson et al., 1984 reported the experiences with an industrial scale 400 m^3 reactor, using channelled ceramic blocks as support material which has been operating on cheese factory effluent for a year and is quoted by van den Berg et al. (1985). The effluent varied widely from day to day and from hour to hour. Total COD varied from about 800 to over 2,500 mg/L and the reactor was designed for a loading rate of about 7 kg COD/m^3-d at an average HRT of 7 hours. A COD removal efficiency of 60-70% at 8 kg COD/m^3-d loading rate was reported.

A couple of 50 m^3 DSFF reactors for piggery waste have been built in Canada and initial findings confirm the results presented above (Hall, 1983). A very large (13,000 m^3) DSFF reactor treating rum stillage waste at a concentration of up to 100,000 ppm is in operation in Puerto Rico (Szendrey, 1983, 1984). The plant operates at a loading rate of about 8 kg/m^3-d and a COD conversion efficiency of 75%. It has been found to be very stable in withstanding variations in loading rate and wastes composition and its ability to start-up after a shut down period (van den Berg et al., 1985).

Table 3.3. Performance data for downflow stationary fixed film reactors (van den Berg et al., 1985)

Type of waste	Waste strength (g/L)	Suspended solids (g/L)	Reactor size (L)	Reactor temperature (°C)	Loading rate (g/L/d)	Conversion (%)
Bean blanching waste	5.5-22 (TVS)	< 0.1	110	35	9.4	75
	10 (COD)	1-3	0.7	10	4.2	88
				35	18.4	88
Chemical industry waste	14 (COD)	0	0.7	25	14.0	81
Cheese plant waste	1-4	< 0.5	1.2	35	5-15	68-83
	1-3	< 0.5	$4x10^5$	30	5	60
Fish processing waste	6-20 (COD)	3-10	1.2	35	2.5-13	70-92
Liquor from heat treated sewage digester sludge	10.5 (COD)	< 1	0.8-1.2	35	29.2	70
	11 (COD)	1	950	35	0-25	60
Mansonite processing waste	9	< 0.1	1.2	35	9	72
Piggery waste	27-51 (COD)	16-33	35	35	6.1	70
					39.2	27
Pear peeling waste	110-140 (COD)	43-55	35	35	6.4	58
					18.9	54
Rum stillage waste	50-70 (COD)	4.5-6.5	35	35	13.3	57
	70-105 (COD)	-	$13x10^6$	37-40	8-10	65-70
Synthetic sugar waste	0.5	0	22.5	35	11.5	56
	1.0	0	22.5	35	9.0	60
	2.0 (COD)	0	22.5	35	4.6	79
Synthetic sewage sludge	55 (COD)	47	35	35	7.4	77
					13.8	71
Tomato peeling waste	11-22 (TVS)	3.8-7.6	110	35	4.5-12.3	50-61
Whey	66 (COD)	< 3	1.2	35	5-20	87-98

Table 3.4. Composition of wastes treated in downflow stationary fixed film reactors (van den Berg *et al.*, 1985)

Type of waste	Waste strength[a] (total COD) (g/L)	Suspended COD (% of total)	Sodium (g/L)	Kjeldahl nitrogen (g/L)	Total phosphate (g/L)	Ratios	
						COD/N	COD/P
Barley stillage waste	53	25	-	1.1	-	48	-
Bean blanching waste	10 (4-40)	10-30	-	0.4	0.1[d]	25	100
Chemical industry waste	14	0	-	2.5	0.3	5.6	47
Citric acid waste	3.6	-	-	0.04	0.025	90	140
Cheese plant waste	1-4	<15	0.15-0.58	0.05-0.2	0.01-0.06	20	67
Fish processing waste	6-20	50	-	0.8-2.5	0.03-0.11	7.5	180
Heat treated sewage digester sludge liquor (HTL)	10.5	<10	-	0.8	0.1	13	100
Pear peeling waste	130 (110-140)	35-50	0.4	2.3[c]	0.47[d]	55	275
Piggery waste	39 (27-51)	60-70	-	2.9	0.8	13	49
Rum stillage waste	60 (50-70)	<10	0.7	1.1[c]	0.21[d]	55	285
Skim milk waste	4	0	0.23[b]	0.2	0.04	20	100
Synthetic sugar waste	10 (5-15)	0	-	0.29	0.09	35	108
Synthetic sewage sludge	55	85	-	2.7	0.32	20	170
Tomato peeling waste	15 (15-30)	<20	2.2	0.28	0.2	54	75
Whey	66	< 5	0.33	0.65	0.65	100	100

[a] Average or most common concentration used; range used in brackets
[b] Sodium hydroxide added to increase alkalinity
[c] NH_4HCO_3 added
[d] Sodium and potassium phosphate added

3.5 Applicability

Stationary fixed film reactors could be changed over from one waste to another with relatively little loss of capacity and could adapt readily to changes in temperature as low as 10°C. This is important for installations where the character of the wastewater changes rapidly due to the season or production schedules (van den Berg, 1982). van den Berg et al. (1981) reported that the reactors could start-up very quickly after a period of starvation (one or two days to reach maximum capacity after 3 weeks of starvation).

This reactor could be used to remove the treated waste water intermittently than continuously and intermittent addition of wastes appears to be feasible (van den Berg, 1982). Intermittent loading increases the rate of methane production and hence the rate of conversion of COD, but decreases the COD removal. The latter may be caused by the short hydraulic retention time for part of the waste (van den Berg et al., 1981).

In stationary fixed film reactors COD removal depends on types of wastes and hydraulic retention times. Waste with hard to digest solids showed lower removals, particularly at short hydraulic retention time. Reactors could handle both low and high nitrogen wastes (pear peeling waste, piggery waste). Further, this fixed film reactor with effluent removal from the bottom has been found to produce methane with high suspended solids contents (van den Berg, 1982).

Due to the self mixing feature of the fixed film reactors, they could treat (this mixing is produced by the rising gas bubbles which causes every channel to act as gas lift pumps) dilute and concentrated wastes equally well. The rapid self mixing distribute wastes quickly through the reactor before local high concentration of volatile acids could develop.

The DSFF reactors could handle severe hydraulic overloading and organic shock loads without serious problems and could be operated at temperatures lower than optimum and still be loaded at high rates without affecting digester gas composition or COD removal efficiency.

Mesophilic DSFF reactors tolerated sudden organic shock loads at constant hydraulic loading (caused by sudden increase in waste strength) and recovered normal performance within a few days, if the alkalinity was sufficiently high to maintain the pH above 6.2 (Kennedy et al., 1984).

Kennedy & van den Berg (1981) reported that for chemical industry waste (TCOD = 14 g/L) the DSFF reactors could be over loaded 8 times their normal rate for a 24-hour period and, recovery to be possible within 12-48 hours while still being loaded normally. COD removal decreased with increasing overloading rates and was temperature dependent. During overloading at a loading rate of 61 kg COD/m^3-d (0.43 kg COD/m^2-d), COD removal efficiencies at 25°C and 35°C were 44% and 61% respectively. Results of overloading tests including wastes other than chemical wastes are given in Table 3.5.

It is evident that the COD removal during overloading decreased with increased rate of overloading while methane production rates increased. Further, Kennedy & van den Berg (1981) stated that repeated overloading improve the reactors as they are more stable and could be loaded at higher steady state rates than before overloading. This may be due to activation of inactive film by the availability of substrate and nutrients (Kennedy & van den Berg, 1981).

Table 3.5. Effect of 2-hour overloading on the performance of stationary fixed film reactor (Kennedy & van den Berg, 1981)

Type of waste	Type of experiment	Temperature (°C)	Loading rate (kg COD/m³-d)	Film loading rate (kg COD/m²-d)	Hydraulic retention time (days)	COD[b] removal (%)	Methane content of digester gas (%)	Volumetric rate of methane production (m³ (STP)/m³-d)
Chemical industry waste	Control	35	10.9	0.078	1.19	83	53	3.0
	Test	35	89.0	0.635	0.15	49	40	14.3
HTL[a]	Control[c]	35	18.2	0.130	0.57	66	73	3.9
	Test[d]	35	94.2	0.673	0.11	30	69	9.2
Chemical industry waste	Control	25	10.9	0.078	1.19	80	51	2.9
	Test	25	61.4	0.438	0.23	44	42	9.1
Bean blanching waste	Control	25	8.7	0.062	1.09	88	59	2.5
	Test	25	36.4	0.260	0.26	42	42	5.1

[a] Liquor from heat treated sewage digester sludge
[b] COD removal determined from methane production rates, 0.33 liter CH_4 (STP)/g COD removed
[c] Control-normal loading
[d] Test-overloading

It is reported that sloughing of the biofilm occurred during organic and hydraulic shock loadings (van den Berg et al., 1985).

Tests to determine the ability of DSFF reactors to handle toxic shock loads are underway. Initial results indicate that the DSFF reactor is able to withstand large toxic shock loads (van den Berg et al., 1985).

3.6 Problems Associated with Downflow Stationary Fixed Film Reactor

3.6.1 Plugging of the Reactor

Non-uniformity of biofilm thickness in DSFF reactor occurs due to excessive growth of biofilm near the top of the reactor. Under certain conditions, this non-uniform growth can cause plugging at the top of the packing (Hall, 1983) and partial plugging of some channels (Samson et al., 1985).

Factors like, width of the channels, smoothness of the packing (will determine the smoothness of occasional sloughing), recirculation rate and the composition of the waste determine whether or not plugging will occur (van den Berg et al., 1985).

van den Berg et al. (1985) reported on several methods which have potential for maintaining a reasonably thin biofilm in DSFF reactors. These are:

- Organic and hydraulic shock loads - the effect on film sloughing will be greatest near the top of the reactor where the load enters. This method will not be of much use for channels already blocked.

- Recirculation of effluent - as already discussed, for wastes with a high suspended solids content, intermittent pumping of liquid from the bottom to the top of the reactor helps to maintain a uniform, relatively thin film.

- Recirculation of gas - large gas bubbles rising in channels should help the sloughing process and may even open blocked channels.

- Reactor configuration - a relatively thin layer of coarse packing on top of the ordinary packing may accumulate the excess biofilm and cope with it.

- Improvement in the flow distribution system on top of the reactor to avoid too low liquid velocity in these channels.

- Horizontal spacing of channels to improve mixing and reduce dead space.

3.6.2 Start-up of the Reactor

Rate of start-up depends on the type of inoculum, the type and strength of waste, level of volatile acids maintained and the characteristics of the support material used. For example reactors were difficult to start-up with chemical industry waste (toxicity could not be demonstrated), while reactors started readily on food processing wastes or sugar waste. Sewage digester sludge generally required a longer time to adapt than inoculum from an active digester fed with food processing waste. The rate of start-up was faster with sugar waste at 5,000 mg COD/L than with a strength of 10,000 mg COD/L or higher. Also, reactors started up faster when volatile acid levels were maintained at about 1,000 mg/L than with levels below 600 mg/L (Kennedy & Droste, 1984). Several factors presumably play a role: concentration of critical types of bacteria, ecological relationships and how close the waste resembled the substrate to which the inoculum was accustomed.

Support material as well as the number of channels in a reactor affect the start-up and ultimate loading rates (van den Berg et al., 1985). This effect was caused by differences in mixing patterns because it affects the amount of dead space and short circuiting. Recirculation rate also improves mixing patterns and rate of start-up (Samson et al., 1985). Horizontal spaces in banks of vertical channels also provide an improvement in the rate of start-up by reducing the amount of dead space (Samson et al., 1985).

With DSFF reactors, the rate of increase in COD loading rate was less consistent. According to the experiments carried out by van den Berg et al. (1980), it was reported that when simulated sewage sludge is fed, doubling of the COD loading rate took generally 30 days or longer while with bean blanching waste the time required to reach a doubling of the COD loading rate was as low as 10-15 days for 0.8 liter reactors and as high as 40-50 days for 35 liter reactors.

In addition, the COD loading rate had got 'stuck' for several weeks at a loading rate well below the ultimate maximum loading rate (Fig. 3.5). This reflected the variable and often less optimum degree of mixing in fixed film reactor, during the start-up. It may be possible that the types of organisms occupying fixed film surfaces during the start-up are not present in a relationship which is quantitatively and qualitatively optimal for high rate performance and readjustment during the start-up may therefore be necessary from time to time, causing a delay in reaching the maximum COD loading rate (van den Berg et al., 1980).

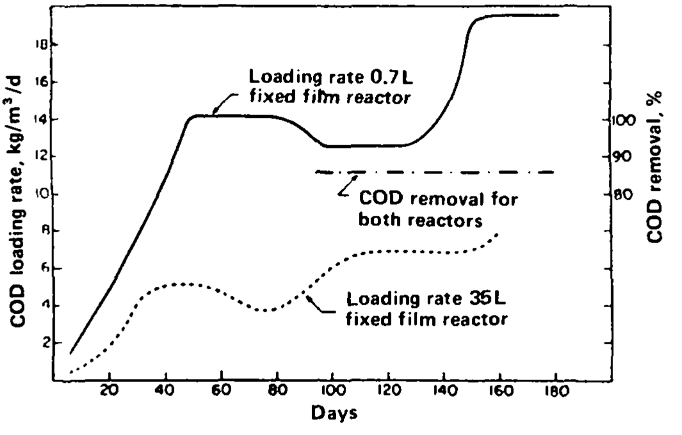

Fig. 3.5 Effect of fixed film reactor size on changes in COD loading rate and COD removal efficiency during start up of reactors fed been blanching waste. Note occurrence of pseudo-steady states in both reactor (van den Berg et al., 1980)

3.7 Advantages and Disadvantages

Advantages

- Elimination of mechanical mixing (mixing in the reactor is provided entirely by the action of rising bubbles)
- Recycling is not necessary
- Better stability at higher loading rates
- Simplicity in construction

Disadvantages

- Lower quality in circumstances where the influent suspended solids concentration is high.
- Requires more care in starting-up of the reactor.

3.8 Conclusion

The DSFF reactors appear to be outstanding in reliability and ability to withstand adverse conditions, but generally do not achieve very high loading rates or COD removal efficiencies. The DSFF reactor is at an infant stage of development and further experiences at pilot plant and full scale level are needed.

