

FS-BIO-007

**TECHNOLOGY FACT SHEETS
FOR EFFLUENT TREATMENT PLANTS
OF TEXTILE INDUSTRY**

FLUIDIZED BED

SERIES: SECONDARY TREATMENTS

TITLE	FLUIDIZED BED (FS-BIO-007)
Last update	December 2014
Last revised	

FLUIDIZED BED (FS-BIO-007)

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Last update	Date	Done by:	Update main topics

INDEX

1.- DESCRIPTION

1.1.- *Specific surface area of carrier material*

2.- DESIGN

2.1.- *Upward velocity*

2.2.- *Removal efficiency*

3.- SECONDARY CLARIFICATION

4.- REQUIRED AREA ESTIMATION

5.- SPECIFICATIONS FOR THE TREATMENT OF WASTEWATER IN TEXTILE INDUSTRY

6.- PARAMETERS AND CONTROL STRATEGIES

7.- OPERATION PROBLEMS

BIBLIOGRAPHY

ANNEX 1.- UNIT PROCESSES GRAPHIC DESCRIPTION



1.- DESCRIPTION

This fact sheet describes the design criteria and operation of fluidized bed reactors in order to oxidize organic matter in waste water from the textile industry.

Fluidized beds are biofilm reactors with a wide range of applications in biological aerobic, anoxic and anaerobic treatment. These systems employ small particulate materials as a support medium for the growth of the attached biocenosis. Support-biofilm assembly (bioparticle) is kept in suspension inside an upward vertical flow whose velocity is high enough to overcome the force of gravity. Bioparticles are in continuous relative movement but are not transported by the flow, that is, they are not washed off from the reactor. Applications of this technology include: anaerobic digestion, oxidation of organic matter, nitrification and denitrification of industrial and urban wastewater.

An upward water flow through a sand bed, granular activated carbon, anthracite, polypropylene particles, etc., when it circulates at high speed, will cause fluidization. The support material of a fluidized bed has an extremely large specific surface, and reaches in minutes the treatment level that any conventional biological treatment process achieves in several hours. Bioparticles suspension maximizes the surface contact between the microorganisms and the waste water.

The most typical fluidized reactor bed consists in a bed of great height, whose lower part introduces water, through a distribution system, at a high enough velocity to fluidize or expand the bed. Anoxic and anaerobic systems are simpler in design. Conversely, aerobic systems require aeration. This aeration is normally conducted in the effluent recirculation line, resulting in a 2-phase system: solid and liquid, as diagrammed in Figure 1. The advantage of introducing the air by the recirculation line is that biofilm is not subject to abrasion resulting in an effluent with low suspended solids concentration.

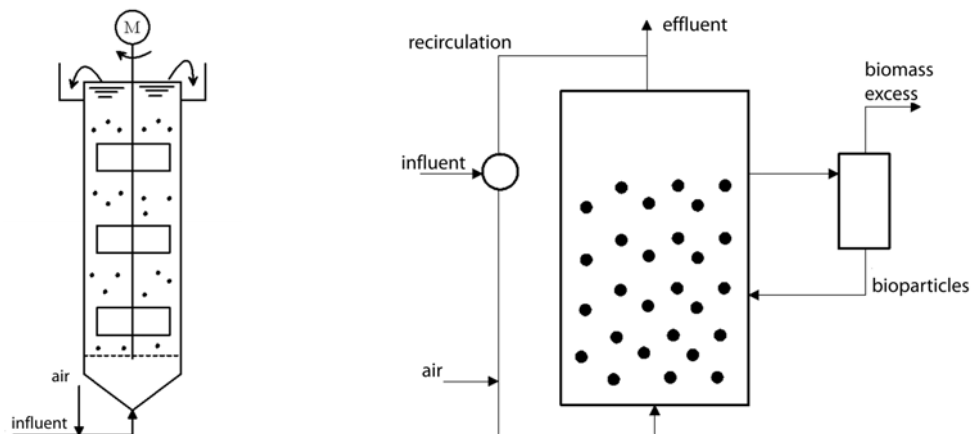


Figure 1.- Typical 2-phase fluidized bed reactor scheme. Plug flow (left) and pseudo-mixed flow reactor (depending on the recirculation rate).

Depending on the expansion degree of the bed, the process is called "expanded bed" or "fluidized bed". The transition between the two systems ranges between 50 to 100% expansion with respect to the fixed bed reactor (WEF 2010). A lesser extent of bed expansion is advantageous because it requires lower flow velocity, less energy and increases the effective biomass concentration (mg SS/L) which reduces the space demand. However, in aerobic processes, the oxygen demand increases as the biomass concentration grows.

