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Hydrodynamics of a Three-phase Fluidized Bed

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ABSTRACT

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 glass beads (2.18mm, 3.05mm and 4.05mm) 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 was varied and the effect on phase hold-up, minimum liquid fluidization velocity, pressure drop and the expansion ratio was studied for different particle size and static bed height. Experimental study based on statistical design has been made to investigate the expansion ratio of fluidized bed and a correlation has been developed for gas hold-up. It is evident from the correlation that gas hold-up is strongly function of modified gas Reynolds number and independent of liquid Reynolds number. The experimental values have been compared with those predicted by the correlations and have been found to agree well.

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).

For chemical processes where mass transfer is the rate-limiting step, it is important to be able to estimate the gas holdup as this relates directly to the mass transfer (Safoniuk *et al*., 2002). The following equations have typically been used to determine the volume fraction (holdup) of each phase in the three phase fluidized bed:

$$\varepsilon_L + \varepsilon_G + \varepsilon_S = 1 \quad (1)$$

$$\Delta P = gH\,(\rho_L\varepsilon_L + \rho_G\varepsilon_G + \rho_S\varepsilon_S) \quad (2)$$

$$\varepsilon_S = M_S / \rho_S AH \quad (3)$$

where the bed height in Eqs. (2) and (3) is obtained either visually or from the measured pressure drop gradient Kim etal., 1975; Bhatia and Epstein, 1974; El- Temtary and Epstein, 1980). A more direct method of measuring ε_G is to simply isolate a representative portion of the test section by simultaneously shutting two quick closing valves and measuring the fraction of the islated volume occupied by the gas (Epstein, 1981). Other most promising methods of measuring the local gas holdup are electroresistivity, electro conductivity methods, γ - ray transmission measurements and radioactive tracer techniques (Dhanuka and Stepanek, 1978; Begovich and Watson, 1978; El- Temtary and Epstein, 1980; Safoniuk *et al*., 2002; Yu and Rittman, 1997).

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. These have been done in order to develop a good understanding of each flow regime in gas-liquid and liquid-solid fluidization. Correlation based on factorial design analysis (Davies, 1978) has been developed for the bed expansion ratio and compared with the experimental values. Also a correlation derived from dimensional analysis has been proposed for gas hold-up and compared with the correlations of Safoniuk *et al*. (2002).

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 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 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 manometers of 1m long (each limb) filled with mercury. The design is to measure the pressure drops at a particular section at three different locations such as at the wall, at the center of the column and at $1/4^{th}$ of the diameter of the column from the wall. So that the wall effects and the gas holdup can be studied clearly.

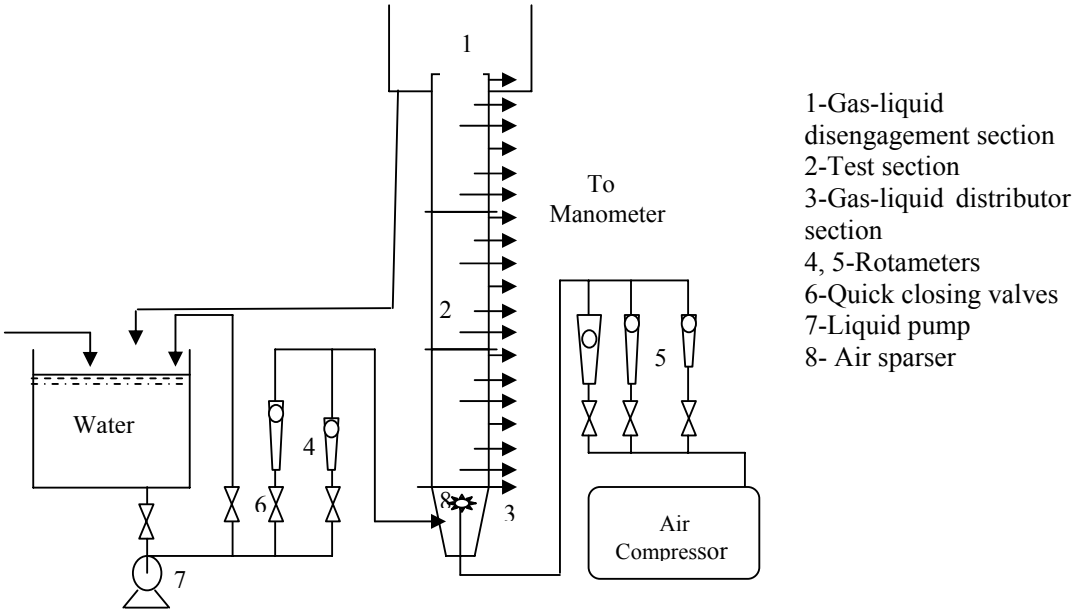

Figure 1: Schematic diagram of the three-phase fluidized bed

The three phases (solid, liquid and gas) present in the column were 2.18, 3.04 and 4.05 mm glass beads, 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. 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.

