Prediction of gas holdup in a three-phase fluidized bed from bed pressure drop measurement

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ABSTRACT

Experiments were conducted to investigate the overall gas holdup characteristics in a co-current three-phase fluidized bed with liquid as the continuous phase using air, water and glass beads as the gas, liquid and solid phases respectively. These studies were carried out in a 0.1 m internal diameter, 1.24 m height vertical Plexiglas column with an antenna type air sparger at different water and air flow rates using four different particle sizes and initial static bed heights. Bed pressure drop measurement was used as the basis for the calculation of fractional gas holdup. From the data obtained through a set of experimental runs, an empirical correlation was developed for the overall gas holdup using dimensional analysis and non-linear regression technique and was compared with some existing correlations. It was found that the gas holdup increased with the flow rate of air and decreased with an increase in the water flow rate.

Key words: Gas-liquid-solid fluidization; Pressure drop; Gas holdup; Dimensional analysis; Multiphase flow

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1. Introduction

A gas-liquid-solid fluidized bed, as defined in this study, is a batch of solid particles which is fluidized by a co-current up flow of liquid as the continuous phase, and gas as the dispersed bubble phase. The gas-liquid-solid fluidized bed has emerged in recent years as one of the most promising devices for three-phase operation. Such a device is of considerable industrial importance as evident from its wide application in chemical, petrochemical and biochemical processing (Muroyama and Fan, 1985). While recent emphasis has been given on the optimal design and the operation of the gas-liquid-solid fluidized bed, it is equally important to examine the characteristics of the three-phase operation with liquid as the continuous phase. Therefore, the hydrodynamic properties such as the bed pressure drop, minimum fluidization velocity, bed porosity, phase holdups, bubble properties and the mixing characteristics have to be studied in order to provide the basic information required for the design of such fluidized bed systems.

The gas holdup is one of the most important characteristics for analyzing the performance of a three-phase fluidized bed. For chemical processes where gas-liquid 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 (Fan et al., 1987; Schweitzer et al., 2001). Considerable work has been carried out on the gas holdup in three-phase fluidized columns. Various aspects of these fluidized beds have been reviewed by several investigators (Catros et al., 1985; Epstein, 1981; Kato et al., 1981; Kim et al., 1975; Muroyama and Fan, 1985; Fan, 1989; Okamura et al., 1989; Han et al., 1990; Lee and Lasa, 1987; Lee et al., 2001; Safoniuk et al., 2002; Song et al., 1989; Yu and Kim, 1986), which include the importance of gas holdup and various factors affecting it.

The gas holdup characteristic depends upon the bubble size and its dispersion in the bed. Thus the generation of fine gas bubble is important which is possible by the suitable

design of the air sparger (Thorat et al., 1998). Although the use of various types of sparger is seen in the literature, but little attention has been made in the precise design of an air sparger which can avoid the high pressure drop in the distributor section. In the present study, an antenna type air sparger has been used, as it is quite efficient in producing fine air bubbles with less pressure drop in the distributor section.

The bed height and the individual phase holdups have been determined from static pressure profiles upto the entire height of the column (Kim et al., 1972,1975).The bed height has been taken as the point at which a change in the slope of the pressure profiles was observed. Previous studies also include the bed characteristics at considerable higher gas velocities and over much wide ranges of liquid surface tension and viscosity. The local liquid holdup has been directly measured by the electro conductivity techniques (Muroyama and Fan, 1985). A capacitance probe technique has also employed to measure the solid and liquid holdups (Yu and Rittman, 1997).

To obtain average gas holdups across the ebulated bed, pressure drops have been measured across the bed height (Dhanuka and Stepanek, 1978; Darton and Harrison, 1975; Dargar and Macchi, 2006). Assuming negligible acceleration and wall friction, the measured pressure drop has been related directly to the density of the individual phases by,

$$\frac{\Delta P}{\Delta H} = g(\rho_g \varepsilon_g + \rho_l \varepsilon_l + \rho_s \varepsilon_s) \tag{1}$$

The solids holdup can be calculated based on the overall bed expansion (Jean and Fan, 1986) and the known solids loading of the bed,

$$\varepsilon_s = \frac{M_s}{\rho_s A H_e} \tag{2}$$

Since the phases present in the reactor are the gas, liquid, and the particles,

$$\varepsilon_l = (1 - \varepsilon_s - \varepsilon_g) \tag{3}$$

Eqs. (1), (2) and (3) constitute the three equations with three unknowns, and, hence, allow the overall gas holdup to be estimated.

The bed height in Eqs. (1) and (2) has been obtained either visually or from the measured pressure drop gradient (Kim et al., 1975, Dargar and Macchi, 2006). 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 isolated volume occupied by the gas (Epstein, 1981). Some other promising methods used for measuring the local gas holdup are electro conductivity methods (Bhatia and Epstein, 1974), γ-ray transmission measurements (Ostergaard, 1977), electroresistivity (Begovich and Watson, 1978), and radioactive tracer techniques (Yu and Rittman, 1997).

In the present investigation, an attempt has been made to study the gas holdup characteristics of a three-phase fluidized bed in a broader range of operation. Experiments have been 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. Spherical glass beads have been used as the solid phase. An antenna type air sparger has been used for the generation of fine bubbles. By the use of such air sparger the pressure drop in the distributor section can be avoided (Meikap et al., 2002). A correlation for overall gas holdup has been developed from dimensional analysis. Comparisons have been made between the overall gas holdup values calculated from the developed and the existing correlations and the experimental ones.

2. Experimental set-up and procedure

A schematic representation of the experimental setup is shown in Figure 1. The experimental fluidized bed consists of three sections, viz., the test section, the gas-liquid distributor section, and the gas-liquid disengagement section. The test section is the main component of the

fluidizer where fluidization takes place. It is a vertical cylindrical Plexiglas column of 0.1 m internal diameter and 1.24 m height. Any entrained particles are retained on the 16-mesh screen attached to the top of the column. 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 Perspex is fructo-conical of 0.31 m in height, and has a divergence angle of 4.5^0. The liquid inlet of 0.0254 m in internal diameter is located centrally at the lower cross-sectional 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 equal to 20 % of the column cross-sectional area with a 16 mesh (BSS) stainless steel screen in between. There is an antenna-type air sparger of 0.09 m diameter just below the distributor plate containing 50 number of 0.001 m holes, for generating uniform bubbles to flow throughout the cross-section of the column. In this section, the gas and the liquid streams are merged and passed through the perforated grid. The mixing section and the grid ensured 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 with 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.

For pressure drop measurement in the bed, the pressure ports have been fitted to the manometers filled with carbon tetrachloride. Pressure ports are available at seven different levels of equal spacing including one at bottom and one at the top of the test section. This has been done to measure the pressure drops at a particular section at three different radial positions, viz., at the wall, at the center of the column and at one-fourth of the diameter of the column from the wall. With this arrangement, the wall effect, expanded bed height, distribution of particle concentration and the gas holdup can be studied clearly.

