21. ANAEROBIC WASTEWATER TREATMENT PROCESSES

21.1 Background

Anaerobic biological treatment is well understood and used frequently as anaerobic digesters to treat complex organic solid wastes such as primary and secondary wastewater sludges. However, it has not been used much in the past to treat low strength organic wastewaters from industrial and domestic applications. Aerobic processes were preferred for treatment of these wastewater streams because they are easy to operate and can tolerate process fluctuations. In comparison, anaerobic reactors were assumed to be less stable under fluctuations, more expensive to install and require long start-up time. This belief was due to limited knowledge of the process and reactor design.

Now the technology advances have significantly reduced the historical weakness of anaerobic treatment. With the work of Young and McCarty in the year 1969, application of anaerobic process for the treatment of industrial and municipal wastewaters has gradually increased in last three decades. Today the anaerobic treatment has emerged as a practical and economical alternative to aerobic treatment due to significant advantages over aerobic treatment.

21.2 Anaerobic Degradation of Organic matter

The factors that determine the removal efficiency of biodegradable organic matter are:

1. The nature and composition of the organic matter to be removed
2. Suitability of environmental factors
3. Sludge retention time in the reactor
4. The intensity of mixing, hence contact between bacterial biomass and organic matter.
5. Specific loading of organic matter with respect to bacterial sludge mass, and retention time.

Factors (1) and (2) are basically dependent on wastewater characteristics, whereas (3) to (5) are related to the type and design of the treatment system. The transformation of complex macromolecules of organic matter present in wastewater into biogas requires several groups of microorganisms. The reaction sequence of the anaerobic digestion of complex macromolecules is presented in Figure 21.1 [Gujer and Zehnder, 1983]. Different steps are necessary for the anaerobic digestion of proteins, carbohydrates, and lipids. Four different phases can be distinguished in the overall conversion process of organic matter to biogas as 1) Hydrolysis, 2) Acidogenesis, 3) Acetogenesis, and 4) Methanogenesis.

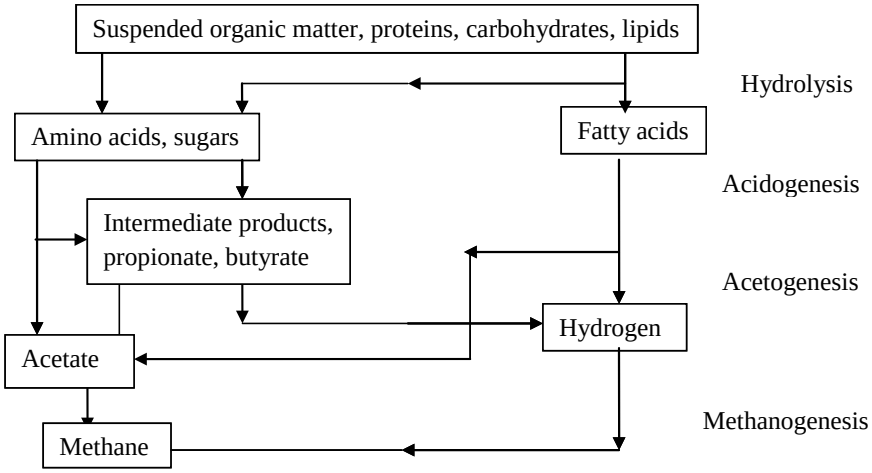

Figure 21.1 Reaction Sequences for the Anaerobic Digestion of Complex Organic Matter

21.3 Overview of Anaerobic Degradation Process

The anaerobic biological conversion of organic waste to methane is a complex process involving a number of microbial populations linked by their individual substrate and product specificities. The overall conversion process may be described to involve direct and indirect symbiotic association between different groups of bacteria. The product of one bacterium is often the substrate for others and hence, a balance between the bacterial numbers and the substrate concentrations must be maintained. The biological conversion of organic matter occurs in three steps. The first step in the process involves transformation of higher-molecular-mass compounds into compounds suitable for use as a source of energy and cell carbon (hydrolysis). The second step (acidogenesis) involves the bacterial conversion of the compounds resulting from the first step into identifiable lower-molecular-mass intermediate compounds. Lower chain volatile fatty acids produced during acidogenesis are utilized by a group of bacteria (acetogens) to produce acetate. The third step (methanogenesis) involves the bacterial conversion of the intermediate compounds into simpler end products, such as methane and carbon dioxide. Several nomenclatures have been proposed for these three steps. Speece and McCarty (1962) called the first and the second steps the constant BOD phase and the third step, the reducing BOD phase, because only the methane formation in the third step brings about the reduction of BOD or COD through the whole process.

According to trophic requirements the bacteria involved can be conveniently divided into three groups as follows.

Hydrolytic bacteria - acidogens: These bacteria hydrolyze the substrate (macromolecule) into short-chain organic acids and other small molecules, which can be taken up and converted into soluble short-chain organic molecules, e.g., carbohydrates are converted into low-chain fatty acids, alcohols, hydrogen and carbon dioxide under anaerobic condition. Strict anaerobes are composed most part of this group of bacteria. The generation time of these bacteria is 2 to 3 hours. The principle intermediate compounds resulting from conversion of the substrate during acid fermentation are acetate (CH_3COOH), propionate (CH_3CH_2COOH), butyrate ($CH_3CH_2CH_2COOH$), hydrogen gas (H_2), carbon dioxide (CO_2), lactate ($CH_3CHOHCOOH$), formate ($HCOOH$), ethanol (CH_3CH_2OH), valeric acid ($CH_3CH_2CH_2CH_2COOH$), isovaleric acid ($(CH_3)_2CHCH_2COOH$), and caproic acid ($CH_3CH_2CH_2CH_2CH_2COOH$). The distribution of final product depends on the species of acidogenic bacteria and on the environmental conditions such as pH and temperature.

Obligate Hydrogen Producing Acetogens (OHPA): This group converts compound formed in the first stage into acetic acid and hydrogen. Low hydrogen pressure favours these reactions [Harper and Pohland, 1986].

e.g. Propionate $CH_3CH_2COOH + 2H_2O \rightarrow CH_3COOH + CO_2 + 3H_2$

$\Delta Go = 76.1$ KJ/mole

Butyrate $CH_3CH_2CH_2COOH + 2H_2O \rightarrow 2CH_3COOH + 2H_2$

$\Delta Go = 48.1$ KJ/mole

From the viewpoint of the thermodynamics, a negative value of free energy change is necessary for any reaction to proceed without input of external energy. This theory apparently suggests that hydrogen producing acetogenic bacteria cannot obtain energy for growth from these reactions. However, the value of free energy change in the actual environment surrounding the bacteria, $\Delta G'$, is different from that of $\Delta Go'$ and depends on the concentrations of substrates and products as follows [Harper and Pohland,1986]:

$$\Delta G' = \Delta Go' + RT_A \ln \frac{[P_1].[P_2]....}{[S_1].[S_2]....} \quad ....(1)$$

Where,

$\Delta G'$ = free energy change at pH = 7 (kJ/mol),

$\Delta Go'$ = standard free energy change at pH = 7 (kJ/mol),

R = gas constant = 0.082 L.atm/mol.^oK,

T_A = temperature (^oK),

$[P_1].[P_2]...$ = product concentration (mol/L or atm), and

$[S_1].[S_2]...$ = substrate concentration (mol/L or atm).

Only low partial pressure of hydrogen can give negative values of $\Delta G'$ in above equations, because substrate concentration cannot be so high and acetate concentration is not so low in anaerobic reactors. This shows that extremely low partial pressure of hydrogen is essential for hydrogen producing acetogenic bacteria although they themselves produce hydrogen. Experimentally it was found that the hydrogen partial pressure higher than 5×10^{-3} atm ceased the degradation of propionate by hydrogen producing acetogenic bacteria [Hanaki et al., 1985]. Based on thermodynamics associated with this reactions Harper and Pohland [1985] indicated that propionic acid oxidation to acetate becomes favourable only at hydrogen partial pressure below 10^{-4} atm, while, butyric acid oxidation becomes favourable at 10^{-3} atm H_2 or below.

Hydrogen utilizing methanogenic bacteria can serve such a thermodynamically favourable conditions for hydrogen producing acetogenic bacteria in anaerobic reactors, thus, the activity of hydrogen producing acetogenic bacteria depends on the existence of methanogenic bacteria. Hydrogen utilizing methanogens receive hydrogen as a substrate from hydrogen producing acetogenic bacteria. The interrelationship between these two groups of bacteria is called interspecies hydrogen transfer, which also exists between acidogenic and methanogenic bacteria. Acidogenic bacteria produce more hydrogen and acetate than propionate or lactate and obtain more energy under low hydrogen partial pressure which is kept by methanogenic bacteria. The interspecies hydrogen transfer is favourable but not essential for acidogenic bacteria, while, it is indispensable for hydrogen producing acetogenic bacteria.

