Student Indian Chemical Engineering Congress-2006 (SCHEMCON-2006), JNTU College of Engineering, Anathapur, AP.

Experimental Study of the Behaviour of a Three-phase Fluidized Bed with Cylindrical Particles

V. Ganesh Aditya[a], P. Panda[b], S. C. Rana[c] and H. M. Jena[d*]

[a]B.Tech Chem. Engg., NIT, Rourkela, India, Email: vgadityanit@gmail.com
[b]B.Tech Chem. Engg., NIT, Rourkela, India, Email: preetam_panda2000@yahoo.com
[c]M. Tech Chem. Engg., NIT, Rourkela, India, Email: subaligit@rediffmail.com
[d*]Dept. of Chem. Engg., NIT, Rourkela, India, Email: hara_jena@rediffmail.com

ABSTRACT
Three-phase fluidized beds have been applied successfully in the bio-oxidation process for wastewater treatment in which various low-moderate density solid particles of different shape and size are used as cell support. Ceramic raschig ring possess moderate density and high surface area due to its hollow cylindrical structure, thus can be used as solid support for microorganisms, where higher mass transfer rate can be achieved. In this study the hydrodynamic characteristics viz. the pressure drop, bed expansion and phase hold up of a co-current gas-liquid-solid three-phase fluidized bed has been studied using liquid as the continuous phase and gas as the discontinuous phase. These have been done in order to develop a good understanding of each flow regime in gas-liquid and liquid-solid fluidization. Air, water and ceramic raschig rings are used as the gas, liquid and solid phases respectively. The experiments were carried out in a 100 mm ID, 2m-height vertical Plexiglas column. The column consists of three sections, viz., the gas-liquid disengagement section, test section and gas-liquid distributor section. Bed pressure measurements have been made to predict the minimum liquid fluidization velocity. By keeping gas velocity at a fixed value, the liquid velocity is varied and the effect on minimum liquid fluidization velocity, pressure drop and the expansion ratio was studied for different particle size and static bed height.

Keywords: three phase fluidization, pressure drop, Minimum Fluidization Velocity, bed expansion

INTRODUCTION
Gas-liquid-solid fluidization also known as three-phase fluidization is a subject of fundamental research since the last three decades due to its industrial importance. Three-phase fluidized beds have been applied successfully to many industrial processes such as in the H-oil process for hydrogenation and hydro-desulfurization of residual oil, the H-coal process for coal liquefaction, Fischer-Tropsch process, and the bio-oxidation process for wastewater treatment. Three-phase fluidized beds are also often used in physical operations (Muroyama

and Fan, 1985). As in the case of fixed bed operation, both cocurrent and countercurrent gas-liquid flow are permissible, and for each of these both bubble flow, in which the liquid is the continuous phase and the gas dispersed, and trickle flow. In which the gas forms a continuous phase and the liquid is more or less dispersed (Epstein, 1981).Gas-liquid-solid fluidization can be classified mainly into four modes of operation. These modes are co-current three-phase fluidization with liquid as the continuous phase (mode I-a); co-current three-phase fluidization with gas as the continuous phase (mode-I-b); inverse three-phase fluidization (mode II-a); and fluidization represented by a turbulent contact absorber (TCA) (mode II-b). Modes II-a, and II-b are achieved with a countercurrent flow of gas and liquid. Amongst which the most striking one is the co-current three-phase fluidization with the liquid as the continuous phase (Muroyama and Fan, 1985). The co-current gas-liquid-solid fluidization is defined as an operation in which a bed of solid particles is suspended in gas and/or liquid upward flowing media due to the net gravitational force on particles. Such an operation generates considerable intimate contact among the gas, liquid and solid particles in these systems and provides substantial advantages for applications in physical, Chemical or biochemical processing involving gas, liquid and solid phases (Dhanuka and Stepanek, 1978).

