

Fluidization Principles for Three-Phase Fluidized Bed Bio-Reactor

K.S.K. Rao Patnaik, *Member, IACSIT* and K. Sriharsha

Abstract—Fluid Bed Reactors have been widely used in Chemical, Bio-Chemical and Petro-Chemical industries. The advantages of Fluid Bed Bio-Reactors (FBBR) are mainly Food, Pharmaceutical and Biological waste treatment sectors. In a fluidized Bio-Reactor, the particles can be much smaller and fine. During fluidization operation, the bed expands to accommodate microbial growth. The high surface area for biomass to grow results in high heat and mass transfer rates. This leads to isothermal and uniform mixing of particles tends to high concentration of active bio-mass per unit volume of reactor. This process also elimination of pollutants and toxins which are slows down the process. The problem of cell washout that aggravates operation of continuous, stirred tank fermenters is less likely in a fluidized bed, indeed it is impossible as long as the superficial liquid velocity is kept below the settling velocity of the solid particles. Aerobic Reactors require oxygen and depending upon this they can be classified as two phase or three phase FBBR. In the present investigation the general fluidization characteristics for Bio-Chemical processes was studied and compared with the literature. Three - phase fluidization is an operation involving a bed of suspended particles in gas and liquid media. This occurs due to a net drag force of gas and liquid in the flow direction on the particles. Gas-Liquid-Solid Fluidization is considered in the present work as operating bed of fluidized solid particles simultaneously with concurrent upward flow of a continuous liquid phase and a gaseous phase dispersed in the forms of bubbles. In this paper mainly the experimental work of the author on FBBR & distributor design are discussed and estimated the Fluidization parameters namely minimum fluidization Velocity, Free Settling Velocity and Bed Stratification for calculating bio-film formation. Hence, the process of bio-film formation, its characteristics and properties need to be well understood to design and operate efficiently a FBBR and also important conclusions are drawn.

Index Terms—fluid bed reactor, minimum fluidization velocity, free settling velocity, biomass and drag coefficient.

I. INTRODUCTION

In a fluidized bed bio-reactor can be classified by P. Ghosh et. al (1996) into 3 categories. In the first case the solid phase contains the enzyme immobilized on its surface or entrapped inside it and the immobilized enzyme catalyses the bio chemical reaction.

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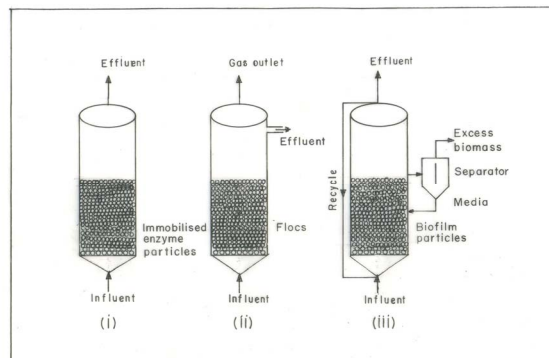


Figure.1.Type of fluidized bed bioreactors: (a) Immobilized enzyme reactor; (b) Tower reactor; (c) Biofilm reactor. (P. Ghosh et. al 1996)

The liquid phase contains the substrate to be converted to product. In the second type of FBBR, the solid phase is a microbial floc which are kept in fluidized state by the upward flow of the medium. In these lower type reactors washout of flocs is minimized by having an expanded sections and an internal settler at top of the reactor. Due to this expanded sections superficial liquid velocity at the top of the reactor decreases and hence the flocs settle back into the reactor without getting washed out. Also increase in floc diameter due to its growth increases its settling velocity and this helps in retaining it the bio-reactor. In the third type, the microorganism grows as a film on media particles. The media particles usually clay or sand or other inert materials offer very high surface area for the biomass to grow as bio film. In this type a supported film FBBR higher upward flow velocity of the fluid can be employed compared to microbial floc bio reactors because the inert solid material increases the composite density of bio particles. However, during operation due to microbial growth, overall density of bio particle

decreases with time which usually leads to washout of the over - grown bio particles. Under these circumstances these over-grown bio particles are intentionally washed out of the bio reactor and by using a mechanical device, the biomass is separated from the inert media and the inert media returned to the bio reactor where as the separated bio mass is washed as excess sludge. FBBR can be classified as aerobic or anaerobic reactors, aerobic reactors require oxygen and depending upon this they can be classified as two phase or three phase FBBR. In a three phase FBBR, bed of solids is fluidized by the co-current flow of a liquid and a gas. The liquid forms a continuous phase and the gas is a dispersed bubble phase.

