FLUIDIZED BED REACTORS FOR ANAEROBIC WASTEWATER TREATMENT

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ABSTRACT

A review of the theoretical basis for the design and operation of fluidized bed reactors for anaerobic treatment of wastewaters is presented. After the description of the characteristics of three-phase fluidized bed and theoretical aspects which could affect the operation of an anaerobic reactor, the design considerations and recommendations based upon characteristics of the wastewater to be treated are presented. A step by step design strategy, including media selection, operative conditions of the reactor, equipment sizing and long-term run effects, as well as mechanical design hints on distribution system, reactor body, piping of recycle lines and gas collection are included.

KEYWORDS

Fluidized bed; expanded bed; anaerobic reactor; anaerobic digestion; fluidization; design; wastewater.

INTRODUCTION

Although the application of the fluidized bed technology to the anaerobic treatment of wastewaters has been studied to a great extent on lab and pilot scale, only a few full-scale reactors have been built. Some of the reasons for this imbalance are based on the advanced technology used on FB reactors but, more commonly, on the anaerobic process itself and its, until now, difficult process control.

The FB technology offers many advantages to the process (Jewell, 1985; Hickey, 1990):

1) High concentration of biomass, attached to a dense carrier, which cannot be easily washed out from the reactor.

2) Very large surface area for biomass attachment.

3) Initial dilution of the influent with effluent, which provides alkalinity and, thus, some neutralization, reduces substrate concentration (important for high COD wastes), and contributes to reduce the shock effect of toxicant spikes.
4) High mass transfer properties. Low concentration gradients around the particles are possible, allowing the treatment of low strength wastes.

5) No plugging, channeling or gas hold-up.

6) Ability to control and optimize biofilm thickness.

7) Biomass carrier can be tailored to a specific application to enhance performance.

This paper will deal with the specific characteristics of the fluidized bed reactors which affect their design, operation and performance, since most of the general aspects will be discussed in other papers at this Workshop.

PREVIOUS STUDIES AT LABORATORY, PILOT AND FULL SCALE

Many studies have been carried out during the last 15 years on the application of FB for the treatment of different substrates, enlightening of the circumstances which provide good efficiencies and description of performance. This paper will describe several, but not all, of these studies since there are too many to list all of them.

Some pilot-scale reactors have been developed around the world, most of them, logical continuation of lab-scale research. The pilot-scale reactor commonly used ranges between 30 and 6000 L volume, with high H/D (length/diameter) ratios, to reduce problems of flow distribution.

Sand and anthracite have been widely used as carriers, more recently, lighter products like ion exchange resin beads, natural or baked clays (kaolinite, sepiolite, Arge (R), Arlita(R)), granular activated carbon, pumice and reticulated polyurethane have been used.

A partial list of examples can be found in Table I.

<table>
<thead>
<tr>
<th>Volume (L)</th>
<th>H/D (m)</th>
<th>Wastewater</th>
<th>Reference</th>
</tr>
</thead>
<tbody>
<tr>
<td>270</td>
<td>6/0.25</td>
<td>Yeast production</td>
<td>Heijnen, 1984</td>
</tr>
<tr>
<td>5380</td>
<td>6.85/1</td>
<td>Soft-drink waste</td>
<td>Yoda et al., 1987</td>
</tr>
<tr>
<td>1000</td>
<td>6/0.5</td>
<td>Sugar beet wastewater</td>
<td>Iza, 1987</td>
</tr>
<tr>
<td>7000</td>
<td>7/1.13</td>
<td>Sewage</td>
<td>Fdz.-Polanco, 1988</td>
</tr>
<tr>
<td>4200</td>
<td>3/1.5</td>
<td>Sewage</td>
<td>Sanz, 1988</td>
</tr>
<tr>
<td>70</td>
<td>2.5/0.144</td>
<td>Synthetic waste</td>
<td>Weiland and Büttgenbach, 1988</td>
</tr>
</tbody>
</table>

Since full-scale development of a FB reactor involves some patented or "secret" devices, engineering firms are very reluctant to provide information. Thus, only a few plants have been reported in technical literature and, when so, with scarce information on meaningful data. This is also related to the use of fancy pseudo-commercial names for the process (such as expanded bed, mobile bed, carrier assisted sludge bed and so on) which are essentially fluidized beds.

The fluidized bed technology, however, has been applied with great success in other biotechnological processes, such as nitrification (Cooper and Williams, 1990), denitrification (MacDonald, 1989), aerobic wastewater
Fluidized bed reactors

Some of the plants in operation are:
- From the same company, there are other systems in Prouvy, France, (2 reactors 125 m$^3$ each) and Monheim, FRG, (125 m$^3$) this one operating as an Expanded granular sludge bed (Borghans et al., 1990).
- Degremont (Cervezas El Aguila), Madrid, Spain. Brewery wastewater (Oliva et al., 1990). 5 reactors 165 m$^3$ each.
- From the same company, other plant in France (Nicol and Prevot, 1986).
- Reliance Industries, Bombay, India. (Dorr Oliver design). 1 reactor 850 m$^3$. (Sutton and Mishra, 1990).
- Lansing Wastewater Treatment Plant, Lansing, Michigan, USA. For heat treatment liquor of sludges. 4 reactors 180 m$^3$, and some other plants in the USA, by Envirex, such as the one in W.C.I., Hatfield, Pennsylvania. (Sutton and Mishra, 1990)

FLUIDIZATION

Description

When an increasing flow of liquid passes through a bed of particles, first the bed expands and then the particles get suspended and free to move with respect to the others. The bed is then fluidized (Richardson, 1971) (Leva, 1959) (Kunii and Levenspiel, 1969).