The successful design and operation of a gas-liquid-solid fluidized bed system depends on the ability to accurately predict the fundamental characteristics of the system. Specially, the hydrodynamics, the mixing of individual phases, and the heat and mass transfer characteristics (Begovich and Watson, 1978). Knowledge of minimum liquid fluidization velocity is essential for the successful operation of gas-liquid-solid fluidized beds. For gas-liquid-solid fluidized systems the minimum liquid fluidization velocity is the superficial liquid velocity at which the bed becomes fluidized for a given gas superficial velocity (Briens and Briens, 1997).The minimum liquid flow rates required to achieve fluidization are determined by a plot of the pressure drop across the bed vs. the superficial liquid velocity at constant gas flow rate. When fluidization, the pressure drop across the bed will no longer change with increasing liquid flow rate. Thus the flow rates at which a break in curve occurs correspond to the MF velocities (Begovich and Watson, 1978). Visual observations determine the liquid minimum fluidization velocity as either the velocity at which the bed first begins to expand or as the velocity at which any particle with in the bed continuously shifts position with neighboring particles (Briens and Briens, 1997). The bed pressure drop is given by the equation (1).

$$\Delta P = gH\,(\rho_L \varepsilon_L + \rho_G \varepsilon_G + \rho_S \varepsilon_S) \qquad (1)$$

Three-phase fluidized beds have been applied successfully in the bio-oxidation process for wastewater treatment in which various low-moderate density solid particles of different shape and size are used as cell support. Ceramic raschig ring possess moderate density and high surface area due to its hollow cylindrical structure, thus can be used as solid support for microorganisms, where higher mass transfer rate can be achieved.

In the present study experiments were conducted to examine the hydrodynamic behavior viz. the pressure drop, minimum fluidization, bed expansion and phase hold up of a co-current gas-liquid-solid three-phase fluidized bed using liquid as the continuous phase and gas as the discontinuous phase. Ceramic raschig rings of different size are used as soild particles. These

have been done in order to develop a good understanding of each flow regime in gas-liquid and liquid-solid fluidization.

EXPERIMENTAL

A schematic diagram of the experimental setup is shown in Figure-1. The vertical Plexiglas fluidizer column is of 100 mm ID with a maximum height of 2m.The column consists of three sections, v.i.z., the gas-liquid disengagement section, test section, and gas-liquid distributor section. The gas-liquid distributor is located at the bottom of the test section and is designed in such a manner that uniform distribution of the liquid and gas can be maintained in the column. The distributor section (Figure-3) is a conical frustum of 12 cm in height, one end 5.08 cm in diameter and the other end of 10 cm diameter having liquid inlets one of 24 cm ID with a perforated plate (Figure-2) made of G.I. sheet of I mm thick, 120 mm diameter, of about 278 numbers of 2, 2.5 and 3mm pores in placed at the top of this section. There is a gas distributor consists of 50 numbers of 1mm pores placed randomly. In this section the gas and liquid streams merged and passed through the perforated grid. The mixing section and grid ensure that the gas and liquid are well mixed and evenly distributed into the bed. Gas-Liquid Disengagement Section is at the top of the column, which allows gas to escape and liquid to be circulated. Any entrained particles retain on the screen attached to the top of this section. For pressure drop measurement the pressure ports are being fitted to the U-tube manometers of 1m & 0.5m long filled with mercury and carbon tetrachloride.

The three phases (solid, liquid and gas) present in the column were raschig rings L=OD= 6.6mm, ID= 3.3mm and L=OD= 11.2mm, ID= 5.6mm, tap water and the oil free compressed air. The properties of the bed material, the fluidizing medium and the manometric fluid are shown in Table-1. The air-water flow were co-current and upwards. Accurately weighed amount of material was fed into the column and adjusted for a specified initial static bed height. Four initial static bed heights Hs=15.4 cm, Hs=21.4 cm, Hs=26.4 cm and 31.4 cm are used in this experiment. Water was pumped to the fluidizer at a desired flow rate. Then air was injected into the column through the air distributor. Approximately five minutes was allowed to make sure that the steady state was reached. Then the readings of each manometer were taken. Also, the bed expansion was noted. For gas hold up measurement, the water and air rotameters valves were quickly closed at same proportion. The values of minimum fluidization velocity for every run have been obtained by plotting pressure drop across the beds versus liquid flow rates at constant air flow rates. The same procedure was repeated for different materials at different static bed height.