Methanogenic bacteria - methanogens: These bacteria produce methane. The doubling time of these bacteria is 2 - 10 days. These are further divided into two groups as:

a) Hydrogen utilisers (lithotrophs)

$$CO_2 + 4H_2 \rightarrow CH_4 + 2H_2O \text{ convert ADP to ATP}$$

b) Acetic acid users (acetotrophs)

$$CH_3COOH \rightarrow CH_4 + CO_2 \text{ produce 0.25 mole of ATP}$$

The methane producing bacteria are strict anaerobes which are extremely sensitive to changes in temperature and pH. These bacteria are active in two temperature zones, namely, in the mesophilic range (30^oC - 35^oC) and in the thermophilic range (50^oC - 60^oC). However,

anaerobic processes have been operated at 15^{o}C successfully when sufficient residence time for these bacteria was provided.

The majority of methanogens in anaerobic wastewater treatment and natural anaerobic environment utilize hydrogen and single carbon compounds as substrates for methane production. In addition there are two known genera of methanogens which can utilize the two-carbon compound, acetic acid. These include species of *Methanosarcina* and *Methanothrix (Methanosaeta*). The ***Methanothrix species*** are unable to use hydrogen in combination with CO_2 and these are **non-hydrogen-oxidizing acetotrophs (NHOA)**. In contrast, ***Methanosarcina*** can utilize **H_2/CO_2 as well as acetate**, carbon monoxide, methanol, and methylamines as growth substrates. Due to their ability to use both H_2/CO_2 and acetate, these bacteria are classified as **Hydrogen Oxidizing Acetotrophs (HOA)**. **Hydrogen-Oxidizing Methanogens (HOM)** do not cleave acetate, but utilize H_2/CO_2 and formate as substrates [Harper and Pohland, 1986]. The HOA are unique in their capability to utilize multiple (one and/or two carbon) substrates. This ability affords a higher potential for survival when competing with sulfur reducing bacteria (SRB) and nitrate reducing bacteria (NRB) for hydrogen and acetate. At hydrogen partial pressure $>10^{-4}$ atm, HOA use H_2/CO_2 in favour of acetate, whereas acetate cleavage by NHOA is unaffected by hydrogen.

NHOA have a much higher affinity for acetate than the HOA. NHOA may outcompete HOA at acetate concentrations below 50 mg/L, while above 250 mg/L acetate, the HOA are more competitive [Speece et al., 1983]. As a result of this comparative kinetics *Methanothrix* (NHOA) may be found in reactors with lower organic loading. *Methanosarcina* are more predominant in low retention time reactors such as in the lower reaches of plug flow anaerobic filters and in two phase reactor system.

Oxidation of reduced organic products to bicarbonate and acetate also occurs due to NRB and SRB. Higher organic waste conversion rates may be available through SRB than through methanogenesis. Moreover, SRB and NRB are not limited to one-and two-carbon substrates, as are methanogens. However, from process engineering perspective, such an approach has disadvantages, including the loss of energy available from methane and the production of hydrogen sulphide or ammonia. Since, sulphide and ammonia are much more soluble than methane, their dissolved components can contribute significantly to effluent COD [Harper

and Pohland, 1986]. However, this approach may hold possibilities for reducing propionic acid and hydrogen, as well as acetic acid in a stressed reactor, in order to more rapidly reestablish the equilibrium with the existing hydrogen removal system.

21.4 Factors Affecting Anaerobic Digestion

Development of anaerobic process technology is dependent on a better understanding of the factors that are associated with the stability of the biological processes involved. Process instability is usually indicated by a rapid increase in the concentration of volatile acids in the first stage of the reaction. Low pH with a concurrent reduction in methane gas production indicates the methanogenesis more susceptible to upset. Acclimatization of the microbes to a substrate may take 3 to 8 weeks. Sufficiently acclimated bacteria show greater stability towards stress-inducing events such as hydraulic overloads, fluctuations in temperature, volatile acid and ammonia concentrations, etc. Several environmental factors can affect anaerobic digestion such as specific growth rate, decay rate, gas production, substrate utilization, etc. The environmental factors of primary importance are discussed below.

21.4.1 pH, Acidity and Alkalinity

Methanogenic microorganisms are susceptible to the minute changes in the pH values. Optimum pH range of 6.6 – 7.6 is considered favourable for the methane producing bacteria, which cannot tolerate the fluctuations. The non-methanogenic bacteria do not exhibit such strong sensitivity for environmental conditions and are able to function in a range of pH form 5 – 8.5. The pH maintained inside the reactor, due to the process results from the interaction of the carbon dioxide-bicarbonate buffering system and volatile acids-ammonia formed by the process. It is necessary to prevent the accumulation of acids to a level, which may become inhibitory to the methanogenic bacteria. For this, it is important that there should be sufficient buffering capacity present in the reactor, which may prevent the reactor from souring. Although, the carbonates and bicarbonates of sodium and calcium are required to be added to the digesters to provide buffering action, lime (Calcium hydroxide) is most commonly used for this purpose. Only the unionized volatile acids in the concentration range of 30 - 60 mg/L are toxic.

21.4.2 Temperature

As in all biological processes, anaerobic processes are affected by temperature. The higher the temperature, higher is the microbial activity until an optimum temperature is reached. A

further increase of the temperature beyond its optimum value results in steep decrease in activity. Anaerobic process can take place over a wide range of temperatures ($4 – 60^0$C). Once as effective temperature is established, small fluctuations can result in a process upset. Although most of the sludge digester are operated in the mesophilic range ($30 – 40^0$C), methanogenesis can occur at temperatures as low as 12 to 15^0C. The effect of increasing temperature on biochemical reaction rate in the range of $4 – 25^0$C is profound.

The optimum temperature for growth of anaerobic microorganisms is 35^0C or greater. Although anaerobic digesters have been reported to operate at substantially lower temperatures, such as 20^0C, anaerobic growth under these temperature conditions is slow requiring prolonged start-up time and difficulties in operation. In situations where reactor's operating temperature is low, start-up will be benefited if initiated at approximately 35^0C. At temperature of less than 25^0C, the digestion rate decreases sharply and conventional anaerobic reactors in operation at ambient temperatures in cooler climates may require detention times of as much as 12 weeks for the treatment of sewage sludges.

The majority of industrial digester systems operate in the mesophilic range of $30 – 40^0$C. It is probable that increase in microbial reaction rates at the elevated temperatures of thermophilic processes ($50 – 60^0$C), and hence decrease in SRT may prove advantageous under some circumstances. However, lack of stability in thermophilic municipal waste treatment can occur. Thermophilic digestion is most practical where wastewater stream to be treated is discharged at an extremely high temperature and the digester is present on site.

In psycrophilic, mesophilic, or thermophilic ranges, uniformity of temperature over the entire vessel contents is of paramount importance to anaerobic digestion. Temperature change of even a few degrees can result in a marked upset in microbial metabolism and rapid alterations in reactions in the reactor and may necessitate several days for the recovery. A consistent temperature throughout the system can be provided by adequate mixing of the reactor by paddle, gas sparging, or flow over heat exchangers.

21.4.3 Nutrients

Anaerobic wastewater treatment processes are often used for industrial waste with only minor amount of nutrients present. This might result in nutrient deficiency, unless additional nutrients are supplemented. Often the COD/N ratio and COD/N/P ratio is used to described

the nutrient requirements. Optimum N/P ratio can be considered to be 7. The theoretical minimum COD/N –ratio is considered to be 350/7. A value around 400/7 is considered reasonable for high rate anaerobic processes (0.8 – 1.2 kg COD /kg VSS.d). For low rate processes (<0.5 kg COD /kg VSS.d) the COD/N-ratio has been observed to be increased dramatically to values of 1000/7 or more [Van den Berg and Lentz, 1980].

Other than nitrogen and phosphorous, trace metals also are essential for anaerobic processes. The presence of trace metals such as molybdenum, selenium, tungsten and nickel is probably necessary for the activity of several enzyme systems. When these trace elements are not present in the wastewater, addition of nickel, cobalt, and molybdenum can increase methane production and allow greater volumes of wastewater to be effectively treated by decreasing the reactor residence time.

21.4.4 Inhibitory Substances

Inhibition of the anaerobic digestion process can be mediated to varying degrees by toxic materials present in the system. These substances may be components of the influent wastewater or byproducts of the metabolic activities in the digester. Inhibitory toxic compounds include sulphides, consequential in the processing of waste from sources such as molasses fermentation, petroleum refining and tanning industries. Volatile acid and other microbial products can accumulate and inhibit reactor-buffering capacity. Inhibition may also arise as the consequence of the increased levels of ammonia, alkali, and the alkaline earth metals, and heavy metals in the system.

Volatile Acids Inhibition: Anaerobic reactor instability is generally evident by a marked and rapid increase in volatile fatty acids concentrations; this is frequently indicative of the failure of the methanogenic population due to other environmental disruptions such as shock loadings, nutrient depletion or infiltration of inhibitory substances. Acetate has been described as the least toxic of the volatile acids, while propionate has often been implicated as a major effecter of digester failure.