After passing the flow threshold which causes the fluidization, two different types of behavior can occur:
- the bed expands, increasing the distances between particles,
- the excess flow passes through the bed forming bubbles.

These fluidization types are known as Particulate and Aggregative, respectively.

Solid-liquid fluidized beds of low density solids such as glass, sand or GAC can be considered ideal homogeneous particulate beds (Ramamurthy and Subbaraju, 1973). The bed itself behaves as a fluid, getting a new set of physical properties (density, viscosity) and follows hydrostatic and hydrodynamic fluid laws. If we add the improved characteristics of thermal (and specially) mass transfer, we find here some of the reasons for the use of fluidized beds for biological process.

In order to get a fluidized bed, an increasing flow of liquid is applied through a settled bed of particles, which form a fixed bed. During the progressive increase of flow, the bed starts to expand. At this moment, all equations which apply to fixed bed describe the situation. If the flowrate is increased, a transition occurs and particles start to move suspended on the upflow liquid separated from the other particles. At this very moment, fixed bed laws are still followed. If the flowrate is increased, particles are more separated from each other; their hydrodynamic behavior resembles the behavior of particles settling. The limit of this phenomenon is called fluid transport, where particles are carried out of the bed by the liquid flow.
Fluidized beds are also the bed of particles lighter than the liquid such as cork and other lignocellulosic products and plastics. The bed is then fluidized downwards by a fluid of lower density. To achieve this lower density fluid, air or, in our case, biogas is injected at the bottom of the column producing a fluid mixture whose overall density is able to drag down the floating particles (Bories et al., 1985) (Paris et al., 1986).

**Fluidized Bed Characteristics**

In this section several variables of fluidized bed systems are discussed in relation to their usefulness for the design.

**Pressure losses.** Pressure losses in a fixed bed can be described by Carman-Kozeny equation. There is a linear relationship between pressure losses and upflow liquid velocity. At the onset of fluidization, the weight of the particles is just equal to the fluid drag force; particles flow freely and the bed becomes fluidized.

\[
P_{mf} = (p_s - p_l)(1-\varepsilon_{mf})H
\]

where the subscript \(mf\) indicates conditions of minimum fluidization, \(P_{mf}\) is pressure loss; \(p_s\) and \(p_l\), specific gravity of solid and liquid, respectively; \(\varepsilon_{mf}\) the porosity of the bed, and \(H\), its height.

A further increase in upflow velocity expands the bed, increasing the average distance between the particles, without affecting the pressure loss.

\[
P = (p_s - p_l)(1-\varepsilon)H
\]

Thus, any expansion of the bed will be described by

\[
(1-\varepsilon)H = \text{constant} = (1-\varepsilon_{mf})H_{mf}
\]

**Figure 1** Pressure loss in a fluidized bed

During the onset of fluidization, some non-ideal behavior can occur when the upflow liquid pass through some clear paths (bed channeling) or when particles are stuck together and adhere to each other (quite common on biological beds); a slight overpressure is needed to separate the particles and lead to a homogeneous particulate fluidization.

Of practical interest is the fact that bed perturbations reach a faster equilibrium when the bed contracts than when the bed expands. This should be noted when analyzing bed behavior (Saxton et al., 1970).

**Minimum fluidization conditions.** The fluidization of a particulate bed occurs very smoothly, with an homogeneous expansion, if particles are uniform, and with a high tendency to segregation, if particles are
heterogenous (Fan et al., 1984).

Of particular importance is the knowledge of expressions which describe the behavior of the bed, in order to make correct decisions about the scale-up of the systems.

The equation most commonly used to describe the expansion of a fluidized bed is the Richardson-Zaki equation, based on experimental data (Richardson, 1971):

\[
\frac{u}{u_t} = \varepsilon^n
\]

where \( u \) is the empty-bed liquid upflow velocity, \( \varepsilon \) is the porosity of the bed (void space) and \( u_t \) is the free-falling terminal velocity of the particle, which should be corrected for the wall effect:

\[
\log u_t = \log u_t^{*} - d/D
\]

where \( d \) and \( D \) are the diameters of the particle and the reactor, respectively.

The \( n \) coefficient is calculated at different Reynolds number (\( Re_t \)) values with the following expressions:

\[
\begin{align*}
n &= 4.65 + 20 \frac{d}{D} \quad \text{for } Re_t < 0.2 \\
n &= (4.4 + 18d/D) Re_t^{-0.03} \quad \text{for } 0.2 < Re_t < 1 \\
n &= (4.4 + 18d/D) Re_t^{-0.1} \quad \text{for } 1 < Re_t < 200 \\
n &= 4.4 Re_t^{-0.1} \quad \text{for } 200 < Re_t < 500 \\
n &= 2.4 \quad \text{for } 500 < Re_t
\end{align*}
\]

A typical log/log plot of \( u \) vs. \( \varepsilon \) shows a straight line. Some interesting values are easily calculated, like \( u_{fr} \) (at \( \varepsilon = \varepsilon_{fr} \)) and \( u_t \) (at \( \varepsilon = 1 \)).

Another approach was presented by Ramamurthy and Subbaraju (1973) using Galileo and Reynolds numbers and a form factor \( \Phi \). The bed is considered as a lattice of particles where the fluid passes through the voids (Ranz, 1953; Turner, 1973; Ruckenstein, 1964). The experimental correlation suggests a change in the form factor to correct the roughness of the model. Wen and Fan (1974) suggested some corrections for the equations at transition and turbulent regimes.