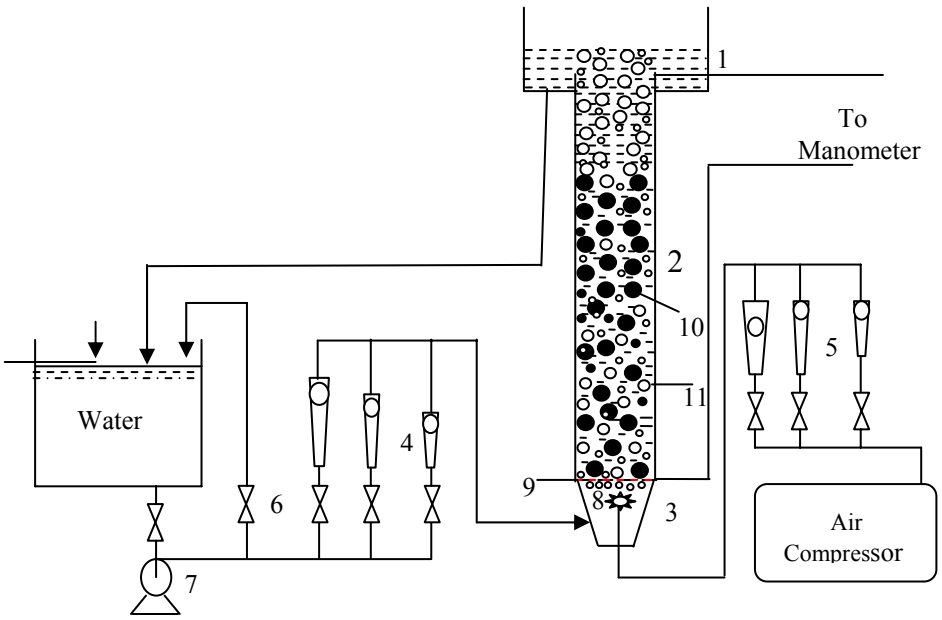

Figure 1: Schematic diagram of gas-liquid-solid three-phase fluidized bed

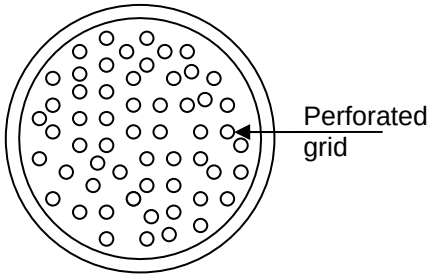

Figure 2: Schematic diagram of the gas and liquid distributor.

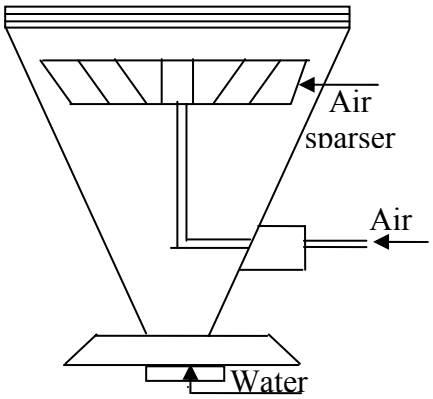

Figure 3: Schematic diagram of conical calming section.

Table 1: Properties of Bed Materials (A), Fluidizing Medium (B), Manometric Fluid (C)

A. Properties of Bed Materials

Particle Notation.	Materials	L = OD, mm	ID, mm	ρ_p (kg.m^{-3})
P1	Ceramic raschig ring	11.2	5.6	1670
P2	Ceramic raschig ring	6.6	3.3	1670

B.Properties of Fluidizing Medium

Fluidizing Medium	ρ (kg.m^{-3})	μ (Ns/m^2)
Air at 30^{0}C	1.168	0.00187
Water at 30^{0}C	999.4	0.095

C.Proprties of Manometric Fluid

Manometric Fluid	ρ (kg.m^{-3})	μ (Ns/m^2)
Mercury	13600	0.15
Carbon Tetra-Chloride (CCl_4)	1600	0.09

RESULTS AND DISCUSSION

Pressure Drop and Minimum Fluidization Velocity

The minimum liquid fluidization velocity (Ulmf) in this study was obtained from the relationship between pressure drop and superficial liquid velocity. Figure-4 shows the variation of pressure drop with superficial liquid velocity for gas-liquid-solid system at various superficial gas velocities. From this it is observed that the minimum fluidization velocity decreases with increase in gas velocity. The variation of minimum liquid fluidization velocity with superficial gas velocity is shown in Figure-5. It indicates sharp decrease in minimum liquid fluidization velocity with gas velocity initially but then the rate of decrease is slow. This indicates the bubble supported fluidization in presence of gas. The values of Ulmf are listed in Table-2a.