The inhibition by the volatile acids at acidic pH values can be attributed to the existence of unionized VFAs in significant quantities in the system. The undissociated nature of these acids allow them to penetrate the bacterial cell membrane more efficiently than their ionized counterparts, and once assimilated, induce an intracellular decrease in pH and hence a decrease in microbial metabolic rate. The resulting VFA concentration in the reactor should

be maintained below 500 mg/L at any point of time and preferably below 200 mg/L for optimum performance.

Ammonia – Nitrogen Inhibition: Although ammonia is an important buffer in anaerobic processes, high ammonia concentration can be a major cause of operational failure. In reactor system that has not previously been acclimated to high ammonia loadings, shock loadings of high ammonia concentration generally caused rapid production of VFAs such that the buffering capacity of the system may not be able to compensate for the decrease in pH. Further depression of alkalinity and reduction of pH may result in reactor failure. Inhibition is indicated by a decrease in gas production and an increase in volatile acid formation.

Sulphide Inhibition: The sulphate and other oxidized compounds of sulphur are easily reduced to sulphide under the conditions prevalent in anaerobic digesters. Sulphur-containing amino acids of protein can also undergo degradation to sulphide. These compounds are of significance when anaerobic treatment is considered for industrial processes which tend to produce large quantities of sulphides in their waste stream. These sulphides formed by the activity of reactor microorganisms may be soluble or insoluble, depending upon their associated cations. When the salts formed are insoluble, they have negligible effects on digestion. Iron addition, for example can suppress sulphide inhibition by removing S^{2-} ion from solution by precipitation.

Desulfovibrio and other sulphate-reducing genera form sulphides from sulphates and some of the fermentative microorganisms utilize the sulphur containing amino acids to produce sulphides. Sulphide concentration in excess of 200 mg/L in a digester at 35 ^{0}C, with continuous feeding and mixing, produced severe inhibitory effects including the complete cessation of gas production [Parkin and Speece, 1983]. All the heavy metals, with the exception of chromium, form insoluble sulphide salts and thus can be removed from solution by sulphide present in the system by precipitation. Free sulphide can also be eliminated as hydrogen sulphide by vigorous gas production.

Heavy Metals Inhibition: The most common agents of inhibition and failure of sewage sludge digesters are identified to be contaminating heavy metals. Heavy metals in the soluble state are in general regarded to be of more significance to reactor toxicity than are insoluble forms. Anaerobic digestion also reduces the valence states of some heavy metals. Both

copper and iron may be reduced from the trivalent to divalent state. This reduces the quantity of the precipitating agent, such as sulphide, necessary for the removal of the metal ion from solution. The heavy metals can be removed from anaerobic systems by adsorption. Those digesters, such as the CSTR configuration; which are tolerant to the wastes containing high levels of suspended solids are effective in metal removal, provided sufficient adsorption sites are present. The metals like copper, chromium, nickel, lead can induce toxicity in the reactor when present in higher concentration, and acceptable concentration in the wastewater to be treated differs from metal to metal.

21.5 Merits of Anaerobic Decomposition Process

It has been recognized that the anaerobic treatment is in many ways ideal for wastewater treatment and has several merits mentioned as below:

- A high degree of waste stabilization;
- A low production of excess biological sludge and this sludge can be directly dried on sludge drying bed without further treatment due to better dewatering ability;
- Low nutrient requirements, hence anaerobic treatment is attractive for the treatment of wastewater where external nutrient addition is required;
- No oxygen requirement, hence saving in power required for supply of oxygen in aerobic methods;
- Production of valuable byproduct, methane gas;
- Organic loading on the system is not limited to oxygen supply hence higher loading rate as compared to aerobic processes can be applied.
- Less land required as compared to many aerobic process.
- Non-feed conditions for few months do not affect adversely to the system and this makes it attractive option for seasonal industrial wastewater treatment.

Quantity of biological solids produced in the anaerobic systems per unit weight of organic material is much less than that in aerobic systems. This is a major advantage of the anaerobic process as the quantity of sludge for ultimate disposal is reduced. This is a result of conversion of volatile solids present, to the high energy level end products such as methane, carbon dioxide and water. Methane has a definite economic value as a fuel, and it is used as a source of energy for both heat and power in many installations.

Another major advantage is the loading potential. Aerobic processes are restricted in maximum organic loading rate by the inability to transfer oxygen at the rate sufficient to

satisfy the oxygen demand of the systems. The stabilized sludge from anaerobic process may be free from strong or foul odours and can be used for land application as ultimate disposal because the digested sludge contains sufficient nutrients required for plants. Pathogens are also destroyed to a high degree during the thermophilic anaerobic process. Due to large retention time and consequent low growth rate, the cell yield is also extremely low; thus, most of the carbon in the waste is available for methanogenesis and under normal circumstances the yield of methane would, on an average, be 0.33 – 0.36 m^3 per kg COD utilized at 35^oC and atmospheric pressure.

However, anaerobic treatment processes are not largely being implemented, because of many factors. Anaerobic microorganisms, especially methanogens have slow growth rate. At lower HRTs, the possibility of washout of biomass is more prominent due to higher upflow velocity. This makes it difficult to maintain the effective number of useful microorganisms in the system. To maintain the population of anaerobes, large reactor volume or higher HRTs with low upflow velocity are required. This may ultimately provide longer SRTs more than 40 days for high rate systems. Thus, provision of larger reactor volume or higher HRTs ultimately leads to higher capital cost. Low synthesis / reaction rate hence, long start-up periods and difficulty in recovery from upset conditions are some of the notable disadvantages. Special attention is therefore required towards controlling the factors that affect process adversely; importantly among them being environmental factors such as, temperature, pH, and concentration of toxic substances. Hence, skill supervision is required for operating anaerobic reactor at optimal performance.

21.6 Anaerobic Waste Treatment Processes

Advantage of anaerobic waste treatment systems as means for recovery of non-conventional energy is increasingly being recognized worldwide. Anaerobic decomposition is a biologically mediated process, indigenous to nature, and capable of being simulated for treating wastes emanating from municipal, agricultural, and industrial activities. Anaerobic digestion as applied in treatment of sewage sludge and other organic wastes, represents the controlled application of a process. Although, anaerobic digesters have traditionally been used for many decades in the stabilization of sewage sludge, their successful and economic employment for the treatment of liquid wastes is only a recent phenomenon, arising from the development of new reactor designs. These concepts have led to development of various reactors, which are capable of retaining a much higher biomass concentration than traditional

digesters. Making the sludge retention independent of the influent retention time makes this possible. The technological approaches to allow this condition of independent sludge retention time can be divided in to the following:

- Attachment of biomass on the media (filters, fluidized systems, and RBC configurations);
- Non-attached biomass concept as suspended growth process (sludge blanket reactors and contact process with sludge recycling).

It is difficult to evaluate the advantages and disadvantages of each system in relation to other concepts, as generalizations are not usually valuable in practice. Considerations such as, purification rates, loading rates, investment cost, energy balance, space requirements, operational costs, and specific long term experience with certain wastewaters are all important, but they will be valued differently from industry to industry. Various anaerobic reactor types in practice are summarized in Table 21.1.

Table 21.1 Basic Types of Reactors used in Anaerobic Process

Type of reactor	Synonyms	Abbreviations
Attached Biomass		
Fixed Bed	Fixed film, filter, submerged filter, /stationary fixed bed	SMAR (Submerged anaerobic reactor)/ ANFIL (upflow anaerobic filter), AUF (Anaerobic upflow filter), ADSR (Anaerobic downflow stationary bed reactor), AF (anaerobic filter), DSFF (downflow stationary fixed film reactor)
Moving Bed	Rotting discs, Rotating biological contactor,	AnRBC, RBC
Expanded bed	Anaerobic attached film expanded bed	AAFEB
Fluidized bed	Anaerobic Fluidized bed reactor, Carrier-assisted contact process	FBBR / IFCR (Immobilized fluidized cells reactor)/ CASBER (Carrier-assisted sludge bed reactor)
Non-attached Biomass		
Recycled flocks, Sludge blanket, Digester	Contact process, Upflow anaerobic sludge blanket reactor (UASB), Upflow sludge blanket (USB), Clarigester type	UASB, USB

(Reference: Henze and Harremoes, 1983)

In suspended growth systems microbial cultures are freely suspended in the reactors. Microorganisms are suspended in the wastewater, as a single cell or as their agglomerates.

Various reactor types developed on the principles of suspended growth are discussed hereunder. The anaerobic contact process essentially involves two phases. A contact phase, where the raw waste is intimately mixed with a previously developed and available anaerobic sludge culture in the reactor; and a separation phase, where the active sludge particles are separated from the treated liquor and recycled to the contact unit. In this process, raw wastewater is mixed with recycled sludge solids and then digested in a digestion chamber. After digestion, the mixture is separated in a clarifier (or vacuum flotation unit), and the supernatant is discharged as effluent. Settled sludge is then recycled to seed the incoming waste.

