

Archived in

dspace@nitr

<http://dspace.nitrkl.ac.in/dspace>

GAS HOLDUP IN A THREE-PHASE FLUIDIZED BED WITH CYLINDRICAL PARTICLES

H. M. JENA^{a*}, G. K. ROY^a, and B.C. MEIKAP^b

^aDepartment of Chemical Engineering, NIT, ROURKELA – 769008 (ORISSA) INDIA

^bDepartment of Chemical Engineering, IIT, KHARAGPUR – 721302 (W. B.) INDIA

International Journal of Chemical Science, Vol 5 (4), P 1712-1722

* Author to whom correspondence may be addressed.

Mr. Hara Mohan Jena, Department of Chemical Engineering, NIT, ROURKELA – 769008 (ORISSA) INDIA
Tel: 0661-2464253, Mob: +919437085967, E-mail address: hara.jena@gmail.com, hmjena@nitrkl.ac.in

ABSTRACT

The gas holdup characteristics in a cocurrent three-phase fluidized bed with liquid as continuous phase and gas as the discontinuous phase have been studied. Air, water and ceramic raschig rings are used as the gas, liquid and solid phases respectively. The experiments were carried out in a 0.1 m internal diameter, 1.88 m height vertical Plexiglas column. The results indicate that gas holdup increases with increasing gas velocity and decreasing liquid velocity. From the correlation developed for gas holdup it is found that gas holdup is a strong function of modified gas Reynolds number and a weak function of the liquid Reynolds number. The experimental values have been compared with those calculated from the developed correlation and have been found to agree well.

Key words: Three-phase fluidized bed, Gas holdup, Multiphase flow, Cylindrical particle.

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 hydrodesulphurization 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 operations¹. 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 intimate contact between the gas, liquid and the solid particles in the system and provides substantial advantages for applications in physical, Chemical or biochemical processing involving gas, liquid and solid phases². The knowledge of hydrodynamic properties such as the phase holdups, bubble properties and mixing characteristics is essential for designing and operating these fluidized beds. Among the hydrodynamic properties, the most important for analyzing the performance of a three phase fluidized bed is the bed porosity and the individual phase holdups. The phase holdup characteristics of these fluidized beds have been reviewed and studied by several investigators⁴⁻¹⁷.

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 gas-liquid mass transfer⁵⁻⁶. Although gas holdup in three-phase fluidized beds have received significant attention as summarized in various reviews, most previous work has utilized air, water, and small glass beads as the gas, liquid, and solids respectively. This combination limits the generality and usefulness of the result. The gas holdup in such systems is often considerably lower than for pilot-plant or industrial-scale units. Most researchers attribute this difference to air, water, and glass beads having physical properties which differ considerably from those typically found in industrial processes. These processes are often operated at much higher temperatures and pressures than those used in the traditional air-water-glass beads experimental systems⁷. Nonhomogeneity of the axial phase holdup is also common in slurry bubble columns and three-phase fluidized bed of low-density particles. Only a few studies have addressed the nonhomogeneity of the phase hold ups with high and moderately high density particles⁸.

* Author to whom correspondence may be addressed.

Mr. Hara Mohan Jena, Department of Chemical Engineering, NIT, ROURKELA – 769008 (ORISSA) INDIA
Tel: 0661-2464253, Mob: +919437085967, E-mail address: hara.jena@gmail.com, hmjena@nitrkl.ac.in

The following equations have typically been used to determine the volume fraction (holdup) of each phase: gas holdup (ε_G) liquid holdup (ε_L), and solid holdup (ε_S) in the three phase fluidized bed:

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

$$\frac{\Delta P}{H_E} = g(\rho_G \varepsilon_G + \rho_L \varepsilon_L + \rho_S \varepsilon_S) \quad (2)$$

$$\varepsilon_S = \frac{M_S}{\rho_S A_C H_E} \quad (3)$$

The expanded bed height, H_E in Eq. (2) and (3) is obtained either visually or from the measured pressure drop gradient¹⁰. A more direct method of measuring gas holdup is to simply isolate a representative portion of the test section by simultaneously shutting two quick closing valves and measuring the fraction of the isolated volume occupied by the gas¹¹. Other most promising methods of measuring the local gas holdup are electroresistivity and electro conductivity methods, γ - ray transmission measurement and radioactive tracer technique⁹⁻¹⁷.