IV. ANAEROBIC EXPANDED/FLUIDIZED BED TREATMENT

4.1 General

The anaerobic fixed bed process was found to be hampered by clogging and inefficient contact of the micro-organisms and the wastewater due to influent suspended solids or excess biomass which creates inactive zones and channeling. Unlike fixed bed reactors, bed fluidization results in little or no short-circuiting and small pressure gradients. Jewell (1974) proposed the attached film expanded bed process as a means of optimizing aerobic systems. This was based on the assumption that large biomass concentrations could be achieved on the large surface area provided by the small sand size particles. The small particles, when fluidized, would minimize diffusional limitations and eliminate clogging problems. Later Jewell & Switzenbaum (1980) demonstrated that it was possible to utilize this concept, using an anaerobic film.

4.2 Anaerobic Expanded/Fluidized Bed Reactor

Expanded/fluidized bed reactors have much larger surface area per unit reactor volume, which increases the reactor micro-organism concentration. The larger specific surface area allows shorter hydraulic retention times or lower operating temperatures for the same degree of treatment in a given volume. A specific surface area of 3000 m^2/m^3 has been reported for fluidized bed reactor and the concentration of the microorganism of 30 g/L have been measured in anaerobic processes (Boening & Larson, 1982).

For anaerobic treatment of wastewater, only a few studies using expanded/fluidized beds have been reported. Switzenbaum (1978) used anaerobic attached film expanded process (AAFEB) and found it to be effective for the treatment of low strength soluble organic wastes at reduced temperatures, at short retention times and at high organic loading rates. Walker (1981) tested pilot plants containing large scale fluidized bed bioreactors as advanced treatment processes for denitrification of industrial wastewaters and he demonstrated long-term stable operation of the unit, as well as the ability to meet stringent discharge limits. Hickey and Owens (1981) used a similar process and have shown this to be effective for the simultaneous generation of methane gas and stabilization of high strength wastewaters including dairy, chemical, food processing, soft drink bottling and heat transfer liquids.

Packed bed reactors such as anaerobic filters often experience problems or increased pressure drop due to the accumulation of biomass. To prevent plugging, relatively large voidage must be maintained which limits specific surface area and biomass concentration. Expanded/fluidized bed reactors overcome these problems allowing the use of low voidage, high surface area of the media.

4.2.1 Principles and Theory of Fluidization

In an expanded bed, the particles remain in stationary contact whilst in a fluidized bed the particles are in free motion. When the liquid is passed upward through an unrestrained bed of particles, the bed will initially expand slightly to take up a loose packed arrangement. If the flow is increased, the pressure drop across the bed increases as shown by the line OA in Fig. 4.1. Eventually the pressure drop equals the force of gravity (corrected for the buoyancy in the liquid) on the particles and the grains begin to move. This is the point 'A' in the figure. During this period, the porosity increases and the pressure drop rises more slowly than before due to the net effect of increased porosity and velocity. When point 'B' is reached, the bed is in extremely loose condition with the grains still in contact. Between points 'A' and 'B' the bed is unstable, the particles begin to loose contact and then adjust their position to present as little resistance to the flow as possible.

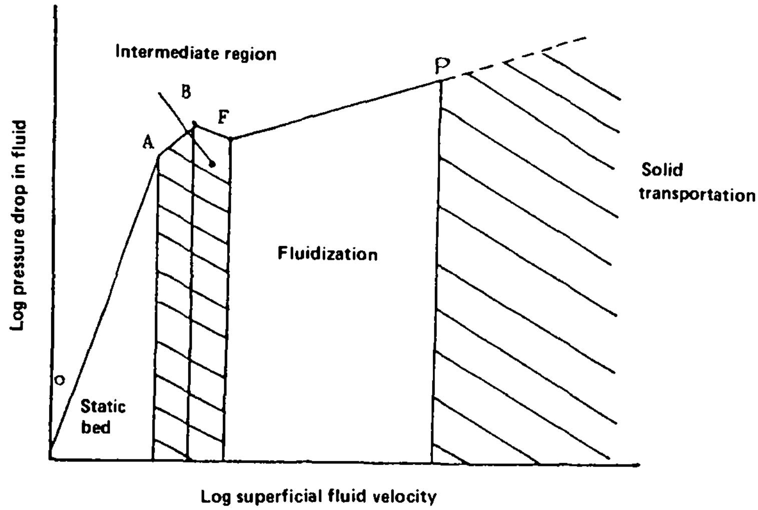

Fig. 4.1 Pressure drops in fludized bed (Cooper & Atkinson, 1981)

As the velocity is further increased, the grains separate and true fluidization begins. This is the point 'F' on the figure. By the time this point is reached, all particles are in motion and beyond that point, the bed continues to expand and the particles move in a more rapid and independent motion. The bed continues to expand as the velocity is increased and maintains a uniform character. As the fluid velocity is increased further, the porosity increases, the bed of solids expands and eventually at point 'P' on the figure, all of the particles have been entrained in the fluid. The porosity approaches one and the bed ceases to exist. From this point on, there exists the simultaneous flow of two phases. At point 'P', the superficial velocity is approximately equal to the terminal settling velocity of the particles.

The systems, as applied to wastewater treatment, consist of inert sand-sized particles in a column which expand with the upward flow of waste through the column. The inert particles act as a support surface for the growth of attached organisms. Fluidization allows the entire surface area of each particle to become available for biological attachment and subsequent reaction.

In anaerobic systems, both the expanded and fluidized beds operate at less than full fluidization (Switzenbaum, 1983).

The biomass hold up is achieved by allowing natural mechanism of flocculation and adhesion to take place in a low shear environment and then to remove excess growth by particle/particle or particle/wall contact.

4.2.2 Design Elements

Like the anaerobic filter, the design of an expanded/fluidized bed consists of a wastewater distributor, a medium support structure, medium, head space, effluent draw off and recycle facilities. For high strength wastes, a device for separating the excess biomass from the support medium and subsequent wasting of this excess growth is generally incorporated into the design.

4.2.3 Recycling

Since the medium should be kept in the fluidized state, upflow velocities must be high enough to keep the particles in suspension, and thus effluent recycle is practised.

In addition to this, a certain amount of recycle is useful as it can:

- help neutralize the pH of the incoming wastewater;
- reduce the amount of alkalinity required;
- reduce the effect of toxic biodegradable compounds;
- minimize the effect of shock loadings; and
- compensate for variability of influent flow rate.

Increasing recycle in effect allows the process to tend towards the results and operational characteristics of a completely mixed system.

4.2.4 Separation Equipment

In fluidized beds, some sort of separation equipment or other biomass retention measure is incorporated in the design as process failure could result in total loss of biomass within 15 minutes. Expanded bed reactors thus have some inherent risk.

4.3 Design Criteria

4.3.1 General

The expanded/fluidized bed reactor can be designed using either the organic volumetric loading rate (OVLR) or solid retention time (SRT) approach. The kinetics of substrate removal in the fludized bed reactor will determine the actual SRT required for a given degree of treatment efficiency. Once the SRT is established together with values for the kinetic parameters, the kinetic equations can be used together with the information from fluidization mechanics to establish values for other design parameters (hydraulic retention time, fluidization velocity, recycle ratio, etc.).

A less rigorous and strictly empirical approach will be to use the organic volumetric loading rate (OVLR) to achieve a given degree of treatment.

4.3.2 Solid Retention Time

The solid retention time (SRT) is the average retention time of organisms in the system. In the expanded/fluidized bed process the SRT is normally defined as:

$$SRT = \frac{\text{volatile suspended solids (VSS) in the reactor}}{\text{volatile solids lost in the effluent or intentionally wasted/day}}$$

In a biological reactor the organism specific growth rate is equal to the reciprocal of the solid retention times of the system.

The organism specific growth rate is expressed according to the kinetic equation given:

$$\mu = \frac{1}{x}\frac{dx}{dt} = YK - b$$

where;

μ = organisms specific growth rate, m^3/m^3-h
x = organisms concentration, kg.VSS/m^3
Y = organisms specific yeild coefficient
K = specific substrate utilization rate, kg.COD/kg.VSS-h
b = organism decay coefficient, h^{-1}

thus

$$\frac{1}{SRT} = YK - b$$

4.3.3 Organic Volumetric Loading Rate (OVLR)

Rudd et al. (1985) found that the biomass concentration was affected by variation in the organic loading, influent substrate concentration, and hydraulic retention time. Schrra and Jewell (1984) observed that the COD loading appears to be the major determinant of biomass concentration.

Numerous authors (Stephenson & Murply, 1980; Sutton et al., 1981; Jewell, 1981) have used this parameter to illustrate the efficiency of fluidized bed reactors in comparison to other systems in treating wastewaters. The OVLR is used when it is difficult to determine the reactor biomass concentration.

The organic volumetric loading rate (OVLR) to the system is defined as:

$$OVLR = \frac{QS_o}{V} = \frac{S_o}{T}$$

where

Q = influent flowrate, m^3/d
S_o = influent substrate concentration, kg/m^3
V = reactor volume, m^3
T = reactor hydraulic retention time, d

Numerous pilot scale tests have shown high COD removal of more than or equal to 80% at COD loading of 10-20 kg/m^3-d for variety of industrial wastes.

4.3.4 Hydraulic Retention Time (HRT)

Hydraulic retention time (HRT) is calculated on the basis of expanded/fluidized bed volume. The HRT is the ratio between the expanded/fluidized bed volume and the influent flow rate of the wastewater.

In the case of wastewater treatment, due to the relative insensitivity of the process performance to HRT, the system should be designed at a low HRT (of the order of several hours). The actual design HRT depend on wastewater organic strength (Switzenbaum et al., 1984).

Laboratory and pilot-scale experiments have been conducted with HRT values varying from a low of five minutes to several days for various industrial wastes. COD loadings have ranged from 0.65 to 60 kg/m^3-d (Switzenbaum, 1983).

Rudd et al. (1985) and Schraa & Jewel (1984) reported that the organic removal efficiency decreased with decreasing HRT at a constant loading rate. Optimum HRT for mesophilic (35°C) anaerobic fluidized bed reactors lie within the range of 6-13 hours (Rudd et al., 1985).

4.3.5 Recycle Ratio

Expanded and fluidized beds operate with very high recycle rates. For concentrated wastes, a very high degree of recycle is needed in order to keep the bed particles in suspension and at the same time to dilute the organic materials present in high concentration. For dilute wastes, like municipal wastewater, the recycle ratio is reduced to reasonably low values.

The recycle ratio (α) is given by the ratio between the influent flow rate (Q) and the recycle flow rate (Q_r).

Thus, recycle ratio (α) $= \dfrac{Q}{Q_r}$

Based on the design values of substrate removal rate, v_x = 0.01 kg.COD/kg.VSS.h; biomass concentration, X = 20 kg.VSS/m^3; hydraulic loading rate, Q = 10 m^3/m^2-h; COD removal efficiency, E = 80%; and total particle density, ρ = 1.02 x 10^3 kg/m^3; the following figure (Fig. 4.2) is developed to show the effect of influent wastewater concentration, C_o upon the degree of recycle required in expanded and fluidized bed reactors.

However, the assumed upflow superficial velocity of 10 m^3/m^2-h is an estimation based on an overall bed particle specific density identical for pure biological flocs. For very thin biofilms on high density media, the particle size will be of great importance in considerations regarding the needed superficial upflow velocity (Jewel, 1982).

4.3.6 Filter Media

The medium used for biofilm attachment is comprised of small diameter inert particles, such as sand, anthracite, or granular activated carbon, which are maintained in the fluidized state.

Various support media have been tested, including sand, PVC particles, granular activated carbon, and diatomaceous earth. A range of particle sizes and densities have been examined. There are trade-offs between size and density of particles and stability of operation with these systems. Smaller particles provide greater specific surface area-to-volume ratio and thus provide greater surfaces for attached biofilms. In addition, lighter particles can be fluidized at lower upflow velocities which reduce the recycle rate necessary to achieve a given HRT.

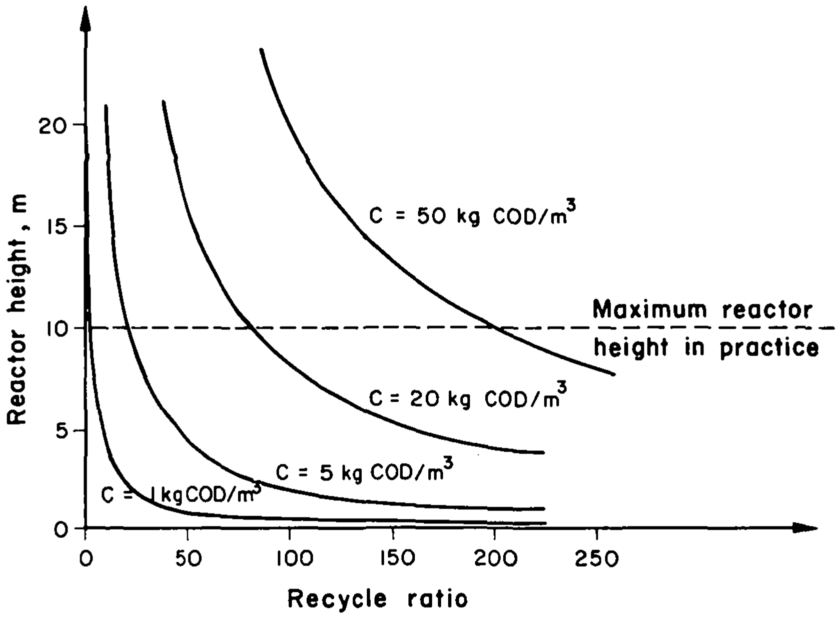

Fig. 4.2 Recycle rates for expanded/fluidized beds

In the design of an anaerobic expanded/fluidized bed, small and light particles which are easy to fluidize and provide a large specific surface are being used. There would exist however, a minimal particle size and/or density in order to prevent carry-over.

If a high shear stage is to be used to disengage the biomass and the medium then it is likely that a tough medium, such as sand, would be chosen in preference to a more fragile one like activated carbon (Cooper & Wheeldon, 1980).

The technical data based on available literature are given in Table 4.1.

Table 4.1. Technical data (Henz & Harremoes, 1982)

	Expanded bed	Fluidized bed
Reactor media		
Inert material type	sand/gravel/plastic	garnet/sand/carbon
Inert material diameter (mm)	0.3-3	0.2-1
Inert material submergence (%)	100	100
Bed expansion (%)	20-40	30-100
Specific surface area, m^2/m^3	1000-3000	1000-2500
Depth of reactor (m)	2-4	4-8
Radius of reactor (m)	2-3	2-3
Vertical velocity, empty bed (including recycle) (m^3/h)	2-10	6-20
Recycle ratio	2-100	5-500

4.3.7 Temperature

Temperature was found to be an important variable affecting process efficiency, but the process was found to compensate well for changes in temperature. However, the optimum temperature for the treatment is around 35°C.