Garside and Al-Dibouni (1977) used a generic equation:

\[
\frac{u_r - A}{B - u_r} = c Re_t^n
\]

where \( A, B, c \) and \( m \) are coefficients and \( u_r \) is a relative liquid velocity, defined as

\[
u_r = \frac{u}{u_t}
\]

To take into account the wall effect, the correction of the terminal velocity of the particle at infinite dilution is:

\[
u_{fr}/u_t = 1 + 2.35d/D \quad (3 < Re < 200)
\]
At low expansion ($\varepsilon < 0.85$), the equation is:

$$\frac{u_r - \varepsilon^{1.14}}{0.8 \varepsilon^{1.28} - u_r} = 0.06 \text{Re}_{\varepsilon}^{(4+0.2)}$$  \hspace{1cm} (9)

Although all these equations can be used to describe the behavior of a fluid bed reactor, the expression more commonly used is the Richardson-Zaki's one (Andrews and Tien, 1979) (Andrews, 1986), (Andrews and Przezdziecki, 1986) (Más and Solá, 1987) (Mulcahy and Shieh, 1987).

The experimental setup required to obtain data is relatively simple and inexpensive and permits the prediction with good accuracy of the upflow velocity/bed expansion relationship. The other equations require many parameters, such as particle diameter, sphericity, density, average diameter, Sauter diameter, etc. Most of the time such parameters are not found on tables and must be experimentally determined or measured. Taking in account the origin of the particles (most of them natural crushed rocks or sand) and its parameter dispersion, and the very important effects which happen inside the reactor, namely, biomass growth and attrition (Sinha et al., 1984), it seems unwise to proceed to a very detailed analysis of the fast-changing particles. This aspect will be studied later.

Three-phase Fluidization

Usually three-phase fluidization is considered as the fluidization of one solid by two fluids (two immiscible liquids or a liquid and a gas) where one of the fluids forms a continuous phase (Østergaard, 1971, 1973, 1979). This is the case of some biotechnological processes such as the aerobic fluidized bed, where biofilm coated particles are fluidized by an upflow liquid stream and air or oxygen is introduced to aerate the culture (Wild et al., 1984).

In our case, the gas is produced inside the bed, most commonly in the active zone close to the inlet port. In low-loaded systems, the influence of gas is negligible. High-loaded reactors, however, have very high gas productions which can affect, even modify the hydrodynamic behavior of the bed.

An interesting, and strange, effect is bed contraction (Dakshinamurthy et al., 1971, 1972). For a small range in particle size, the introduction of a gas stream reduces the bed porosity, contracting the bed (Turner, 1964; Stewart and Davidson, 1964; Østergaard, 1965). Østergaard and Theisen (1966) found that a bed of particles 0.28-2.2 mm in diameter suffered a contraction up to 48% when gas was injected. The effect is caused by the different behavior of both fluids. With this small size particles, the gas bubbles coalesce and form big bubbles which travel upward faster than the liquid and cause a big wake. Many theories have been postulated to explain the contraction (Stewart and Davidson, 1964; Bathia et al., 1972; Rigby and Capes, 1970; Darton and Harrison, 1975).

The effect has been described in anaerobic reactors by Iza (1987) and in aerobic reactors by Hatzifotiadu et al. (1989).

Although many studies have been carried out, the resulting equations predict the behavior of three-phase fluidized beds, but not the reduction to two-phase (no gas production) (Dakshinamurthy et al., 1971; Soung, 1978; Epstein, 1976). It is thus recommended to use one of the classic two-phase equations.
**Bed Segregation**

The size distribution of the particles used on a fluidized bed reactor is usually very narrow. If broader ranges are used, the smallest particles are highly fluidized, even washed out from the reactor, whereas the bigger ones remain non-fluidized, forming a fixed bed (Shi et al., 1984).

Due to the biological nature of the process, biofilm growth affects the size, overall density, shape and roughness of the particles, as well as its chemical and adsorptive characteristics. For narrow ranges, some authors use the average between minimum and maximum mesh gauge (Dakshinamurty, 1971). For broader distributions, it is common to define three different parameters (Cooper and Weldon, 1981):

- **mean diameter** $d_{60}$, diameter of the screen which allows the passage of 60\% (w/w)
- **effective diameter** $d_{10}$, defined as $d_{60}$
- **uniformity coefficient** defined as $d_{60} / d_{10}$.

It is also needed to consider the shape of particles. For some of the most elaborate equations describing the bed behavior, a factor accounting for the sphericity is needed (Andrews and Tien, 1979).

The presence of different shape and size particles normally causes a segregated bed: the heavier particles move down to the bottom of the bed, whereas the lighter ones rise to the top. In most of the cases, where the size distribution of the particles is a continuum, there is a linear distribution by sizes from bottom to top. The cause is the interaction between segregation and diffusion (mixing).

For solid-liquid beds of two size particles, Wen and Yu (1966) found total segregation when the ratio between the minimum fluidization velocity of the particles is bigger than 2.0 or when the diameter ratio is bigger than 1.3. In this case, the global expansion of the bed can be calculated applying the Richardson-Zaki equation to each section of homogeneous particles and adding the partial expansion of each section (Sinha et al., 1984).

For three-phase fluidized beds, the tendency to segregate is greater than mixing, when the size ratio is bigger than 2 (strictly 2.2) Depending on the hydrodynamic regime of the system, the gas release (in general, gas injection) can cause either mixing or segregation (Garside and Al Dibouni, 1974).