Generally, the speed required for fluidization is much higher than that required to achieve the retention time for the biological reaction, so the bed effluent has thereby to be recycled, increasing the upward flow velocity.

Recent developments in the design process have enabled the incorporation of a gas phase, allowing transfer of oxygen directly into the bioreactor (Fig. 2). If aeration is carried into the bed, disruption of fluidization occurs and increases collisions between bioparticles producing biofilm detachment. In these types of reactors the fluidization of bioparticles is achieved by aeration (3-phase systems: solid, liquid and gas), enabled by the use of support material with specific weight similar to water's. Although the popularity of these systems is rising, 3-phase systems process analysis and modeling is very complex. Additionally, the fact of increased erosion of the biofilm by collision necessitates a liquid-solid phase separation unit as biofilm detachment can increase the concentration of suspended solids in the effluent. Thus, the present work is focused on the bi-phasic systems.

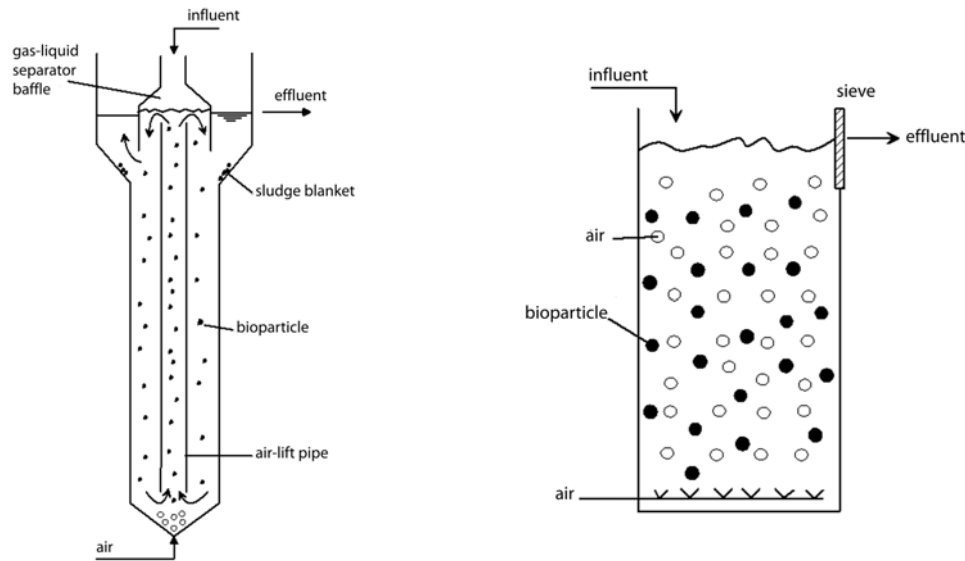


Figure 2.- 3-phase fluidized bed reactors schemes. Air-lift fluidized bed with integrated sedimentation chamber (left). Fluidized bed with bottom aeration diffusers and screening unit so as to separate bioparticles (right).

In general, the sedimentation rate of the bioparticles is far superior to that of activated sludge flocs. Moreover, bioparticles are retained in the reactor and the effluent has a very low concentration of suspended solids what allows its discharge without clarification. However, as the biomass grows, bioparticles may turn larger, increasing bed expansion. To avoid excessive expansion of the bed, which could increase the suspended solids in the effluent, the bioparticles are routinely removed, and a separating unit, for example a sieve, purges excess biomass and clean support particles are returned to the reactor. Thus, a stable amount of biomass can be maintained in the system while the effluent contains low concentration of suspended solids.

1.1.- Specific surface area of support material

As support material silica sand (diameter 0.3 to 0.7 mm) or granular activated carbon (0.6 to 1.4 mm) are commonly used. Other materials may also be used. For example, vitreous coke was used at pilot scale (0.7 to 1.0 mm). In any case, since they are small sized particles (1 mm), a large surface area is obtained (up to 2400 m²/m³ when the expansion is 50%, WEF 2010), which is one of the key factors of this technology. When the bed is expanded, the surface area of the support particle can be estimated by the following formula (Rittmann and McCarty, 2001):

$$a_s = 1000 \frac{6 (1 - \varepsilon)}{d \Psi}$$

Where:

a_s = specific surface area (m⁻¹)

ε = expanded bed porosity (dimensionless, typically around 0.4 - 0.6)

d = support particle diameter (mm)

Ψ = form factor (dimensionless, equal to 1 if considered as pseudo-spherical particle)

The specific surface area in fluidized beds depends on the size of granular media (0.3 to 1.4 mm) and the bed expansion degree (from 50 to 100%), but generally ranges from 1000 to 3000 m²/m³. This large surface area allows biomass concentrations of 15,000 mg VSS/L in aerobic beds or reaches up to 40,000 mg VSS/L in anoxic beds (Grady et al., 1999; Grady et al., 2011) (Table 1). Thus, the volumetric efficiency of a fluidized bed is 10 times higher than that of an activated sludge system. Due to this high concentration of biomass, a fluidized bed with very low retention time (about 3 minutes) can be very effective as denitrification process, even with high nitrate loads (>7 kg N-NO₃/m³/d) (Green et al., 1994).