The three-phases solid, liquid and gas are glass beads, tap water and the oil free compressed air, respectively. The scope of the experiment is presented in Table 1. The air-water flow was 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 using water rotameter. The air was then injected into the column through the air sparger at a desired flow rate using air rotameter. Three calibrated rotameters with different ranges each for water as well as for air have been used for the accurately record of the flow rates.

All experiments have been started with the column completely filled with water and glass beads, the initial level of manometer were adjusted to have zero drop value. For liquid-solid experiment the liquid flow rate was gradually increased. Approximately five minutes were allowed to make sure that the steady state was reached. Then the readings of the manometers and the expanded heights of the bed were noted. For gas-liquid-solid experiment with little flow of liquid close to zero, the air was slowly introduced and gradually increased to the desired flow rate afterwards the liquid flow rate increased and the readings were noted down, as above. From the total pressure drop in the bed the gas holdup was determined. The procedure was repeated for different values of initial static bed height, particle size and gas velocity.

3. Results and discussion

Experiments were conducted with the gas and liquid flow rates which varied from 0 to 0.10615 m s^{-1} and from 0.02123 – 0.16985 m s^{-1}, respectively. The temperature was maintained at 30 ± 2^0C. To ensure steady state in operation at least five minutes were allowed. Readings for bed expansion and pressure drop were noted down. Each experiment

was repeated thrice to have an accurate reading. The experimental results have been presented graphically in this section. Empirical equations have been developed.

The variation of bed pressure drops in the fluidization regime (i.e. at higher fluid flow rates than at minimum fluidization) with superficial liquid velocity at constant values of gas velocity has been shown in Figure 2. The dotted line in the figure shows the effective bed weight (i.e. buoyant weight) per unit area. The experimental pressure drop in the fluidization regime for liquid-solid system agrees well to that of the calculated buoyant weight per unit area. For gas-liquid-solid system the observed bed pressure drop recorded in manometer is found to be less than that for liquid-solid bed. With higher values of gas flow rate, the observed bed pressure drop decreases further. This is due to the increased gas holdup in the bed which decreases the hydrostatic pressure. For superficial gas velocity of 0.10615 m s^{-1}, negative bed pressure drop has been measured by the manometer. The equivalent liquid column of the difference in bed pressure drop of the liquid-solid and gas-liquid-solid bed has been considered to be the region of the column filled with gas. The gas holdup has been calculated using the following equation.

$$\varepsilon_g = \frac{(\Delta p^{ls} - \Delta p^{gls})/(\rho_l g)}{\Delta H} \tag{4}$$

In Eq. (4) ΔH is the height of the pressure tapping just above the expanded bed height from the bottom. The measured bed pressure drop ΔP reported has been taken to be the pressure drop between the same two tapings. This has been done in order to avoid the interference of the gas holdup of the bubble column region at the top on the gas-liquid-solid bed.

Figure 3 represents the variation of fractional gas holdup with superficial gas velocity, at constant liquid velocities. 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. This corroborates the findings of Begovich and Watson (1978), Dhanuka and Stepanek (1978),

Lee and Lasa (1987), Briens et al. (1997), Safoniuk et al. (2002), and Dargar and Macchi (2006). In the lower range 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 the gas velocity increases, the bubble size grows due to the bubble coalescence, and relatively the gas holdup decreases. As the experiments have been carried out for the distributed to coalesced bubble regime, the decrease in slope is not significant. This behavior has been observed for the transformation of bubble regime to the slug flow regime.

Figure 4 shows the variation of fractional gas holdup with superficial liquid velocity at different values of constant 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 small. It has been reported by Safoniuk et al. (2002) that the fractional gas holdup is practically unaffected by liquid velocity except at very high liquid superficial velocity range. According Breins et al. (1997) the gas holdup decreases with liquid velocity but at higher liquid velocity range it is almost constant. Dhanuka and Stepanek (1978), Begovich and Watson (1978), and Lee and Lasa (1987) have reported a slight decrease in gas holdup with liquid velocity over large range of the later. At higher liquid velocity large number of fine bubbles are possible as the flow regime is completely distributed or dispersed, for which the gas holdup should be more. But the decrease in gas holdup with liquid velocity may possibly be due to the fact that at higher liquid velocity the bubbles are fast driven by the liquid. The residence time of the bubbles decreases with the liquid velocity and hence the gas holdup is likely to decrease.

In Figure 5 the variation of fractional gas holdup with superficial liquid velocity at a constant gas velocity for different particle sizes has been represented. The gas holdup decreases with liquid velocity like the above finding. But the measured gas holdup is found to

increase with particle size. For glass beads of 0.00218 m and 0.00258 m size, the gas hold up is almost the same. Higher gas holdup for bigger size particles may be attributed to their enhanced bubble breaking capacity. Kim et al. (1975) used glass beads of 0.001 m and 0.006 m in their study and have reported the existence of two distinct types of three-phase fluidized beds, viz., "bubble coalescing" and "bubble disintegrating" beds. According to them 0.006 m glass beads in air-water system exhibited bubble disintegrating behaviour. However 0.001 m glass beads exhibited bubble coalescing behaviour. They have reported the existence of critical particle size which separates the "bubble coalescing regime" from the "bubble disintegrating regime". The critical size for particles with a density similar to that of glass has been reported by them to be about 0.0025 m in diameter for the air-water system. This seems to be a well fit to the present experimental finding. In the present investigation for particles of size less than 0.00258 m, the measured gas holdup shows similar trend. However for particles bigger than 0.00258 m, an increase in the gas holdup which has obtained in this investigation has not been reported in the literature. The difference in gas holdup for different glass beads is more in the higher liquid velocity range. This may be due to the better fluid particle interaction and higher mass of particles adds up to its bubble breaking behaviour in such a situation, thus resulting in large number of small size bubbles. The investigation of Dargar and Macchi (2006) also indicates the same behaviour for air-water system. They have used 0.0012 m and 0.005 m glass beads. Higher gas holdup has been reported by them for 0.005 m size glass beads over the other sizes. Fan et al. (1987) have shown opposite behaviour for 0.001, 0.003, 0.004, 0.006 m glass beads in aqueous solution of 0.5-wt% of t-pentanol. With increase in particle size, reduced gas holdup has been reported by them.

The variation of fractional gas holdup with liquid velocity for different static bed inventories (initial static bed heights) at a constant gas velocity are shown in Figure 6. It is clear from the figure that at low liquid velocity range the gas holdup is almost same for

different static bed heights. But at high liquid velocity range, i.e. higher bed voidage there is slight change in the gas holdup. In the higher liquid flow range, a marginal increase in the gas holdup is observed with higher bed heights. This may possibly be due to the gas-liquid-solid interaction for a longer time in the bed for higher initial static bed height. Study on the effect of bed inventory on gas holdup is not seen in the literature.

Figure 7 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 at a faster rate. 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.