It has been found that, for the same liquid velocity, solids mixing increases with gas velocity (Fan et al., 1984, 1987). The effect of bed contraction is also found in this kind of segregated beds (Sinha et al., 1984).

**Axial (Vertical) Dispersion**

One of the phenomena easily visualized on laboratory-scale reactors is the formation of big vertical particle streams, some ascendent, some descendent. The streams are not related to the typical random motion of particles in a particulate fluidized bed but rather to tilting of the reactor walls. The effect has been found in reactors with less than 1° of tilting (Van der Meer et al., 1984). This internal recycling motion cannot be described mathematically and, thus, the description of the axial dispersion is not feasible. The practical effect of this behavior is important, contributing to a better homogenization of the reactor content and exposing the particles to different environments (Dorgelo et al., 1985).
As a conclusion, segregation and mixing in fluidized bed reactors are phenomena not well known (Andrews, 1986). In case of total mixing, the reactor would show a flat profile of biomass concentration, and particle size distribution. Field studies using broad size distribution and low expansion ($\varepsilon < 0.7$) have shown the segregation effect is bigger than the mixing (Iza, 1987). All these comments are related to the mixing/segregation of particles, not the behavior of the liquid phase which usually, due to the recycle effect, is closer to a completely mixed regime (Bull et al., 1983).

**DESIGN CONSIDERATIONS AND RECOMMENDATIONS**

**Waste Characteristics**

As with any other anaerobic system, optimal wastewaters are warm, biodegradable, pH neutral, low salinity and free of toxic or inhibitory compounds. The presence or absence of particulate material is dependent on the degree of bed expansion.

On low expansion FB reactors, particles are retained by entrapment or adhesion and destroyed or reduced in size by attrition or (bio)chemical attack (Yoda et al., 1985).

On high expansion FB reactors, this effect is greatly reduced due to higher upflow velocities which carry away all particulate material from the bed.

**Alkalinity**

FB reactors usually operate with moderate to high recycle ratios: effluent is recycled to achieve the liquid upflow velocity needed for fluidization. Due to the anaerobic conversion, where CO$_2$ is formed, the effluent has a relatively high alkalinity, which, when mixed with the incoming influent, eases the pH control of the process, and reduces the caustic needs.

However, FBR systems are very high-rate processes, operating at very low retention times. An organic spike can lead to an imbalance and a very fast acidification of the system. Provision for equalization tanks may be warranted, as well as an alkalinity (caustic) dosage system.

**Nutrients and Trace Minerals**

Due to the high upflow velocities, precipitation of mineral sulfides or carbonates and accumulation of inorganic solids is highly unlikely to occur, as is common in all other anaerobic systems. An early screening of the waste and a nutrient dosage determination are needed, thus, to avoid nutrient deficiency during operation. With an adequate blending of wastewaters, these dosage requirements can be greatly reduced or even avoided.

As it is well known, most of these compounds are also toxic when applied in high concentrations (spikes) or sustained medium to low concentrations, and also when pH, temperature and chemical conditions change. Microorganisms show a high adaption capacity to some of the toxicants, including population shifts as a way to overcome the problem.

One of the interesting aspects of FB reactors is the presence of carrier
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117 particles. Although sand and other dense carriers present a low adsorption capacity, all clays (natural or baked), pumice and activated carbon show a high adsorptive capacity. Some authors indicate the whole particle is covered by biomass and, if not, by glyocalyx or extracellular polymers, covering all adsorption niches. Studies carried out by Suidan et al. (1988) showed the adsorption capacity of granular activated carbon was still available after 200 days of operation. This effect contributes to the elimination of toxics from the bulk liquid and fix them over a surface where chemical and biochemical attack can occur with success (Henze and Harremoës, 1983).

Biological Solids Yield and Sludge Wasting

One of the main problems in FBR technology is the control of the biomass growth. Due to high liquid upflow velocity (greater than the settling velocity of suspended growth biomass), non-attached biomass usually leaves the reactor with the effluent. This can cause a problem if the amount of biomass lost is significant, since it reduces the effluent quality. Since in the initial design, provision for an internal settler cannot be done, in some cases, a small external settler (also used as a carrier (sand) trap) should be introduced into the final design.

The effect caused by biomass growth is the formation of thicker biofilms, which has two side-effects:

1) Thicker biofilms are not as well attached to the support carrier as thinner ones. Collisions can cause major damage and detachment of big portions of the biofilm, which, in turn, can be washed out from the reactor or can promote the formation of granules without carrier particle, depending on the operating conditions.

2) Bioparticles with different biofilm thicknesses have different physical properties (volume, density, cross-sectional area), and, thus, different fluidization properties (terminal velocity, minimum fluidization velocity, hydraulic drag coefficient, etc.) (Hermanowicz and Cheng, 1990). These differences can lead to bed segregation (or, in some extreme cases, mixing) changing the solids profile along the bed, which can be of microbial or (bio)chemical importance. Usually, beds tend to segregate, with denser (less covered) particles at the bottom and lighter particles (thick biofilm, bioparticles with different carrier material or granules without solid carrier) on top of the bed. These differences produce a non-homogeneity on the fluidization of the bed which can cause particle washout (if high upflow velocity is applied) or bed compaction and stagnant areas (if low upflow velocity is used). Neither of the cases is good from an operational point of view.

The technique commonly used to overcome this problem is to keep the expansion of the bed constant, by wasting the excess of thicker biofilm particles. In practice, some patented systems are used, which separate the biofilm from the carrier and recycle the scrubbed carrier to the reactor.