Table 1. Concentration of biomass in fluidized beds (WEF 2010).

Treatment	MLVSS (mg/L)
Organic oxidation	12.000 – 15.000
Nitrification	8.000 – 40.000
Denitrification	30.000 – 40.000

2.- DESIGN

In fluidized bed reactors, both the amount of biomass retained and the total biofilm area are very large because biofilm support particles are of small diameter (0.5 to 1.4 mm). Therefore, they are able to treat wastewater with an extremely short hydraulic retention time or at high organic load. Thus, fluidized bed reactors can be operated with a retention time of just over 10 minutes, or with a volumetric load of about 10 kg BOD₅/m³/day.

On the other hand, careful operations are required to form and maintain a stable fluidized bed. To maintain a fluidization state in a longitudinal reactor as shown in Figure 1, it is necessary to thoroughly evaluate the upward flow velocity. If the upward velocity is greater than the free settling of a single particle, then the particle will be ejected from the biotower. Therefore, the upward speed needed to a proper fluidized bed operation will be within a narrow range between the minimum fluidization velocity and the free settling velocity of the particles composing the bed.

Minimum fluidization velocity

One of the expressions for the minimum fluidization velocity is as follows (modified from Iwai and Kitao, 1994):

$$V_{mf} = 0.09843 \frac{d_{60}^{1.82} [\rho(\rho_s - \rho)]^{0.94}}{\mu^{0.88}}$$

Where:

- V_{mf} = minimum fluidization velocity (m/h)
- d_{60} = 60% finer size of the filter media (mm)
- ρ_s, ρ = specific weight of the filter medium and the water, respectively (lb/ft³)
- μ = viscosity of water (centipoise) (at 20 °C = 1,0020)

Another formula to estimate V_{mf} (in m/h) is given by the following equation (modified from Iwai and Kitao, 1994):

$$V_{mf} = 16,50 \frac{d^2(\rho_s - \rho)g}{\mu}$$

Where:

- d = diameter of the filter medium (mm)
- ρ_s, ρ = specific weight of the filter medium and the water, respectively (g/m³)
- μ = dynamic viscosity of water (g/m/h)
- g = acceleration of gravity (m/h²)

Settling velocity of the bioparticles

Bioparticles, and clean support carrier materials, settle down by free sedimentation according to Stokes' law:

$$V_s = 200 \frac{g}{\mu} (\rho_s - \rho) d^2$$

Where:

- V_s = free settling velocity (m/h)
- d = diameter of the filter medium (cm)
- ρ_s, ρ = specific weight of the filter medium and the water, respectively (g/cm³)
- μ = dynamic viscosity of water (a 20 °C = 0.01002 g/cm/s)
- g = acceleration of gravity (= 9.8 m/s²)

2.1.- Hydraulic loading rate

The hydraulic loading rate results from:

$$HLR = \frac{F(1 + R)}{S_t}$$

Where:

HLR = upward velocity (m/h)

F = inflow (m³/h)

R = recirculation rate (fraction)

S_t = cross sectional area (m²)

Generally, the acceptable range of upward velocity is 30 to 60 m/h. The table presents examples of values depending on the filling material.

Table 2.- Hydraulic loading rate to fluidize different carrier materials (Adapted from WEF 2010).

Carrier material	Size (mm)	HLR (m/h)	HLR _{50%} (m/h)
Coke glass	0.7-1.0	9.0	30
GAC	1.7	10.2	--
Silica sand	0.5-1.0	22.2	90 (typical range: 24 to 36)

HLR_{50%}: Hydraulic loading rate for 50% of bed expansion

GAC: granular activated carbon

As biomass adheres to the support material, the apparent density decreases. Apparent density can be calculated from the amount of adhered biomass density assuming a microbial film is 1,005 g/cm³.

Despite the difference in densities between the fillers used (sand, anthracite, PVC, porous glass beads, polypropylene, nylon), the fact is that once developed biofilm, the bioparticle tends to a specific weight value similar and stable at around 1.1 (WEF 2010).

2.2.- Removal efficiency

Yields close to the upper limit of the reachable performance in biological treatments are attributed to fluidized bed reactors. This is due to the high biofilms microbial concentrations and efficiency factors developed in these processes (which are of limited thickness offering little resistance to mass transfer from the liquid phase).