In the above cited literature the solid phase used are spherical particles: like glass beads, steel balls, plastic beads and other spherical catalyst particles, cylindrical particles: like aluminum cylinders and pvc cylinders other cylindrical catalyst particles and irregular particles like: sand, irregular gravel, quartz particles etc. having sphericity ranging from 0.7 – 1.0 approximately. Three-phase fluidized beds have been applied successfully in the bio-oxidation process for wastewater treatment in which various low-to-moderate density solid particles of different shape and size are used as cell support. In such reactors high surface area of the particle is desirable, which can be used as solid support for microorganisms, thus resulting in higher mass transfer rate. This can be achieved by the use of hollow cylindrical particles as, these possess very high surface to volume ratio i.e. of low sphericity.

In the present study experiments were conducted to examine the gas holdup of a co-current gas-liquid-solid three-phase fluidized bed with a modified air sparger using liquid as the continuous phase and gas as the discontinuous phase. Ceramic raschig rings having sphericity of 0.58 have been used as the solid phase as it is of moderate density and high surface to volume ratio due to its hollow cylindrical structure. These have been done in order to develop a good understanding of the gas holdup in low-moderate Reynolds number range. Correlation for gas holdups has been developed from experimental data by dimensional analysis approach and compared with the correlation of Safoniuk et al. as they have used cylindrical particle as the solid phase⁷.

EXPERIMENTAL

A schematic diagram of the experimental setup is shown in Fig. 1. The experimental fluidized bed consists of three sections, v.i.z., test section, gas-liquid distributor section, and gas-liquid disengagement section. The test section is the main component of the fluidizer where fluidization takes place, is a vertical cylindrical plexiglas column of 0.1 m internal diameter and 1.88 m long. Any entrained particles retain on the 16 mesh screen attached to the top of this section. The gas-liquid distributor is located at the bottom of the test section and is designed in such a manner that uniformly distributed liquid and gas mixture enters the test section. The distributor section made of plexiglas is fruit-conical of 0.31 m in height, had a divergence angle of 4.5° with one end of 0.0508 m in internal diameter and the other end of 0.1 m in internal diameter having liquid inlet of 0.0254 m internal diameter located centrally at the lower cross-section end. The higher cross-section end is fitted to the test section, with a perforated plate made of G.I. sheet of 0.001 m thick, 0.12 m diameter having open area 20 % of the column area (higher than that used in gas-solid fluidized bed, which uses optimum open area of about 10 %) in between covered with 16 mesh stainless steel screen at the top. Totally 288 numbers of 0.002 m, 0.0025 m and 0.003 m holes in triangular pitch made in 10 circles of nearly 0.005 m gap from centre, so that the large circle of hole is just inside the inner wall of the column. This is done to have less pressure drop at the distributor plate and uniform flow of fluid to the test section. There is an antenna-type air sparger of 0.09 m diameter just below the distributor plate containing 50 numbers of 0.001 m holes, for generating bubbles uniformly throughout the entire cross-section of the column. In this section the gas and liquid streams are 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. The gas-liquid disengagement section at the top of the column is a cylindrical section of

0.026 m internal diameter and 0.034 m height, assembled to the test section having 0.08 m of the test section inside it, which allows gas to escape and liquid to be circulated through the outlet of 0.0254 m internal diameter at the bottom of this section.