Figure-6 shows the variation of bed pressure drop with superficial gas velocity for different static bed height. From this pressure drop plot minimum liquid fluidization velocity has been calculated. Observed pressure drop increases with the initial static bed height which indicates higher energy requirement to fluidize higher bed mass. The observed minimum liquid fluidization velocities are plotted against the initial static bed height and are shown in Figure-7. The plot shows that the minimum liquid fluidization velocities are approximately same for all static bed heights (the values are listed in Table-2b).

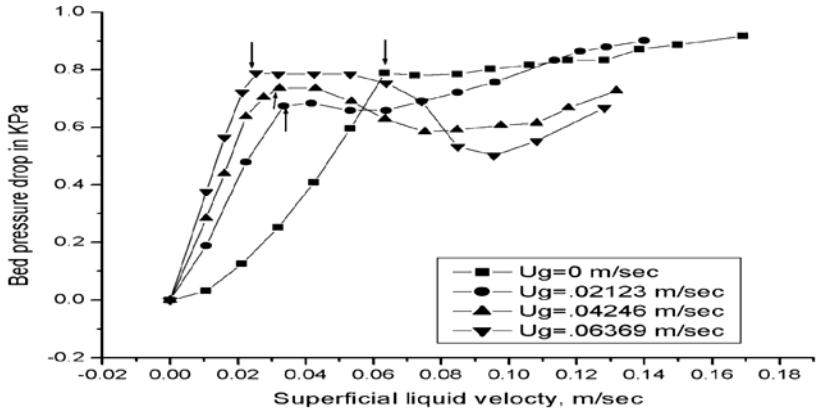

Figure 4: Variation of bed pressure drop with superficial liquid velocity for different superficial gas velocities at static bed height 21.4 cm for 6.6 mm raschig ring.

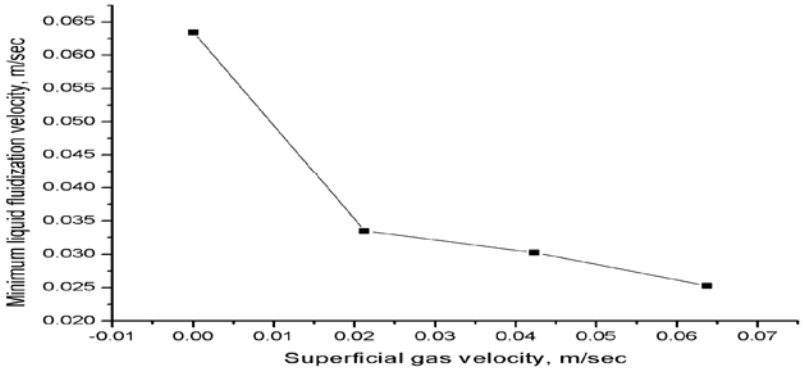

Figure.5: Variation of minimum liquid fluidization velocity with superficial gas velocity at constant static bed height 21.4 cm for 6.6 mm raschig ring.

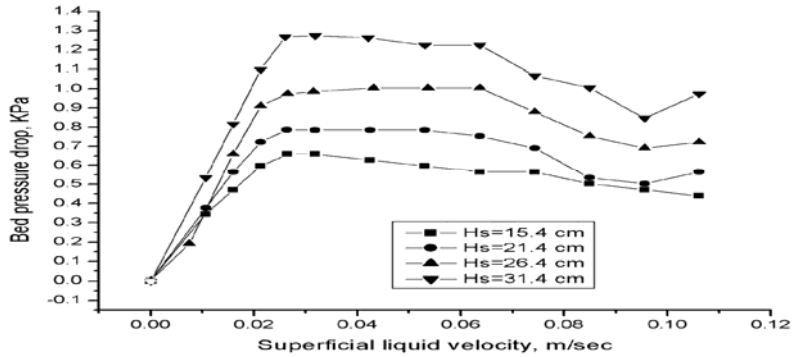

Figure 6: Variation of bed pressure drop with superficial liquid velocity for different static bed height at superficial gas velocity 0.06369 m/sec for 6.6 mm raschig ring.