To sum up, the organic matter removal performance will range from 70 to 95%. For nitrogen removal by nitrification-denitrification process, yields above 95% were achieved (Iwai and Kitao, 1994; WEF 2010).

3.- SECONDARY CLARIFICATION

Considering the internal aeration variant (air-lift and/or diffusers) the reactor effluent can have a concentration of suspended solids that do not comply with discharge common limits. In this case, a solid-liquid separation can be done by simple sedimentation; among other processes like: filtration through granular bed (sand, anthracite, etc.); cyclones; accelerated or lamellar settling; sieving; etc.

For simple sedimentation static circular or rectangular clarifiers can be used. Fluidized beds bioparticles concentration at the reactor outlet can reach or exceed 200-400 mg/L, being applicable a zonal sedimentation theory.

3.1.- Design variables

- Surface hydraulic loading rate: based on the real flow rate through the unit, that is, which goes by the discharge weir (outflow).

$$HLR = \frac{Q}{A_{HLR}}$$

Where:

HLR = surface hydraulic loading rate (m/h)

Q = outflow (m³/h)

A_{HLR} = horizontal surface of clarification (m^2)

- Hydraulic retention time:

$$HRT = \frac{V}{F}$$

Where:

HRT = hydraulic retention time (hours)

H = water depth side-wall (m)

V = useful volume for clarification (m^3)

$F = F_{max}$ (m^3/h)

- Weir overflow rate:** corresponds to the effluent flow rate per linear meter of outlet weir.

$$WOR = \frac{F}{W_L}$$

Where:

WOR = Weir overflow rate ($m^3/h/m$)

W_L = weir length (m)

$F = F_{max}$ (m^3/h)

3.2.- Summary of design values

The following table summarizes the typical values for design parameters.

Table 3. - Design values for secondary sedimentation effluent from fluidized beds.

Parameter	Value
HLR (m/h)	< 0.6 (F_{av}) < 1.5 (F_{max})
HRT (h)	> 2 (F_{max})
WOR ($m^3/h/m$)	< 10 (F_{max})
Sludge concentration (%)	≤ 1
H (m)	≥ 2.5

When the clarifier unit diameter is less than 5 meters, it is recommended to use truncated cone shaped clarifiers without scrapers, also called vertical flow clarifiers. In these decanters, the effective horizontal surface is set at the midpoint of the distance between the elevation of water entering the unit (ie, leaving the central baffle) and the elevation of the free water level (see figure below) .

In order to facilitate the real sludge sedimentation, the slope of the conical zone wall will respond to an inclination angle greater than or equal to 60°.

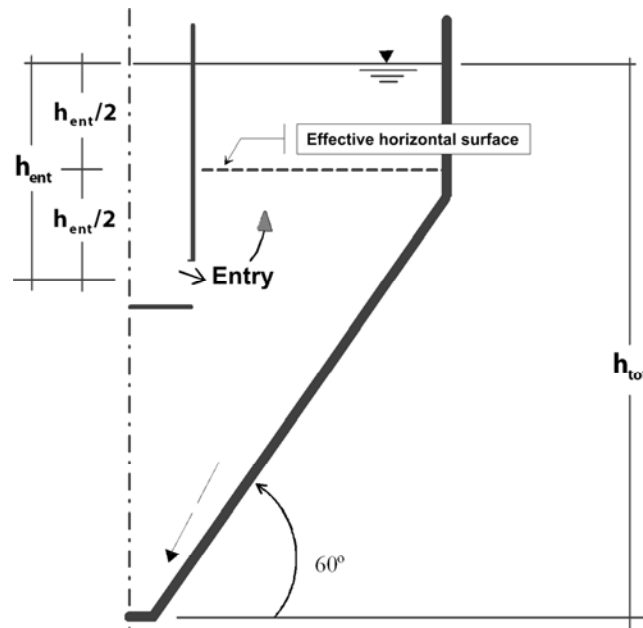


Figure 3.- Truncated cone shaped clarifier.

4.-REQUIRED AREA ESTIMATION

4.1.- Area requirements for biological fluidized bed reactor

The following table presents minimum area demand for a biological fluidized bed reactor filled with granular activated carbon (GAC) of 1.5 mm in diameter.

The area was calculated for different sizes of textile factories in terms of water treatment flow rate.

It is considered that the treatment prior to fluidized bed reactor includes: flow rate and concentration tank homogenization, screening, sieving and primary sedimentation. In this approach influent BOD₅ concentration to fluidized bed is theoretically 300 mg/L.

The main design criterion is the upward velocity, which shall not exceed 30 m/h, for a maximum expansion of 60%.