The loading rates permissible in an anaerobic waste treatment process are primarily dictated by the sludge retention in an anaerobic reactor. The maintenance of high Sludge Retention Time (SRT) has been the major problem in the practical application of the process, especially for waste with Chemical Oxygen Demand (COD) below about 3000 mg/L. Obviously, a waste treatment process for low-strength wastes is an economical one if large volume of waste can be forced through the system in a relatively short time period. For this purpose process are required in which the biomass retention time can be controlled independently of the wastewater flow rate. Conventional anaerobic treatment processes of the flow-through type are therefore inadequate to treat low-strength wastes.

Advances in the understanding of how anaerobic system function, improved understanding of mixing and mass transfer, and anaerobic reactor design, has led to the evolution of a new generation of high-rate anaerobic processes *e.g.* Anaerobic Filters (AF), Anaerobic expanded / Fluidized bed rectors, Upflow Anaerobic Sludge Blanket (UASB) Reactor, *etc.* These systems have been schematically presented in Figure 21.2 and Figure 21.3.

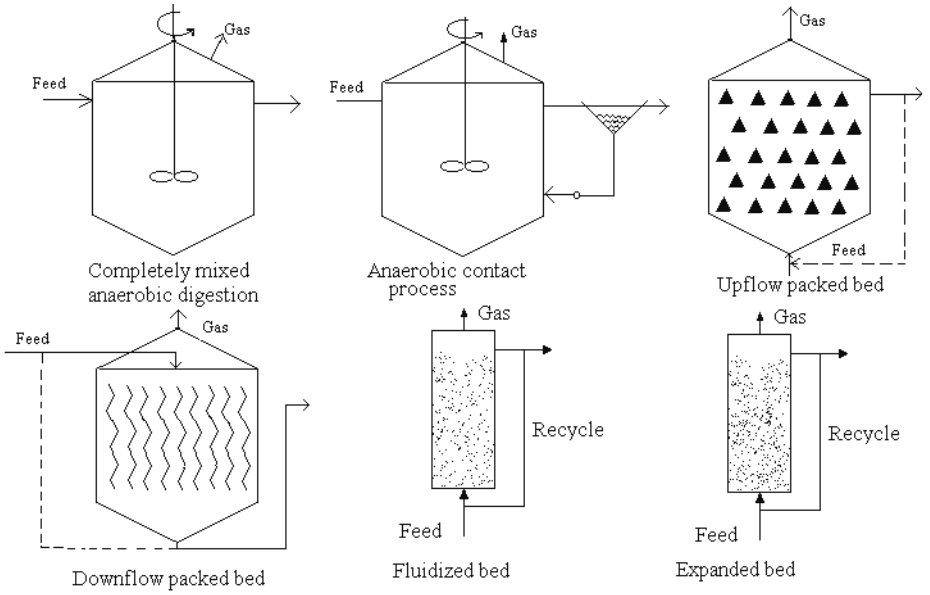

Figure 21.2 Typical reactor configurations used in anaerobic wastewater treatment

One common feature offered by all the high-rate processes is their ability to provide high SRT in relation to hydraulic retention time (HRT). High biomass concentration is maintained in a reactor with relatively low treatment time. In anaerobic filter and expanded / fluidized beds, this is accomplished by development of bio-film on support surface. In UASB systems, this is accomplished by the development of granules or flocs that have extremely good settling properties. Among the other improved high rate anaerobic treatment methods, UASB has secured an important place.

21.6.1 Anaerobic Filter Systems

The anaerobic filter (AF) and more recent variation of the filter process, the downflow stationary fixed film (DSFF) reactor, are packed with a fixed support media. In the DSFF reactor, the biomass is present as a biofilm attached to the support media. In the anaerobic filter, most of the biomass is present as suspended and/or entrapped biomass in the interstitial pore volume of the support media. The other major difference between these reactors is the direction of liquid flow through the packing. In the AF, the feed enters at bottom of the reactor. In the DSFF reactor, influent is applied in downward direction from the top of the rector. Both reactors can be used to treat either diluted or concentrated soluble wastewaters. Because of the relatively large clearance between the channels in the vertically oriented media used, the DSFF reactor is able to treat wastewater with relatively high suspended solids while the AF cannot.

Because the bacteria are retained on the media and not washed off in the effluent, mean cell residence times of the order of 100 days can be obtained. Large values of θc can be achieved with short hydraulic retention times, so the anaerobic filters can be used for the treatment of low-strength wastewater at ambient temperature. In the AF, most of the biological activity is due to the biomass in suspension (entrapped) rather than to the attached biofilm. The media with a high capacity to entrap and prevent washout of the biomass from the reactor is more important than the specific surface area (surface area to volume ratio) of the media. The biofilm thickness of 1 to 3 mm has been observed in fixed-bed reactor.

In the DSFF reactor systems, virtually all of the active biomass is attached to the support media. Different types of support media such as needle punched polyester (NPP) and red drain tile clay, PVC or glass can be used. For NPP, this attachment is probably associated with its surface roughness. The leaching of minerals from the clay could potentially stimulate bacterial activity and adhesion to this media support.

Selection of proper inoculum source is important to obtain rapid reactor start-up and minimize the time required for the initial biofilm establishment. Usually a bacterial flora adapted to the target wastewater should be used. In general, the volume of inoculum used should at least 10% (v/v) to obtained good result. During the start-up period the initial organic load applied should be maintained at levels less than 0.1 kg COD /kg VSS.d and HRT greater than 1 day should be maintained to prevent wash out of the inoculated biomass. Typical organic loading rates generally between 1.0 and 10 kg COD /m^3.d can be applied with 75 – 85 % removal efficiency. The hydraulic retention time is generally kept in the range of 18 to 24 h, but lower HRT values can also give fairly good removal efficiency depending on the type of organic matter present in the wastewater.

21.6.2 Expanded Bed Process

In the expanded bed process, the wastewater to be treated is pumped upward through a bed of appropriate medium (*e.g.* sand, coal, expanded aggregate, plastic media) on which a biological growth has been developed. Effluent is recycled to dilute the incoming wastewater and to provide an adequate flow to maintain the bed in an expanded condition. Biomass concentrations exceeding 15,000 to 40,000 mg/L can be developed. Since more biomass can be maintained, the expanded bed process can also be used for the treatment of low strength

wastewater, such as municipal sewage, at very short hydraulic retention times. Organic loading in the range of 5 to 10 kg COD / m^3.d can be applied with COD removal efficiency of 80 to 85 %. The hydraulic retention time generally will be in the range of 5 to 10 hours.

21.6.3 Anaerobic Contact Process

The essential feature of the anaerobic contact process is that the washout of the active anaerobic bacterial mass from the reactor is controlled by a sludge separation and recycles system. The major problem in the practical application of the contact process has always been the separation (and concentration) of the sludge from the effluent solution. For this purpose several methods have been used or were recommended for use, *e.g.* plain sedimentation, settling combined with chemical flocculation, with vacuum degasification, floatation and centrifugation. A basic idea underlying the contact process is that it is considered necessary to thoroughly mix the digester contents e.g., by gas recirculation, sludge recirculation, or continuous or intermittent mechanical agitation. This is generally used for concentrated wastewater treatment such as distillery wastewater.

21.6.4 Upflow Anaerobic Sludge Blanket (UASB) Reactor

It is somewhat modified version of the contact process, based on an upward movement of the liquid waste through a dense blanket of anaerobic sludge. No inert medium is provided in these systems. The biomass growth takes place on the fine sludge particles, which then develop as sludge granules of high specific gravity.

The reactor can be divided in three parts (Figure 21.3), sludge bed, sludge blanket and three phase separator (gas-liquid-solid, GLS separator) provided at the top of the reactor. The sludge bed consists of high concentration of active anaerobic bacteria (40 – 100 g/L) and it occupies about 40 to 60% of reactor volume. Majority of organic matter degradation (> 95%) takes place in this zone. The sludge consists of biologically formed granules or thick flocculent sludge. Treatment occurs as the wastewater comes in contact with the granules and/or thick flocculent sludge. The gases produced causes internal mixing in the reactor. Some of the gas produced within the sludge bed gets attached to the biological granules. The free gas and the particles with the attached gas rise to the top of the reactor. On the top of sludge bed and below GLS separator, thin concentration of sludge is maintained, which is called as sludge blanket. This zone occupies 15 to 25% of reactor volume. Maintaining sludge blanket zone is important to dilute and further treat the wastewater stream that has bypassed the sludge bed portion following the rising biogas. The GLS separator occupies

about 20 to 30% of the reactor volume. The particles that raise to the liquid surface strike the bottom of the degassing baffles, which causes the attached gas bubbles to be released. The degassed granules typically drop back to the surface of the sludge bed. The free gas and gas released from the granules is captured in the gas collection domes located at the top of the reactor. Liquid containing some residual solids and biological granules passes into a settling chamber, where the residual solids are separated from the liquid. The separated solids fall back through the baffle system to the top of the sludge blanket.