Since the thicker biofilm particles seem to be the most active particles in the bed, it appears unwise to waste all this biomass but it is the only way to keep a fluidized bed of particles under hydrodynamic control.

Another possibility, not yet applied, involves the wasting of medium size biofilm particles, leaving a hybrid reactor composed of a small bed of particles barely covered and a bed of non-carrier granules, transforming the Fluidized Bed system to an Expanded Granular Sludge Bed (de Man et al., 1988). The lower part will perform as a flow distributor, separating...
the granules from the flow inlet area, where turbulence and shear forces can destroy their structure.

The extreme case is to replace all carrier bioparticles by anaerobic granular sludge. The system is operated as a fluidized bed at very low upflow velocities (4-10 m/h) with recycle ratios as low as 2:1 (Borghans et al., 1990)

In conclusion, solids wasting in FB is not mandated because of the possibility of clogging but by hydrodynamic control of bed expansion.

**Temperature**

Until now, all known full-scale FB systems operate in mesophilic and submesophilic conditions. Studies have been carried out on thermophilic, mesophilic and psychrophilic FB for many different substrates, and their performance with respect to temperature is similar to those obtained with other reactor systems.

Thermal shock studies are not widespread, but the thermal momentum of the system, mainly based on the thermal conductivity of the carrier solids, seems to provide some thermal buffer capacity. However, this system is expected to provide high efficiency on highly loaded wastewaters. If the overall temperature decreases, so does the efficiency and it can cause a temporary overload, which can be fatal for the system.

**Equalization**

FB systems require the recycle flow to be kept running even during non-operating periods, such as weekends. The possibility to use a large equalization tank offers many advantages, such as hydraulic and organic buffering, and provides the influent required to run the system continuously. The continuous flow of the same strength wastewater is the optimal technique to use with any biological system. Due to the high rate applied to these FB systems, the importance of this equalization/buffering should be emphasized.

As a rule of thumb on equalization tanks, the more and bigger, the better. It also should be noted that the presence of equalization tanks can provide not only peak smoothing but another (bio)chemical (even microbial) reactor, where further decomposition of the wastewater is achieved (mainly hydrolysis and acidification), lowering also the buffering capacity.

**Influent Waste Concentration**

FB reactors have been used to treat all kind of wastewaters, ranging from sewage (low COD) to beer and sugar production wastewater (high COD). The dilution effect of the recycle and the excellent mass transfer characteristics of the FB reactors allows the systems to operate under such conditions. Inlet concentrations (after blending) range between less than 100 mg COD/L to 2000 mg/L without adverse effect.

The concentration of suspended solids in the influent is also affected by the blending process: due to chemical reactions and, depending on the kind of wastewater, a change in suspended solids nature can occur.

The precipitation of solids in the bed, when it happens, does not affect the overall performance, since the bed expands to hold the extra material. These precipitates can remain in the bed, promoting the formation of
granular sludge and serving as inorganic buffers, or react (chemically or biochemically) and redissolve, leaving the reactor with the effluent.

The formation of new bioparticles using inorganic precipitates as nuclei can affect the hydrodynamics of the bed, since the characteristics of these particles are different from the ones with the added material (size, shape, density, biofilm coverage, etc.)

Organic Loading

FB reactors have been operated with different substrates (sewage to vinasses) under widely dispersed loading rates (a few kg COD/m³.d to 150-180 kg/m³.d).

On full-scale plants, loadings are lower than in lab-scale reactors but, still, values of 30 to 60 kg COD/m³.d are common. Some differences are encountered in the literature when calculating these values. Some authors use the bed volume as a basis, while others use the full volume of the reactor (which seems to be more correct, because the whole reactor must be built, not only the bed section).

Grady (1985), after analyzing data from various full-scale plants, extracted two conclusions related to the scale-up of the systems:

1) Removal efficiency on full-scale reactors is lower than in lab-scale reactors
2) At lab scale, there is a direct correspondence between organic loading rate and removal efficiency, which is not followed in full-scale systems.

The special mass transfer properties of the systems allow the treatment of very diluted wastewaters, where organic loads are very small, with good efficiency and low hydraulic retention time.

DESIGN STRATEGY

The design strategy of a fluidized bed can be described as follows: first consider the carriers in their initial state (stationary bi-phase reactor) and then manage the controllable variables to correct the effect of biofilm growth over the hydrodynamic behavior of the particles (non-stationary bi-phase reactor) and the added effect of biogas production (non-stationary three-phase fluidized bed).

To proceed with the design, the following steps are recommended:

a) Selection of support material and its size,
b) Select the operating expansion and calculation of the superficial velocity,
c) Bed (D,H) and equipment sizing,
d) Evaluation of the effects caused by biofilm growth, and
e) Impact of the gas production/release.

For a correct design or scale-up of a biological reactor, data obtained previously (kinetic and hydraulic) are needed. If a literature review does not provide reliable information for this kind of substrate, or for new wastes, it is needed to perform a lab-or pilot-scale study to characterize the effluent and evaluate the kinetic parameters in conditions close to real or, at least, determine the minimum hydraulic residence time needed to get a reliable operation of the system. These parameters will serve
later as a basis for a correct design.

During start-up, the fluidized bed is a non-stationary system; initially, the support particles have the characteristics defined by the designer. Once the biological process starts, first a conditioning layer and then a biofilm grows over the surface of the particles altering most of their characteristics (physical, chemical & biological), which also affects the operating conditions. As a result of the biological process some gas is produced, changing again the characteristics of the fluidized bed from a solid-liquid two-phase bed to a gas-liquid-solid three-phase fluidized bed system.