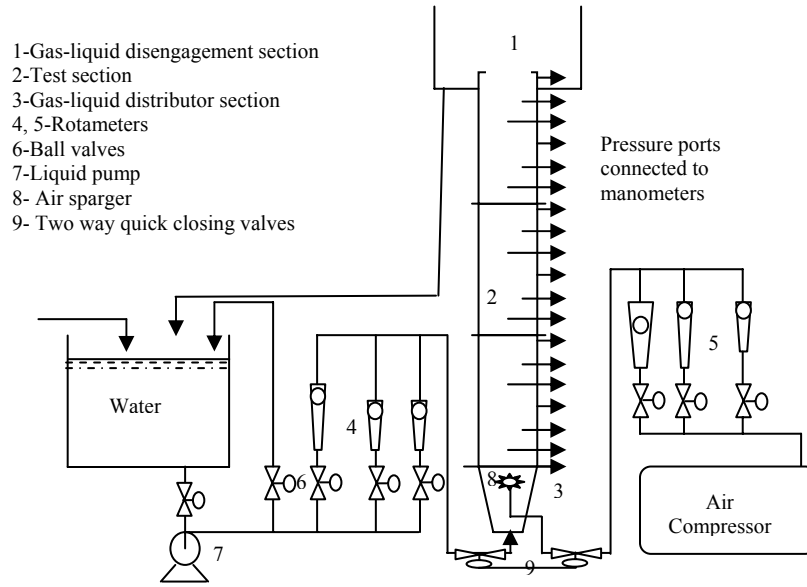


Figure 1. Schematic diagram of the three-phase fluidized bed.

Table 1: Properties of Bed Materials (A), Fluidizing Medium (B).

A. Properties of Bed Materials.

| Materials | L = OD, m | ID, m | Spherical volume-equivalent diameter, m | ρ_s (kg.m ⁻³) |
|----------------------|-----------|--------|---|--------------------------------|
| Ceramic raschig ring | 0.0066 | 0.0033 | 0.006864 | 1670 |

B. Properties of Fluidizing Medium.

| Fluidizing Medium | ρ (kg.m ⁻³) | Viscosity, μ (Ns/m ²) |
|----------------------------|------------------------------|---------------------------------------|
| Air at 30 ⁰ C | 1.168 | 0.0000186 |
| Water at 30 ⁰ C | 998.4 | 0.0008032 |

The three phases present in the column were ceramic raschig rings, tap water and the oil free compressed air. The properties of the bed material, the fluidizing medium 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 reproducible initial static bed height. Water was pumped to the fluidizer at a desired flow rate using water rotameter. Then air was injected into the column through the air sparger at a desired flow rate using air rotameter. Three different calibrated rotameters for each of water and air were used to cover the complete range of flow. Approximately five minutes was allowed to make sure that the steady state was reached, after which the bed expansion was noted by visual observation. For gas holdup measurement, the quick closing valves (9, Fig. 1) in the water and air line were closed simultaneously. At first free board experiment with wide variation of gas and liquid flow were conducted to calculate the two phase gas holdup using the following equation:

$$\varepsilon_G = \left(\frac{H - H_L}{H} \right) * 100 \quad (4)$$

For three-phase fluidization the gas holdup in the bed was calculated using the equation (4). The gas holdup in two-phase condition is observed to be less than the corresponding values at same gas and liquid flow rate (as seen in Figure 4). Thus the gas holdup in the three-phase region was calculated by subtracting the gas

holdup in the two-phase region above the three-phase zone taking the volume of three-phase and two-phase region into account.

RESULTS AND DISCUSSION

Figure 2 shows the variation of fractional gas holdup with superficial liquid velocity at different values of fixed superficial gas velocity. It is seen from the figure that with increasing liquid velocity, the gas holdup decreases. However the variation of fractional gas holdup with liquid velocity is very small. Several workers have reported that the fractional gas holdup is practically unaffected by liquid velocity except at very high liquid superficial velocities⁷. Many others have reported a slight decrease in gas holdup with liquid velocity^{2, 4}. This may possibly be due to at higher liquid velocity the bubbles are fast driven by the liquid.

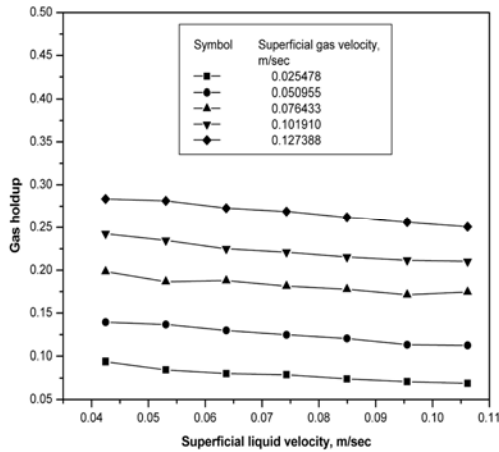


Figure 2. Effect of liquid velocity on gas holdup at constant gas velocity.