The floor area requirements only depend on the upward velocity and inflow. Thus, the following results were obtained:

a.- Starting variables	Reference	Adopted
Carrier material:		Sand
d (mm):	0.4-1.0	0.8
Expanded bed porosity	0.4-0.7	0.5
Form factor	> 1	1.8
Organic load (g/m ² /d)	5-10	5
Upward velocity (m/h)	24-36	36
Influent BOD (mg/L):		400
H minimum (m):	2	2
H recommended (m):	6	5
b.- Specific surface area of the expanded sand		
As (m ⁻¹):		2083

Table 4.- Area, volume, height and recirculation rate of expanded bed reactor

Q	Area	Volume	Height	R
(m ³ /d)	(m ²)	(m ³)	(m)	
400	3.07	15.36	5.0	0.01
1000	7.68	38.40	5.0	0.01
2000	15.36	76.80	5.0	0.01
4000	30.72	153.60	5.0	0.01

4.2.- Area needed for the secondary clarification

So as to estimate the required clarification area, the hydraulic loading rate 0.6 m/h at average flow rate is used as design criterion.

Extracted results are presented in the following table:

Table 5.- Area requirements estimation for the secondary clarification.

Inflow	Area
(m ³ /d)	(m ²)
400	28
1000	69
2000	139

4.3.- Total area required for secondary treatment

Finally, the area needed for the "secondary treatment" is obtained by adding the area of reactor over the settling. Results are presented in the table below:

Table 6.- Total area necessary for the secondary treatment (reactor + clarifier).

Inflow	Total area
(m ³ /d)	(m ²)
400	30
1000	71
2000	142

5.- SPECIFICATIONS FOR THE TREATMENT OF WASTEWATER IN TEXTILE INDUSTRY

In the textile sector, the majority of applications are developed by anaerobic beds. It takes advantage from the high organic matter concentration to generate biogas. Moreover, depending on the bed material, it can achieve color removal objectives. The following paragraphs provide a collected chronological review of fluidized beds applications in textile industry.

Kim et al. (2002) evaluated the effectiveness of a pilot plant designed to treat real wastewater from a textile factory whose average pollutant concentrations were COD 870 mg/L, color 1340 mg/L Pt-Co and SS 240 mg/L. The pilot plant treatment line comprised: aerobic fluidized bed + chemical coagulation + electrochemical oxidation. The reactor had a volume of **1.8 m³** and diffusers aeration inside the tank, that means, it was a **3-phase system**. Support materials were polyurethane 1.3 cm edge cubes, with a density of 0.21 g/cm³. The reactor reached a concentration of 570 mg MLSS/L, and 6 days of cellular retention time (CRT). Dissolved oxygen (DO) is always maintained between 3 and 4 mg/L in the reactor. With these features yields observed in the removal of COD and color were 68.8% and 54.5% respectively.

Sen and Demirel (2003) evaluated the treatability of a real wastewater from cotton textile industry in Ankara with an anaerobic fluidized bed bioreactor at laboratory scale. The support material consisted on pumice stone particles with a diameter from 0.25 to 1.4 mm and a density of 1764 kg/m³. With a HRT of 24 hours and an organic load of 3 kg COD/m³/d; COD, BOD₅ and color maximum removal performances (in %) were 82, 94 and 59%, respectively. In order to achieve such performance, it was necessary to provide glucose at a dose of 2 g/L as external organic material. A higher dose of 2 g glucose/L did not result in improved process efficiency.

Georgiou and Aivasidis (2006) evaluated the laboratory scale treatability of real textile industry wastewater with anaerobic fluidized bed. This treatment had two specific objectives: wastewater discoloration and the transformation of non-biodegradable reactive azo dyes into amino-aromatic aerobically degradable dyes. With a **HRT of 6 hours** the process was effective for complete discoloration of the wastewater together with obtaining an effluent highly biodegradable by activated sludge process aerobic microorganisms. The anaerobic fluidized bed support material consisted in **porous sintered glass beads** (Siran®) of **1 to 2 mm in diameter**. This is a porous glass, with a 55% porosity and pore diameters between 60 and 300 microns.

Haroun and Idris (2009) also performed a similar evaluation of an anaerobic fluidized bed at laboratory scale to treat wastewater from a fiber dyeing factory in Malaysia. The difference between both investigations, apart from raw water characteristics, is that the bioreactor used **activated carbon particles of 0.02 to 0.25 mm** in diameter as support material. With a **HRT of 12 hours** and an organic load of **4.4 kg COD/m³/d**, maximum achieved yields in COD, BOD₅ and color removal were (in%) 98, 95 and 65, respectively. Also, glucose was employed as external carbon source, with an optimal dose of 0.6 g/L.