Schraa & Jewell (1984), reported high rate conversions of soluble organics with anaerobic fluidized bed reactors at thermophilic (55°C) temperatures, after a 5 month accumulation of biomass. A mature microbial attached film was developed successfully at 55°C over a short time period and high solids concentrations and film depths (60 g/L and 170 cm respectively) were achieved with these thermophilic films. Medium strength (1.5 to 3 g/L COD) and high strength (5-16 g/L COD) soluble wastes were treated with a 70% removal efficiency at a volumetric loading rate of 30 g/L.d COD.

Results obtained by Rudd et al. (1985) indicate that thermophilic anaerobic fluidized bed reactors are inferior to mesophilic reactors in several ways. The most important was the inferior organic removal efficiencies achieved by thermophilic anaerobic fluidized bed reactors under a number of operating conditions.

4.3.8 Process Design

Since anaerobic fluidized bed treatment technology is relatively a new approach, process design criteria is not available at present time.

Pilot plant studies were completed by Dorr-Oliver T.M. of U.S.A. for industrial wastewater treatment as reported by Sutton and Li (1982) in order to derive information for process design of single phase and two phase anaerobic fluidized bed systems. Sutton & Li (1982) presented the range of operating conditions and the range of process design parameter values for single- and two- phased fluidized bed systems (Tables 4.2, 4.3 and 4.4).

The design values presented by Sutton and Li (1982) for single phase and two phase fluidized bed reactors are summarized in Table 4.5.

Table 4.2. Operating conditions for fluidized bed reactor of single phase system (Sutton & Li, 1982)

Characteristic	Reactor value
Mean sand size, mm	0.5
Hydraulic loading rate, m^3/m^2-h	25-33
Controlled bed expansion, %*	90-110
pH range	6.7-7.2
Temperature, °C	30-35

* Percent expansion of settled sand bed.

Table 4.3. Operating conditions for fluidized bed reactors of two phase system (Sutton & Li, 1982)

Characteristic	First stage reactor value	Second stage reactor value
Mean sand size, mm	0.5	0.5
Hydraulic loading rate, m^3/m^2-h	25-33	25-33
Controlled bed expansion, %*	40-60	90-110
pH range	5.7-6.2	6.7-7.2
Temperature, °C	30-35	30-35

* Percent expansion of settled sand bed.

Table 4.4. Design parameter values for single phase and two-phase fluidized bed reactors operated at 30°C to 35°C (Sutton & Li, 1982)

	Reactor biomass concentration g VSS/L	Reactor organic loading rate kg COD/kg VSS-day	Reactor volumetric loading rate kg COD/m^3-d
Single phase			
Organics to methane	15-25	0.4-1.0	10-20
Two phase			
Organics to acetic acid	15-25	2.0-4.0	30-40
Acetic acid to methane	8-15	2.0-4.0	25-35

4.4 Different Design Configurations

4.4.1 Dorr-Oliver's Anitron System

Fig. 4.3 shows a pilot plant that was used by Dorr-Oliver and reported by Sutton & Li (1982). The units were skid-mounted, self contained units. The fluidized bed reactor consisted of a clear PVC column having a diameter of 16.2 cm. The height of the reactor to the effluent out-let port was approximately 3 m. The fluidized bed height was controlled below 2.44 m. Additional pilot plant components included a refrigerated feed tank, feed and recycle pumps, a means for foam control, gas liquid separator, a wet-test meter for gas flow measurement, and various other instrumentation. A refrigerated holding tank was installed between the two pilot plants to help balance the feed flow rate during two phased operation. Temperature control of each reactor was achieved by adjusting the temperature of the recycle stream using an electrically controlled heat exchanger. The pH of each reactor was controlled by the addition of either sodium bicarbonate or sodium hydroxide. Ammonium dibasic phosphates were added to ensure a proper nutrient balance for biological growth.

Table 4.5. Single- and two-phase fluidized bed design values (Sutton & Li, 1982)

Design characteristic	Single phase reactor design value	Two phases design values	
		First stage	Second stage
Volumetric loading rate, kg COD/m^3-d*	15	40	30
Controlled fluidized bed height, m	10.5	10.5	10.5
Hydraulic loading rate, m/h	24	24	17
Reactor area, m^2	127	48	51
Fluidized bed volume, m^3	1333	504	536

* A 20% COD reduction is assumed in the first stage reactor of the two-phase system.

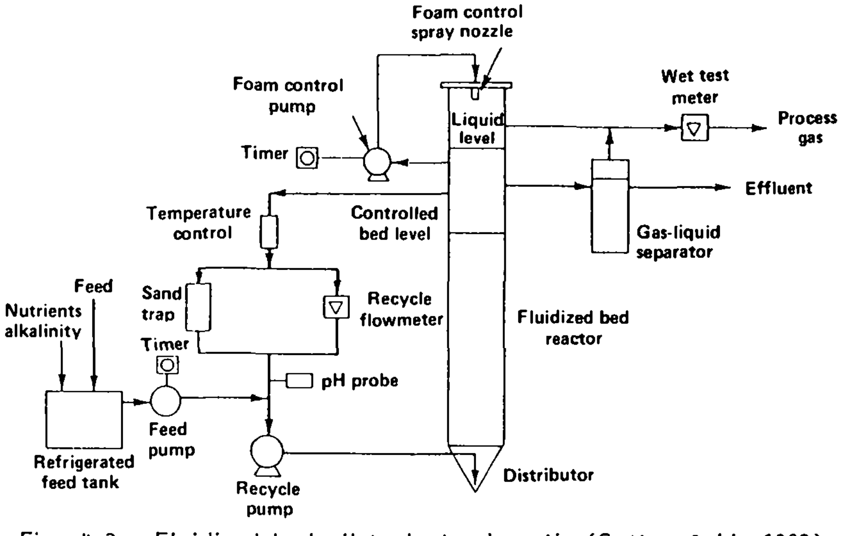

Fig. 4.3 Fluidized bed pilot plant schematic (Sutton & Li, 1982)

4.4.2 Tapered fluidized bed

In tapered configuration (Fig 4.4), the cross sectional area gradually increases from bottom to the top of the reactor. This configuration provides the flow patterns a minimal backmixing, especially at the feed entry point and prevents plugging by maintaining a high inlet velocity. Relatively stable flow through the reactor can be achieved when the entry cross section is sufficiently small and the expansion is gradual (an angle of few degrees). Since the fluid velocity decreases with the reactor, the height of the column allows a wide range of flow rates without the loss of bed material. Hence, the tapered fluidized bed can effectively operate over a wide range of feed flow rates. This system can maintain a stable operation over a wide range of volumetric flow rates than the conventional fluidized bed.

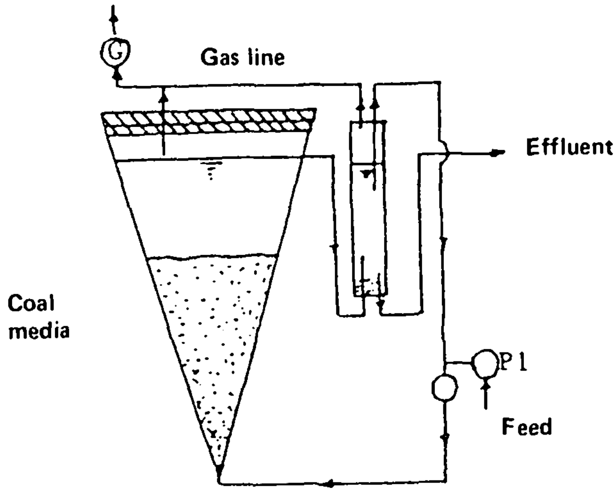

Fig. 4.4 Schematic of tapered coal bed reactor (P1) for pump, (P2) recycle pump, (ST) solids traps, (G) gas matter (Boening & Larsen, 1982).

4.5 Application Status

4.5.1 Municipal Applications

The initial research on the anaerobic expanded/fluidized bed was conducted on primary effluent from the Ithaca (New York) Sewage Treatment Plant, using a process called anaerobic attached film expanded bed process. The work reported by Jewell et al. (1981) found the process to be capable of achieving COD removals of 60-80% at HRT of 0.5-8 h at 20°C. Effluent quality was 40-50 mg/L COD and 5-15 mg/L SS at an HRT of 1 to 5 h.

A later laboratory study conducted by Froster (1984), for settled sewage obtained daily from a sewerline running along the perimeter of the University of Birmingham campus, U.K., showed an unsatisfactory performance of this process under unsteady conditions. The efficiencies were significantly lower than those quoted previously.

A study on anaerobic fluidized beds was conducted at U.S. EPA's Test and Evaluation Facility in Cincinnati, Ohio, investigating the treatment of a municipal wastewater (containing industrial wastes) with an average sulfate concentration of 250 mg/L (Bowker, 1983). COD removal efficiencies ranged from 40 to 50% for hydraulic retention times of 6 to 24 hours. Methane production was found to be low due to the effect of the thermodynamically favored reduction of the influent sulfate that was occurring in the reactors.

A pilot-scale testing of the anaerobic fluidized bed process was conducted by Switzenbaum et al. (1984) for the primary effluent from the Amherst wastewater treatment plant at the University of Massachusetts/Amherst, U.S.A. Sand was used as the support material. This study was conducted over a five-month intensive testing period and over a wide range of organic volumetric loading rates, (0.18-0.41 g BOD/L-d) and at a hydraulic retention time varying from 1.67 to 6.37 hrs; a mean effluent 5 day BOD of 47.2 mg/L (standard deviation = 15.5 mg/L) and a mean suspended solids (SS) concentration of 30.5 mg/L (standard deviation = 16.6 mg/L) were achieved. The influent BOD_5 and SS were 74.2 mg/L and 35.5 mg/L respectively. Relatively low percentages of removal were obtained. A comparison of pilot-scale data (Switzenbaum et al., 1984) with the laboratory scale data (Jewell et al., 1981) is presented in Table 4.6.

Table 4.6. Comparison of laboratory and pilot scale units

	Influent (mg/L)		Effluent (mg/L)	
	Mean	Range	Mean	Range
Pilot scale*				
TCOD (Total COD)	171.4	(70-1106)	101.7	(43-225)
SCOD (Soluble COD)	130.4	(55-225)	76.7	(11-178)
SS	35.5	(7-116)	30.5	(7-144)
BOD_5	74.2	(28-199)	47.2	(13-74)
Lab scale**				
TCOD (Total COD)	186	(88-306)	49.2	(22-126)
SS	86	(40-186)	16.5	(3-90)

* Data of Switzenbaum et al. (1984).

** Data of Jewell et al. (1981).

Culp/Wesner/Culp consultants (1980) concluded that the fluidized bed SMAR appeared to have its greatest potential as a secondary treatment unit process due to its ability to treat low-strength soluble wastes and produce high quality effluents. They found the fluidized bed SMAR comparable to conventional aerobic systems in the treatment of dilute organic wastes in terms of organic removal efficiencies, detention times and organic loading rates. They also developed process flow sheets and cost details for treatment schemes incorporating fluidized bed SMAR for wastewater flows of 1 mgd and 25 mgd, and found that these costs compared favorably with costs for conventional aerobic treatment processes.

So far no full scale application of the anaerobic expanded/fluidized bed process is being reported on sanitary wastewater.

4.5.2 Industrial Applications

For the treatment of industrial wastewater, Ecolotrol Inc., of U.S.A. has operated a number of pilot scale anaerobic fluidized bed reactors in order to establish the design parameters (Jeris, 1982).

Results of a pilot study conducted by Ecolotrol Inc., U.S.A. (Jeris 1982) for a food processing waste is given in Table 4.7. Temperatures were maintained at 35-37°C. For the four organic loading rates used, the BOD_5 and COD removals varied between 86 to 93 and 75 to 86 percent respectively showing a relatively low spread of removal efficiency over the organic load range studied. Further evidence of the stability of this operation are the volatile acids which rarely exceeded 200 mg/L as acetic acid. Also, the suspended solids removal averaged close to 50% with an influent concentration of 1,140 mg/L and an effluent of 550 mg/L.

Table 4.7. Anaerobic fluidized bed reactor treatment - food processing waste(Jeris, 1982)

Run	Organic load kg COD/m³-d	HRT (h)	COD mg/L		% COD rem.	BOD_5 mg/L		% BOD rem.
			Inf.	Eff.		Inf.	Eff.	
1	3.5	49.4	7210	1040	86	4370	320	93
2	8.3	21.4	7390	1430	81	4700	440	91
3	16.8	13.5	9450	1910	80	5900	630	89
4	24.1	7.5	7530	1900	75	4775	690	87

The percentage COD reduction for five waste categories at different organic loading are summarized in Fig 4.5 (Jeris, 1982). It is evident that the removal efficiency is slightly affected upto 16 kg COD/m³-day for most wastes and that the efficiency remained high for some of these wastes to loadings in excess of 36 kg COD/m³-day.

Numerous pilot-scale tests have shown high removal percentages (80%) at loading rates ranging from 10 to 20 kg/m³-day for a variety of industrial wastes. The results are summarized in Table 4.8.

The COD removal efficiency of whey waste at different loading rates and at two temperatures were studied (24°C and 35°C). About 10% reduction in removal efficiency was caused by the temperature difference (Fig. 4.6).

Greater efficiencies were reported for two stage application of fluidized bed reactors in series and the results for different types of waste are given in Table 4.8. In single stage application of influent COD in excess of 50,000 mg/L and for 80% removal, an effluent COD of 10,000 mg/L was reported, where as in the two stage application an influent COD of 52,200 mg/L was reduced to 3,200 mg/L to give an overall COD removal of 94%. A five day retention period was required.

Only two full scale applications of the anaerobic expanded/fluidized bed process for treating industrial wastes have been reported (Switzenbaum, 1983) (Table 4.9).

Extensive literature on denifrification in anaerobic fluidized and expanded beds are available but as yet there is little full scale operating experience. All the work that has been reported on anaerobic denitrifying fluidized bed systems are summarized by Cooper & Wheeldon (1980) and is presented in Table 4.10.