II. PROBLEM FORMULATION

A. Characteristics of fluidization

Once the bed is fluidized, the pressure drop across the bed remains constant but the bed height continues to increase with increasing flow (Figure.2). More over, biochemical reactions are slow and require high residence time. This control the linear velocity of the fluid to lower ranges, where it sufficient to fluidize the particle and prevents agglomeration. Generally in the supported film FBBR, the superficial velocity is always maintained slightly above the minimum fluidization velocity or media used. This indirectly creates shear on the bio particles and hence allows the development of bio-film on the inert particle.

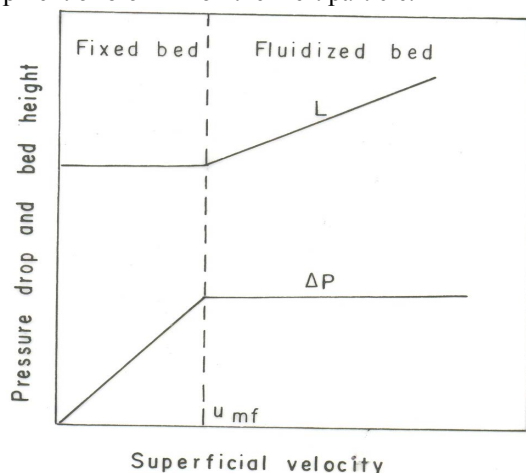


Figure.2. General Fluidization Characteristics

B. Bioparticles

Physical properties of Bioparticles such as size, shape and density, change dramatically as biomass accumulates on the support particle surface. These changes directly affect the hydrodynamic behavior of the percales in a biological fluidized bed reactor and, subsequently, reactor performance. There fore, knowing the biofilm properties is vitally important for design and operation of biological fluidized beds. Biofilm dry density is the bulk dry density of the attached biomass, calculated as the attached dry mass per unit wet biofilm volume. The biofilm dry density depends on the type of microorganism, type of substrate and substrate loading rate, thickness of biofilm and other environmental conditions.

C. Objectivities and aim

Following the problem formulation the objectivities of the work are:

- 1) Effect of Minimum fluidization velocity (U_{mf})
- 2) Effect of Terminal or Free settling velocity of a particle
- 3) Effect of Bed Stratification characteristics i.e., change in terminal settling velocity of bioparticle with film growth.
- 4) Experimental correlation of the dimensionless group CD_{NRep} versus $NRep$

III. EXPERIMENTAL

A. Experimental set-up

The experimental set up for investigating the fluidization characteristics for bio chemical process has been presented as a line diagram in Figure 3. The Test fluidization column consists of copper tube of 120 cm length and having a diameter of 10.16 cm. The column was fitted with a calming section having the same diameter as that of the test column and a distributor sandwiched between the column and calming section by means of flanges. The test column was heated externally by means of a kanthal resistance coil (6 kilowatt rate & 18 gauge) wound on the column and controlled through a voltage regulator. The 18 gauge kanthal wire having a total capacity of 6 k.w was wound with equal spacing over six slotted asbestos r. The strip were placed vertically over the test column, so as to keep the coil away from the column. The input and output terminals on the column winding were so selected as to divide the coil in the sections with a provision to yield 3 kw energy in each section. This is also provides uniform distribution of the kanthal energy generated. Three copper constantan thermocouples fixed on the test column by brazing them in the grooves etched on the surface , at 3 different places, so as to measure at 3 wall temperature check the uniformly of the wall temperature. These thermocouples were connected to a Honey-Well Brown strip chart recorder with ¼ sec pen speed through a selector switch. The reference point in all case was ice point, provided by placing the ice in thermoflask.