Figure 8 shows the variation of the fractional gas holdup obtained from the experiment and those predicted from the correlations proposed by various researchers with gas velocities at a constant liquid velocity. The experimental data is almost the same as predicted by the correlation of Begovich and Watson (1978). There is also a very close agreement between the experimental and the predicted gas hold up from the correlation of Yu and Rittman (1997) at low gas velocity, but as the gas velocity increases the gas velocity deviates negatively from the experimental. Correlation of Safoniuk et al. (2002) for all the ranges predict high values for the gas hold up. Where as correlations of Catros et al. (1985) and Song et al. (1989) predict the gas holdup values around the experimental ones, which is differs with the variation in gas velocity. For gas velocities less than 0.04 m s^{-1} and 0.06 m s^{-1} correlations of Catros et al. (1985) and Song et al. (1989) respectively, predict gas holdup

more than the experimental, whereas for the higher gas velocity than these, the predicted gas holdup is less than the experimental ones for liquid velocity of 0.07431 m s^{-1}.

The flow conditions influence mass and heat transfer, and determine the mixing quality. While 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 has been found in the literature to predict the gas holdup in a three-phase fluidized bed which shows the simultaneous effect of gas and liquid velocities, particle size, column dimension and bed mass, the present experimental data have been analyzed on the basis of dimensional analysis so as to predict gas holdup in systems similar to the present system and operating nearly the same conditions.

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). The large number of possible variables on which the dispersed phase holdup depends has been reduced to a pertinent few, since many of these variables are interrelated or are maintained as constant.

Therefore, if a theoretical relation exists between the true fractional gas holdup, ε_g, and the physical characteristic, and flow variables of the system, then ε_g may be written in the following form:

$$\varepsilon_g = f_1(V_g,\ V_l,\ \rho_g,\ \mu_g,\ \rho_l,\ \mu_l,\ d_p,\ H_s,\ D_c,\ g) \tag{5}$$

The dimensional analysis carried out indicates that the fractional gas holdup may be simplified to Eq. (6) as,

$$\varepsilon_g = f_2[Fr_g]^a\ [Re_l]^b[H_r]^c[d_r]^d \tag{6}$$

In order to establish the functional relationship between ε_g and the various dimensionless groups in Eq. (6), multiple linear regression analysis has been used to evaluate the constant and coefficients of the equation, assuming power law functional relationship. It can be seen that the following equation, which yields the regression coefficient of 0.994 and a standard deviation of percentage error 2.57, presents the best possible correlation among the family of equations.

$$\varepsilon_g = 5.53(Fr_g)^{0.4135}\,(Re_l)^{-0.1808}\,(H_r)^{0.0597}\,(d_r)^{0.0873} \tag{7}$$

The values of gas holdup predicted by Eq. (7) have been plotted against the experimental values of fractional gas holdup in Figure 9. Very close agreement between the experimental and calculated values from the developed correlation is seen. This has been possible due to the repeated experimentation and rejection of odd data points. The correlation (Eq. (7)) is highly significant at 99% confidence level.

4. Conclusions

Gas hold-up characteristics are one of the important design parameters for gas-liquid-solid three-phase fluidized bed system since the rate of gas-liquid mass transfer is influenced by the gas holdup. In this paper an attempt has been made to predict gas holdup from measurement of bed pressure drop for air-water-solid particles in the fluidization regime. Detailed experimental investigations have been carried out to study the effect of gas velocity, liquid velocity, particle size, and initial static bed height on gas holdup in a gas-liquid-solid fluidized bed. A specially designed antenna type air sparger has been used for the generation of fine size bubbles. It is interesting to note that the gas hold-up is a strong function of gas to liquid velocity ratio and the gas hold-up drastically reduces upto the ratio of 1.5. Measurement of gas holdup has confirmed the known fact that the structure of the bed is different for small and large particles, with a transition taking place at particle size of 0.00258

m. A correlation has been developed to predict the gas hold-up for three-phase fluidized bed with a correlation co-efficient of 0.994. It has been found that the gas hold-up predicted was in good agreement with experimental values. A maximum of 6.5 % deviations was found in all flow conditions.

Nomenclature

A	cross-sectional area of the column, m^2
D_c	diameter of the column, m
d_p	particle diameter, m
d_r	particle diameter to column diameter ratio $= d_p/D_c$
Fr_g	Froude's number $= V_g^2/(Dc.g)$
g	acceleration due to the gravity, ms^{-2}
H_e	height of expanded bed, m
H_r	bed aspect ratio $= H_s/D_c$
H_s	initial static bed height, m
ΔH	height between pressure tapping, m
M_s	mass of the solid in the bed, kg
Δp^{ls}	pressure drop for liquid-solid fluidization, Pa
Δp^{gls}	pressure drop for gas-liquid-solid fluidization, Pa
$\dfrac{\Delta P}{\Delta H}$	pressure gradient in the bed, Pa m^{-1}
Re_l	liquid Reynolds number $= \rho_l\, V_l\, D_c/\mu_l$
V_g	gas velocity, m s^{-1}
V_l	liquid velocity, m s^{-1}

Greek Symbols

$\mu_{l,}\ \mu_g$	liquid and gas phase viscosity, Pa s
$\varepsilon_{g,}\ \varepsilon_{l,}\ \varepsilon_s$	gas, liquid and solids holdups
$\rho_{g,}\ \rho_{l,}\ \rho_s$	gas, liquid and particle density, kg m^{-3}

Subscripts

g	gas phase
l	liquid phase
s	solid phase

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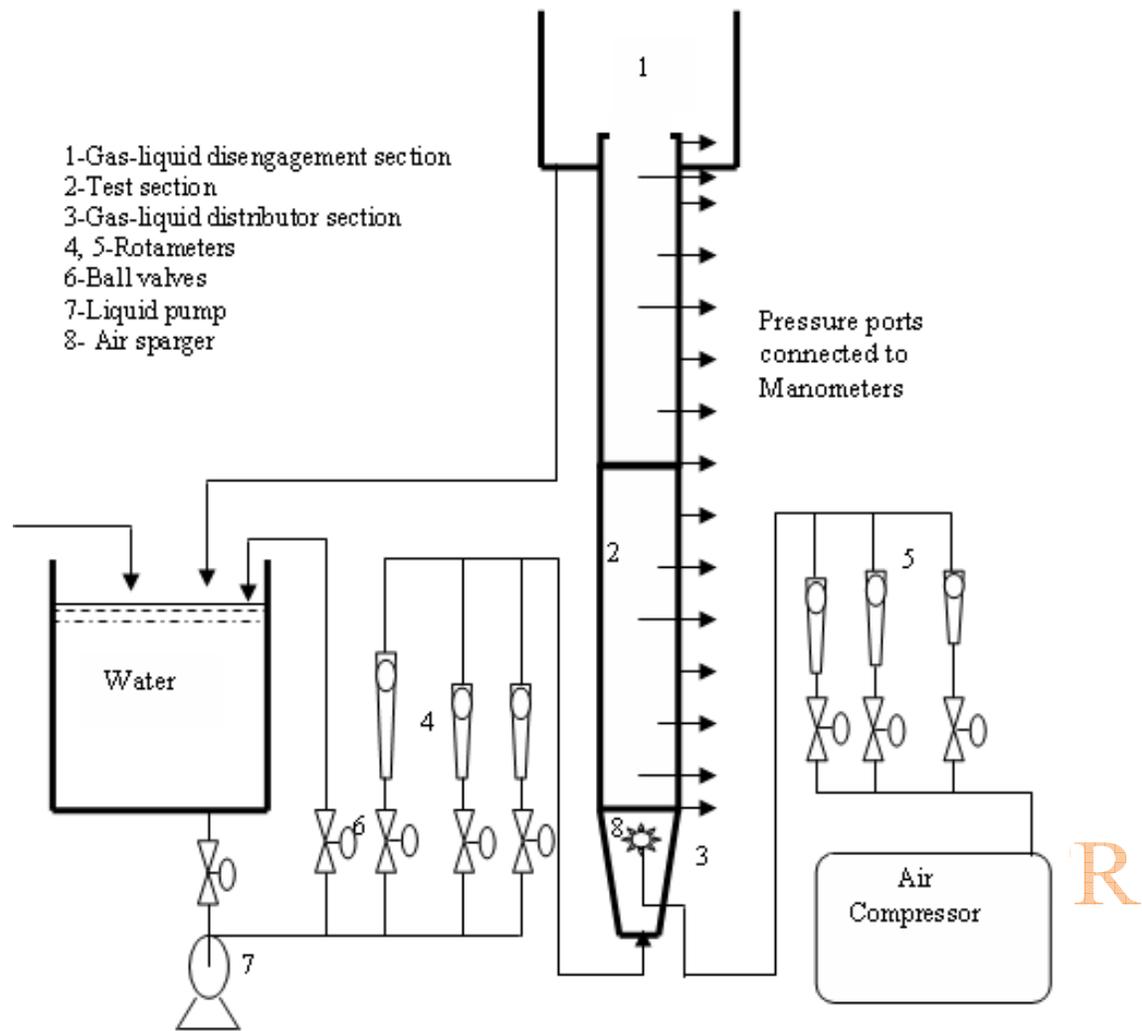
FIGURE CAPTIONS