4.6 Applicability

Shock loadings (in terms of temperature and loading strength) had relatively little influence on the process (Jewell et al., 1981). Expanded bed reactor was found to be efficient for the treatment of particulate wastes (Morris & Jewell, 1982)

Table 4.8. Expanded/fludized bed studies for industrial applications

Waste	Influent	Effluent	Removal (%)	Loading rate (kg/m³-d)	HRT	Temp. (°C)	References	Remarks
Acid whey	50.3-56.1 g/L COD	8.3-14.6 g/L COD	72.0-83.6	13.4-37.6	1.4-4.9[a] days	35	Hickey & Owens (1981)	One stage fluidized bed
Acid whey	52.2-55.4 g/L COD	15.1-19.2 g/L COD	65.2-71.0	15.0-36.8	1.5-3.5[a] days	24	Hickey & Owens (1981)	One stage fluidized bed
Acid whey	52.2 g/L COD	3.2 g/L COD	94	10.5	5.0 days[a]		Hickey & Owens (1981)	Two stages in series (Fluidized beds)
Food processing waste	7.2-9.4 g/L COD	1.0-1.9 g/L COD	75-86	3.5-24.1	7.5-49.4[a] hrs	35	Hickey & Owens (1981)	Fluidized bed
Chemical waste	12 g/L COD 8.4 g/L BOD_5[b]	-	79-93 (COD) 81-98 (BOD_5)	4.1-27.3	-	35	Hickey & Owens (1981)	Fluidized bed
Soft drink bottling waste	6 g/L COD[b] 3.9 g/L BOD_5[b]	-	66-89 (COD) 61-95 (BOD_5)	4-18.5	-	35	Hickey & Owens (1981)	Fluidized bed
Heat treatment liquor	10 g/L COD[b] 5 g/L BOD_5	-	52-75 (COD) 66-95.5 (BOD_5)	4.3-21.4	-	35	Hickey & Owens (1981)	Fluidized bed
Whey premeate	6.8 g/L COD	2.2 g/L COD 1.3 g/L BOD_5	68 (COD)	8.6-10.4	-	30-35	Sutton & Li (1982)	Fluidized bed
Whey premeate	27.3 g/L COD	5 g/L COD 3/g/d BOD_5	82 (COD)	5.3-7.4	-	30-35	Sutton & Li (1982)	Fluidized bed
Sweet whey	10 g/L COD	1374-6071 mg/L SCOD	36.9-93.1	8.9-60.0	4.1-27.1 hrs[c]	25	Switzenbaum & Danskin (1982)	Expanded bed
	5-20 g/L SCOD	417-7958	58.9-92.3	8.2-29.1	15-35 hrs[c]	35	Switzenbaum & Danskin (1982)	Expanded bed
Black liquor condensate	1.4 kg/m³ COD	0.16 kg/m³ COD	89	10.0	3.5 hrs	22	Norrman (1982)	Expanded bed
Black liquor	1.4 kg/m³ COD	0.28 kg/m³	80	13.0	0.8 hrs	22	Norman (1982)	Fluidized bed

[a] empty bed retention time
[b] average values
[c] based on expanded bed volume

Table 4.9. Full-scale anaerobic fluidized bed installation (Switzenbaum, 1983)

Location	Type	COD (mg/L)	Flow (m^3/d)	HRT (h)	COD load (kg/m^3-d)	Media	COD removal (%)
Birmingham AL	Soft-drink bottling waste	6900	380	6	9.5	0.6 mm E.S. sand	77
Midwest	Soy process waste	9000	-	less than 24	13	0.4 mm E.S. sand	-

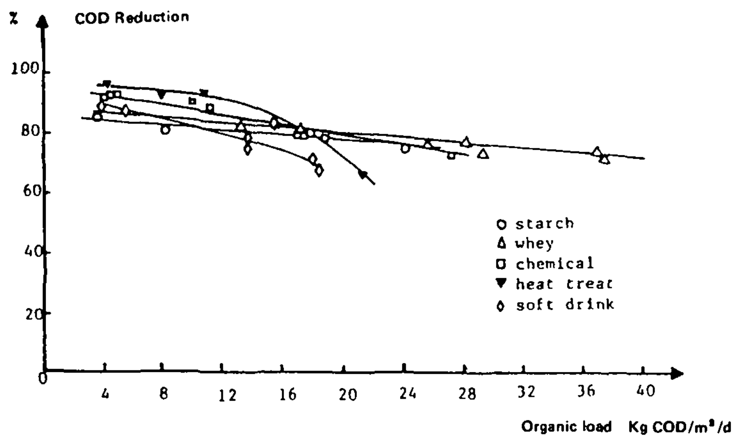

Fig. 4.5 Anaerobic fluidized bed reactor treatment of high strength waste (Jeris, 1982)

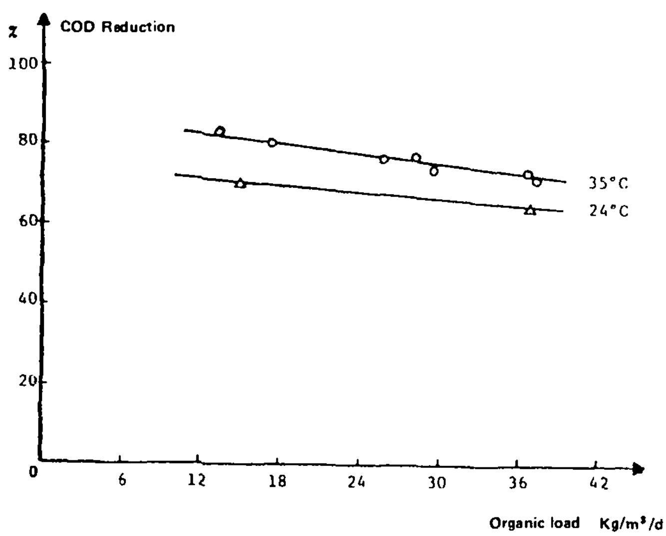

Fig. 4.6 Acid whey - % COD removal vs. organic loading (Jeris, 1982)

Table 4.10. Summary of performance data for anoxic fluidized and expanded bed system (Cooper & Wheeldon, 1980)

Medium	Particle size (mm)	Upflow velocity (m/h)	Influent type	Oxidized nitrogen conc. in influent (mg N/L)	Superficial retention period (min.)	Temp. (°C)	Biomass concentration[a] (g/L)	Volumetric rate of denitrification[b] (kg N/m³-d)	Mass rate of denitrification (mg NO_3-N/ g BTS/h)	Carbon source	Comments
Activated carbon	0.65	29.3	$NaNO_3$ solution	NR	6.4	24.0	NR	6.8	NR	Methanol	Fluidized bed
"	"	28.8	"	25-40	7.0	22.0	"	4.7	"	"	" "
"	"	28.8	"	NR	7.1	20.5	"	4.9	"	"	" "
"	"	30.3	"	NR	5.4	16.2	"	4.1	"	"	" "
"	"	57.3	"	34-50	3.5	22.0	"	6.2	"	"	" "
"	"	19.5	"	NR	10.5	26.0	"	4.7	"	"	" "
Sand	0.6	37.2	Secondary effluent	21.5	6.5	20.0	30-40	5.5	NR	"	" "
"	"	58.8	"	18.7	4.0	19.0	"	7.2	"	"	" "
"	"	36.8	Secondary effluent + $NaNO_3$	55.0	6.5	25.0	"	13.5	"	"	" "
"	1.0	4.6	Secondary effluent	36	20	12-14	NR	2.7	NR	"	Expanded bed lab.-scale
"	1.0-2.0	11	Secondary effluent	NR	9.5	16-17	NR	4.0	NR	"	Expanded bed full-scale
"	"	15	"	19-32	7.0	12	NR	4.0	"	"	" "
"	0.25	25	River water	10-15	8.3	6-16	15-20	2.2-3.0	NR	"	" "
"	"	"	"	11-17	8.3	2-7	16-21	1.5-2.9	"	"	" "
"	"	"	"	15-21.5	8.3	7-14.5	NR	3.4-4.2	"	"	
"	0.2-0.5	20	River water	8	NR	10	10-14	3.8	13	"	" "
"	"	20	"	"	NR	20	10-14	7.4	26	"	" "

Table 4.10. Summary of performance data for anoxic fluidized and expanded bed system (Cooper & Wheeldon, 1980) (cont'd)

Medium	Particle size (mm)	Upflow velocity (m/h)	Influent type	Oxidized nitrogen conc. in influent (mg N/L)	Superficial retention period (min.)	Temp. (°C)	Biomass concentration[a] (g/L)	Volumetric rate of denitrification[b] (kg N/m³-d)	Mass rate of denitrification (mg NO_3-N/ g BTS/h)	Carbon source	Comments
Anthracite	2-3	0.1-8.0	Industrial effluent	226-2260	120	22-34	NR	3.9-27.5	NR	"	Tapered fluidized bed
Sand	1.0	78.6	Industrial effluent	1450	NR	20	NR	10.6	NR	Molasses	Fluidized bed 10:1 recycle
"	"	"	"	1450	NR	38	NR	38.4	NR	"	Fluidized bed 10:1 recycle maximum rate
Activated sludge	1-2	0.39	Industrial effluent	7550	1143	NR	25	6	Average 10	Petro-chemical waste	Fluidized bed (stirred) exogenous and endogenous respiration
Activated sludge	NR	0.12	Secondary effluent	11	72	NR	29.5	NR	NR	Sewage	Fluidized bed endogenous respiration
Sand	0.45	42.2	Secondary effluent	3-10	2.8	13	5.5-11.0	4.6	Average 10 Maximum 35	Methanol	Fluidized bed

[a] in fluidized state
[b] based on volume of fluidized bed
NR Not reported

The anaerobic expanded bed process has been demonstrated to be effective for treating low strength wastes (COD less than 600 mg/L) at short retention times (several hours) and at high organic loading rates (up to 8 kg COD/m^3-day) even at low temperatures (10°C, 20°C) (Switzenbaum & Jewell, 1980).

Due to the relative insensitivity of the process performance to hydraulic retention time, the system could be designed at a low HRT (on the order of several hours). The actual HRT will depend on wastewater organic strength. Little gas is produced when treating low strength wastes and as a result anaerobic treatment of sanitary wastewater cannot be regarded as a large energy producer.

However, for high strength industrial wastes (COD range 5,000-50,000 mg/L) BOD and COD removal varies between 60-95 and 65-85 percent respectively in 0.3 to 4.9 days HRT over a wide range of organic loading (4-25 kg COD/m^3-day) with a considerable amount of methane production.

The experiments carried out on anaerobic fluidized beds by Jewell et al. (1981) to study the effect of shock loads, showed that the process was unaffected by large instantaneous fluctuations in temperature, flow rate, organic concentrations, and organic loading rate.

However, the optimum temperature for the treatment is around 35°C and for lower temperatures, reduced removal efficiencies are observed. The energy produced could be used to heat the influent wastewater depending on its temperature. Typically 0.4 liters CH_4 were produced per gram of COD removed at 35°C. Methane content averages approximately 70% of the biogas with a range between 65- and 75% (Table 4.11)

Table 4.11. Summary of methane produced (Jeris, 1982)

Wastes	COD gm/L	LCH_4/g COD removed	% CH_4	Kg COD/m^3-d
Food process	7-10	0.4	70	3.5-24.1
Chemical	12	0.41	82	3.5-5.7
Soft drink	4-18	0.41	60	-
Zimpro supernatant	7.8*	-	72	3.4-16.7*

* Based on ultimate BOD, 35 days

Expanded/fluidized beds are able to withstand severe hydraulic overloading. They are less suited for organic overloading compared to fixed film reactors (Henz & Harremoes, 1982).

Laboratory scale experiments carried out by Norrman (1982) for the treatment of black liquor evaporator condensate from a kraft mill, showed that the process in the expanded bed was upset after exposure to the toxic wastes while the fluidized bed showed no signs of toxicity influence even at high volumetric loads. The fixed bed reactor survived the toxicity effects. Generally, expanded/fluidized beds are believed to have a relatively poor ability to withstand the effect of toxic compounds and other environmental factors compared to fixed bed reactors. (Henz & Harremoes, 1982).

Results of a few pilot-plant scale experiments conducted indicate the feasibility of using expanded/fluidized bed reactor for denitrification.

4.7 Problems Associated with Expanded/Fluidized Bed Reactors

4.7.1 Non-attached Biomass

In fluidized bed reactors, the high vertical velocity will take the free swimming organisms with the flow and deteriorate the effluent quality (Henz & Harremoes, 1982).

4.7.2 Suspended Organics

Expanded and fluidized beds have their biomass structure damaged by significant amounts of suspended organics in the influent (Henz & Harremoes, 1982).

4.7.3 Foaming

Foaming problems were reported in expanded/fluidized beds (Henz & Harremoes, 1982).

4.7.4 Gas bubbles

Problems with gas bubbles are frequently met with in expanded/fluidized beds as in fixed beds and sludge blanket reactors. Gas bubbles may adhere to flocs/bed particles and cause these to rise in the reactor, and may result in wash-out of biomass or deterioration of the effluent quality (Henz & Harremoes, 1982).

4.7.5 Start-up

Although all anaerobic reactors have start-up problems, expanded and fluidized bed reactors are believed to be the most troublesome in this aspect. (Henz & Harremoes, 1982).

4.8 Advantages and Disadvantages

4.8.1 Advantages

All the advantages claimed for the anaerobic expanded/fluidized bed reactors are derived directly or indirectly from the high concentration of biomass. Generally 10-40 kg/m^3 of volatile solids loading can be achieved (Switzenbaum, 1983).

Expanded and fluidized beds have several important advantages over anaerobic filters. These include the following (Switzenbaum 1982):

- no danger of clogging;
- small headloss;
- easier removal or addition of active material;
- better hydraulic circulation (avoidance of short circuiting);
- ability to operate at lower required detention times and/or higher organic loading rates;
- greater surface area available per unit of reactor volume;
- greater efficiency; and
- better mitigation against toxic shocks and dilution of high strength wastes.

Further, Cooper and Wheeldon (1980) have claimed the following advantages for biological fluidized bed systems:

- the high biomass concentration achievable leads to a small plant design which could reduce the land area requirement by up to 80 per cent; and
- a large reduction of capital cost is achieved due to the greatly reduced reactor volumes.

4.8.2 Disadvantages

One disadvantage of this system is that recycling of effluent may be necessary to achieve bed expansion, and the system is more complex.

4.9 Conclusion

An anaerobic fluidized bed may be a good pretreatment process for sanitary wastewater, followed by some post-treatment process like aerobic treatment.