B. Distributor design

The distributor was constructed by brazing 2 copper plates of 0.32 cm thick on either side of a 1.9 cm long copper tube of 10.16 cm diameter to form a chamber having the same diameter as that of the test column. After drilling holes in the distributor section uniformly 0.32 cm copper tube of 2.54 cm length were inserted and brazed to allow the fluid from the calming section to pass through to the test column without entering into the distributor section. Holes of diameter 0.06 mm were drilled on the upper plate of the distributor in between the spacing left after fixing of the 2.5 cm long copper tubes. 4 pots were provided at the side of the distributor chamber for admission of air into the test column. The arrangement enables independent of each fluid directly to the test column. A brass wire screen of 100 mesh was spread over the distributor so as to prevent the particals weeping in to the calming section through the water inlets. An overflow chamber 15.24 cm diameter was provided at top of the test column a wire mesh cylinder was used to prevent any solid paricals coming with exit stream from a test column. The out coming fluid collected in the chamber and led to a storage time through a 5.08 cm outlet tube.

A Beckman thermometer was placed the test column just at the termination of the heating section for measuring the outlet fluid temperature in the test column.

A glass column having the same diameter and length as the test column was provided to observe the fluid bed conditions existing in the test column. The glass column has a similar distributor and a calming section as the test column.

An overflow chamber of 10.16 cm diameter and 50 cm

long, made of copper was fitted on the top of the glass column to allow for the variations in the bed height during the adjustment of flow rates of the fluids. The overflow chamber

has a provision to introduce solids in the glass column. The outgoing fluids from the glass column were let to the storage tank.

Table I: The following range of parameters are covered in the present authors data⁺

Particle sizes	0.17, 0.20, 0.25, 0.36, 0.46, 0.50, 0.66mm
Bed material	Fine river sand ⁺
Fluid mass	Water, Air ⁺
Superficial liquid velocity	$0.30 < u_L < 5$ cm/sec ⁺
Superficial gas velocity	$0 < u_G < 3$ cm/sec ⁺
Air pressure	0.65 kg/cm^2 (gauge) ⁺
Diameter of the fluidized column	10.16 cm ⁺
Viscosity of aqueous Glycerol	$0.01 < \mu_r < 0.0355 \text{ pa.s}$ ⁺
Solid hold up	$0.10 < \epsilon_s < 0.3$ ⁺
Liquid hold up	$0.16 < \epsilon_L < 0.95$ ⁺
Gas hold up	$0.08 < \epsilon_G < 0.65$ ⁺