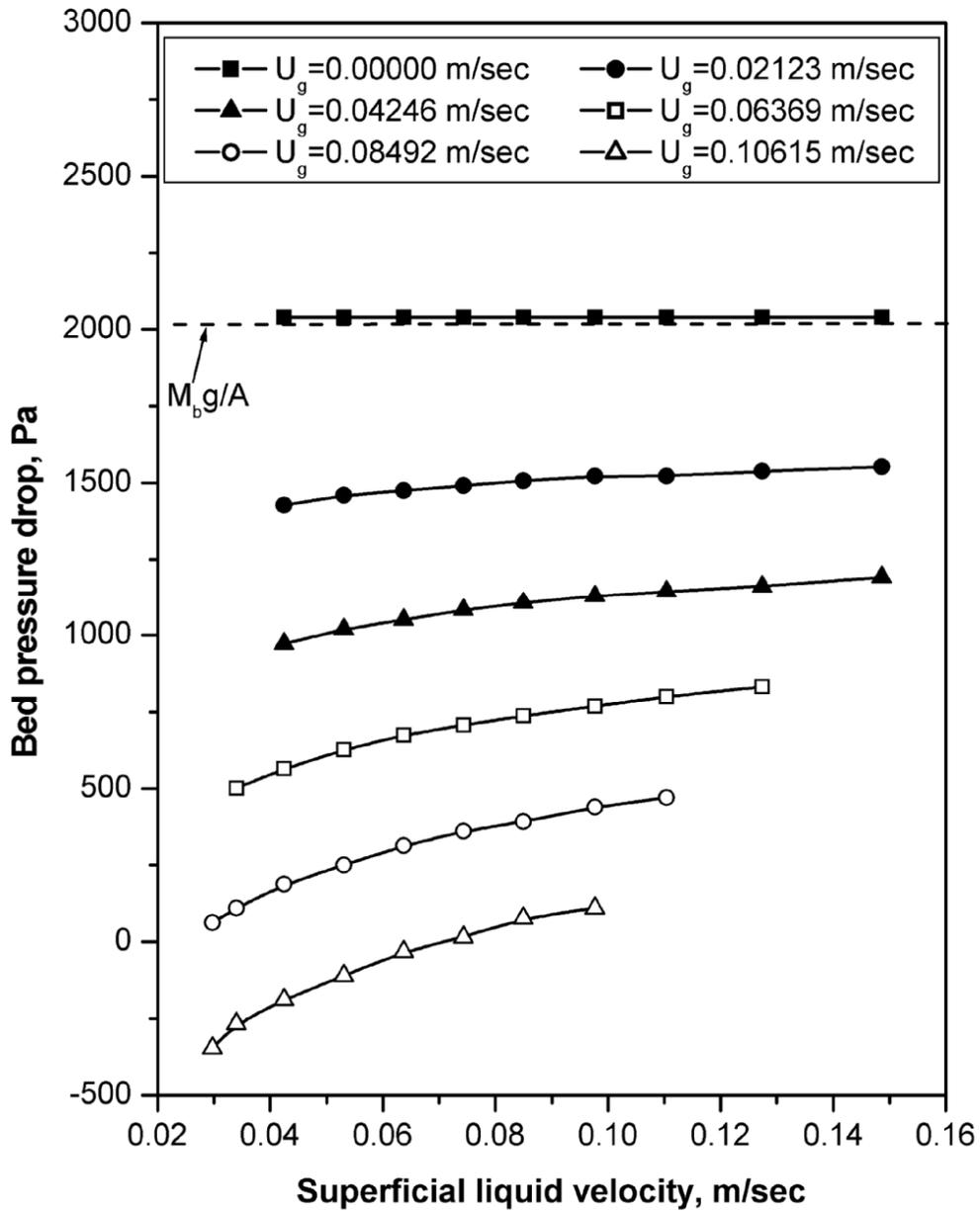*Figure 1*.   Schematic representation of the three-phase fluidized bed.