Desai and Kore (2011) carried out the control and monitoring of a biological treatment system comprising two aerobic fluidized bed bio-reactors in series for the treatment of textile waste water which had previously passed through pH regulation and coagulation-flocculation units. One of the interesting things about this study is that it was a full-scale plant, located in Kolhapur (Maharashtra, India). The average observed yields (considering the full treatment line) in **COD and BOD₅ removal were 85% and 82%**, respectively. Unfortunately, the article does not incorporate the operational variables of bioreactors or the characteristics of the support material.

Nzila et al. (2011) evaluated the color and COD reduction of textile wastewater with an anaerobic fluidized bed. Raw water was subjected to coagulation and flocculation treatment prior to the bioreactor. The bed was operated with an **HRT of 12 hours** and an organic load of **2.6 kg COD/m³/d**. The bio-reactor reached yields of 40 and 68% for COD and color, respectively. The overall system performance averaged a 72 and 87% reduction in COD and color respectively.

Cases from other industry sources

In 1999, more than 80 full-scale facilities were in operation in the US and Europe (WEF 2010). Two thirds of these were dedicated to the treatment of industrial wastewater, and the rest to urban wastewater.

But, the boom of this technology is recent. In particular, the anaerobic systems have found a market for the treatment of industrial wastewater strong concentration, with COD values greater than 1000 mg/L, since the use of oxygen is prevented and can take advantage of the biogas produced.

Jeris et al. (1977) reported the results of a research pilot nitrification, denitrification and BOD removal of urban wastewater from fluidized bed units with surface areas greater than 3,300 m²/m³ level. The process employed pure oxygen aeration in the fluidized bed influent and effluent recirculation. Nitrification and denitrification processes reached a maximum 99% removal yield.

Shieh and Li (1989) treated wastewater from cornstarch industry using a fine sand support medium in the fluidized bed reactor. The reactor was operated with mass loads of 0.42 to 1.61 g BOD₅/g TVS/day. A good yield of coupled organic matter and ammonium oxidation were reached (removal efficiencies higher than 90%), when mass load and CRT were set at 1.0 BOD₅ g/TVS g/day, and 5 days, respectively. The biomass concentration in the reactor was high and ranged between 5000-16000 TVS mg/L.

6.- PARAMETERS AND CONTROL STRATEGIES

The maintenance of these facilities is not as simple as that of fixed bed reactors (submerged bed biological reactor, trickling filters, etc.), since it is difficult to maintain the reactor bed at a stable fluidized state.

Controlling the height of the bed requires a continuous purge of biomass. Otherwise, bioparticles without proper size can be generated and of the expanded bed height may become excessive. The most common procedure is to purge the biomass from the bed top zone. This induces to bed stratification. The bed is maintained in a dynamic state by continuous purge of large bioparticles on top and a return of clean support material, which migrates to the bottom of the bed where it contacts with a high concentration substrate. This assumes that the support material has good size uniformity. If the material has poor size distribution, then the stratification conditions are not related with the bioparticles size. In the latter case, the larger support media would remain near the bottom of the bed and the smaller would accumulate on top. Consequently, this material could be always purged and recycled, which is clearly inadequate. Thus, it is very important that the support material has excellent size uniformity.

6.1.- Air supply control

In biological reactors, dissolved oxygen (DO) concentration control is a point of great importance. For this purpose, a fixed or portable DO probe is commonly employed. The optimum DO concentration in order to reach an efficient oxidation of organic matter ranges between 3 and 4 ppm. Aeration in large plants aeration is usually automated, so that, depending on the DO probe measuring values, the aeration equipment will start or stop. In addition, air supply flow will be regulated if there are frequency inverters available

6.2.- Clarification and purge control

When more than one clarification line is installed, a proper distribution of reactor effluent among all clarifiers must be ensured.

Furthermore, sludge on the secondary clarifier should be prevented from a long retention time. With this purpose, sludge purge pumping times should be regulated. In general, an hourly purge is recommended.

Even though an installation is working properly, a certain amount of detached biofilm and/or low density flocs will float to the top of the clarifier. A surface baffle prevents these materials from flowing out the unit together with the treated water.

6.3.- Daily maintenance operations in the reactor and the secondary clarifier

Tasks to control and perform are:

- Observe water appearance in reactors and decanters.
- Adequate maintenance and lubrication of aeration unit.
- Brush clarifier's outlet weirs.
- Removal of oil and grease and other floating materials such as pieces of rubber and plastic.

7.- OPERATION PROBLEMS

The key advantage of the fluidized bed is its high surface area for biofilm growth. This produces a high concentration of active biomass, high reaction rate and reduced area requirements. However, due to the high concentration of biomass, aerobic processes may be limited by the oxygen demand.