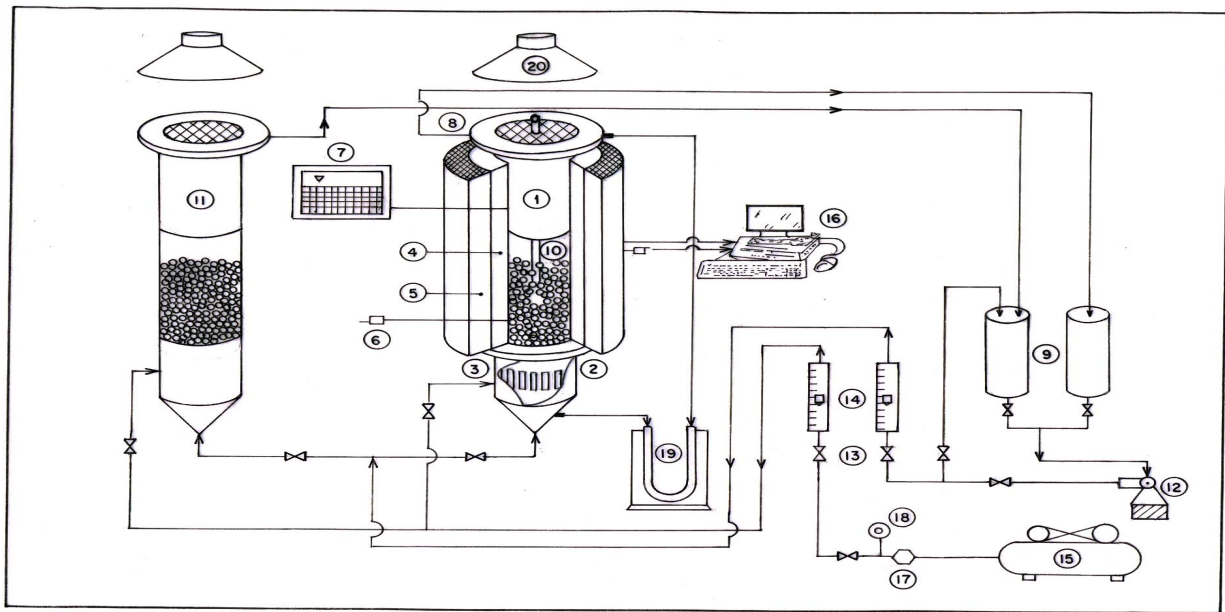


Figure.3. A Schematic Diagram of the experimental setup

- | | | |
|---------------------|-----------------------|--------------------|
| 1. Fluidized bed | 8. Chamber | 15. Air compressor |
| 2. Calming section | 9. Storage Tank | 16. Computer |
| 3. Distributor | 10. Probe/Thermometer | 17. Regulator |
| 4. Electric heaters | 11. Glass column | 18. Pressure gauge |
| 5. Insulation layer | 12. Pump | 19. Manometer |
| 6. Thermocouples | 13. Bypass valves | 20. Solid port |
| 7. Recorder | 14. Rotameter | |

IV. RESULTS AND DISCUSSION

A. Minimum Fluidization velocity (U_{mf})

Minimum Fluidization velocity (U_{mf}) has been calculated by making a force balance across the fluidized bed. For biological flocs and supported bio-film, the factors namely ρ_s , ϵ_{mf} and Φ_s varies with shape of the flocs, diameter of the flocs and nature of the bio-particals are observed. At the incipient fluidization, the minimum porosity ϵ_{mf} . Thus, from the literature by P.Ghosh et.al(1996) as follows.

$$\frac{\Delta P}{L_{mf}} = \frac{g}{g_c} (1 - \epsilon_{mf}) (\rho_p - \rho_l) \quad (1)$$

Ergun Equation for pressure drop across the packed bad gives

$$\frac{\Delta P_{gc}}{L_{mf}} = \frac{150\mu u_{mf}}{\phi_s^2 d_p^2} \frac{(1 - \epsilon_{mf})^2}{\epsilon_{mf}^3} + \frac{1.75\rho_l u_{mf}^2}{\phi_s d_p} \frac{(1 - \epsilon_{mf})}{\epsilon_{mf}^3} \quad (2)$$

Combining Equations (1) and (2) gives a quadratic equation for minimum fluidization velocity, u_{mf} as,

$$\frac{150\mu u_{mf}}{\phi_s^2 d_p^2} \frac{(1 - \epsilon_{mf})}{\epsilon_{mf}^3} + \frac{1.75\rho_l u_{mf}^2}{\phi_s d_p} \frac{1}{\epsilon_{mf}^3} = g(\rho_p - \rho_l) \quad (3)$$

For very small particles ($N_{Rep} < 1$) Equation (3) can be simplified and the minimum fluidization is given by the following equation,

$$u_{mf} = \frac{g(\rho_p - \rho_l)}{150\mu} \frac{\epsilon_{mf}^3}{(1 - \epsilon_{mf})} \phi_s^2 d_p^2 \quad (4)$$

For Large particles ($N_{Rep} > 10^3$), U_{mf} can be calculated from

$$u_{mf} = \left[\frac{\phi_s d_p g (\rho_p - \rho_1) \varepsilon_{mf}^3}{1.75 \rho_1} \right]^{1/2} \quad (5)$$

In case of biological flocs and supported biofilms, since the factors Φ_s , U_{mf} and ρ_s changes with diameter, shape and composite nature of the bioparticle, the minimum fluidization velocities are often found to have large deviations. An empirical correlation for minimum fluidization velocity proposed by Wen and Yu (1966) is expressed as,

$$u_{mf} = \frac{\mu}{\rho_1 d_p} \left[(33.7)^2 + \frac{0.04075 d_p^3 (\rho_p - \rho_1) g}{\mu} \right]^{0.5} - 33.7 \quad (6)$$

The minimum fluidization velocities are compared with the general equations $N_{Re} < 1$ and $N_{Re} > 10^3$. Particles are found to have large deviations.