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

A. Properties of Bed Materials

Particle Notation.	Materials	Mesh size	d_p,mm	ρ_p (kg.m^{-3})
P1	Glass Beads	-7+8	2.18	2,216
P2	Glass Beads	-5+6	3.05	2,253
P3	Glass Beads	-4+5	4.05	2,470

B.Properties of Fluidizing Medium

Fluidizing Medium	ρ (kg.m^{-3})	μ (Ns/m^2)
Air at 25^{0}C	1.168	0.00187
Water at 25^{0}C	1,000	0.095

C.Proprties of Manometric Fluid

Manometric Fluid	ρ (gm/cc)	μ (Ns/m^2)
Mercury	13.6	0.15
Carbon Tetra-Chloride (CCl_4)	1.59	0.09

RESULTS AND DISCUSSION

Pressure Drop And Minimum Fluidization Velocity

The minimum fluidization velocity in this study was obtained from the relationship between pressure gradient and superficial liquid velocity. Fig.2 and Fig.3 shows the variation of pressure drop with superficial liquid velocity for liquid-solid system at various bed heights and particle size. From this it is observed that bed mass has no effect on minimum fluidization velocity, but minimum fluidization velocity increases with increase in particle, which is listed in Table-2.

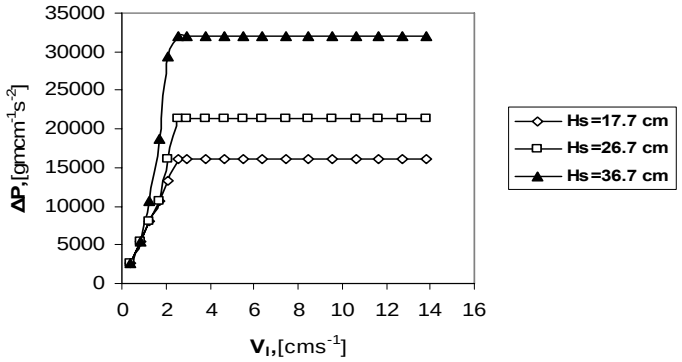

Figure.2: Variation of pressure drop with liquid velocity for different bed height at Vg=0cm/s for 2.18 mm glass beads.

Fig.4 shows the variation of pressure drop with superficial liquid velocity for gas-liquid-solid system for different superficial gas velocities. The minimum fluidization velocity decreases with increase in gas velocity. Fig.5 shows the variation of minimum fluidization velocity with superficial gas velocity for different particle size. Minimum fluidization velocity decreases with gas velocity, but more for particles of higher sizes. Finally a comparison of minimum fluidization velocity is listed in Table-2.

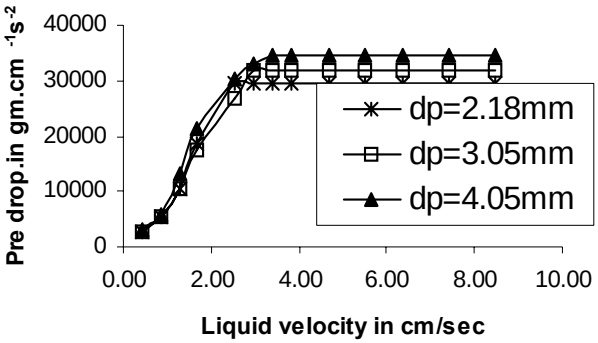

Figure.3: Variation of pressure drop with liquid velocity for different particle size at V_g=2 cm/s for H_s=36.7 cm.

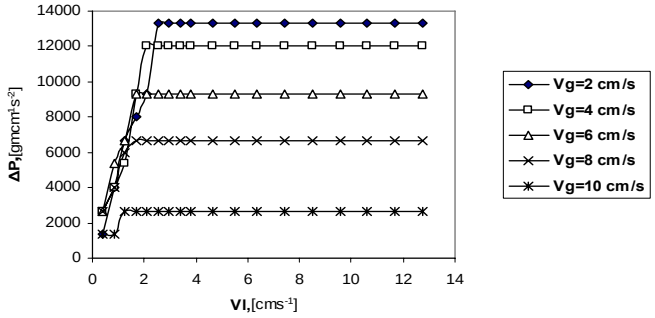

Figure.4: Variation of pressure drop with liquid velocity at different gas velocity for H_s=26.7 cm and 3.05 mm glass beads.