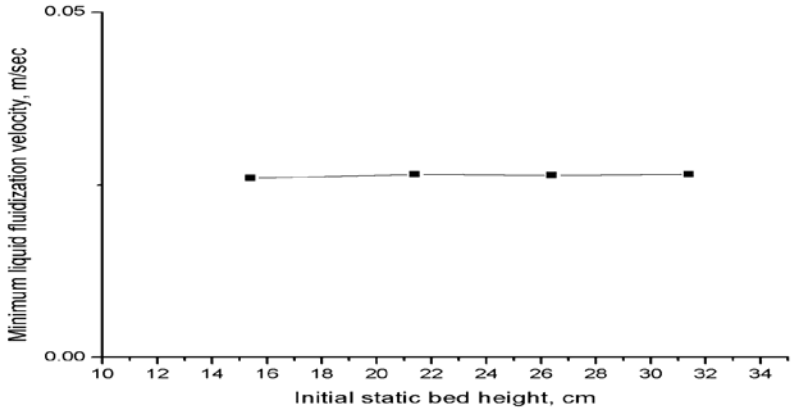

Figure 7: Variation of minimum liquid fluidization velocity with initial static bed height at superficial gas velocity 0.06369 m/sec for 6.6 mm raschig ring.

Figure-8 shows the variation of bed pressure drop with superficial gas velocity for different particle size. The plot shows higher total pressure drop for the higher size particle and higher

minimum liquid fluidization velocity also. But initially for the same superficial liquid velocity the pressure drop is low for higher size particle due to higher bed voidage. The values of Ulmf for different particle size are listed in Table-2a.

Table 2a: Ulmf in m/sec for different particle sizes, gas velocities for Hs=21.4 cm.

Dp, mm	Ug = 0 m/sec	Ug = 0.02123 m/sec	Ug = 0.04246 m/sec	Ug = 0.06369 m/sec
11.2	-	-	-	0.41728
6.6	0.06344	0.03348	0.030247	0.02527

Table 2b: Ulmf in m/sec for different Hs at Ug=0.06369 m/sec & Dp=6.6mm.

Hs = 15.4	Hs = 21.4	Hs = 26.4	Hs = 31.4
0.026539	0.02527	0.026539	0.02599

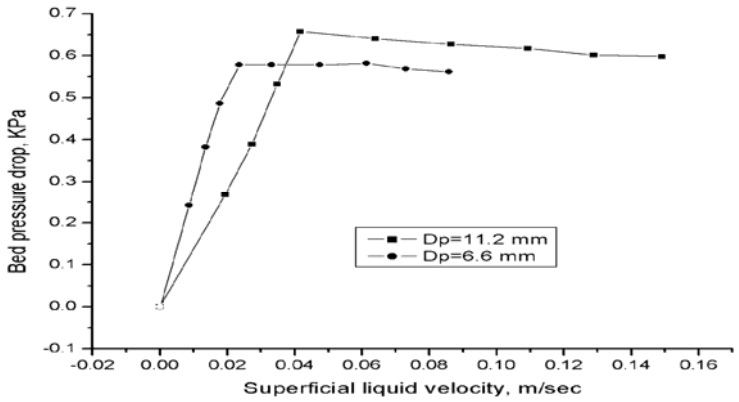

Figure 8: Variation of bed pressure drop with superficial liquid velocity for different particle size at superficial gas velocity 0.06369 m/sec for static bed height.

Bed Expansion

The bed voidage increases with both increasing liquid velocity and gas velocity as shown in Figure-9. For same superficial liquid velocity the bed expansion is more for higher superficial gas velocity which indicates higher extent of bubble supported fluidization. Initially the bed voidage decreases to some limited value of superficial liquid velocity then increases. The slope of the curve is more for higher gas velocities. When at Ulmf the gas velocity is varied (Figure-10) bed expansion ratio increases sharply, then the slope decreases. This indicates that after certain value, the visual observation of the bed height is subjective due to high concentration variation of the particles along the length of the bed.