Figure 2.   Variation of measured bed pressure drop with liquid velocity for different gas velocities at [H_s = 0.256 m, d_p = 0.00307 m].
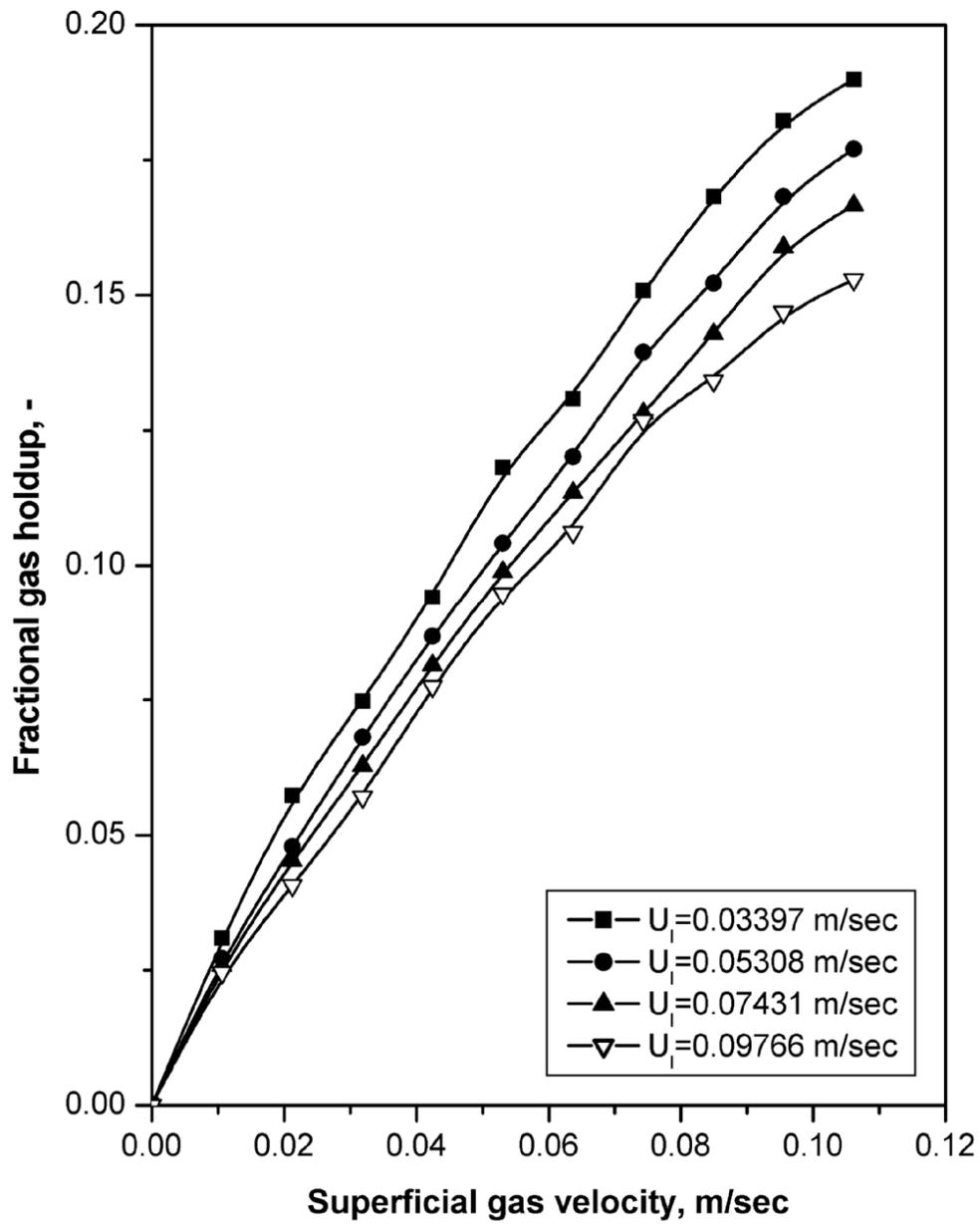
Figure 3.   Variation of gas holdup with gas velocity for different liquid velocities at [H_s = 0.256 m, d_p = 0.00307 m].
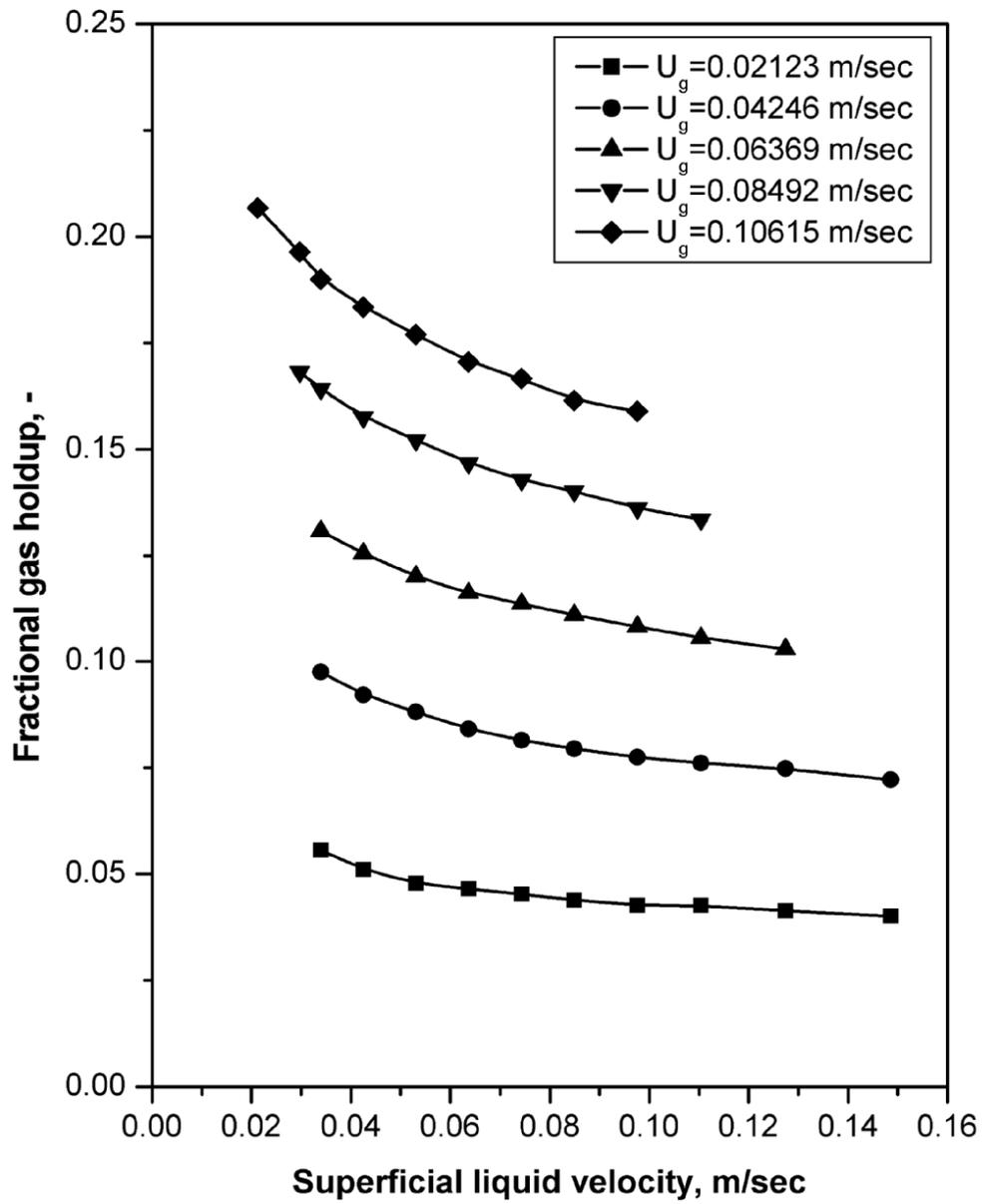
Figure 4.   Variation of gas holdup with liquid velocity for different gas velocities at [H_s = 0.256 m, d_p = 0.00307 m].
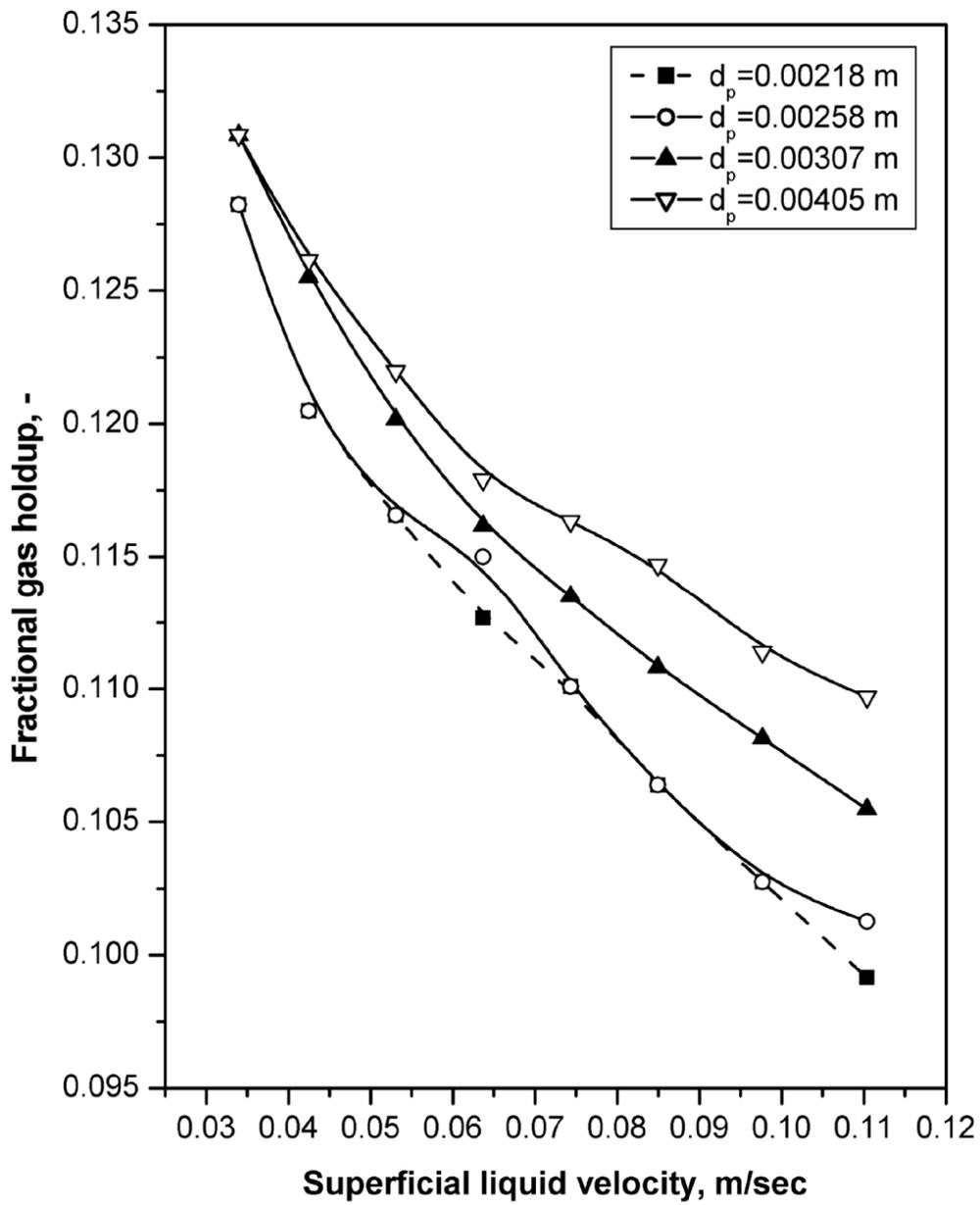
Figure 5.   Variation of gas holdup with liquid velocity for different particle sizes at [H_s = 0.256 m, U_g = 0.06369 m s^{-1}].