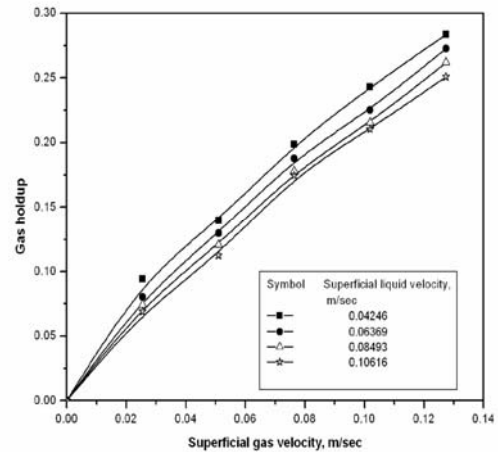


Figure 3. Effect of gas velocity on gas holdup at constant liquid velocity.

Figure 3 represents the variation of fractional gas holdup with superficial gas velocity, at a constant liquid velocity. As seen from the figure, the fractional gas holdup increases monotonically with the gas velocity with higher value of the slope at low gas velocities. Which corroborates the findings of Dhanuka and Stepneck, Safoniuk et al.^{2, 7}. At lower values of gas velocity, an increase in gas velocity results in the formation of a larger number of gas bubbles without appreciable increase in the bubble diameter. Therefore an increasing fractional gas holdup is observed. As gas velocity increases the bubble size grows due to bubble coalescence, and relatively the gas holdup decreases. As the experiment has been conducted for the gas velocity range pertaining to the distributed bubble regime, the decrease in slope is not significant which is observed for the transformation to the slug flow regime.

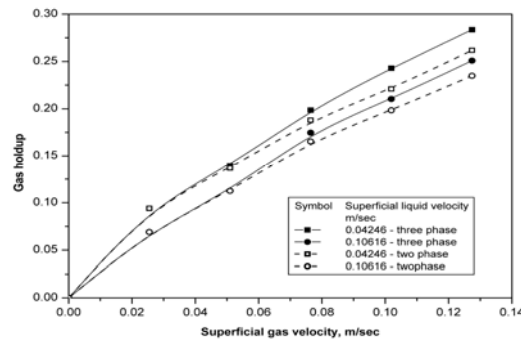


Figure 4. Effect of gas velocity on gas holdup at constant liquid velocity for two-phase and three-phase flow.

Figure 5 shows the effect of liquid to gas velocity ratio on the fractional gas holdup. It is seen from the figure that the holdup fraction versus L/G plot show two distinct regimes. As the liquid to gas velocity ratio decreases, the fractional gas holdup initially increases very slowly but below a certain L/G ratio the holdup fraction starts increasing very fast. Hence it can be concluded that there is an optimum L/G ratio below which it is advantageous to operate the three-phase fluidized bed system with hollow cylindrical particles. By drawing a tangent (shown by a broken line) to the holdup fraction versus L/G plots, the optimum values of L/G have been obtained for the different values of liquid velocity.

Flow conditions influence mass and heat transfer, and determine the mixing quality. When scaling up a gas-liquid-solid reactor with hollow cylindrical particles it is important to maintain the flow regime identical to that existing in the experimental column. As no correlation have been found in the literature to predict the gas holdup in a three-phase fluidized bed with hollow cylindrical particles, the experimental data have been analyzed on the basis of dimensional analysis so as to predict holdup in the present system.