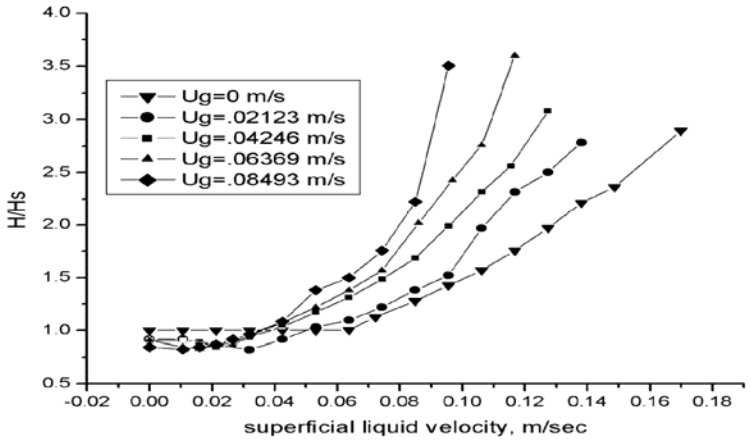

Figure 9: Variation of bed expansion ratio with superficial liquid velocity for different superficial gas velocities at Hs=21.4 cm for 6.6 cm raschig ring.

Figure-11 shows the variation of expansion ratio with superficial liquid velocity for different static bed heights. The figure shows that the bed expansion ratio is almost independent upon the static bed height. Similarly initial contraction in the bed height observed with increase in superficial liquid velocity.

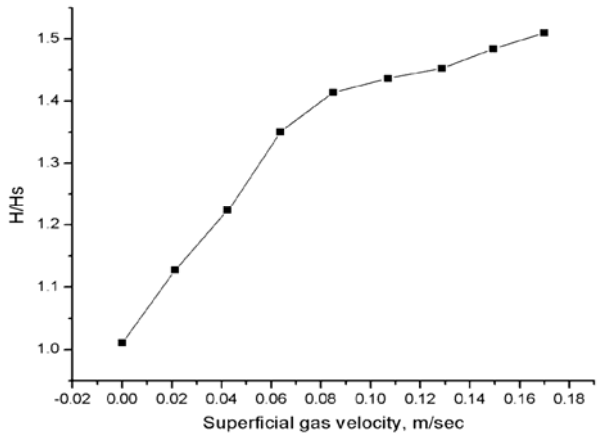

Figure 10: Variation of bed expansion ratio with superficial liquid velocity for different static bed heights at Ug=0.06369 m/sec for 6.6 cm raschig ring.

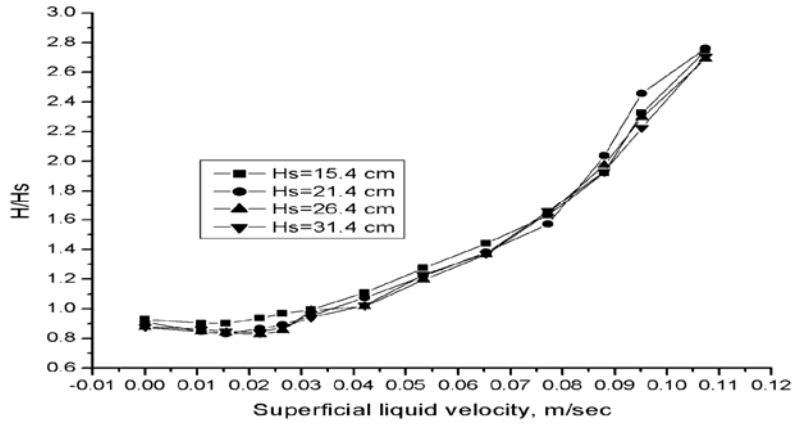

Figure 11: Variation of bed expansion ratio with superficial liquid velocity for different static bed heights at Ug=0.06369 m/sec for 6.6 cm raschig ring.