To correlate the authors data with Cleasby and Bauman (1977) empirical equation with $\pm 5\%$ deviations.

Cleasby and Bauman (1977) have suggested an alternate empirical equation for calculating minimum fluidization velocity with $\pm 5\%$ deviations.

$$u_{mf} = \frac{0.00381 d_p^{1.82} g^{1.88} [\rho_1 (\rho_p - \rho_1)]^{0.94}}{\mu^{0.88}} \quad (7)$$

It is also suggested that fluidized beds can be operated at velocities much higher than the minimum fluidization velocity but is limited by the terminal or free fall velocity of the particles.

B. Terminal or Free Settling Velocity

Free settling velocity of a particle is given by

$$u_t = \left[\frac{4 g d_p (\rho_p - \rho_1)}{3 \rho_1 C_D} \right]^{1/2} \quad (8)$$

For spherical particles, the drag coefficient C_D is given by the following equations, depending upon the particle Reynolds number N_{Rep} ,

$$C_D = \frac{24}{N_{Rep}} \quad \text{For } N_{Rep} < 0.4 \quad (9)$$

$$C_D = \frac{10}{N_{Rep}^{1/2}} \quad \text{For } 0.4 < N_{Rep} < 500 \quad (10)$$

$$C_D = 0.43 \quad \text{For } N_{Rep} > 500 \quad (11)$$

Replacing the values of C_D in equation (8), depending upon N_{Rep} gives an analytical expression for u_t . For non-spherical particles the terminal settling velocity can be obtained from the experimental correlation of the dimensionless group $C_D N_{Rep}$ versus N_{Rep} (Kunii and Levenspiel, 1969). An alternate correlation for C_D in the intermediate range of N_{Rep} is given by Perry (1963) as

$$C_D = \frac{18.5}{N_{Rep}^{0.6}} \quad (12)$$

In FBRR, the shape of bioparticles is far from spherical and also they are not rigid. This may result in drag coefficient more than that predicted by the conventional equation. Hence, Hermanowicz and Ganczarczyk (1983) modified the drag coefficient as

$$C_D = \frac{17.1}{N_{Re, p}^{0.47}} \quad (13)$$

C. Bed Stratification Characteristics

It is observed that the microbial flocs and supported bio-film, the terminal or free settling velocity of the Bioparticle changes with time. This is attributed for due to merely biomass growth. The authors calculated the $N_{Re, p}$ lies in between 1.0 to 400, where most particles of behavior is observed.

In searching for alternate and appropriate method under these circumstances, the authors compared with Andrews equation (1982) and successful results are reported. The terminal velocity at any time to the terminal velocity of the solid as follows.

$$\frac{u_t}{u_{ts}} = \frac{(1 + \beta x)^{1/1.5}}{(1 + x)^{1/3}} \quad (14)$$

Where

$$\beta = \frac{\rho_b - \rho_1}{\rho_p - \rho_1} \quad (15)$$

This equation is plotted in figure 4.

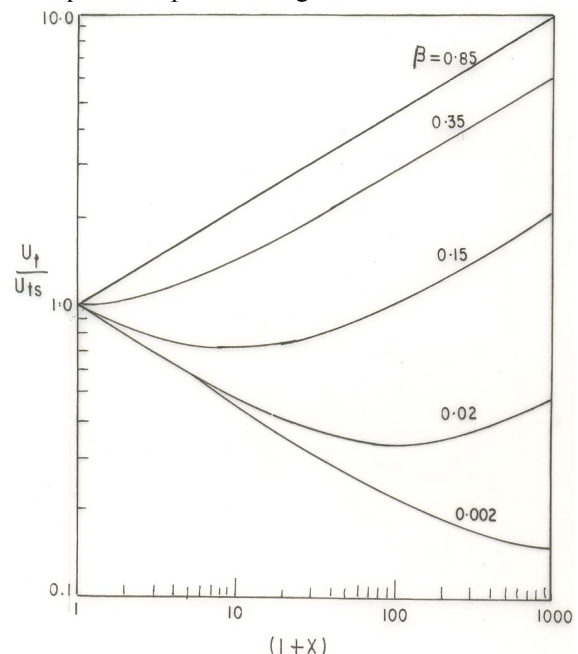


Figure 4. Bed Stratification Characteristics

It is concluded that for heavy support particles ($\beta \ll 1$), growth of biomass decreases the terminal settling velocity. This results in bed stratification. However, for flocs ($\beta = 1$) growth always increases settling velocity.