Numerous pilot tests on anaerobic expanded/fluidized beds have shown high removal efficiencies at greater loading rates and low retention times for a variety of industrial wastes. For high strength wastes, a considerable amount of methane is produced.

V. UPFLOW ANAEROBIC SLUDGE BLANKET TREATMENT

5.1 General

In contrast to the previous high solid retention time processes, the upflow anaerobic sludge blanket (UASB) operates entirely as a suspended growth system and consequently utilizes no packing material. A sludge blanket reactor is basically a dense blanket of granular or flocculated sludge placed in a reactor, which is designed to allow the upward movement of liquid waste through the blanket.

5.2 Upflow Anaerobic Sludge Blanket (UASB) Reactor

The UASB takes the advantage of the fact that with proper physical and chemical conditions, anaerobic sludge can be flocculated and formed into granules which have excellent settling properties. The use of upflow anaerobic sludge blanket reactors for industrial wastewater treatment is well documented. Henz and Harremoes (1982) listed investigations of the sludge blanket reactors treating various types of industrial wastes. Lettinga et al. (1980) summarized the most relevant results of laboratory and pilot plant experiments with various types of wastes. Lettinga et al. (1984), has summarized the full scale treatment experience with this system. Lettinga & Vinken (1980) and Lettinga et al. (1983, 1984) investigated the UASB reactor in domestic wastewater treatment. Recently considerable efforts have been made in the Netherlands towards the application of UASB reactor for low strength wastes with a high percentage of insoluble COD.

5.2.1 Principles

The UASB reactor is initially seeded with digester sludge and then fed in the upflow mode. After few months of operation a very concentrated sludge bed (40-100 kg VSS/m^3) develops near the bottom. This sludge is very dense and granular in nature with high settling velocities. Individual particles may grow to diameters of 1-2 mm in the absence of shearing forces induced by mechanical mixing. Development of this pelletized sludge depends on the characteristics of the wastewater and on the seed sludge used when starting the reactor.

Above the sludge bed is a blanket zone of more diffuse growth with lower particle settling velocities at concentrations between 15-30 kg VSS/m^3. COD removal occurs throughout the entire bed and blanket reaction zone and the system is self-mixed by rising gas bubbles. The bubbles which are produced in the reactor are removed by submerged gas collector. During the start-up, when gas production is minimal, it may be desirable to provide additional mixing by gas recirculation.

In the region around and above the gas-solids separator, solid-liquid separation takes place in a quiescent settling zone. All surfaces in the settling zone are constructed with steep slopes to allow settled solids to return to the bed-blanket region.

5.2.2 System Design

Like the other systems discussed, proper influent distribution, head space, and effluent draw off facilities must be designed. The key to successful operation of the UASB is to keep the sludge within the system (i.e., maintaining the solids without any support material). This is accomplished with the internal gas-solids separator and by minimization of mechanical mixing and/or sludge recirculation for the sake of improving the ability of sludge to settle.

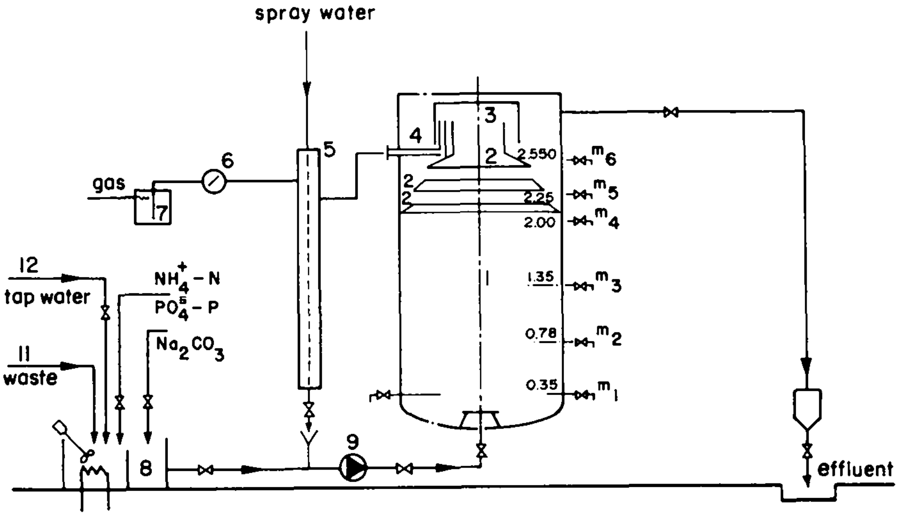

Fig. 5.1 Schematic diagram of the 6 m³ UASB pilot plant as employed in the experiments with sugar-beet and potato-processing wastes, although the gas-solid separator was slightly different in construction in the latter case. (1) digestion compartment (4.65 m³); (2) gas seal; (3) gas bowl; (4) settler compartment (1.15 m³); (5) foam separator; (6) gas meter; (7) water seal; (8) influent buffer with heating and supply of chemicals and - if desired - of dilution water. Height of the reactor; 3 m; 1.59 m i.e. (Lettinga et al., 1980)

A schematic diagram of a 6 m³ UASB pilot plant is given in Fig. 5.1.

5.2.2.1 Gas-Solids Separator

The gas-solids separator (GSS) located in the upper part of the reactor is particularly important. Mainly it serves for retaining the anaerobic sludge within the anaerobic reactor. This is accomplished by separating as effectively as possible the gas bubbles from the system at approximately 2 m beneath the effluent weir. Thus, a quiescent zone is created in the upper part of the reactor where sludge flocs or particles can flocculate, settle out and/or can be entrapped in the sludge blanket present.

In order to combat buoyancy of the sludge, the dimensions of the GSS-device should be such that the gas liquid interface in the gas collector will be well mixed by the up-flowing gas bubbles. The surface area of the gas-liquid in the gas collector should not be too small, because it might lead to severe scumming (Lettinga et al., 1982).

Stringent design criteria for the GSS-device is not yet provided. Lettinga et al. (1982) provided the following guidelines:

- The first main objective of the GSS-device is to accomplish an effective separation of the gas. For this purpose proper baffle plates should be installed beneath the aperture between the gas collectors.

- The second objective is entrapping sludge flocs (granules) conveyed into the settler compartment with the upflowing solution and returning them back into the digestion compartment. For this purpose the inclined wall of the settler should be approximately 50°C from the horizontal. The surface load of the reactor should not exceed 2 m/h in flocculant sludge bed systems.

- The height of the settler compartment should not be lower than 1.5-2 m and perhaps even should exceed 2 m.

5.2.2.2 Feed Inlet System

In a UASB reactor, short circuiting of the liquid flow could be reduced (if necessary) by increasing the number of feed inlet points.

Acceptable results can be achieved in highly loaded processes (i.e. at greater than 1-2 kg COD/m^3-d) with one feed inlet point per 5-10 m^2, although a secondary start-up of the process may proceed rather slow in that case. For purpose of shortening a secondary start-up, a larger number of feed inlet points should be pursued. This applies particularly for low loaded systems. In that case one feed inlet point per m^2 may be sufficient particularly in granular sludge and/or dense flocculant sludge bed systems (Lettinga, et al., 1982). In rather voluminous flocculant sludge bed systems (10-15 kg VSS/m^3) quite satisfactory results have been obtained with only one feed inlet point per 2 m^2 in treating raw sewage at liquid retention times of 8 hrs at temperatures in the range of 15°C-20°C (Grin et al., 1981). Lettinga et al. (1984), presented some rough guidelines for the required number of feed-inlet points. (Table 5.1).

Table 5.1. Rough guidelines for the number of feed-inlet nozzles required in a UASB reactor (Lettinga et al., 1984)

Type of sludge	Area (m^2) per nozzle
1. Dense flocculant sludge (exceeding 40 kg DS/m^3)	One at loads less than 1 - 2 kg COD/m^3-d
2. Thin flocculant sludge (less than 40 kg DS/m^3)	Five at loads exceeding approximately 3 kg COD/m^3-d
3. Thick granular sludge	One at loads of approximately 1 - 2 kg COD/m^3-d

5.2.2.3 The Height-Area Ratio

The amount of wastewater to be treated, the design capacity of the reactor and the maximum permissible surface load (tentatively 1-1.5 m^3/m^2-h for flocculant sludge systems) roughly dictates the height-area ratio of the reactor.

A schematic diagram of a large-scale UASB plant is given in Fig. 5.2.

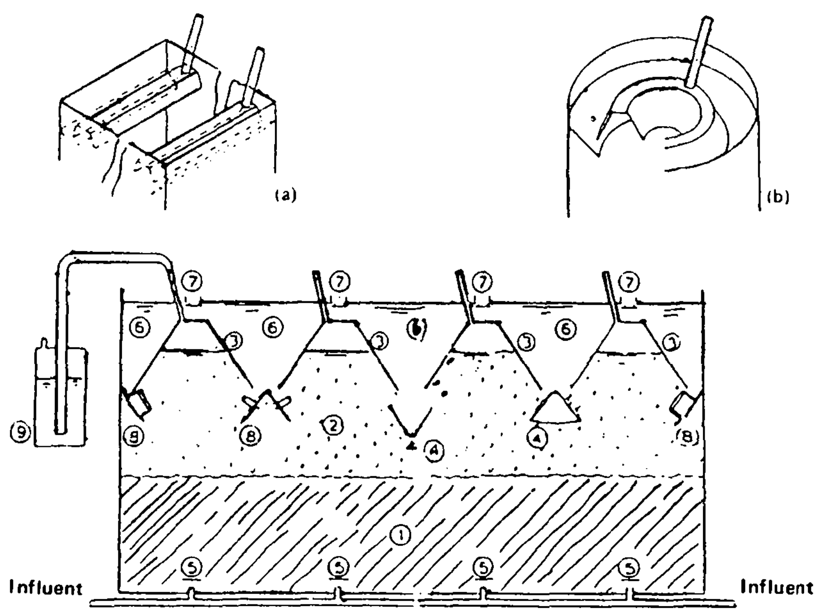

Fig. 5.2 Schematic diagrams of a large-scale UASB plant. (a) Rectangular reactor; (b) cyclindrical reactor. (1) Sludge bed; (2) bulk of liquid with suspended sludge; (3) bowl (collector); (4) gas seal; (5) feed inlet; (6) settler compartment; (7) launderer; (8) gas collector with gas outlet pipe to (3); (9) water seal. (Lettinga et al., 1980)

5.2.3 Recycling

With a sophisticated feed distribution system installed in a UASB reactor, effluent recycle (to fluidize the sludge bed) is not necessary as sufficient contact between wastewater and sludge is guaranteed even at low organic loads. (Lettinga et al., 1982). So far, effluent recycle is not generally applied in UASB reactors.

The results of a 12.5 liter (2 m height) UASB reactor is shown in Fig. 5.3. High organic and hydraulic loads have been applied.

On the other hand effluent recycle certainly will result in a high treatment efficiency as illustrated by results shown in Fig. 5.4, which were obtained with a volatile fatty acid's (VFA's) solutions in the same 12.5 liter (2 m height) UASB reactor as used in the experiment in Fig. 5.3

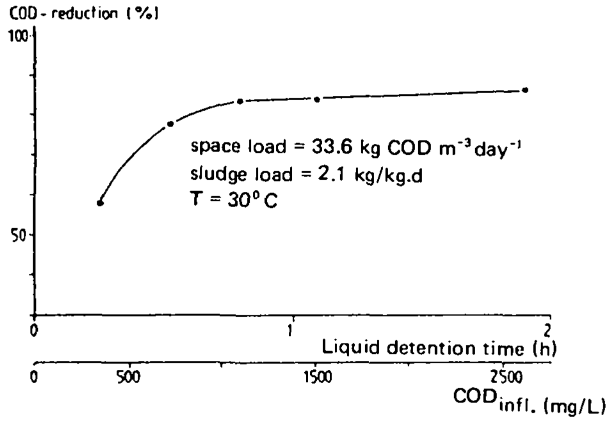

Fig. 5.3 Results of a 12.5 liter UASB-exp. with 210 g VSS (4 liter granular seed sludge and VFAs-feeds of various concentrations. (No effluent recycle) (Lettinga et al., 1982)

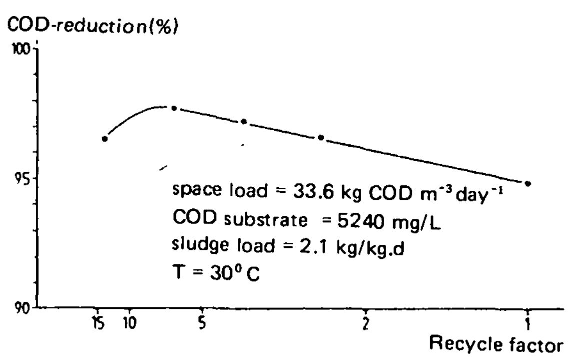

Fig. 5.4 Results obtained in a 12.5 liter (2 m high) USAB-reactor treating a medium strength VFA-feed in the case effluent recycle is applied. (Lettinga et al., 1981)

5.2.4 Mechanical Mixing

In order to prevent difficulty with buoying sludge, the gas liquid interface in the gas collector has to be kept well stirred.

At high organic loading rates the agitation caused by gas evolution will usually be sufficient to prevent an excessive buoying of sludge. However, at low organic loading rates - depending upon the design of the gas-solids separator and the type of waste treated, some form of mechanical agitation of the gas-liquid interface in the gas collector may be required. This applies in particular to wastes that can readily form a scum layer, i.e. dairy wastes and raw sewage (Lettinga et al., 1980).

5.3 Design Criteria

5.3.1 Organic Loading

None of the full scale plants built so far has been designed for space loads exceeding 15 kg COD/m³-d at 30°C, despite the fact that considerably higher loads have been applied in small and large pilot plants. The conservative design criteria are chosen due to the lack of full scale experience of UASB operating reactors (Lettinga et al., 1982).

5.3.2 Temperature

A treatment temperature of 30-35°C is being used in most of the current operating full scale plants. Lettinga et al. (1984), reported that recent experimental results conducted by Hulshoff Pol. (unpublished work) indicate that the rate of start-up can be maintained significantly by increasing the operational temperature (from 30°C to 38°C).

Lettinga and Vinken (1980) conducted experiments to study the effect of temperature on the specific activity (g COD/g VSS-d) using two types of feeds, a mixture of volatile fatty acid (VFA's) and dry solids from potato. Temperatures exceeding 45°C were found to be detrimental, which results in a sharp reduction of the specific activity amounting to 50-60% of the activity over the whole temperature range 20-40°C. Otherwise the system adapts very rapidly to temperature changes in the 10-40°C range.