Figure 6.   Variation of gas holdup with liquid velocity for different static bed heights at [d_p = 0.00307 m, U_g = 0.06369 m s^{-1}].

Figure 7.   Variation of gas holdup with liquid to gas velocity ratio for different liquid velocities at [H_s = 0.256 m, d_p = 0.00307 m].

Figure 8. Comparison of different correlations with present investigation.

Figure 9. Comparison of experimental and calculated values of fractional gas holdup from Eq. (7).

TABLE CAPTIONS

Table 1.   Properties and variables used in the experiment

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

Figure 2. Variation of measured bed pressure drop with liquid velocity for different gas velocities at [H_s = 0.256 m, d_p = 0.00307 m].

Figure 3. Variation of gas holdup with gas velocity for different liquid velocities at [H_s = 0.256 m, d_p = 0.00307 m].

Figure 4. Variation of gas holdup with liquid velocity for different gas velocities at [H_s = 0.256 m, d_p = 0.00307 m].

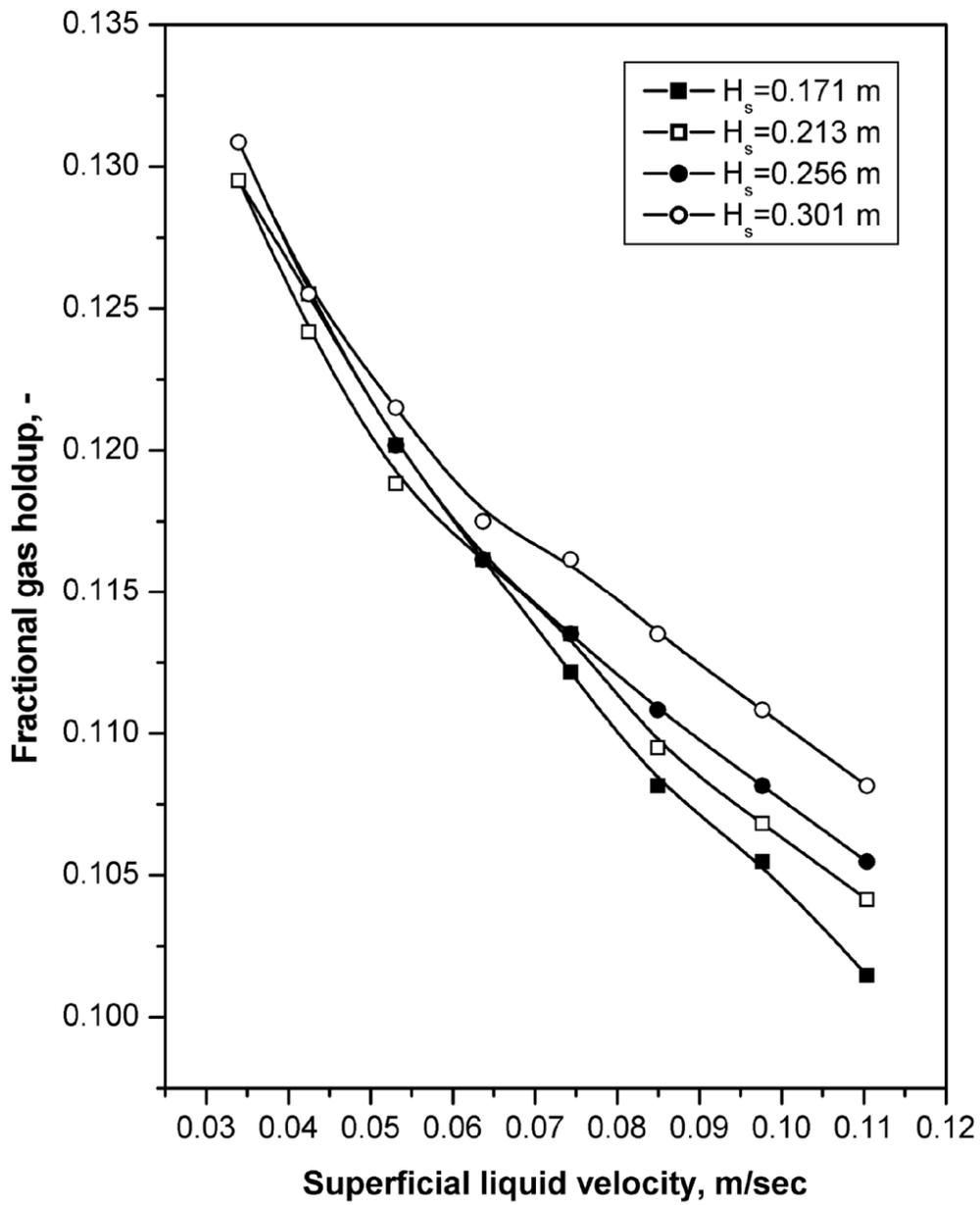*Figure 5*. Variation of gas holdup with liquid velocity for different particle sizes at [H$_s$ = 0.256 m, U$_g$ = 0.06369 m s^{-1}].