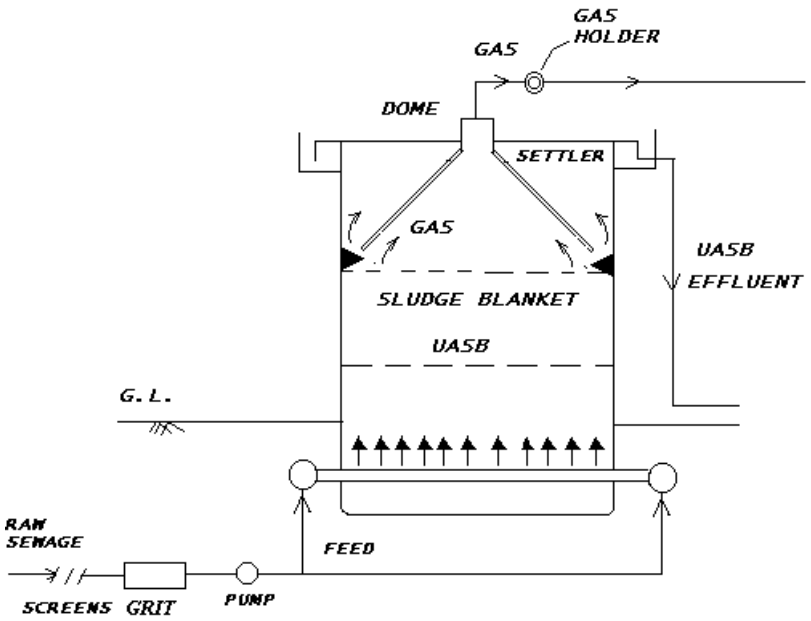

Figure 21.3 Upflow Anaerobic Sludge Blanket Reactor

The granular biomass from the existing UASB reactor can be used as inoculum material to start-up new UASB reactor. When such material is not available, non-granular material such as anaerobic digested sludge, waste activated sludge and cow dung manure can be used as inoculum. Granular sludge can be developed using non-granular material for inoculation.

Although, there are reports of wastewaters containing high-suspended solids being successfully treated in UASB reactors without primary sedimentation, the separation of suspended solids is still suggested, especially for reactors having non-granular configuration. Pretreatment such as sedimentation, neutralization of wastewater is normally desirable in treating waste in UASB reactor. Organic loading in the range of 1-20 kg COD /m^3.d can be applied with removal efficiency of 75 to 85 % and HRT of 4 to 24 h.

21.6.5 Modification of the Anaerobic Process

The efficient removal of organic matter from sewage can be accomplished by using conventional UASB reactor at mesophilic temperature *i.e.* temperature exceeding about 20 ^{o}C. For suitability of anaerobic process for wastewater treatment at lower temperature and for low strength wastewater some modification in reactor is necessary. It is possible to give treatment at lower temperature and at low strength with certain modification in conventional UASB reactor. The modification like Expanded Granular Sludge Bed (EGSB) reactor and UASB hybrid reactors are discussed below.

21.6.5.1 The EGSB reactor

The sludge present inside the UASB reactor can be either flocculent or granular form. The sludge in granular form (size 1 – 5 mm) exhibit distinct advantages over flocculent sludge form. The granular form of sludge offers maximum microorganism to space ratio due to high density, it has high settling velocity, high methanogenic activity, and excellent mechanical strength. The granular sludge form develops mainly on soluble type of wastewaters. The formation of the granules is mainly dependent on operating conditions inside rector and the characteristics of the wastewater to be treated [Ghangrekar *et.al.,* 1996]. So far, granulation has not been reported in any of the existing full-scale UASB reactor treating sewage. Granulation of biomass is reported in the laboratory scale and pilot scale UASB reactor when appropriate mixing conditions are maintained in the reactor [Bhunia and Ghangrekar, 2010]. In all cases, although flocculent sludge configuration is reported while treating raw sewage, excellent BOD and TSS removal efficiencies can be achieved.

In EGSB reactor very high upflow velocity is maintained to keep sludge bed biomass in expanded form. It was reported that, the EGSB reactor was efficient in removal of the soluble organic matter even at low temperature [Handel and Lettinga, 1994]. This can be attributed to the intensive contact between the incoming organic matter and sludge granules as a result of high upflow velocity (6-12 m/h), against less than 1 m/h in conventional UASB reactor. The EGSB reactors are useful for treatment of wastewater, particularly at low temperature and relatively low strength wastewater, when the production rate of biogas and consequently the mixing intensity induced by it are relatively low. Under these conditions the higher kinetic energy content of the influent and extended height of the expanded granular bed contribute to better performance compared with conventional UASB reactors. The EGSB reactor is inadequate for removal of particulate organic matter due to high upflow liquid velocity used.

21.6.5.2 UASB hybrid reactor

In this reactor instead of GLS separator a filter or plate settler is provided at the top of UASB reactor. Improved reactor performance can be obtained with this modification with improved sludge retention in the reactor. This reactor is taller than the UASB reactor.

21.6.5.3 Anaerobic baffle reactor

This reactor consists of 3 to 5 chambers and wastewater is allowed to flow upward direction in each compartment or upflow and downflow mode in alternate compartments. This reactor can give reliable treatment efficiency particularly for the treatment of low strength wastewater containing particulate organic matter.

21.7 Application of UASB Reactor for Wastewater Treatment

21.7.1 Suitable Wastewater Characteristics

Granulation of biomass is indicative of successful operation of UASB reactor. Although, acceptable efficiency from the reactor can be obtained when sludge is in flocculent form, existence of granulation sludge configuration offers distinct advantages. The composition of wastewater plays an important role in granulation process. Substrates that support granulation are carbohydrates or proteins mainly in soluble, and possibly in colloidal form. Industrial wastewaters from sugar industry, breweries, apple juice, yeast factory, and grape wine satisfy this criterion and give granulation in UASB reactor.

For wastewater containing mainly proteins, granulation proceeds satisfactory; but problems may arise from foaming and protein precipitation under conditions of overloading or low pH less than 6.0 [Souza, 1986]. More importantly, they release ammonia upon degradation, which may exert an inhibition effect on microbial activities. High SS concentration in influent can adversely affect granulation and performance of the reactor. The influent SS concentration shall be less than 1 g/L and SS to COD ratio shall be less than 0.5 for successful operation of the reactor [Souza, 1986].

For wastewater that contains substrates, which do not yield hydrogen in the fermentation process (short chain fatty acids), granulation will not takes place. No granulation in the UASB reactor was reported for acetate only as substrate. For the waste where H_2 generated is preferentially utilized by other organisms such as sulphate reducers, granulation is limited

because of reduced amount of H_2 available to the hydrogenotrophic methanogens e.g., paper pulping waste. However, the granular yield does not reduce to zero even when SO_4 supplementation is in excess [Russo and Dold, 1989].

Typical industries where UASB reactors are reported to be most successful for wastewater treatment are beet sugar, cane sugar, starch, breweries, dairy, tannery, food processing industries and paper and pulp. This process is also proved to be feasible for the treatment of domestic wastewater. The feasibility of this process has already been proved for this wastewater in wide COD range, from 500 to above 10,000 mg/L. The treatment is feasible under both mesophilic and thermophilic conditions but temperature above 15 ^{0}C is essential for proper treatment [Bogte et. al., 1993]. When wastewater is mostly in biodegradable form and COD is in the range of 1000 to 5000 mg/L, efficiency of COD removal of 85 to 90% can easily be achieved, with short HRT of 6 to 12 h. When the wastewater is complex, or COD is lower or higher than the above mentioned range, COD removal efficiency of 60 to 80 % can still be achieved. Once, the proper start-up of the reactor is achieved with generation of good quality of granular sludge, having good settling characteristics and activity, very high Organic Loading Rates (OLR) greater than 20 kg COD/m^3.d can be applied.

21.7.2 Modes of Operation

UASB reactor is successful for industries listed earlier, where the wastewater coming out of the industry is being continuously treated. It is also successful for wastewater treatment when mode of operation is intermittent. For example, in the case of dairy wastewater treatment, the wastewater is generated only for few hours a day, and not continuously. The process is reported to perform well even under such intermittent mode of operation. Also, this has been experienced that [Ghangrekar, 1997] the intermittent operation is useful during initial days of operation to overcome problem of sludge buoying due to poor quality of inoculum used. In case where excessive volatile acids production occurs in UASB reactor, reducing pH lower than 6.5, intermittent mode of operation could be resorted to reduce volatile acids concentration and increase pH in the reactor.