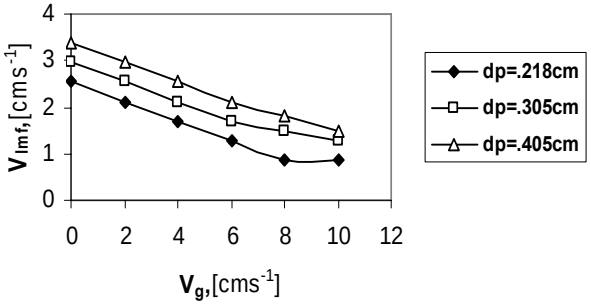

Figure.5: Variation of minimum liquid fluidization velocity with gas velocity for different particle size at constant static bed height.

Table 2: Comparison of minimum fluidization velocity for different particle size at different gas velocities.

d_p,mm	V_g=0 cm/s	V_g=2 cm/s	V_g=4 cm/s	V_g=6 cm/s	V_g=8 cm/s	V_g=10 cm/s
2.18	2.55	2.12	1.70	1.27	0.85	0.85
3.05	2.97	2.55	2.12	1.70	1.49	1.27
4.05	3.40	2.97	2.55	2.12	1.81	1.49

Bed Expansion

The bed voidage increases with both increasing liquid velocity and gas velocity as shown in Fig.6. Correlation based on factorial design analysis (Davies, 1978) has been developed for the bed expansion ratio. The method of Factorial Design Analysis bring out the interaction

effects of variables, which would not be found otherwise by conventional experimentation and to explicitly find out the effect of each of the variables quantitatively on the response.

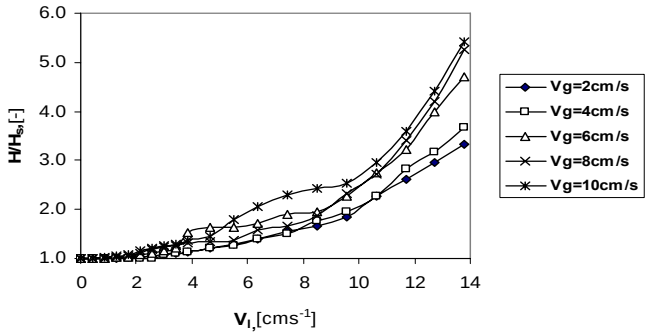

Figure.6: Variation of expansion ratio with liquid velocity at different gas velocity at H_s=26.7 cm for 3.05 mm glass beads.

The scope of the factors consider for factorial experimentation is presented in Table-3.
The variables which affect bed expansion ratios in fluidization are static bed height, particle diameter and gas velocity. Thus total numbers of experiments required at two levels for the three variables is eight for responses expansion ratio at minimum fluidization velocity.

Table 3: Scope of the factors for hydrodynamics

Sl. No.	Name of the variables	Factorial variables (General symbol)	Factorial design symbol	Min. level(-1)	Max. level(+1)	Magnitude of variables
1	Static bed height(cm)	H_s	A	17.7	36.7	17.7,26.7,36.7
2	Particle dia(cm)	d_p	B	0.218	0.405	0.218,0.305,0.405
3	Gas velocity(cm/s)	V_g	C	2	10	2,4,6,8,10

Development of model equation

The model equations are assumed to be linear and the equations take the general form,

$$Y= (b_0+b_1A+b_2B+b_3C+\ldots+b_{12}AB+b_{13}AC+\ldots+b_{123}ABC) \quad (4)$$

Coefficients are calculated by the Yates technique; $b_i= \Sigma (\alpha_i Y_i)/N$)

The following equation has been obtained,

$$Y= (1.05 + .0225*B - 0.0175*c + 0.01*A*B + 0.01*B*C) \quad (5)$$

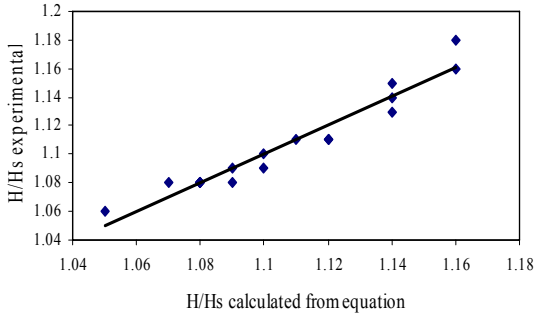

Figure.7: Comparison of experimental values of expansion ratio with those calculated by equation (5).

The value of the coefficients indicates the magnitude of the effect of the variables and the sign of the coefficient gives the direction of the effect of the variable. That is a positive

coefficient indicating an increasing in the value of the responses with increase in the value of the variable and a negative coefficient showing that the response decreases with increase in the value of the variable. The comparison of the experimental values with that of calculated values from equation-5 shows good agreement as shown in Fig 7.