Figure 6. Variation of gas holdup with liquid velocity for different static bed heights at [d_p = 0.00307 m, U_g = 0.06369 m s^{-1}].
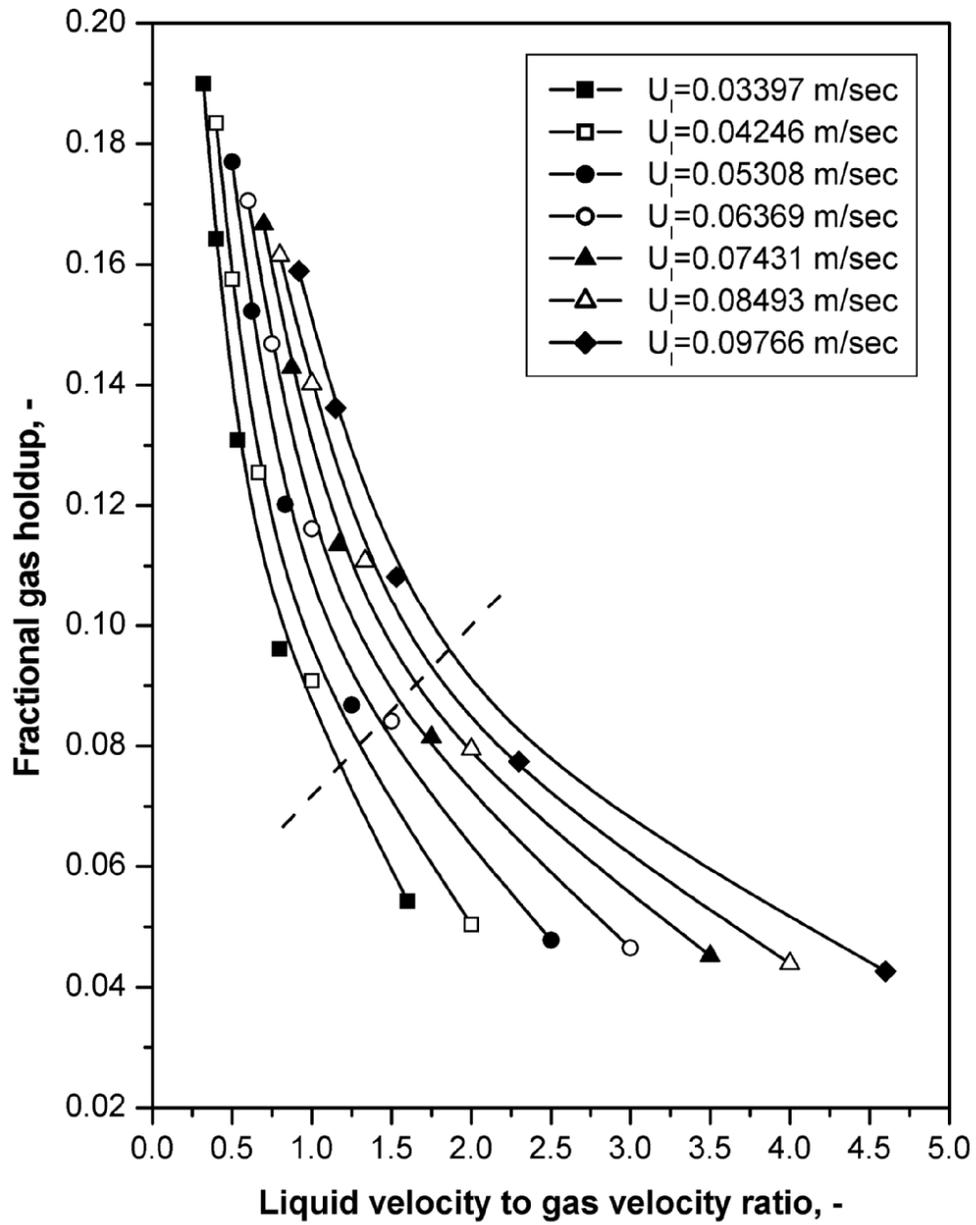
Figure 7. Variation of gas holdup with liquid to gas velocity ratio for different liquid velocities at [H_s = 0.256 m, d_p = 0.00307 m].
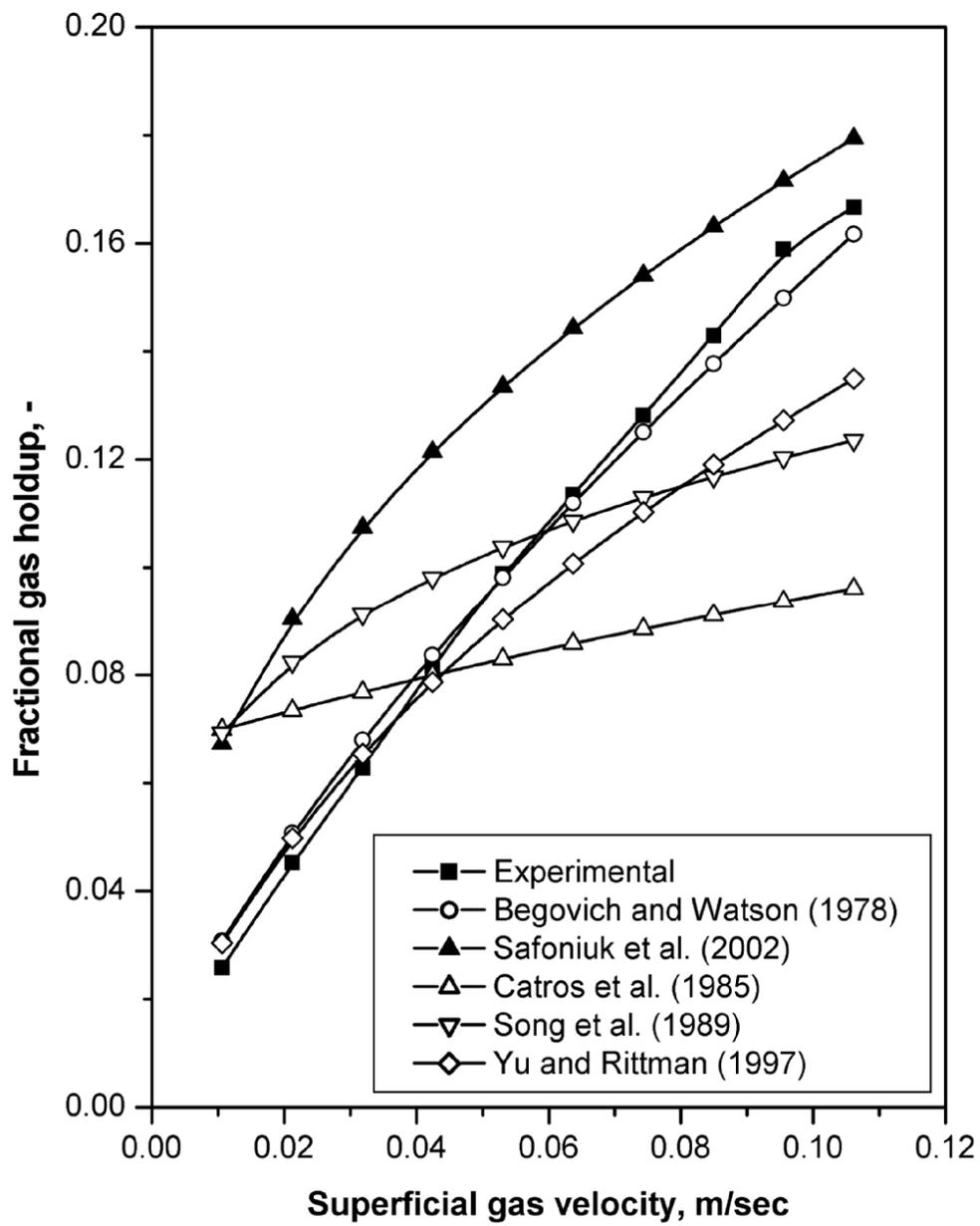
Figure 8. Comparison of different correlations with present investigation.
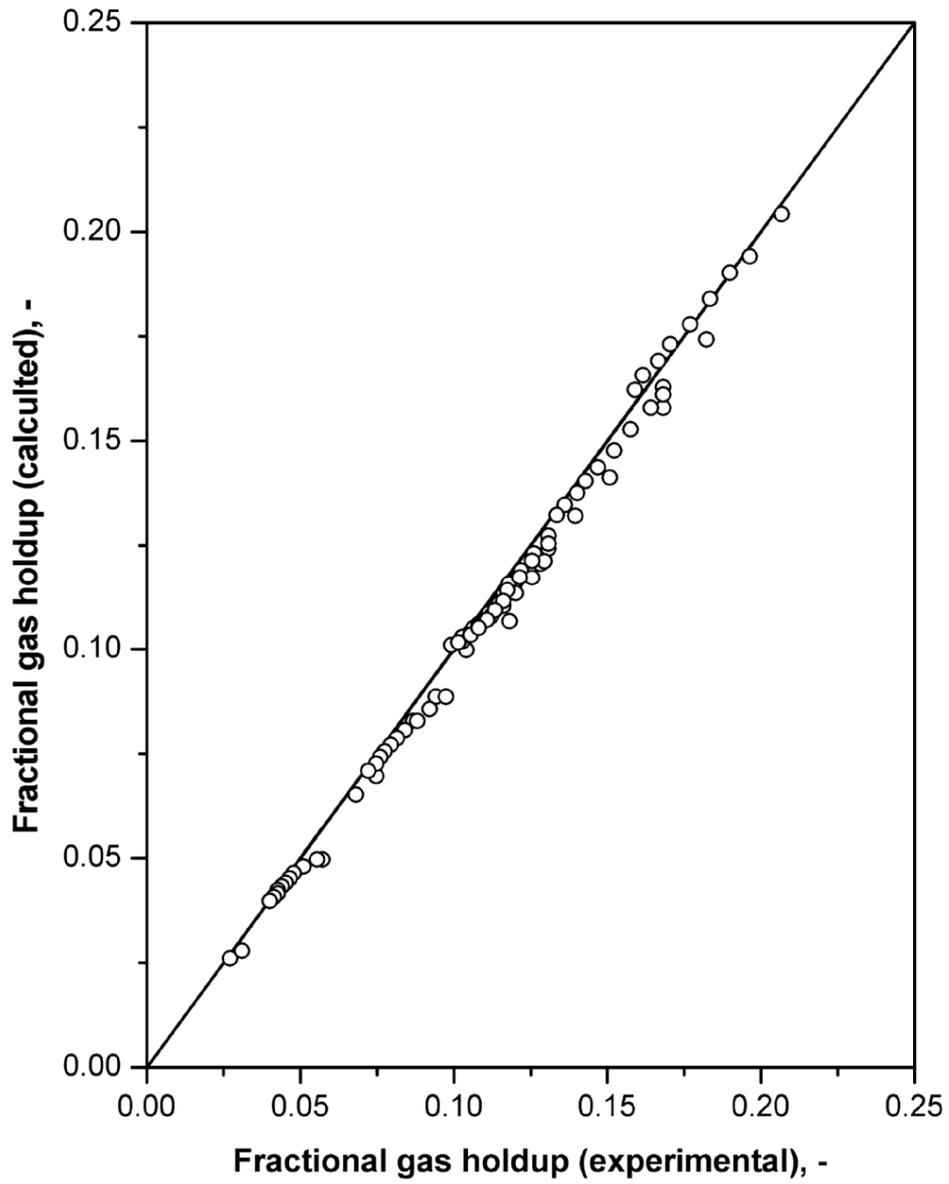
Figure 9. Comparison of experimental and calculated values of fractional gas holdup from Eq. (7).

Table 1
Properties and variables used in the experiment

A. Properties of bed materials

Materials	Mesh size (BSS)	d_p, m	ρ_p, kg m^{-3}	Initial static bed height, m
Glass beads	-7+8	0.00218	2,216	0.171
Glass beads	-6+7	0.00258	2,253	0.213
Glass beads	-5+6	0.00305	2,253	0.256
Glass beads	-4+5	0.00405	2,270	0.301

B. Properties of fluidizing medium

	ρ, kg m^{-3}	μ, Pa s
Air at 30^{0}C	1.168	0.0000186
Water at 30^{0}C	998.4	0.0008032

C. Properties of manometric fluid

	ρ, kg m^{-3}	μ, Pa s
Carbon tetra-chloride (CCl_4)	1,600	0.000942