UASB reactor is also applicable for the treatment of wastewater from the industries, which are seasonal in origin, like food processing industries. Once, the primary start-up of the reactor is over, with development of good quality of granular sludge, the shutdown of the reactor is possible when the season is over. The reactor put into operation in new season takes very less time (1 to 2 weeks) for this secondary start-up, to restore its COD removal

efficiency [Ghangrekar, 1997]. For short duration of shut down less than a month reactor can capture its original COD removal efficiency within a week.

21.7.3 Treatment Flow Sheet

The typical units required for UASB type wastewater treatment plant are as follows:

1. Screening,
2. Grit removal (Optional),
3. Skimming Tank,
4. Pumping,
5. UASB reactor,
6. Gas collection system,
7. Post-treatment such as aerobic processes or settling tank, depending on the disposal mode of effluent, and
8. Sludge drying beds.

The provision of screens and grit chamber is necessary for the treatment of municipal wastewater as required in conventional wastewater treatment plants. For certain industrial wastewaters provision of screens and grit chamber may not be necessary. When the wastewater contains floating matter such as, oil, grease, soap, pieces of cork and wood, vegetable debris and fruit skins, it is advantageous to have a skimming tank to remove these materials. The presence of oil and grease, if gets adsorbed on the sludge surface, can hinder transport of metabolites and mass transfer, ultimately causing reduction in process efficiency. This may be accomplished in a separate tank or can be combined with primary sedimentation when wastewater also has high suspended solids of inorganic origin.

After the primary treatment, it is required to provide pumping unit to pump the wastewater in upward direction in UASB reactor. Location at which topography of the site suits for utilization of gravity head, choosing appropriate site for UASB reactor may not require pumping. The separate gas collection system can be provided if the gas produced is desired to use for combustion or power generation. Generally, the production of gas in UASB reactor is in the range of 0.25 to 0.35 m^3 CH_4/kg COD removed. The utilization of biogas for power generation is economical for larger treatment plants.

21.7.4 Post Treatment

The UASB reactor is an efficient process for removal of organic material and suspended solids from sewage or industrial effluents. Particularly, this process is more attractive for treatment of sewage in warm climate. However, the UASB reactor can hardly remove macronutrients (nitrogen and phosphorous), and pathogenic microorganisms are only partially removed. Hence, depending on the final disposal of the effluent quality, post-treatment may be required for removal of suspended solids, organic matter, nutrients, and pathogens present in the raw wastewaters.

After UASB reactor, some form of post treatment is generally desirable depending upon source of effluent discharge. UASB reactor can hardly remove any nitrogen from the wastewater. Hence, effluent from UASB reactor is suitable for irrigation purposes. UASB reactor when followed by post treatment such as aeration and/or sedimentation could conveniently achieve irrigation standards. The aeration can be obtained to the effluent flowing through a channel to an irrigation area. Where, the treatment efficiency is adequate to meet the discharge standards, further treatment such as, aeration is only necessary to destroy anaerobicity. In such cases simple cascade type aerator can serve the purpose. In some cases where treatment efficiency is meeting the discharge standards for organic matter but the effluent is high in suspended solids, the use of secondary settling tank becomes essential.

When stricter effluent standards have to be met (as for river discharge) some better form of post-treatment may become necessary. The use of aerobic biological treatment is generally preferred for this polishing treatment. The aerobic process such as, biotowers, conventional activated sludge process, or extended aeration can be employed as a second stage treatment. Where the effluent from UASB reactor is expected to be high in nutrient such as nitrogen and phosphorous, the post treatment need to be designed for removal of these nutrients to meet discharge standards for surface water. Shallow oxidation ponds can also be used after UASB reactor for complete treatment of wastewater.

21.8 Design Procedure for UASB Reactor

The UASB reactor can be designed as circular or rectangular. Modular design can be preferred when the volume of reactor exceeds about 400 m^3. It is necessary to select proper range of operating parameters for design, such as OLR, SLR, superficial liquid upflow velocity

(referred as liquid upflow velocity), and HRT. The literature recommendations for all these parameters and design procedure to account these recommendations are given below.

21.8.1 Organic Concentration and Loading

For COD concentration in the range 2 to 5 g/L, the performance of the reactor depends upon the loading rate and is independent of influent substrate concentration. For COD concentration greater than 5 g/L, it is recommended to dilute the wastewater to about 2 g COD/ L during primary start-up of the reactor. Once, the primary start-up of the reactor is over with granulation of sludge, loading rates can be increased in steps to bring the actual COD concentration of the wastewater. The loading above 1 - 2 kg COD/ m^3.d is essential for proper functioning of the reactor. For primary start-up the optimal loading rates for getting high COD removal efficiency (about 90%) within short start-up time, coupled with generation of good quality granular sludge, are OLR between 2.0 and 3.6 kg COD/ m^3.d and SLR between 0.15 and 0.25 kg COD/ kg VSS.d (Ghangrekar et al., 1996). The OLR to be used for design of UASB reactor for different temperature is provided by Lettinga and Hulshoff (1991). In general, for temperature between 15 and 35^0C, the reactor can be designed for loading between 1.5 to 18 kg COD/ m^3.d. Lower OLR should be preferred for low temperature and higher OLR can be adopted for high temperature.

For sewage treatment, the design of reactor at higher loading rate is not possible due to limitations of upflow velocity, and maximum loading of about 2 to 3 kg COD/m^3.d can be adopted for design. Similarly, for high strength wastewater, such as distillery, satisfying minimum velocity criteria and maximum HRT limit is difficult. Therefore, categorization of wastewater based on COD concentration is necessary for generalizing the design procedure of UASB reactor to meet the recommended operating conditions to the maximum extent. Thus, the COD concentration of the wastewater is suitably divided in four categories. It has been proposed to adopt loading conditions as recommended in the Table 21.2, for design of UASB reactor depending on the average COD concentration of the raw wastewater. These loading rates recommended are suitable for temperature about 30^oC. For higher temperature, the loading rates can be slightly increased and for low temperature these design loading rates can be reduced.

Table 21.2 Recommended loading range for design of UASB reactor based on COD concentration at average flow

Category of wastewater	COD concentration, mg/L	OLR, Kg COD/ m^3.d	SLR, Kg COD/ kg VSS.d	HRT, Hours	Liquid Upflow Velocity, m/h	Expected Efficiency, %
Low strength	Up to 750	1.0 - 3.0	0.1 - 0.3	6 – 18	0.25 – 0.7	70 –75
Medium strength	750 – 3000	2.0 – 5.0	0.2 – 0.5	6 – 24	0.25 – 0.7	80 – 90
High strength	3000 – 10,000	5.0 – 10.0	0.2 – 0.6	6 – 24	0.15 – 0.7	75 – 85
Very high strength	> 10,000	5.0 – 15.0	0.2 – 1.0	> 24	---	65 – 75

(Source: Ghangrekar et al., 2003)

21.8.2 Reactor Volume

Based on the higher suitable value of OLR, for given COD concentration, the volume of reactor required is to be worked out as:

Volume = (Flow Rate x COD concentration) / OLR ……. (2)

For the suitable SLR values for that COD range (Table 21.2), the volume of sludge required can be worked out considering the average concentration of VSS between 25 and 35 g/L for medium and high strength wastewater, and 15 to 25 g/L for low strength wastewater. This volume of sludge should be less than 50% of the reactor volume, worked out based on OLR, to avoid overloading of the reactor with respect to SLR. If the volume is not meeting the requirements, the OLR can be reduced to increase the volume. The volume of the reactor is thus, finalized to meet both the requirements. For this volume, the HRT should not be allowed to be less than 6 h for any type of wastewater and generally, it should be less than 18 h to reduce volume and hence, cost of the reactor. For very high strength of the wastewater, COD greater than 10,000 mg/L, it may not be possible to meet this requirement, hence, under such situation the HRT may be allowed to exceed even 24 h and as high as 200 h.

21.8.3 Superficial Liquid Upflow Velocity

Higher upflow velocities, favors better selective process for the sludge and improve mixing in the reactor. However, at very high upflow velocity, greater than 1.0 to 1.5 m/h, the inoculum may get washed out during start-up or during normal operation granules may get disintegrated, and the resulting fragments can easily wash out of the reactor. The maximum liquid upflow velocity allowed in design should not exceed 1.2 – 1.5 m/h. Upflow velocities as 0.25 to 0.8 m/h are favorable for granule growth and accumulation, during normal operation of the reactor and maximum upflow velocity up to 1.5 m/h at peak flow conditions for short duration can be used in design.

21.8.4 Reactor Height and Area

The reactor should be as tall as possible to reduce plan area and to reduce cost of land, GLS device, and influent distribution arrangement. The height should be sufficient to provide enough sludge bed height to avoid channelling and to keep liquid upflow velocity within maximum permissible limits. In order to minimise channelling the minimum height of the sludge bed should be about 1.5 to 2.5 m. For this reason, the minimum height of the reactor should be restricted to 4.0 m, to conveniently accommodate sludge bed, sludge blanket and GLS separator. The maximum height of the reactor can be about 8 m. The height of the reactor adopted in practice is usually between 4.5 and 8 m and 6 m is the typical height used for UASB reactors.