Another disadvantage is the recirculation degree required to maintain an upward velocity which allows bed expansion and fluidization of bioparticles, increasing the energy costs for pumping. However, the recirculation pump must overcome only the friction losses and the density difference between the fluid in the recirculation line (air and water) and expanded bed (water and bioparticles). The pressure loss in the recirculation pump is significantly smaller than the height of water in the reactor.

Most problems in the distribution elements can be attributed to clogging issues. This can be avoided by removing solids in the influent and designing an element that prevents the return of the support material. However, whenever the formation of struvite is possible (anaerobic digestion systems), removable systems must be designed for maintenance operations.

Granular activated carbon (GAC) has some properties better than those of silica sand, for example, low density, high porosity, good adsorption, biofilm is developed uniformly throughout the bed and easy start or re-start of the process. On the contrary, facing upward speed variations, the stability along the bed height increases with the size and density of the support material (that is, in such case sand has a better performance).

Moreover, anaerobic processes involve the following problems and/or disadvantages:

- Anaerobic bacteria are slow-growing, and therefore at any event, like hydraulic and/or contaminant overload, the reactor treatment capacity has a slow increase.
- It may require heating equipment when biogas production decreases.
- Most part of the process elements, if not all, require to be installed inside a shed.
- Although low COD effluent concentrations are achievable (up to 100 mg/L, depending also on influent COD), it is always safer to get better quality effluent with an aerobic process.
- The operation of the anaerobic process should be performed by highly trained and experienced staff.

Stratification of the bed produced when the size of the support material is not sufficiently uniform is a common trouble. Smaller bioparticles will be accumulated near the top of the bed, being less exposed to detachment, what generates biofilm accumulation resulting in a density decrease in relative terms. As a consequence, they get a higher fluidization and the situation continues over the time. The final problem occurs when this light bioparticles are carried out the reactor (i.e. with an hydraulic overload). The use of a medium with good uniformity coefficient minimizes and/or effectively prevents stratification of the bed. Other control measures include the design of a conical section at the top of the reactor so as to settle down light particles, sieve installation, etc.

In terms of costs, fluidized beds investment can result in savings of 50% compared to activated sludge, but operating costs are much higher (WEF 2000).