Media Selection

The selection of a material for fluidized bed reactors should consider many aspects of vital importance for the sizing of equipment, for the biological process itself and for the operation of the system. One of the aspects to consider is the cost of the material since, as the material becomes more specific for the application, the cost increases influencing the economy of the system.

Physical characteristics: Size. The size of the particle influences the available surface for attachment as well as many characteristics of fluidization (and mass transfer). This is one of the design variables which can be easily modified. Figure 2 is a plot of the specific area of spherical particles vs. diameter and the influence of the size on the superficial liquid velocity needed to achieve a fluidized bed with a 20% expansion over the fixed bed.

In order to reduce the operating costs the superficial velocity should be kept at low values, forcing the use of small size particles, which also provides greater surface area available for colonization. Recommended sizes range between 0.1 and 0.7 mm (100-700 μm), since smaller sizes are difficult to operate.

Most of the materials used are natural (crushed rocks, sand...). That means the size as well as the shape are not regular. Size distributions must also be considered. Broad distribution material is cheap and easily available but it adds some problems to the operation of the system. Very narrow cuts are expensive and a compromise should be made. The influence of the size distribution will be discussed later.

Physical characteristics: Shape. All fluidization equations were calculated or derived from equations applied to spherical particles. When particles are not spherical, a correction factor called sphericity (Φ) (defined as ratio between the surfaces of a sphere of equal volume and the actual surface of the particle) is used. The natural origin and the initial porosity of the material affect this value as well as the biofilm, covering mature bioparticles. The experimental determination of this
parameter is difficult, even more so when the gelatinous biofilm is adhered to the particle. As a common approach, the value is estimated at around \( \Phi = 0.75 \).

Other authors (Chen et al., 1985) use a model where particles are considered ellipsoids where their cross-section is circular.

The Dynamic Shape Factor DSF from Briggs (Fdz.-Polanco and Díez, 1988) is less often used, although it is directly related to fluidization/settling processes. DSF is defined as a relation between terminal settling velocities of the particle and a sphere of equal equivalent diameter.

Another aspect of particle shape is the spatial distribution of the biofilm. One of the assumptions taken during model development is that the biofilm is uniformly distributed along the carrier, forming a layer of equal thickness. This is a very rough approach since visual observation shows biofilm accumulation filling the crevices and holes (where shear forces are smaller) and with bald areas where exposure is greater. With developed biofilms, particles show a round aspect, closer to a sphere than the initial particle.

Physical characteristics: Density. One of the important variables for design is material density since it affects the hydrodynamics of the fluidized bed and has a direct connection with power consumption. Since most of the materials are porous, some differences should be established between the densities measured with wetting and non-wetting fluids, and the bulk density.

Water usually gets through the pores of the material, reducing the effective density of the particle. In order to release all air trapped at the pores it is recommended to perform the measurement under vacuum, and tapping the container to help releasing the air (Hulshoff Pol et al., 1986). This trapped air/water also affects the values of initial porosity, although after colonization most of the pores get clogged with biomass and are not available anymore.

The density affects the fluidization of the bed and its effect is shown in Figure 3. It can be seen that for particle density closer to the density of the fluidizing liquid, the superficial velocity for minimum fluidization conditions and for 20% expansion become closer. The hydrodynamic control of the bed is, thus, more difficult.

Another aspect to be taken into account is that the mass transfer coefficients have a strong dependency not on the density of the carrier itself but on the difference of density with the fluidizing liquid. Riba (1978) defined a dimensionless number \( (p_i - p_f)/p_i \) and found an experimental relationship between this number and the Sherwood number as an exponent equal to 0.37 (Riba et al., 1978).
Physical characteristics: Hardness (Carrier strength). It is important to remember that the particles in a fluidized bed are in continuous motion, suffering collisions with other particles and the walls of the reactor. If the material used is brittle, the particles can break and the newly formed particles have different fluidization characteristics, which can make the control of the expansion even more difficult.

Hardness is, though, not directly correlated with density. Most of the materials currently used for beds are highly porous, showing an overall density much lower than non-porous particles of the same material.

Physical characteristics: Superficial area (specific area). One of the advantages of fluidized bed reactors is the large surface available for attachment, where biofilm can grow, increasing the concentration of biomass.

The use of porous materials such as GAC, sepiolite, pumice, kaolinite, Arlita (R) offer also the advantage of internal pores which, depending on their size, can also be colonized. The amount of available surface grows then by two or three orders of magnitude, although it is needed to make a clear distinction between pore sizes. Messing (1982) studied the colonization of ceramic materials and found the bacteria usually attach by their minor dimension, leaving maximal bacterial surface available for mass and heat transfer. The optimal pore size found was around five times the major dimension of the cell, leaving space for two cells attached and two cells growing by fission, and some free space between them for substrate transfer.

Physical characteristics: Roughness. The roughness of the solid is important in the initial steps of biofilm formation. The bacteria are able to form bonds with the coating film around the particle. If the surface is smooth, the chances of the particle to remain attached after 'shocks with other particles or caused by shear forces of the liquid are scarce. When the surface is rough, the bacteria attached in crevices of the material are protected from shocks and shear forces and the probability of success is high. The presence of inhomogeneous biofilms is a constant in fluid bed reactors (Heijnen, 1984).