Lettinga et al. (1982), provided guidelines for the design capacity of UASB plants treating mainly soluble wastes in relation to temperatures; details are presented in Table 5.2.

Table 5.2. Tentative rough design-capacities for UASB-reactors treating mainly liquid wastes in relation to the temperature (conservative figures) (Lettinga et al., 1982)

Temperature (°C)	Design-capacity kg COD/m³-d
40	15-25
30	10-15
20	5-10
15	2-5
10	1-3

5.3.3 Hydraulic Retention Time

Hydraulic retention times (HRT) are in the range of 4 to 24 hrs. The HRT depends upon the effluent characteristics and the treatment objectives. Very short hydraulic retention times (3-8 hrs) could be applied with medium concentrated wastewater

(1-3 kg/m^3 of soluble COD). With more concentrated wastewater (10-50 kg/m^3 of soluble COD), longer hydraulic retention times (approximately 1 day) have to be applied. In both cases, however, a reduction of 80-98% of a soluble COD load at approximately 15 kg COD/m^3-d could be obtained.

5.3.4 Sludge Bed Height

So far no sound arguments have been provided to restrict the total amount of granular sludge in a UASB reactor. Experiments carried out in a 6 m^3 pilot plant in which a dense-mainly granular-sludge bed occupied the lower 1-2 m of the reactor (with a more flocculant sludge blanket above it) have shown that space loads up to 45 kg COD/m^3-d with potato processing waste and 30 kg COD/m^3-d with sugar beet waste (Lettinga et al., 1982) can be well accommodated. Any evidence that a maximum should be set for the height of the granular bed was not obtained from these experiments.

The basic design criteria for reactors of 30 m^3 and 800 m^3, treating liquid sugar wastes and beet sugar wastes respectively are given in Table 5.3. The composition of the wastewater is presented in Table 5.4.

Table 5.3. Basic design features of the anaerobic treatment plants in operation in 1979 (Pette et al., 1980)

	30-m^3 plant	800-m^3 plant
Tank configuration	Cylindrical	Rectangular
Building material	Steel	Concrete
Height (m)	6	4.5
Bottom surface (m^2)	5	178
Depth of digesting zone (m)	4.9	3.3
Depth of settling zone (m)	1.1	1.2
Type of wastewater	Liquid sugar	Beet sugar
Organic loading (kg COD/d)	400	13,000
(kg COD/m^3-d)	13.3	16.25
Influent concentration (mg COD/L)	17,000	3,000
Purification efficiency (%)	94	88
Average hydraulic flow (m^3/h)	1	180
Peak flow (m^3/h)	1.5	240
Hydraulic settler surface load (m^3/m^2-h)	0.5	1.5
Gas Production (m^3/m^2-h)	1.7	1.2

5.4 Application Status

5.4.1 Municipal Applications

Lettinga et al. (1980) presented experimental data (120 liter reactor) for the upflow anaerobic sludge blanket (UASB) reactor treating domestic raw sewage with a COD greater than or equal to 400 mg/L operating at a 12 h HRT, throughout the year. During the winter period the temperature ranged from 5 to 10°C and COD removal

Table 5.4. Composition of wastewater (Pette et al., 1980)

Parameter	Beet sugar, 800-m^3	Liquid sugar, 30-m^3
COD (mg/L)	1,000-2,600	8,000-56,000
BOD_5 (mg/L)	550-1,450	
Sucrose (% of COD)		50
Acetate (% of COD)	40	20
Propionate (% of COD)	45	5
Butyrate (% of COD)	4	5
Suspended solids (g/L)	0.1-1.0	0
Kjeldahl-N (mg/L)	20-55	10-40
pH	6.8-7.8	3.2-4.0
Alkalinity (meq HCO^{-3}/L)	10-40	

efficiencies of 65-85% are reported. Sugar beet waste adapted granular sludge was used as the seed for the reactor.

In a 6 m^3 pilot plant (reactor height 3 m) experiment for raw sewage with a COD of 400 mg/L using digested sewage sludge as the seed, at a temperature of 20°C with a HRT of 22 hours, COD removal efficiency of 60-80% was obtained. (Lettinga et al., 1980). Slight mixing (by gas recirculation) was applied.

Lettinga et al. (1983) presented pilot scale data for the upflow anaerobic sludge blanket treating domestic wastewater. At hydraulic retention times as low as 12 hours, 65-85% COD reduction was reported, but with heavy rainfall (i.e. low influent CODs), COD reduction dropped to 50-70%, and at very low CODs, to less than 50%. Over the course of the experiments, the average total COD concentration was 163.2 mg/L. Lettinga et al. (1984) reported the results of experiments with raw domestic sewage at lower temperatures and the results are summarized in Table 5.5.

Table 5.5. Main results obtained in a 6 m^3 UASB plant with raw domestic sewage as influent (temperature range 9.5 - 19°C, liquid-retention time 8 hours) (Lettinga et al., 1984)

Measurement	Week number		
	1-11	16-22	24-26
Temperature (°C)	19-15	11-12	9.5-10
Effluent $COD_{filtr.}$ (mg/L)	100-200	150-175	175-250
COD reduction $(\%)_{filtr.\ effl.}$	65-80	55-70	55
COD reduction $(\%)_{raw\ effl.}$	40-55	30-50	30
CH_4 production (m^3/kg $COD_{infl.}$)	0.130	0.090	0.050
Excess sludge production (kg DS/kg $COD_{infl.}$)	0.195	0.172	0.271

Fernandes et al. (1985) concluded that the performance data on two small UASB reactors treating settled domestic wastewater obtained at the Water Research Centre, Stevenage Laboratory, essentially confirmed the studies by Lettinga and others at the University of Wageningen. Anaerobic treatment removed 50 to 80% of the BOD and SS from settled wastewater at hydraulic retention periods of 3 to 12 h at 20°C, with only minimal production of sludge.

Results obtained in a 50 m^3 UASB with raw domestic sewage in Cali, Colombia, have shown a treatment efficiency of 60-80% despite the fact that the wastewater COD was less than 300 mg/L and the wastewater temperature was around 26-28°C. (Lettinga et al., 1984)

Lettinga et al. (1982) reported results obtained with granular sludge UASB reactors and wastewater containing high fraction of suspended matter. As seen from Table 5.6, high loading rates were reported.

Table 5.6. Results obtained with granular sludge UASB-reactors and wastewaters containing a high fraction of suspended matter (Lettinga et al., 1982)

Origin	Waste water		Reactor		Amount of seed sludge	COD-load	Temp.	COD-reduction
	COD (g/L)	Fraction dissolved (%)	Volume (liter)	Height (m)	(gram)	kg/m^3-d	(°C)	(%)
Animal carrion	3-7	75	60	2	1500	30	30	92
destruction	3.5	75	60	2	1500	60	30	78
Slaughter	1.5-2.2	50	30	1.3	1000	10	30	87
house	1.5-2.2				1000	6	20	91
Raw sewage	0.2-0.9	5-35	120	2	3300	0.7-2.7	8-20	50-85

5.4.2 Industrial Applications

Considerable research work has been carried out in recent years in the Netherlands towards the application of UASB reactor for the treatment of medium strength wastes such as slaughterhouse and meat processing wastes. Sayed et al. (1984), recently presented the application of UASB reactor for treating slaughterhouse waste. A pilot plant of 30 m^3, seeded with 13 m^3 of digested sewage sludge from a municipal treatment plant was used. The characteristics of the slaughterhouse wastewater is presented in Table 5.7.

The main part of the pollutants present was insoluble and non-biodegradable or poorly biodegradable suspended matter. The wastewter contained 40-50 percent of suspended solids as a percentage of total COD. Initially the pilot plant was operated at a temperature of 30°C, but, 20 weeks after the start-up, the temperature was reduced to 20°C. Table 5.8 summarizes the applied overall organic loading rates, the liquid retention times and the temperature over the entire experimental period for the pilot plant reactor.

Table 5.7. The characteristics of the wastewater (Sayed et al., 1984)

COD-total	1,500-2,200 mg/L
BOD-total	490-650 mg/L
Nitrogen (N) Kjeldahl	120-180 mg/L
Phosphate (P) - total	12-20 mg/L
pH	6.8-7.1
Suspended solids as a percentage of COD-total	40-50
Acidification (percentage COD-VFA of COD-total)	12-15
Grease	50-100 mg/L (5% of the total solids)
Temperature	20°C

Table 5.8. The organic loading rates, the liquid detention times and the process temperature over the entire experimental period (Sayed et al., 1984)

Phase	Number of days	Organic space load (kg COD/m³-d)	Liquid detention time (h)	Temperature (°C)
I	0-142	0.5-1.5	40-20	30
II	159-296	1.5-2.5	18-12	20
III	297-324	2.5-3	10	30

The results of the loading rates and treatment efficiencies are shown graphically in Fig. 5.5. Organic space loads up to 3.5 kg COD/m³-d and a liquid detention time of 7 hours at temperature as low as 20°C were achieved at fairly satisfactory treatment efficiencies; after 3 weeks, up to 70% total COD reduction and 90% soluble COD reduction were achieved. The soluble BOD_5 was reduced from an average value of 650 to 25 mg/L, corresponding to a 96% treatment efficiency. As expected, the anaerobic digestion procedure was inefficient in nitrogen reduction, with only approximately 24% of the total N being removed.

The methane yield amounted to 0.28 m³ per kilogram of COD removed; the methane content of the biogas from the wastewater varied between 65 and 75%.

Numerous laboratory studies of the UASB process indicate successful treatment of a number of industrial wastes with suggested design loading rates of 10-14 kg COD/m³-d corresponding to a maximum sludge loads in the range of 1 kg COD/kg VSS-d. The most relevent results are summarized in Table 5.9 (Lettinga et al., 1980).

The first pilot experiments have been carried out with sugar-beet wastes (Lettinga, 1980). Initially a 6 m³ reactor with a height of 3 m was used. In the second stage a 30 m³ reactor with a height of 6 m was tested in order to obtain additional data for the design of a full-scale plant.

Table 5.9. Results of some laboratory UASB experiments with various type of wastes[a] (Lettinga *et al.*, 1980)

Origin	Waste solution: COD Total (mg/L)	COD Dissolved[b] (%)	VFA (mg/L)	Maximum COD load applied: Space load (kg/m²-d)	Sludge load (kg/kg VSS-d)	HRT (h)	COD reduction (%)	Sludge yield factor (kg/kg COD)	Temp. (°C)	USAB reactor: Volume (liter)	Height (cm)
Sugar-beet sap, unsoured	5000-6000	95	nil	4-5 (diss.)	0.5-0.8	48-24	95	0.15	30	61	105
Sugar-beet sap, soured (closed circuit)[c]	6000-9500	80-60	4000-5000	8-10 (diss.) 10-14 (total)	0.7-1.1	12-24	84-95 (diss.) 65-75 (total)	0.09-0.07 (diss.) 0.1 (total)	30	18 30	70 100
Sugar-beet sap, soured (two-stage)	6000-9000	95	5400-8000	8-9 (2nd stage)	0.8-1.0 (2nd stage)	24	90-97	0.04-0.03	30	18	70
Bean blanching	5200	90	500-1500	8-10	0.6-0.8	13-15	90-95	-	30	2.7	30
Sauerkraut	10000-20000	97	400-1500	8-9	0.8-1.2	24	88-93	0.05-0.07	30	2.7	30
Dairy (skimmed milk)	1500	75	nil	7-8	0.4-0.6	5	90	0.18[d]	30	18	70

[a] Values mentioned for the COD load concern the maximum values that could be applied in the specific experiment.
[b] COD remaining after filtration over a filter SS 520 b.
[c] Experiments conducted in a closed simulated wastewater circuit: a known amount of sugar-beet sap solution is supplied continuously in the circuit water.
[d] Higher values are obtained in case of substrate precipitation (i.e. at a pH fall or in case of overloading).

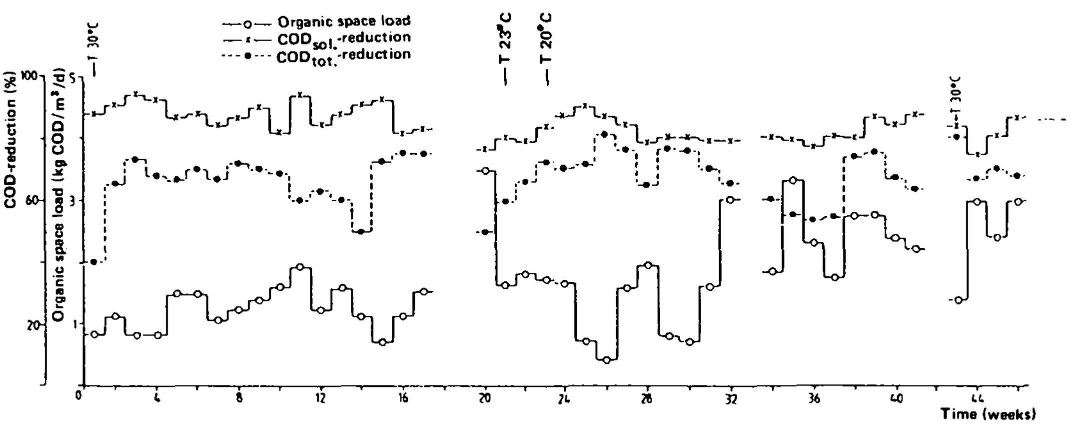

Fig. 5.5 The results of the organic space load and the reduction rates, during the entire experimental period (Sayed et al., 1984)

Pilot-plant experiments with 6 m^3 reactors for the treatment of both sugar-beet and potato processing wastes were started up with sludge obtained from a municipal digester. The results indicated that with both type of wastes a well-adapted sludge could be obtained within a period of 8-12 weeks. Space loadings up to 10 kg COD/m^3-d at 30°C and hydraulic loads upto 3 m^3/m^3-d were experienced.

The 30 and 200 m^3 reactors were seeded with well-adapted sludge fostered during the former experiments, and no difficulties were reported to start the process. The 200 m^3 reactor seeded with 1800 kg sludge (84% VS) could digest a COD load of 8.5 kg COD/m^3-d after only 14 days of operation. Exceptionally high organic and hydraulic loading rates were possible in combination with satisfactory treatment efficiencies. Moreover, this ability was observed for both soured and unsoured waste. The results as summarized by Lettinga et al. (1980) is presented in Table 5.10.