While designing, initially suitable height of the reactor (about 6m) can be chosen, and superficial liquid upflow velocity is to be worked out as height/ HRT. It is recommended to adopt upflow velocity of 0.7 m/h at average flow and 1.0 m/h to 1.2 m/h at peak flow. Accordingly, if the upflow velocity exceeds the maximum limits height of the reactor can be reduced in steps up to minimum of 4 to 4.5 m. If this is not possible in the applicable range of height, HRT shall be modified and fresh reactor volume and OLR shall be worked out. For low strength wastewater, the maximum liquid upflow velocity becomes limiting and for very high strength wastewater very low velocity (less than 0.1 m/h) is required while designing the UASB reactor. Under certain situations, the revised OLR may be less than the initial OLR recommended. It is advisable to allow lowering of OLR in such situations to control upflow velocity in the reactor for proper performance of the reactor.

After these iterations for volume and height, the plan area can be worked out and suitable dimensions of the reactor can be adopted. Generally, the maximum diameter or side length of single reactor should be kept less than 20 m. Before finalizing the dimensions of the reactors, it is necessary to consider the dimensions required for GLS separator, because to accommodate the GLS separator meeting all requirements, it may be necessary to alter height and plan area of the reactor.

21.8.5 Gas-Liquid-Solid (GLS) Separator

In order to achieve highest possible sludge hold-up under operational conditions, it is necessary to equip the UASB reactor with a GLS separator device. The main objective of

this design is to facilitate the sludge return without help of any external energy and control device. The guidelines for shapes and design of GLS separator are given by Lettinga and Hulshoff (1991). The GLS should be designed to meet the requirements such as, provision of enough gas-water interface inside the gas dome, sufficient settling area outside the dome to control surface overflow rate; and sufficient aperture opening at bottom to avoid turbulence due to high inlet velocity of liquid in the settler, and to allow proper return of solid back to the reactor. Due attention has to be paid to the geometry of the unit and its hydraulics, to ensure proper working of the GLS separator.

Design of GLS separator: The shape of the GLS device considered in design is presented in Figure 21.4. The gas-water interface inside the dome is considered at the depth Δh from top of the dome. In the beginning, the height of GLS separator can be considered as 25% of the total reactor height. For estimating initially the number of domes required the angle of dome with horizontal can be assumed as 45^o, and base width of dome (Wb) can be calculated as $2(h+\Delta h)/ \tan \theta$. The Δh is to be calculated as $(Wt/2) \tan \theta$, and initially the top width (Wt) can be considered as 0.2 to 0.3 m. The number of domes required for given diameter (or width for rectangular reactor) can be calculated by dividing width or diameter by W_B, and rounding this number. Where, W_B=Wb+Wa, and Wa can be considered as 0.2 m initially. After deciding the number of domes, the flow rate shared by each dome, is to be estimated in proportion to the base area of each dome, including aperture width, to the total area of the reactor.

Aperture width at bottom of gas dome: The area of aperture (Ap) required can be computed based on the maximum inlet velocity of liquid to be allowed. This area can be estimated as flow rate per dome for rectangular reactor (or central dome in case of circular) divided by maximum velocity to be allowed. The maximum inlet velocity of 3 m/h is safe for medium and high strength wastewater and for low strength wastewater lower inlet velocity should be preferred. The width of aperture (Wa) is to be calculated as aperture area divided by length (or in case of circular reactor by diameter) of the reactor. It is recommended to use minimum aperture width of 0.2 m and if the width required is greater than 0.5 m, then increase the number of dome by one and repeat earlier steps till it is less than 0.5 m.

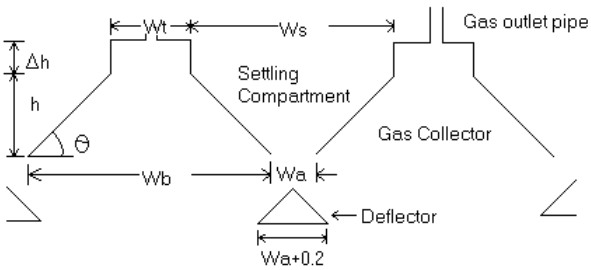

Figure 21.4. Details of the Gas-Liquid-Solid (GLS) Separator

Width at gas-water interface: The gas production expected in the reactor can be estimated based on the OLR selected for the design and expected COD removal efficiency in the range 70 to 90 percent. The methane production can be estimated as 0.35 m^3 /kg COD removed at ambient temperature and methane content of 70% in biogas. From this gas production the biogas collection per dome is to be worked out in proportion with percentage of area covered by the dome. The biogas loading at gas-water interface can be calculated as gas collection per dome divided by product of top width of gas collector (Wt) and length of the gas collector dome. The loading of biogas at gas-water interface should be kept less than 80 m^3 gas/ m^2.d (about 3 m/h) (Ghangrekar et al., 2003). Initially the top width can be assumed as 0.3 m and for this width if the biogas loading is less than 3.0 m/h then adopt 0.3 m as top width. If the biogas loading is greater than 3.0 m/h, calculate the top width required. Generally, top width of 0.3 to 0.7 m can be adopted in design with maximum of 1.0 m. When even with maximum top width, if biogas loading is greater than 3.0 m/h reduce the height of GLS separation device to 20% and repeat the earlier steps of GLS separator design, with fresh number of domes. Even with reduction in height of GLS separator if these checks are not satisfying, provide additional layer of gas collector dome. When two or more layer of gas collectors are used the height of each layer can be 15 to 20% of the overall reactor height, with minimum height of each layer as 1.2 m and maximum up to 1.5 to 2.0 m. The fresh biogas collection per dome is to be worked out and further steps are repeated until all design conditions are satisfied.

Check for Surface overflow rate: The width of the water surface (Ws) available for settling of solids for each gas dome, at top of the reactor, can be calculated as difference of base width of dome ($W_B = W_b + W_a$) and Wt. The corresponding surface overflow rate is calculated as hydraulic flow rate per dome divided by product of length (or diameter) and Ws. It is recommended that the surface overflow rate for effective settling of solids back to

the reactor should be less than 20 m^3/m^2.d at average flow and should be less than 36 m^3/m^2.d under peak flow conditions. If the calculated surface overflow rate is meeting these criteria the design of the GLS separator is final. When it is exceeding the limits recommended, it is advisable to reduce the height of the reactor, thus, for same volume of the reactor more plan area will be available. When the height of the reactor is reduced all earlier steps for design of GLS separator should be repeated to satisfy all design criterions. The minimum height of the reactor should be restricted to 4.0 m (preferably 4.5 m). Once, all the design criteria are satisfied the angle of inclination of the gas collector dome with horizontal (θ) can be calculated as $\theta = \tan^{-1}[\ 2h/\ (Wb - Wt)]$.

Baffle of sufficient overlap (0.1 to 0.2 m) should be provided below the gas collector in order to avoid entry of biogas in the settling compartment. The diameter of the gas exhaust pipes should be sufficient to guarantee easy removal of the biogas from the gas collection cap, particularly in case of foaming. Generally, lower reactor height is required for UASB reactor treating sewage. Under certain situation, particularly for very low strength of wastewater, even with reduction of height to the minimum may not meet all design requirements. In such cases the OLR adopted for design can be reduced to provide greater volume of the reactor and hence more plan area to meet the entire design criterion.

21.8.6 Effluent Collection System

The effluent has to leave the UASB reactor via number of launders distributed over entire area discharging to main launder provided at periphery of the reactor. The effluent launders can be designed in such a way that the weir loading (m^3/m.d) should not exceed the design criteria of secondary settling tank (*i.e.*185 m^3/m.d). The width of the launders may be minimum 0.20 m to facilitate maintenance. The depth of the launder can be worked out as open channel flow. Additional depth of 0.10 to 0.15 m shall be provided to facilitate free flow. On both sides of the launders 'V' notches shall be used. When effluent launders are provided with scum baffles, the 'V' notches will be protected from clogging as the baffles retain the floating materials. A scum layer may form at the top of reactor and sludge accumulation can occur in the launder hence, periodical cleaning of launders and removal of scum should be carried out.

21.8.7 Design of Feed Inlet System

It is important to establish optimum contact between the sludge available inside the reactor

and wastewater admitted, and to avoid channeling of the wastewater through sludge bed. Hence, proper design of inlet distribution system is necessary. Depending on topography, pumping arrangement, and likelihood blocking of inlet pipes, one could provide either (i) gravity feed from top (preferred for wastewater with high suspended fraction), or (ii) pumped feed from bottom through manifold and laterals (preferred in case of soluble industrial wastewaters). The rough guidelines for the number of feed inlet points required in UASB reactor is presented by Lettinga and Hulshoff (1991) for different concentration of the sludge inside the reactor and applicable loading rates. In general, the area to be served by each feed inlet point should be between 1 and 3 m^2. Lower area per inlet point (1 m^2) is to be adopted for reactor designed for OLR of about 1 kg COD/m^3.d, and higher area (2 to 3 m^2) per inlet point can be provided to the reactor designed for OLR greater than 2 kg COD/m^3.d. Apart from the number of feed inlet points, the minimum and maximum outflow velocity through the nozzles should also be given due consideration while designing. This outflow velocity through nozzles can be kept between 0.5 and 4.0 m/s. The equation of 'condition for maximum power transfer through nozzle' can be used for working out nozzle or inlet pipes diameter. The clogging of the nozzles may represent serious problem resulting in uneven distribution of the wastewater over reactor bottom, particularly when treating partially soluble wastewater. Hence, arrangements should be made for cleaning or flushing the inlet system.