Chemical Properties

Chemical inertia. The material should be able to withstand the chemical attack of the common products found in the wastewater. The nature of the material has been found to be important in some studies, since the material (Sanz, 1987) can provide some trace nutrients to bacteria. Other effect is the impact of the biofilm upon the material. On some occasions the biofilm growth causes cracks on the particles which eventually lead to breaks (Oakley et al., 1985).

Adsorption. Although the main cause of adsorption is physical, particles are usually covered by a biofilm and thus the adsorption properties change dynamically. It has been found that the nature of the material clearly influences the start-up period. The velocity of colonization of different support materials increases from glass to zeolites, activated carbon and sepiolite (Huysman et al., 1983). The amount of macro- and micro-pores on the last of the supports as well as its roughness are important factors for its ability to get a fast colonization (Beefink and Staaugard, 1983) (Shimp and Pfaender, 1982).
Selection of Operating Expansion and Calculation of the Liquid Superficial Velocity

Once the material nature and size are defined, the values of fluidization characteristics ($u_m$, $u_t$, $\varepsilon_m$) can be measured or estimated.

In order to calculate the upflow velocity needed to achieve a fixed bed expansion, we can use a very simple graphical method based on easily measured variables in the lab, or using one of the equation previously described (Richardson-Zaki, Garside-Al Dibouni).

Bed and Equipment Sizing

Once the parameters which define the operating conditions are known it is straightforward to make calculations and obtain some other design values. These parameters are:
- Fluidization characteristics (superficial velocity and expansion),
- Influent flowrate and concentration, and
- Estimated hydraulic residence time.

The Volume of reactor is defined by the expression:

$$\tilde{t} = \frac{V}{Q}$$  \hspace{1cm} (10)

where $\tilde{t}$ is the hydraulic residence time,
$Q$ is the influent flowrate, and
$V$ is the reactor volume.

The second calculation, H/D ratio possesses some problems which should be addressed.

One of the problems associated with fluidized bed reactors is flow distribution, not only for its contribution to the homogeneity (and, thus, efficiency) of the system but its more important role as producer of the fluidization. The mechanical design of distribution systems will be discussed later. Due to the difficulty of the problem, most researchers use very long and thin reactors for lab and pilot scale, which are very difficult, if not impossible, to scale-up. The H/D ratio affects also the recycle ratio, because the superficial velocity ($u$) should be maintained and, thus, has a direct economic impact.

In order to reduce pumping costs, the recycle should be kept at a minimum, taking into account for that minimum, the dilution factor. In other words,
there is a compromise between dilution factor and recycle factor, affected by the H/D ratio of the reactor, then affected by the capability to achieve a good flow distribution/ fluidization. An abstract of this discussion is outlined on Table 2.

<table>
<thead>
<tr>
<th>Dilution</th>
<th>Pumping</th>
<th>Distribution</th>
<th>Cost</th>
<th>Footprint</th>
</tr>
</thead>
<tbody>
<tr>
<td>High H/D</td>
<td>high</td>
<td>less</td>
<td>easy</td>
<td>more</td>
</tr>
<tr>
<td>Low H/D</td>
<td>low</td>
<td>more</td>
<td>difficult</td>
<td>less</td>
</tr>
</tbody>
</table>

'Footprint: Bottom surface area.

The calculations involved on the sizing are simple.

Since the fluidization flow rate is composed of the influent stream $Q$ and a complementary stream of effluent $E$

$$R = Q + E \ (m^3/h)$$  \hspace{1cm} \text{(11)}

As the superficial velocity is a design parameter,

$$\frac{R \ (m^3/h)}{u \ (m/h)} = \text{Reactor surface} \ (m^2)$$  \hspace{1cm} \text{(12)}

and from this,

$$\frac{V \ (m^3)}{S \ (m^2)} = \text{Height of the reactor} \ (m)$$  \hspace{1cm} \text{(13)}

Soung (1978) and Euzen et al., (1981) found the bed porosity is affected by the size of the reactor, and the way it is scaled-up. To avoid this effect, pilot- and full-scale reactors should be taller than 4 m and wider than 0.15 m. The same recommendation is found for solid-gas fluidized bed reactors, where diameter should be wider than 0.5 m (Werther, 1980).

Evaluation of the Effects Caused by Biofilm

Once the reactor is started-up, the particles are covered partially or totally by a biofilm which is growing continuously. The biofilm is affected by two opposite phenomena:

- biofilm growth, and
- biofilm detachment

which are interrelated.

When the biofilm grows to a thick size, chances are that collision with other particles, internal gas production or diffusion limitations (which will lead to substrate deficiency in the inner layers of the biofilm) will cause biofilm detachment from the carrier. The detached biofilm can be washed out from the bed and the reactor, can stick to other particles or can lead, if conditions are adequate, to the formation of a non-carrier granule. Both non-carrier granules and thick biofilm particles have different fluidization characteristics than the remaining particles. The size can be different but the density is definitely different and, for particles with carriers denser than water, lighter than the other particles. As already discussed on the basics of fluidization, the difference in density can produce, under some conditions, segregation of the bed. These particles travel up to the top of the bed, where shear forces caused by the violent gas eruption can lead to biomass sloughing and a subsequent change in characteristics.
This self-regulatory mechanism is in many cases not able to cope with the biofilm growth and should be helped by mechanical equipment, which separates the carrier and the biofilm and re-introduces the uncovered material back to the bed. With this system hydrodynamic control of the bed expansion is greatly eased and the system performs closer to design specifications.

**Impact of Gas Production/Release**

Two main effects should be noted.