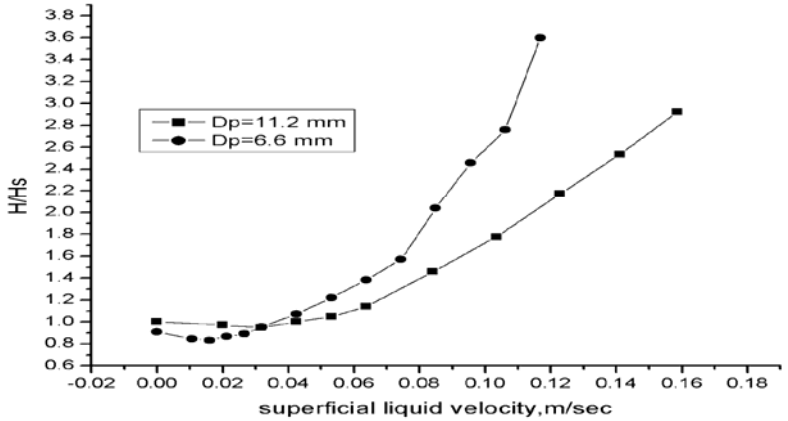

Figure 12: Variation of bed expansion ratio with superficial liquid velocity for different particle size at Ug=0.06369 m/sec for static bed height 21.4 cm.

The effect of particle size on bed expansion ratio is shown in Figure-12. The figure is the plot of bed expansion ratio vs superficial liquid velocity for different particle size at same superficial gas velocity and static bed height. Higher initial bed contraction observed for lower size particle, but after a critical value the expansion ratio for lower size particle steeply increases. The bed voidage at minimum fluidization is same for both the particles.

CONCLUSIONS

The hydrodynamic study of the three-phase fluidized bed with cylindrical particles reveals that the minimum liquid fluidization velocity (V_{lmf}) increases with increase in particle size, decreases with increase in gas velocity but not affected by the bed mass. The expansion ratio increases with increase in liquid and gas velocity, but decreases with increase in particle size. Bed mass has no effect on the expansion ratio. The bed voidage at minimum fluidization

condition is slightly less for higher gas velocities, but it is almost same for different bed mass and particle sizes.

NOMENCLATURE

Dp	Particle diameter, [mm], the outer diameter (OD) of particle
H	Average height of expanded bed, [cm]
Hs	Static bed height, [cm]
Ms	Mass of the solid in the bed, [kg]
ΔP	Pressure drop, [KPa]
Ul	Liquid velocity, [m/sec]
Ug	Gas velocity, [m/sec]
Ulmf	Minimum liquid velocity for a three-phase system, [m/sec]
U	Minimum liquid fluidization velocity for liquid-solid system, [cms^{-1}]
ρ	Phase density,[kgm^{-3}]
ε	Phase volume fraction [dimensionless]
s	Solid phase
l	Liquid phase
g	Gas phase

REFERENCES

Jena, H. M., G. K. Roy and B. C. Meikap (2005), "Hydrodynamics of a Three-phase Fluidized Bed",Chemcon 2005, 14-17 December 2005, IIT Delhi.

Begovich, J. M. and J. S. Watson (1978), "Hydrodynamic characteristics of Three-Phase Fluidized Beds", Fluidization, Cambridge University Press, 190-195.

Bhatia, V. K. and Epstein, N., 1974, Proc. Int. Symp. On Fluidization and its Applications, Toulouse, P. 380.

Briens, L.A., C.L.Briens, A.Margaritis and J.Hay (1997), "Minimum Liquid Fluidization Velocity in Gas-Liquid-Solid Fluidized Beds of Low-Density Particles", Chem.Eng.Sci. 52 (21-22), 4231-4238.

Davies, O.L (1978), "Design and Analysis of Industrial Experiments", Second ed., Longman Publishers.

Dhanuka, V.R., and J.B. Stepanek (1978), "Gas and Liquid Hold-up and Pressure Drop Measurements in a Three-Phase Fluidized Bed", Fluidization, Cambridge University Press, 179-183.

Kim, S. D., Baker G. C. J., and Bergounou, M. A., 1975, Phase holdup characteristics of three phase fluidized beds, Can. J. Chem. Eng., 53, 134 - 39.

Epstein, N., 1981, Three phase fluidization: Some knowledge gaps, Can. J. Chem. Eng., 59, 649 – 57.

Muroyama, K. and L.S.Fan (1985), "Fundamentals of Gas-Liquid-Solid Fluidization", AIChE.J. 31 (1), 1-34.