21.8.8 Other Requirements

It is necessary to keep provision for removal of excess sludge from the reactor. Although, the excess sludge is wasted from about middle height of the reactor, it is also necessary to make arrangement at bottom of the reactor. In addition, 5 to 6 numbers of valves should be provided over reactor height to facilitate sampling of the sludge. For treating high strength wastewater it is recommended to apply effluent recycle, in order to dilute COD concentration and to improve contact between sludge and wastewater. For treating wastewater with COD concentration greater than 4 - 5 g/L, it is recommended to apply dilution during start-up, for proper granulation of sludge inside UASB reactor. Auxiliary equipment has to be installed for addition of essential nutrients, and alkalinity for control of pH of the influent. The other equipments to be provided are for measurement of pH, temperature, influent flow rate, and gas production rate.

Example: 1

Design an UASB reactor for treatment of 4 MLD sewage having BOD of 200 mg/L and COD of 500 mg/L. The average minimum temperature of wastewater in winter is about 20 ^{o}C and maximum temperature in summer is 35 ^{o}C. The wastewater contains 80 mg/L sulphate.

Solution

Reactor Volume (V)

Range for HRT is 6 to 18 h

$V = Q \times HRT$

Provide HRT of 8 h

$V = 4 \times 10^3 \times \frac{1}{24} \times 8$

$= 1333.33\ m^3$

Check for OLR

Range for OLR = 1 to 3 kg COD/ m^3.day

$$OLR = \frac{\text{flow per reactor X COD}}{\text{volume of each reactor}}$$

$$= \frac{4 \times 10^3 \times 500 \times 10^{-3}}{1333.33}$$

$= 1.5$ kg COD/ m^3.day

Check for SLR

Range for SLR = 0.1 to 0.3 kg COD/kg VSS/day

$$SLR = \frac{\text{flow per reactor X COD}}{\text{volume of sludge x VSS in reactor}}$$

Volume occupied by sludge bed should be about 50% of reactor volume

Assume VSS in reactor sludge = 25 g/l

$$SLR = \frac{4 \times 10^3 \times 500 \times 10^{-3}}{1333.33 \times 0.5 \times 25}$$

$= 0.12$ kg COD/kg VSS.day

Check for MCRT

Range for MCRT = 40 - 100 days

$$\text{MCRT} = \frac{\text{mass of inoculum sludge}}{\text{mass of sludge wasted per day}}$$

Assume sludge in effluent to be 100 mg/L

$$= \frac{1333.33 \times 0.5 \times 25}{4 \times 10^{3} \times 100 \times 10^{-3}}$$

$$= 41.67 \text{ days}$$

Height of the reactor

Let us provide height of the reactor H = 4.5 m.

Check for upflow velocity

Upflow velocity = H/HRT = 4.5/8 = 0.562 m/h (less than 0.7 m/h)

Area of the reactor

$$A = \frac{\text{volume of reactor}}{\text{height of reactor}}$$

$$= \frac{1333.33}{4.5} = 296.296 \text{ m}^2$$

Provide length of the reactor L = 19 m, hence width B = 15.6 m

Design of GLS separator

Height of dome = 0.25 x ht. of reactor

$$= 0.25 \times 4.5$$

$$= 1.125 \text{ m}$$

Provide 1.20 m height of the dome and 0.3 m free board above the water surface for gas collection.

Provide max liquid velocity at aperture i.e. inlet of the settler = 3 m/h

$$\text{Area of opening at inlet of settler} = \frac{4000}{3 \times 24}$$

$$= 55.56 \text{ m}^2$$

$$\text{Total width of opening required} = \frac{\text{Area of opening}}{\text{width of reactor}}$$

$$= \frac{55.56}{15.6} = 3.56 \text{ m}$$

Provide width of each gap = 0.4 m

No of gaps = $\frac{3.56}{0.4}$ = 8.9 say 9, provide 9 number of domes which will make 8 openings in the middle of the domes and two opening along the side wall.

Width of each aperture opening will be 0.395 m and along the wall it will be 0.198 m.

Hence provide deflector beam of 0.59 m below the aperture opening and 0.3 m base width

Provide 0.3 m width at top of the dome

Total width of base of domes = length – width of each gap x no. of gaps – top width x no. of domes

$$= 19 – 0.395 \text{ x } 9 – 0.3 \text{ x } 9$$

$$= 12.745 \text{ m}$$

Width of base of each dome = $\frac{12.745}{9}$ = 1.416 m

Angle of inclination = $\tan^{-1} \left(\frac{1.2}{0.71 – 0.15}\right) = 64.98^0$

Gas production

Methane production in litres = 1.28 x T (^{0}K) per kg of COD removed

= 1.28 x (273+30) per kg of COD removed

= 387.84 L per kg of COD removed

Let the COD efficiency of the system be 75%.

So kg of COD removed = flow per reactor x COD x 0.75

$$= 4 \text{ x } 10^3 \text{x } 500 \text{ x } 10^{-3} \text{x } 0.75 \Big/ \text{day}$$

= 1500 kg/day

The total COD removed in the reactor = 1500 kg/day

But not all the organic matter present in the influent is carbonaceous. The influent also consists of sulphates which are reduced to sulphides and consume about 0.67 kg of COD per kg of sulphate

$$SO_4^{2-} \rightarrow S^{2-}$$

Assuming sulphate removal of 80%, the total sulphate reduction

$= 0.8 \times 4000 \times 80 \times 10^{-3}$

$= 256$ kg/day

COD consumed in sulphate reduction = 256 x 0.67 = 171.52 kg/day

Hence COD available for methane production = 1500 – 171.52 = 1328.48 kg/day

Also some portion of biogas will remain in soluble form in the reactor effluent due to high partial pressure of biogas inside the reactor. Typically about 16 mg/L of methane will be lost along with the effluent.

Methane that can be collected = $1328.48 \times 0.38 - 4000 \times 16 \times 10^{-3} = 440.822$ m^3/day

Also gas collection efficiency of the domes will be about 85 to 90%, hence actually methane collected at 85% efficiency will be = 374.69 m^3/day

Check for biogas loading at gas-water interface

Total biogas produced assuming methane content to be 70% = $\frac{1328.48 \times 0.38}{0.70}$ = 721 m^3/day.

Max gas loading rate = 3 $m^3/m^2/h$

Area required = $\frac{\text{Volume of biogas produced}}{\text{Max gas loading}}$

$= \frac{721}{3 \times 24}$

$= 10$ m^2

Total top width = $\frac{\text{Area required}}{\text{width of UASB}}$

$= \frac{10}{15.6} = 0.64$ m. Hence, the width required for each dome = 0.07 m which is less than 0.3 m provided.

Check for surface overflow rate (SOR)

Width available in the settling compartment i.e. outside the domes

$= (19 - 0.3*9) \quad = 16.3$

Hence, SOR $= 4000/(16.3*15.6) = 15.73$ $m^3/m^2.d$ (Less than 20 $m^3/m^2.d$)

Questions

1. Explain the reaction sequence involved in the anaerobic treatment of wastewater.

2. Describe different types of bacteria and their role in anaerobic degradation of organic matter to final end product.
3. Discuss the factors that can affect the anaerobic reactor performance adversely.
4. Describe advantages and disadvantages of anaerobic treatment of liquid waste.
5. Describe different types of reactors used for anaerobic treatment of wastewaters.
6. What is high rate anaerobic process? Name different high rate anaerobic reactors.
7. Explain different types of filters used in anaerobic treatment.
8. With the help of schematic explain UASB reactor and its working.
9. What is GLS separator? What are the design guidelines for GLS separator?
10. Describe advantages of sludge granulation in UASB reactor.
11. Why post treatment is necessary for anaerobic reactor effluents? What post treatment you will recommend after treatment of sewage in UASB reactor?
12. Describe organic loading rates used for design of UASB reactor. How reactor height is important for proper functioning of UASB reactor?
13. Design a UASB reactor for treatment of 2 MLD of sewage having COD of 500 mg/L and BOD of 250 mg/L. Make suitable assumptions for the design.

Answer

Q. 13.

Assume HRT = 8 hr

Volume required = 666.67 m^3; provide 2 reactors of height of 5.6 m and L x W = 10 x 6 m.