1.) The increase in turbulence (and, thus, mixing) caused by the rising bubbles which can also affect the biofilm sloughing, and

2.) The bed contraction effect. This effect should be taken into account in order to keep the hydrodynamic conditions of the bed, if bed-level meters are used to control the recycle ratio, since it can produce deviations from the design set-points.

**MECHANICAL DESIGN**

**Distribution System**

The design of the distribution system is the key point of a fluidized bed reactor. The performance and behavior of the reactor relies on the capability of distributing the incoming flow through the active surface of the bed. Most of the details of full-scale reactors are covered by patents and/or proprietary design policies and these are not commonly known. Although for small diameter reactors an inverted cone with a downward pipe producing a downward-and-then-upward flow provides satisfactory results, large diameter reactors cannot use this system (with the exception of full-tapered reactors). The most common system is composed of an annular, star-shaped or squared piping network with nozzles directed towards the bottom of the reactor, providing jets which cause high turbulence areas and help prevent the formation of dead zones.

During the development of industrial fluidized bed reactors, a design based on trenches and longitudinal pipes was created. A great advantage of this system is that the design is easily scaled-up as needed, extending or incrementing the number of trenches. However, this system increases the costs of the system and it is not used anymore.

The distribution system causes some head losses, which are important from the economic side of the operation. The literature on gas/solid fluidized bed reactors is clear on this aspect and indicates that a good distribution systems should produce a head loss at least equal to the total head loss caused by the bed (Leva, 1959). For liquid/solid reactors, this does not apply and head losses found on pilot reactors are a fraction of the total losses caused by the bed (Iza, 1987).

**Reactor Body**

Due to the problems associated with flow distribution and the need to reduce recycle ratio (due to its economic impact), fluidized bed reactors are usually long thin columns instead of fat short tanks. Most of the fluidized bed reactors used for municipal sewage are made of concrete, based on square or rectangular tanks. The reactors for industrial waste waters are essentially cylindrical, and usually made of stainless steel (304L or 316L, depending on the aggressiveness of the raw influent) or reinforced plastics.
One of the most viable alternatives to this configuration is the tapered fluidized bed (Shi et al., 1984; Pitt et al., 1978). This type of reactor produces a natural gradient of upflow velocity and thus allows a perfect segregation of the particles along the vertical axis. The slope of the reactor is very important, since it can cause the particles to get stuck together and produce a spouted bed, with a high turbulence and shear area close to the bottom of the reactor. Another problem of this design is a mechanical aspect since tapered reactors are more difficult to build than cylindrical tanks.

One of the aspects commonly forgotten during the design is the fact that the distributor liquid jets are directed towards the bottom plate of the reactor. Since this area has a extremely high turbulence, shear forces are important and the particles found there are almost clean particles or covered with a very thin biofilm. The continuous impact of the liquid and particles on the reactor bottom plate produces an important erosion effect which can also be enhanced by chemical attack and eventually leads to a plate perforation and bed leakage.

**Recycle Piping**

One of the aspects to consider is how to connect the influent and recycle pumps. Until now, since most of the beds work with solids much denser than water, the recycle flowrate is usually several times bigger than the influent flowrate. It seems to be adequate to introduce the influent stream just before the recycle pump. In this way, the pump consumption on the influent pump is greatly reduced since it works against a very low overpressure (even underpressure) and the recycle flow (sum of influent and effluent pick-up) can be easily managed to keep the expansion of the bed under control. This configuration also permits changing the influent flowrate without affecting the hydrodynamic control of the bed and to manage hydraulic overloads.

This system will work properly even with low recycle ratios. When using low density carrier particles or granules, and depending on the influent concentration, the system can even work without any recirculation. This possibility should also be considered during design and adequately addressed with the incorporation of check or shutter valves. It is also a good practice to install a "sand" trap along the recycle line to avoid particles entering the recycle pump and causing abrasion of the pump parts. This is more important when the recycle flow rate is high and can cause some suction on the inlet ports at the top of the reactor. Stainless steel or rubber-lined impellers should be used for recycle pumping.

**Gas Collection System**

Due to its intrinsic high-rate system characteristic and its (usually) high H/D ratio, fluidized bed reactors have an extremely high superficial gas flowrate. The evolved gas of a highly loaded reactor resembles a pot of boiling water. The presence of any kind of surfactant causes big foaming problem which should be corrected to avoid clogging of the gas pipes and deterioration of the effluent quality. Systems with bells, like the Gas-Solid-Separator (GSS) device from the UASB have been used, with special provision for showers (sprays to break up the foam) or antifoaming agents dosage.

**CONCLUSIONS**

The fluidized bed reactor offers as an advantage in respect to other systems a very high concentration of active biomass, attached to a dense carrier, which cannot be easily washed out.
The mass transfer properties of the fluidized bed allow the treatment of very dilute influents.

As the energy required to keep the bed fluidized is a key economic factor, the trend is to use lighter particles, which make the hydrodynamic control of the bed more difficult and reduce the mass transfer characteristics of the system.

There are no problems of plugging or channeling due to solids accumulation, since the bed expands to hold the excess material, but an adequate distribution of the influent is a rather difficult task.

The use of recycle for fluidization purposes also offers other advantages for the process: it provides alkalinity to neutralize the influent, reduces its concentration and smooths spikes of toxicants or inhibitory compounds.

The scale-up of the systems is very difficult and many compromises between technical and economic aspects must be made.

The very high loading rate achieved on these systems is in many cases related to very low retention times. This means that these systems must be closely monitored.

The restart-up of the systems is very easy and long seasonal shut-downs do not affect the capacity of the system.

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