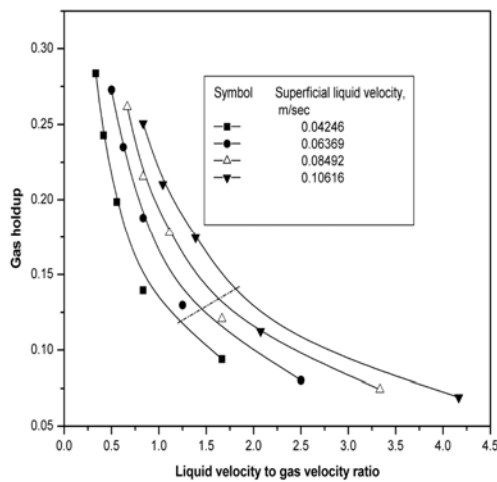


Figure 5. Effect of liquid velocity to gas velocity ratio on gas holdup.

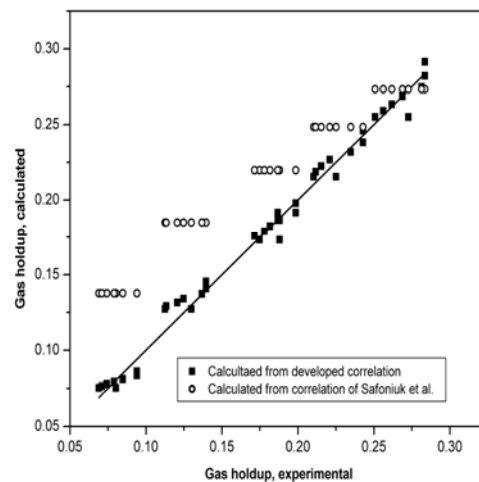


Figure 6. Comparison of experimental values of gas holdup with those calculated from correlation.

Conceivable variables on which the gas holdup in the present system may depend are: gas velocity (V_G), liquid velocity (V_L), particle size (D_p), sphericity of particle (ϕ_p), column diameter (D_C), expanded bed height (H_E), static bed height (H_S), diameter of the sparger orifice (D_O), density of gas (ρ_G), density of liquid (ρ_L), density of solid (ρ_S), viscosity of gas (μ_G), viscosity of the liquid (μ_L), surface tension of liquid (σ_L) and gravitational constant (g). In the present study except V_G , V_L , and H all other parameters are maintained constants (all the experiments were conducted at $30 \pm 2^\circ\text{C}$). Further H depends upon V_G and V_L and can be neglected. Thus the correlations can be developed in the form:

$$\varepsilon_G = f_1[V_L]^a[V_G]^b \quad (4)$$

or

$$\varepsilon_G = f_2[Re_L]^c[Re_G]^d \quad (5)$$

The modified gas Reynolds number, Re_G is used here instead of simply the gas Reynolds number as the role of gas properties like viscosity and density are negligible. Correlations in both the form have been developed. The results have been fitted to power-law equation passing through origin (zero gas holdup at zero gas flow). For the range of $0.042463 \leq V_L \leq 0.106157$ ($362.3 \leq Re_L \leq 905.7477$) and $0.025478 \leq V_G \leq 0.127389$ ($217.3795 \leq Re_G \leq 1086.897$), this leads to:

$$\varepsilon_G = 0.8791(V_L)^{-0.14497}(V_G)^{0.7582} \quad (6)$$

and

$$\varepsilon_G = 0.0034(\text{Re}_L)^{-0.14497}(\text{Re}_G)^{0.7582} \quad (7)$$

(with a standard deviation of 0.00785 and a correlation factor of 0.9821). Many empirical correlations are available in the literature, but most of them are with spherical particles or other irregular particles. The latest correlation with cylindrical particles of Safoniuk et al.⁷ is given by,

$$\varepsilon_G = 0.0139(\text{Re}_G)^{0.426} \quad (8)$$

As no work is available with hollow cylindrical particles, the developed correlation is compared with the correlation of Safoniuk et al.⁷ and shown in Figure 6. Figure 6 shows the comparison of experimental values of gas holdup with those calculated from equations (7) and (8). Higher holdup is seen for equation (8) as the effect of liquid Reynolds number is neglected here. But in the present work there is decrease in gas hold of with increase in liquid velocity or liquid Reynolds number. The developed correlation can be used universally for any type of hollow cylindrical particles.