Gas Holdup

Fig.8 and Fig.9 shows the variation of gas holdup with liquid velocity at various fixed gas velocities and with gas velocity at different fixed liquid velocities. It is seen that with increase in liquid velocity the gas holdup decreases but remains constant after reaching a moderate value of liquid velocity, with increase in gas velocity at constant liquid velocity the gas holdup increases monotonically.

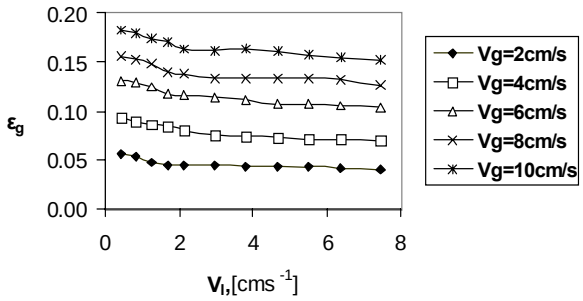

Figure.8: Variation of gas hold-up with liquid velocity at different gas velocity at H_s=26.7cm for 2.18 mm glass beads

The avg. gas holdup was plotted against modified gas Reynolds number (Re_g) for $12.64 \leq Re_l \leq 309.60$ .The results were fitted to a power-law equation passing through the origin (at zero gas flow rate) as, $\varepsilon_g = 0.0023 * Re_g^{0.7293}$ (6)

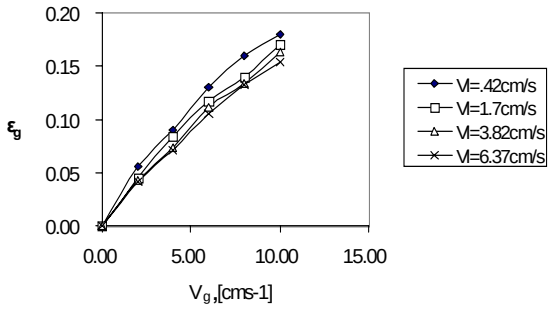

Figure.9: Variation of gas hold-up with gas velocity at different liquid Velocity at H_s=26.7cm for 2.18mm glass beads.

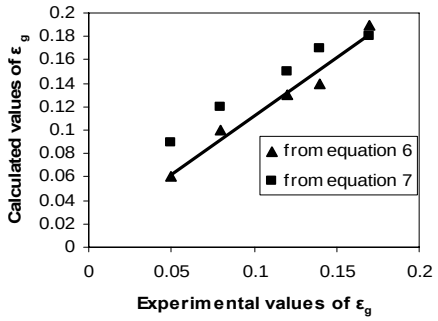

Figure.10: Comparison of experimental values of gas hold-up with those calculated from equation (6) and (7).

The Safoniuk etal.2002 correlation is given by, $\varepsilon_g = 0.0139 * Re_g^{0.426}$ (7)

Fig.10 shows the comparison of experimental values of gas holdup with those calculated from equation (6) and (7). Higher holdup is seen for equation (7)

CONCLUSIONS

The hydrodynamic study of the three-phase fluidized bed reveals that the minimum liquid fluidization velocity (V_{lmf}) increases with increase in particle size at constant gas velocity but decreases with increase in gas velocity at constant liquid velocity. The expansion ratio increases with increase in liquid and gas velocity and decreases with increase in particle size

and static bed height. The gas hold-up increases monotonically when the gas velocity is increased. At a fixed gas velocity, at low liquid velocity gas hold-up decreases and remains constant with further increase in liquid velocity. Gas hold-up increases with increase in particle size. It is evident from the correlation that gas hold-up is a strong function of modified gas Reynolds number and independent of liquid Reynolds number.

NOMENCLATURE

d_p Particle diameter, [mm]
H Average height of expanded bed, [cm]
H_s Static bed height, [cm]
M_s Mass of the solid in the bed, [kg]
ΔP Pressure drop, [$gm.cm^{-1}s^{-2}$]
Re_l Liquid Reynolds number, [dimensionless]
Re_g Modified gas Reynolds number, [dimensionless]
V_l Liquid velocity, [cms^{-1}]
V_g Gas velocity, [cms^{-1}]
V_{lmf} Minimum liquid velocity for a three-phase system, [cms^{-1}]
V^{ls}_{lmf} Minimum liquid fluidization velocity for liquid-solid system, [cms^{-1}]
ε_g Gas holdup, [dimensionless]
β_u Ratio of superficial velocities= (V_g/ V_l) , [dimensionless]
ρ Phase density,[kgm^{-3}]

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