V. CONCLUSIONS

In case of biological flocs and supported biofilms, since the factors sphericity, minimum bed porosity and density of solid changes with diameter, shape and composite nature of the bioparticle, the minimum fluidization velocities are often found to have large deviations.

Fluidized beds can be operated at velocities much higher than the minimum fluidization velocity but is limited by the terminal or free fall velocity of the particles.

Microbial flocs and supported biofilm, the terminal settling velocity of the bioparticle changes with time due to biomass growth.

It can be seen for heavy support particles ($\beta \ll 1$), growth of biomass decreases the terminal settling velocity. However, for flocs ($\beta = 1$), growth always increases settling velocity.

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REFERENCES

- [1] Andrews. G.F (1982) Fluidized bed fermenters. A steady state analysis. BioTech mol, Bio Engg., vol.24, pp2013 – 2030
- [2] Bryers, J.D and Characklis, W.G (1982) Processes governing primary biofilm formation, Biotech mol., Bio engg., vol pp 2451 – 2477
- [3] Characklis, W.G (1981) Fowling bio-film development. A process analysis. Biotech mol., Bio Engg., vol – 23, p.p. 1923 -1960
- [4] Cleasby J.L and Bauman, E.R (1977) Backwash of granular filters used in waste water filtration, Environmental Protection Technology Series, EPA-600/12-77-016
- [5] Ghosh, P., Ramachandran. K.B and Praveen. V.V (1996) Application of fluidization principles for bio processes, volume 8, Gordon and Breach publishers, pp 303 -338.
- [6] Hermanowicz, S.W and Ganczarczyk, J.J (1983). Some fluidization Characteristics of biological beds, Biotechnol. Bioeng., vol 25, pp1321-1330
- [7] Hoehn, R.C and Ray. A.D (1973) Effects of thickness on bacterial film. J. water pollution control fed., vol. 45, pp 2302 – 2320
- [8] Kunii, D and Levenspiel, O (1969) Fluidization Engineering, John Wiley & Sons, New York.
- [9] Shieh, W.K. and Keenan. J.D (1986) Fluidization bed biofilm reactor for waste water treatment. Adv. Bio chem. Engg., vol.33 – 167.
- [10] Wen, C.Y and Yu Y.H. (1966) Mechanics of fluidization in fluid particle technology edited by S. Lee, Chem. Eng. Prog. Ser., vol. 52, pp 100 -111.
- [11] Perry, J.H. Ed (1963) chemical Engineers Hand book, Mc.Graw-hill, New York.
- [12] Kyoung Sin Ro, J.B. Neething, (1991) Bio-Film desity for biological fluidized beds Research Jowrnal WPCF, Vol.63 No.5, pp615-619.
- [13] Kian-Fab Ngian and William R.B. Martion (1980), Bed expansion charactstics of liquid fluidized particals with attached microbial growth val XXII, pp1843 to 1856 John Wiley and sons, Inc.
- [14] Kyoung Sin Ro, J.B. Neething (1990), Treminal settling charchtristics of bio particals, Reasearch journal WpC F, vol.62 No7 pp 901-906.
- [15] Cristiano Nicolella, Mark M.C Van Loosdrecht, Renzo Di Felice, Mauro Rovalli (1999.), Treminal settling and Bed expansion charactstics of Bio filim coated particals, Bio Tech and Bio Enginnering, Vol.62, No.1, .
- [16] Mehran Mdalib, Jasse Zhee and George Nakhla (2010) Terminal Settling velocity and Drag Coefficient of Bio film- Coated particals at high Reynolds Numbers. AIChE Journal, Vol. 56, No.10, Pp 2598-2606.



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