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ANNEX 1

GRAPHIC DESCRIPTION OF UNITS OF PROCESS

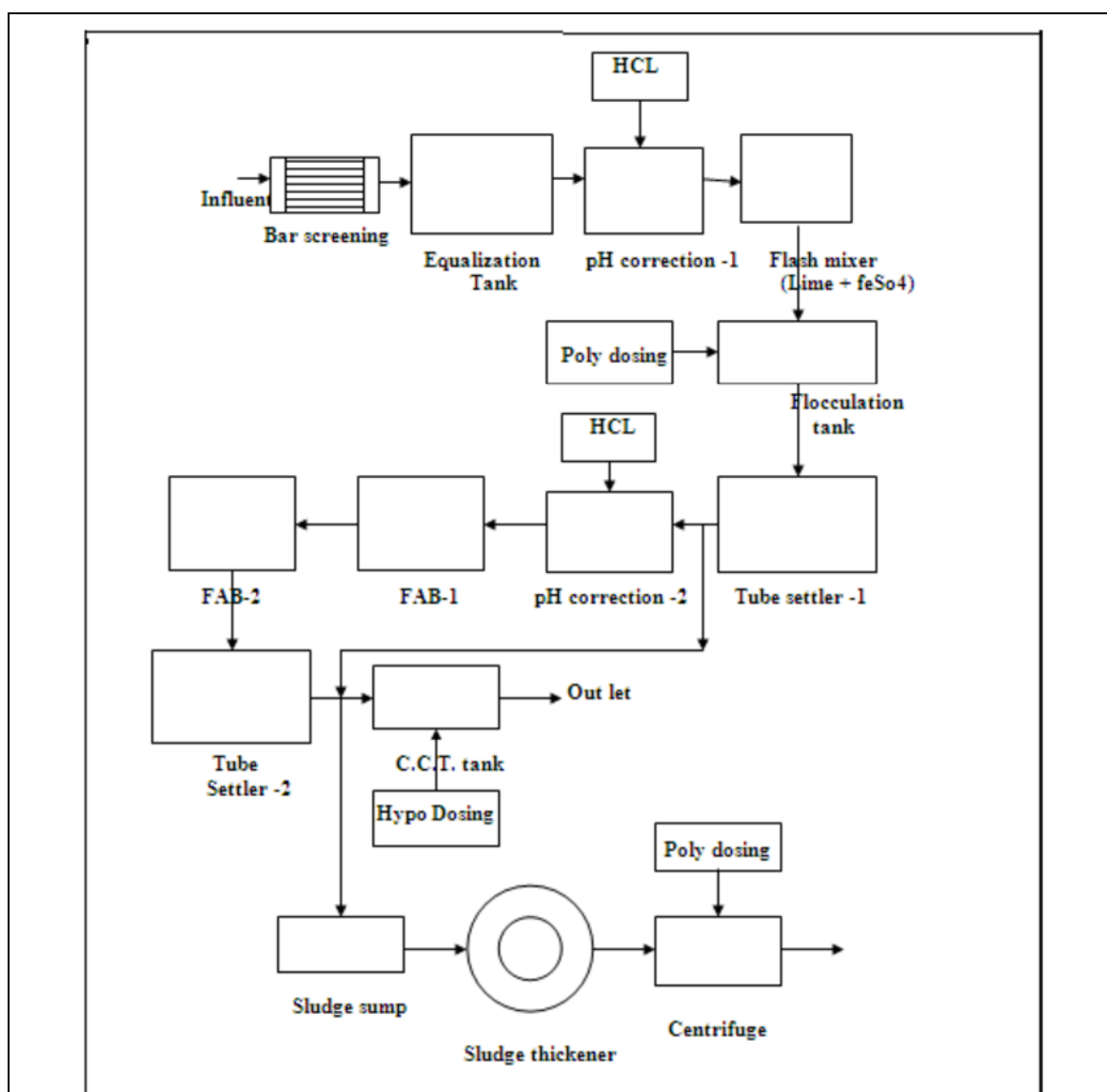


Figure 1

Flowchart of the textile effluent treatment plant Raymond Zambaiti Ltd, Kolhapur (Maharashtra, India). The water line comprises: screening, homogenization, coagulation-flocculation with pH correction, lamellar primary settling, pH correction again, two aerobic fluidized bed reactors in series (FAB FAB-1 and -2), lamellar secondary settling and chlorination. The primary and secondary sludge is just thickened and centrifuged. The annual average flow is 2,000 m³/d, with a COD and BOD average of 1200 and 300 mg/L, respectively (Reproduced from Sen and Demirer, 2003).



Figure 2
Anaerobic fluidized bed reactor (Source: ACRP 2013)

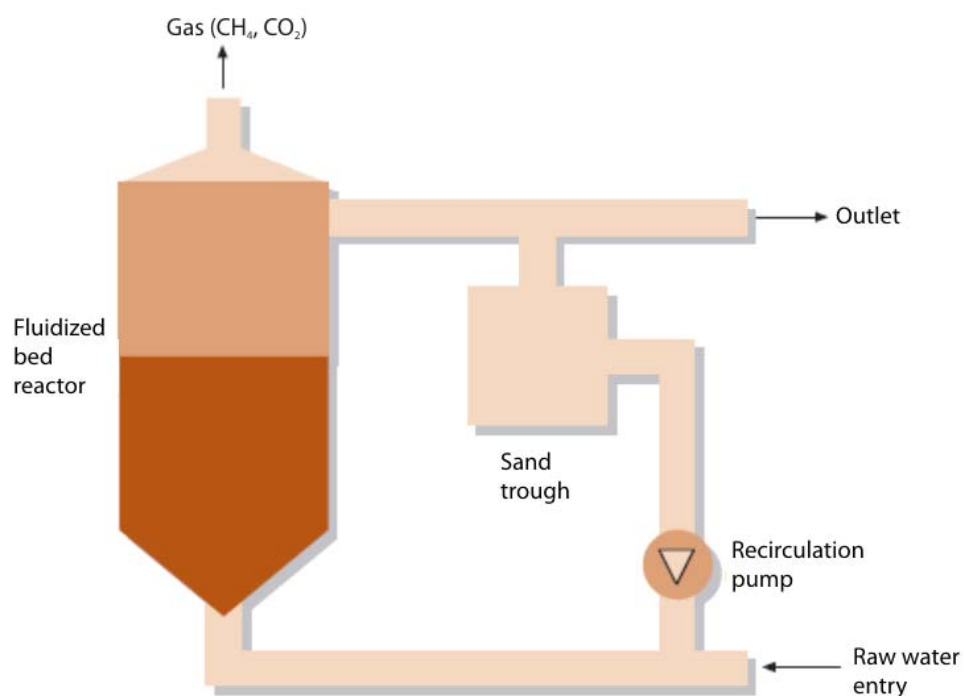


Figure 3
Anaerobic fluidized bed reactor scheme (Source: Rodriguez et al., 2006)



Figure 4
Aerobic fluidized bed reactors for urban wastewater treatment (Thermax system).



Figure 5
Fluidized bed reactors for treatment of industrial and urban wastewater (Envirogen system Technologies).



Figure 6
Fluidized bed reactors for treatment of urban or industrial wastewater, prior to activated carbon filling (Framax blog).