CONCLUSIONS

Investigations have been carried out to study the effect of liquid and gas velocities on gas holdup in a gas-liquid-solid fluidized bed with hollow cylindrical particles. It may be concluded that the overall gas holdup is a strong function of gas velocity and modified gas Reynolds number. The gas holdup is a weak function of liquid velocity and liquid Reynolds number. Correlation for the estimation of the gas holdup has been proposed. The values calculated from the developed correlation have been compared with the experimental ones and also with those obtained by the correlation of Safoniuk et al.⁷. Fairly good agreement has been obtained thus emphasizing the validity of the developed correlation over the range of the operating parameters investigated.

NOMENCLATURE

| | |
|-----------------|---|
| A_C | cross-sectional area of the column, (m ²) |
| D_p | particle diameter, (m) |
| g | acceleration due to gravity, (m/s ²) |
| H | height of the test section, (m) |
| H_E | expanded bed height, (m) |
| H_L | height of liquid or solid-liquid mixture remained in the column after escape of the gas, (m) |
| M_S | mass of the solid particles in the bed, (kg) |
| ΔP | pressure drop, (Pa) |
| Re_G | modified gas Reynolds number = $(V_G/V_L) \text{Re}_L = \rho_L D_p V_G/\mu_L$, (dimensionless) |
| Re_L | liquid Reynolds number = $\rho_L D_p V_L/\mu_L$, (dimensionless) |
| V_L | liquid velocity, (m/s) |
| V_G | gas velocity, (m/s) |
| ε_G | fractional gas holdup, (dimensionless) |
| ε_L | fractional liquid holdup, (dimensionless) |
| ε_S | fractional solid holdup, (dimensionless) |
| ρ_G | gas density, (kg/m ³) |
| ρ_L | liquid density, (kg/m ³) |
| ρ_S | solid density, (kg/m ³) |

REFERENCES

1. K. Muroyama and L.S. Fan, *AICHE J.*, **31**, 1 (1985).
2. V.R. Dhanuka and J.B. Stepanek, *Fluidization*, Cambridge University Press, UK, (1978)p. 179.
3. J.M. Begovich and J.S. Watson, *Fluidization*, Cambridge University Press, Cambridge, (1978), p.190.
4. R.C. Darton and D. Harrison, *Chem. Eng. Sci.* **50**, 581 (1975).
5. J.M. Schweitzer, J. Bayle and T. Gauthier, *Chem. Eng. Sci.*, **56**, 1103 (2001).
6. L.S. Fan, F.R. Bavarian, I. Gorowara and B.E. Kreischer, *Powder Technol.*, **53**, 285 (1987).
7. M. Safoniuk, J.R. Grace, L. Hackman and C.A. Mcknight, *AICHE J.*, **48**, 1581 (2002).
8. A. Catros, J.R. Bernard, C. Briens and M.A. Bergougnou, *Can. J. Chem. Eng.*, **63**, 754 (1985).

9. N. Epstein, *Can. J. Chem. Eng.*, **59**, 649 (1981).
10. S.P. Kim, C.G.J. Baker and M.A. Bergougnou, *Can. J. Chem. Eng.*, **53**, 134 (1975).
11. N. Epstein, *Can. J. Chem. Eng.*, **59**, 649 (1981).
12. Y.H. Yu and S.D. Kim, *Chem. Ind. Techno.*, **4**, 14 (1986).
13. L. S. Fan, *Gas-Liquid-Solid Fluidization Engineering* (Butterworth's Series in Chemical Engineering) Butterworth Publishers, Boston, MA (1989).
14. J.H. Han, G. Wild and S.D. Kim, *Chem. Eng. J.*, **43**, 67 (1990).
15. D.H. Lee, A. Macchi, J.R. Grace and N. Epstein, *Chem. Eng. Sci.*, **56**, 6031 (2001).
16. S.L.P. Lee and H.I.D. Lasa, *AIChE J.*, **33**, 1359 (2004).
17. P. Dargar and A. Macchi, *Chem. Eng. and Proc.*, **45**, 764 (2006).