It has been quoted by Lettinga et al. (1980) that considerably higher loads can be applied in UASB systems with potato starch wastes (COD approximately 16 g/L), viz. up to 30 kg COD/m^3-d at 30°C and presumably even higher. The COD reduction was approximately 92% corresponding to a BOD-reduction of 97%. Up to 30 kg VSS/m^3 of well pelletized sludge could be retained in a 2 m^3 UASB reactor (diameter 0.8 m, height 4 m).

So far over 50 full-scale UASB plants have been installed and Table 5.11 shows an up-to-date list of the UASB plants in operation. As reported by Lettinga et al. (1984) the performance of these plants are very satisfactory with a few exceptions. The La Cross, Wisconsin, UASB, reactor which treats wastes from a brewery, is the largest in the world.

Denitrification

The feasibility of using the UASB reactor feasibility for denitrification was investigated in 24 and 41 liter reactors both approximately 1 m in height (Lettinga et al., 1980). Acetate solution as well as alcoholic wastes have been used as the carbon source in these experiments. The main results are contained in Table 5.12. Very satisfactory results were obtained with respect to nitrate and/or COD reduction, even at extremely high organic and hydraulic loading rates. It is reported that with alcoholic waste, a granular sludge (pellets of 1-3 mm in diameter) developed in the course of 6-8 weeks.

Table 5.10. Results obtained in a 6-m^3 pilot-plant and 200 m^3 full-scale plant experiments (Lettinga et al., 1980)

Type of waste	UASB reactor (m^3)	Influent characteristics: COD range[a] (mg/L)	Influent characteristics: Soured (%)	Maximum loading rates applied: Organic (kg COD/m^3-d)	Maximum loading rates applied: Hydraulic (m^3/m^3-d)	Treatment efficiency based on: Temp. (°C)	Treatment efficiency based on: $E_{COD_{cen.}}$[b] (%)	Treatment efficiency based on: $E_{COD_{diss.}}$[c] (%)	Treatment efficiency based on: $E_{COD_{cen.}}$ (%)
Liquid sugar	6	4000-6000	15-25	20-25	4	28-32	92-95	93-98	-
Campaign waste	6	3500-4000	75	30-32	4-6	28-32	95-80	95-98	-
Campaign waste	200	4000-5200	70-90	14-16	3-4	30-34	87-95	-	90-95
Potato processing	6	2000-5000	25	3-5	1.2	19	95(92)[d]	-	94
(lime used as	6	2000-5000	30	10-15	3	26	95(89)	-	98
neutralizing	6	2000-5000	12	15-18	4	30	95(89)	-	97
agent)	6	4000-16500	8	25-45	6-7	35	93(89)	-	96

[a] COD value based on centrifugated samples.
[b] $E_{COD_{cen.}}$: based on centrifuged effluent samples and raw influent COD values.
[c] $E_{COD_{diss.}}$: based on centrifuged influent COD values and effluent COD values determined after flocculation of the sample with 200 mg Fe^{3+}/L.
[d] Values in parentheses refer to effluent samples that have been allowed to settle for 30 minutes.

Table 5.11. UASB plants installed and commissioned by July 1984 (Lettinga et al., 1984)

Type of waste water	No. of plants	Country of installation	No. of plant in country cited	Design* Capacity (kg COD/m^3-d)	Reactor Volume (m^3)
Sugar beet	10	Netherlands	7	12.5-17	200-1700
		Germany	2	9, 12	2300, 1500
		Austria	1	8	3000
Liquid sugar	1	Netherlands	1	17	30
Potato processing	10	Netherlands	8	5-11	240-1500
		USA	1	6	2200
		Switzerlands	1	8.5	600
Potato starch	3	Netherlands	2	8.5, 15	1700, 5500
		USA	1	11	1800
Maize starch	1	Netherlands	1	10-12	900
Wheat starch	3	Netherlands	1	7	500
		Ireland	1	9	2200
		Australia	1	11	4200
Alcohol	2	Netherlands	1	16	700
		Germany	1	9	2300
Yeast	1	USA	1	12	4400.
Brewery	2	USA	1	14	4600
		Netherlands	1	5-10 (23°C)	1400
Shellfish	1	Netherlands	1	10	2 x 50
Slaughterhouse	1	Netherlands	1	3-5	600
Dairy	1	Canada	1	6-8	450
Paper	2	Netherlands	2	8-10	1000, 740
				4 (20°C)	740
Vegetable canning	1	Netherlands	1	10	375
White spirit	12	Thailand	12	15	3000

* At treatment temperature 30-35°C, unless otherwise stated.

Table 5.12. Results of denitrification experiments using the USB process (Lettinga *et al.*, 1980)

COD load		Influent						
Space load (kg/m³-d)	Sludge load (kg/kg-d)	COD (mg/L)	NO_3^--N (mg/L)	HRT (h)	COD reduction (%)	NO_3^--N reduction (%)	Amount of sludge in reactor (g)	TS in sludge bed (g/L)
1) Experiments with acetate as C source[a]								
5.0	0.3-0.33	288	68	1.5	98	100	380	20
4.8	0.3-0.31	194	44	0.97	90	99	388	25-22
11.1	0.6-0.65	178	36	0.38	55	42	450	42-33
2) Experiments with alcoholic-waste[b] as C source[c]								
19.7	1.1	2600	858	3.2	98	89	700	60
18.8	1.1	1590	426	2.0	97	96	650	47
19.7	1.1	495	163	0.6	93	85	700	31
19.7	1.1	370	123	0.5	91	82	700	-
19.7	1.8	270	89	0.3	92	77	420	-

[a] USB reactor: 31.2 liter volume; d_1 = 19 cm, 110 cm height; stirring: 2 sec at 45 rpm every 1-2 min.
[b] Composition of the undiluted waste; methanol: 380 g COD/L; ethanol: 201 g COD/L; propanol: 81 g COD/L; butanol: 76 g COD/L.
[c] USB reactor: 41 liter volume; d_1 = 19 cm; 140 cm height; stirring: 6 sec at 21 rpm every 30-45 sec.

5.5 Applicability

Although the UASB process was originally developed for the treatment of mainly soluble low- and medium-strength wastewater, satisfactory results have already been achieved with complex wastes, at optimal and suboptimal mesophilic temperatures.

The potential of the UASB concept for treating mainly soluble liquid wastes has been demonstrated in both full-scale and pilot plant UASB reactors, as well as in numerous bench-scale UASB experiments with various types of wastes, e.g., from sugar beet (soured as well as unsoured), bean blanching, sauerkraut, alcholic fermentations, potato processing, as well composite VFA wastes etc.

5.5.1 Soluble Wastes

In treating mainly soluble wastes, generally very high loading rates can be applied. Some results achieved with VFA substrates, and alcoholic waste are summarized in Table 5.13.

In applying very high sludge loads, appropriate adjustments have to be made to the design of the GSS because, under these conditions, a considerable fraction of the sludge granules will be redispersed in the liquid medium above the sludge bed, because of the marked turbulence brought about by rising gas bubbles, as well as the increasing tendency of the granules to float (Lettinga et al., 1984).

5.5.2 Complex Wastes:

In treating complex (i.e. partially insoluble) wastes, generally low loading rates are applied. High loading rates are possible with complex wastes only when employing granular sludge-bed reactors. In flocculant sludge-bed UASB reactors the presence of poorly or non-biodegradable suspended matter in the wastewater will result irrevocably in a sharp drop in the specific methanogenic activity, because the dispersed solids will be trapped in the sludge. Moreover, any significant granulation will not occur under these conditions. The maximum loading potential of such a flocculant sludge-bed system is in the range of 1-4 kg COD/m^3-d. (Lettinga et al., 1984). This applies particularly to low-strength wastes in which the insoluble fraction is less than about 50%, but is also true for medium-and high-strength wastes which after hydrolysis (and acidogenesis), do not allow an easy separation of the remaining solids from the liquid. Lettinga et al. (1984) presented (in Tables 5.14 and 5.15) some relevant results for complex wastes with flocculant and granular sludge-bed UASB reactors respectively.

The performance of the UASB system with respect to the removal of SS is fairly poor, particularly at low temperatures. This presumably is due to channelling in the sludge bed, resulting from the very low gas production at these low temperatures. At higher temperatures, finely dispersed matter present in the raw sewage is removed considerably more efficiently, allowing the system to be exposed to higher hydraulic and organic loads. The maximum loading potential of flocculant sludge bed reactors for raw sewage has not been established for temperatures exceeding 20°C.

Significantly higher loading rates can be accommodated in granular-sludge UASB reactors compared with flocculant-sludge bed reactors, and it is recognized that the SS reduction in granular-sludge bed systems becomes very poor at high loads, mainly because of the considerable turbulence resulting from the vigorous gas evolution. A primary or secondary settler therefore has to be installed in line with the anaerobic reactor. (Lettinga et al., 1984)

Table 5.13. Results obtained with UASB reactors using granular or mainly granular seed sludge, and VFA solutions, alcoholic wastewater and potato-processing waste as feed (Lettinga et al., 1984)

Substrate characteristics			Experimental conditions						COD reduction	
Type	COD (mg/L)	Soured (%)	Medium used for growth of seedsludge inoculum	Reactor volume	Volume of sludge bed	COD sludge load (kg/kg VSS-d)	Temp. (°C)	COD load (kg/m³-d)	Total (%)	Filtered effluent (%)
VFA	$1000C_2$ $1000C_3$	100	VFA	30L	15-20L	2-3	30	62		80-90
Alcoholic*		0	Sugarbeet waste	2.7L	~1.5	0.6-0.7	30	22		>95
				28L	~10	0.7	80	14		85-90
Potato	2.5-4.2		Digested	6m³	< 2m³	0.27	19	3-5	88	95
processing	3.3-5	6-16	sewage	6m³	< 2m³	0.65	26	10-15	86	95
	3.5-4.5		sludge	6m³	< 2m³	0.97	30	15-18	83	95
	3.5-7.1			6m³	~ 4m³	1.45	35	25-45	84	93

* Methanol 51%; ethanol 27%; propanol 12%; butanol 10%.

Table 5.14. Results obtained with three types of complex wastes using flocculant-sludge UASB reactors (Lettinga *et al.*, 1984)

Waste	Influent COD* total (mg/L)	Influent COD* soluble (mg/L)	Temperature applied (°C)	COD load applied (kg/m^3-d)	COD reduction achieved** Filtered (%)	COD reduction achieved** Unfiltered (%)	Volume of reactor
Domestic sewage	322-950	235-460	15-20	max. 2.0	30-80		6 m^3
			9-12	max. 2.0			
Calf-fattening	9500	3800	30	4	93	90	25 L
			25	2	90	85	25 L
Slaughterhouse	1500-2200	600-1100	30	2.5-3.0	75-85	65-80	30 m^3
			20	1.5-2.5	78-85	55-75	

* COD values for domestic sewage comprise average values of 5-15 composite daily samples.

** Filtered: COD reduction based on filtered effluent and raw influent; unfiltered: COD reduction based on raw effluent and influent samples.

Table 5.15. Results obtained with three types of complex wastes using granular sludge UASB reactors (Lettinga et al., 1984)**

Origin	Waste water COD (g/L)	Waste water Fraction dissolved (%)	Reactor Volume (L)	Reactor Height (m)	Amount of seed sludge (g)	COD load (kg/m³-d)	Temp. (°C)	COD reduction* Filtered effluent (%)	COD reduction* Unfiltered effluent (%)
Rendering wastes	5500	70	60	2	1500	27	30	94	64
	3000	85	60	2	1500	63	30	80	63
Slaughterhouse	1.5-2.2	50	30	1.3	1000	10	30	87	
	1.5-2.2				1000	6	20	91	
Raw sewage	0.2-0.9	5-35	120	2	3300	0.7-2.7	8-20	60-89	54-72

* Based on filtered effluent samples and unfiltered influent samples.
** Averaged values measured over periods of 5-12 days.

The UASB reactor was found to perform well at an intermittent feeding. Start-up difficulties were not reported for feed interruptions of even for few weeks. It is of practical importance for situations where large daily and/or weekly variations occur in the pollution load; as a result installation of an equalization tank can be avoided or the size of the tank could be reduced significantly.

The process accommodates fairly well to hydraulic and organic shock loads, temperature fluctuations, and low influent pH values, provided that the reactor pH remains well above 6.0 and that the sludge load applied is below the maximum specific COD removal rate of the sludge at the temperature prevailing in the reactor.

The UASB reactor (as in the case of an anaerobic filter) seems to be relatively insensitive to inhibition by sulphide due to its high sludge retention time (Parkin and Speece, 1982). In the UASB system, an attractive volumetric loading rate could be achieved when the methanogenic activity of the sludge is still very low. In UASB reactor treating yeast-production waste, loading rates up to 14 kg COD/m³-d could be applied at hydrogen sulphide concentrations of approximately 90 mg/L. As quoted by Lettinga et al. (1984), a COD removal efficiency of 60-80% could still be reached under these circumstances.

The UASB reactor is able to tolerate extremely high organic and hydraulic loading rates up to 30 kg COD/m³-d and 8 m³/m³-d respectively once the sludge has been adapted to the waste. None of the full scale plants built so far has been designed for space loads exceeding 15 kg COD/m³-d at 30°C.

Results of a few pilot scale experiments conducted indicate the feasibility of the UASB reactor's use for denitrification and acid fermentation. As in the case of other anaerobic wastewater treatment systems, the UASB treatment may require further post treatment in the form of cascade aeration or the effluent could be discharged into the sewer system for further treatment.

5.6 Problems Associated with Anaerobic Upflow Sludge Blanket Reactor

5.6.1 Start-Up

To achieve high treatment efficiencies at high loading rates the formation of a highly settleable and active sludge in the UASB reactor is of utmost importance. Granular type of sludge is reported to have these properties (Lettinga, et al., 1980).

It has been found (De Zeew and Lettinga, 1980, Hulshoff Pol. et al., 1982) that space loads exceeding 5 kg COD/m^3-d can already be accommodated within 6-7 weeks in using VFA's (Volatile fatty acids) mixtures as substrate, provided the start-up of the process is managed well.

Lettinga et al. (1984) proposed the guidelines given in Table 5.16 to cultivate a granular sludge during the start-up of the process. Following the guidelines given in Table 5.16, it is possible to cultivate a granular sludge on VFA feeds. The first granules have been observed 4-6 weeks after the start of the experiment, but as pointed out by Hulshoff Pol et al. (1982), various factors are involved in the granulation process.

Lettinga et al. (1984), reported that the rate of start-up can be enhanced significantly by supplying a suitable carrier material to the seed sludge, e.g. anthracite particles (Hulshoff Pol, unpublished work).

Table 5.16. Tentative guidelines for the first start-up of a UASB plant using digested sewage sludge as seed (Lettinga et al., 1984)

1.	Amount of seed sludge: 10-15 kg VSS/m^3
2.	Initial sludge load: 0.05-0.1 kg COD/kg VSS/day
3.	No increase of the sludge load unless all VFA's are more than 80% degraded.
4.	Permit the wash-out of voluminous (poorly settling) sludge
5.	Retain the heavy part of the sludge

Factors Affecting the Granulation Process (Hulshoff Pol et al., 1982).

As bacterial granulation must be primarily governed by bacterial growth, the granulation process will be affected by the following:

a) Environmental conditions, such as:

- the availability of essential nutrients, because growth conditions should be optimal;

- the temperature, since the specific activity of methanogenic sludge is highly temperature dependent;

- the pH, which should be in the optimal range (6.5-7.8); and

- the type of wastewater, with regard to the composition of the waste, the biodegradability of the organic matter, the presence of finely dispersed non-biodegradable organic and inorganic matter, the ionic-composition (concentration of uni- and divalent-cations) and the presence of inhibitory compounds.

b) The type of the seed sludge, i.e. with respect to its specific activity, its settleability and the nature of the inert fraction.

c) The process conditions applied during the start-up, such as:

- the procedure followed in increasing the loading rate, e.g. the extent of overloading and the allowed wash-out of suspended solids;

- the amount of seed sludge used.

Distillery waste and corn-starch wastes are reported to have problems with the formation of granular sludge although ultimate granulation of the sludge occurred. (Hulshoff Pol et al., 1982)

Hulshoff Pol et al. (1982) conducted experiments to verify the influence of sludge loading rate, the addition of small amount of granular sludge to the seed sludge, and to investigate the effect of NH_4^+ and Ca^{++} concentration in the influent on the pelletization process of sludge granules. Further investigations are necessary to establish exact criteria and the reader is referred to an article by Hulshoff Pol et al. (1982) for a more complete discussion of factors affecting the granulation process.

5.6.2 Sludge Washout

The results of all start-up experiments carried out with UASB reactors reveal a very significant wash-out of sludge during the initial start-up phase of the process and consequently a deep depression is developed in the retained (heavier) amount of the seed sludge. Due to the wash-out of the sludge-ingredients, a considerable fraction of the net bacterial growth occurring during the initial phase of the start-up is also lost with the effluent.

To prevent the formation of a deep depression in the retained sludge, the following measures are proposed by Hulshoff Pol et al. (1982).

- The selection of the right type of seed sludge.

  There is evidence that a thick, relatively inactive seed sludge is beneficial to a good sludge retention (De Zeeuw, 1980).

- Adjusting the wastewater composition.

  The addition of nutrients or vitamins if necessary to assure optimal growth conditions.

- Proper management.

  During the start-up, long periods of over- and under-loading must be prevented. Under-loading leads to the development of voluminous sludge. Over-loading is detrimental because of the gas production that will occur in the gas-solids separator (GSS), which will hamper the settlement of the sludge in the GSS.

Sludge wash-out during the operation of the process is closely connected to the finely dispersed sludge present in the reactor. When the top of the sludge blanket remains well below the effluent weir of the reactor, the sludge washout is considerably less than the sludge accretion from growth. In this case the sludge lost in the effluent may consist of a considerable amount of suspended solids which may originate from the suspended solids in the influent. When the sludge blanket reaches the effluent weir under steady loading conditions, the wash-out of sludge and the accretion of sludge by growth will range over a similar order of magnitude per unit time (Lettinga et al., 1980).

Temporary drastic wash-out of the sludge may occur under an excessive expansion of the sludge blanket due to the result of a shock loading or due to suddenly deteriorating conditions (i.e. nutrient deficiency, high concentration of finely flocculating matter, etc.). This may last until a new steady bed could be established.

Some relevant data concerning the wash-out of sludge as summarized by Lettinga et al. (1980) are presented in Table 5.17.

5.6.3 Foaming

Excessive foaming has been encountered in the event of relatively poor treatment efficiencies, due to overloading or nutrient deficiency, and at very high gas production rates. Difficulties may be expected in the wastes that are relatively rich in proteins, such as potato starch wastewater (Lettinga, 1980). Foaming could be effectively depressed by adding anti-foaming agents to the feed solution.

5.7 Advantages and Disadvantages

5.7.1 Advantages

- Simple construction.

- No need for any form of mechanical mixing except, if applied, at low organic loads (2-4 kg COD/m^3-d), at high hydraulic loads (3-4 m^3/m^2-d) and in treating wastes containing a significant amount of undissolved solids.

Table 5.17. Suspended solids washout as found in 6 m³ pilot-plant and full-scale experiments with the UASB process (Lettinga *et al.*, 1980)

Period of experiment	COD space load (kg/m³-d)	HRT (hour)	SS washout (mg SS-COD/g COD infl. [a]) Total	SS washout After 30 min. settling	Total amount of sludge in reactor (kg TS)	Total amount of sludge in reactor (kg vol TS)
1) Potato waste (6 m³ pilot plant)[b]						
408-413	10-13	6.3	70-120	40-80	110-120	89-97
423-427	15-17	6.0	70-90	-	125-130	110-114
429-434	22-30	3.9	60-90	30-50	145-155	127-136
436-440	33-43	3.5	40-110	30-50	200-210	174-183

Day	COD space load (kg/m³-d)	HRT (hour)	SS washout (mg SS-COD/g COD infl. [a])	Total amount of sludge in reactor (kg TS)	Total amount of sludge in reactor (kg vol TS)
2) Sugar-beet waste (200 m³ full-scale plant)[c]					
54	11.2	6.7	100	6500	2280
61	13.9	7.6	60	6650	2530
68[d]	12.9	7.5	200	6650	2530
75	14.0	8	74	6500	-

[a] SS COD-content in potato wastes varied between 100-150 mg COD/g COD infl. and in the sugar beet waste between 70-150 mg COD/g COD infl. during the experimental periods considered.

[b] Sludge growth; 0.18-0.20 g sludge-COD/g COD infl. (30°C); 0.26-0.30 g sludge-COD/g COD infl. (20°C)

[c] Sludge growth; 0.08-0.16 sludge-COD/g COD infl.

[d] Only 20% of COD was present as VFA-COD, whereas this was 80-85% during the other days.

- Comparatively less investment when compared to an anaerobic filter or a fluidized bed system.

- Higher loading capacities and treatment efficiencies compared to that of anaerobic filter process.

- No real need for applying effluent recycle.

5.7.2 Disadvantages

- High wash out of suspended matter.

- Significant wash-out of sludge during the initial phase of the process and consequently a deep depression is formed in the retained amount of seed sludge.

- Necessity to develop dense and preferably a granular type of sludge in the reactor on the wastewater submitted to the reactor.

- Requirement of sufficient amount of granular seed sludge when the formation of dense, and granular sludge is not possible.

5.8 Conclusion

Upflow anaerobic sludge blanket process is a very suitable treatment process for soluble industrial wastes. A high concentration of very active methane forming sludge could be maintained in the reactor (30-40 kg/m^3). This makes the system highly resilient regarding fluctuations in COD-loading (0-20 kg COD/m^3-d), pH (6-8) and temperature (25-40°C).

VI. COMPARISON OF HIGH-RATE ANAEROBIC WASTEWATER TREATMENT REACTORS

The four reactors reviewed in this book are similar in that each is a means of maintaining the active biomass independently of the hydraulic retention time in the fermentation process. Thus high biomass concentrations, for systems stability and high efficiency, are achieved with low HRT, which is necessary for system economy. Each represents an advancement of conventional anaerobic digestion technology.

As reported by Lettinga et al. (1984), so far no comparison study have been made at relevant scales between the various high-rate processes, although such a study is in progress at the Waste Water Technology Centre, Burlington, Canada (Hall, 1981) for a UASB process, an anaerobic stationary fixed film reactor, an anaerobic filter and a small-scale fluidized bed reactor.

The main factors determining the hold-up of viable biomass of various high-rate systems, as compiled by Lettinga et al. (1984) is given in Table 6.1.

Other important features of the various high-rate anaerobic wastewater treatment systems, such as the rate of start-up, the capacity of the process to remove suspended solids, the risk of clogging, the need for effluent recycle, the installation of a sophisticated feed inlet distribution system, a gas-solids separator (GSS) and the use of packing materials are listed in Table 6.2.

All of the reactors have significant problems with start-up. It is essential to have the proper culture acclimatized for a given wastewater for a rapid start-up since methonogenesis have slow growth rates. However, for fixed film processes, microbial adsorption must occur, and then granulation or attachment to form the proper biofilm. These processes are not well understood and hence there is no clear rational means of starting up anaerobic reactors. However, all the systems have in common the necessity for optimal conditions for bacterial growth as no high loading rates can be accommodated if a sufficient amount of adapted sludge has not been formed. This means that the composition of each wastewater should be examined for the presence of macro-nutrients (N, P and S) and trace elements (Fe, Co and Ni seem to be particularly important). Furthermore, the possible presence of toxic or inhibitory substances in the wastewater should be monitored (Lettinga et al., 1984).

Temperature and pH are also of great importance. The temperature has a major effect on the bacterial growth rate and activity and 35°C is regarded as the optimum for mesophilic treatment. With respect to pH, it may be necessary to add a buffering agent to the influent to keep the pH at a desired level of 6.5 or above.

Under good conditions, conversion capacities of 8-10 kg COD/m^3-d should be achievable within a period of about 3 months (Lettinga et al., 1984).

It has been quoted by Lettinga et al. (1984) that unfed adapted methanogenic sludge can be preserved surprisingly well. Once an active adapted population has been obtained, the restart of a reactor will in general proceed within a few weeks without any major problems.

A very important aspect is the contact between the microorganisms and the wastewaters. The upflow anaerobic filter system can suffer from clogging (channelling) problems. In UASB reactors, channelling problems occur only at low loading rates and

Table 6.1. Positive and negative factors determining the active biomass retention of the various high-rate anaerobic wastewater treatment process under very high loading conditions (Lettinga et al., 1984)*

Factors	Upflow sludge reactors (UASB, 'IRIS', tower reactors)	Upflow anaerobic filter (AF)	Downflow anaerobic stationary fixed-film reactor	Anaerobic fixed-film expanded-bed processes (AFFEB)	Fluidized-bed systems
Re-dispersion of the sludge due to: high turbulence flotation	--- (at very high sludge loads)	-			- (?)
Disintegration of the sludge aggregates	- (?) (possibly at very high loads)				
Detachment of the bio-film		- (?)	- (?)	-	--- (at incomplete acidogenesis)
Sludge bed expansion	±				
Space occupied by the packing/carrier material		- (---) (depending on the packing)	- (---) (depending on the packing)	---- (70-80%)	-- (10-20%)
Surface area of the packing/carrier		+	++++ (100 m^2/m^3)	+++ (100 m^2/m^3)	+++++ (2000-5000 m^2/m^3)
Bed expansion required				-- (10-20%)	---- (approx. 50%)
Film thickness		+	+++	+++++	++

* It is assumed that all primary conditions for the optimal application of the various processes are met.

Table 6.2. Important features of the various high-rate anaerobic wastewater treatment systems (Lettinga et al., 1984)

Features	Upflow sludge bed reactors (UASB, 'IRIS', tower reactor	Upflow anaerobic filters (AF)	Downflow anaerobic stationary fixed-film reactor	Anaerobic fixed-film expanded-bed (AFFEB)	Fluidized-bed systems
Rate of start-up:					
first start-up	4-16 weeks	> 3-4 weeks	> 3-4 weeks	> 3-4 weeks	Approx. 3-4 weeks
secondary start-up	0-2 days	0-2 days	A few days?	A few days?	Uncertain
Performance with respect to the removal and the stabilization of SS	Satisfactory at low and moderate loading rates	Fairly good at low SS concentration and when the filter is not clogged	Very poor	Rather poor	Very poor
Risk of channelling	Small, unless a poor feed inlet distribution system was installed	Great at high SS concentration and in clogged filters	Small	Small	Almost non-existent
Extent of effluent recycle required	Generally not required	Generally not required	Slight	Moderate	High recycle factor generally required
Sophisticated feed inlet distribution system required	For low-strength wastes and with dense sludge beds	Presumably beneficial	Not	Necessary	Essential
Gas-solids separator device required	Yes, essential	Could be beneficial	Not required	Could be beneficial	Beneficial
Carrier packing required	Can be beneficial in specific cases	Essential	Essential	Essential	Essential
Height-area ratio	Can be fairly high for granular sludge beds	Moderate	Moderate	Moderate (?)	Very high

when a poor feed-inlet distribution system has been installed in the reactor. In fluidized-bed reactors, a good contact between micro-organisms and wastewater is guaranteed provided that a sophisticated feed-inlet distribution system is installed.

On the other hand fluidized bed reactors require a high recycle factor, which may result in a distinct drop in substrate utilization rate by the active biomass because of the relatively low substrate levels prevailing in the reactor.

In attached film processes the maximum sludge retention depends mainly on the surface area for sludge attachment, the film thickness, the space occupied by the carrier material and the extent to which dispersed sludge aggregates are retained. As indicated in Table 6.1, there are considerable differences between the various fixed-film processes.

In upflow anaerobic filters and downflow anaerobic stationary fixed film reactors the voidage of the packing material is a factor of prime importance with regard to sludge retention (Lettinga et al., 1984).

Lettinga et al. (1984), summarized their qualitative evaluation of the four reactor systems with respect to the maximum achievable loading rates for mainly soluble wastes, with the following diminishing sequence:

Granular sludge UASB ----------→ Fluidized/Expanded bed Systems ------------→ Upflow Anaerobic filter --------→ Downflow stationary fixed film reactor

In summary, the four reactors each have distinct advantages and disadvantages. The selection of any of the four depends to a large extent on the wastewater which is to be treated and such factors as strength, composition, presence of toxic materials, inert materials, variation in loadings, etc. With more operational experience, the selection process for these reactor systems will become easier.

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