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Some Hydrodynamic Aspects of Gas-Liquid Fluidized Beds

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ABSTRACT

Hydrodynamic characteristiccs of up-ward flow gas-liquid fluidized bed for air-water-glass bead system was investigated using a 19 cm dameter perspex column. The gas and liquid superficial velocities in the range of 0.0-1.1 cm/s and 0.0-3.64 cm/s respectively, and glass beads of average dimeters 2.50, 2.56 and 3.71 mm were studied. The effect of gas and liquid velocities and that of particle size on the quality of fluidization is related to various phase hold-ups in the bed. The gas holdup increased with increasing gas velocity and it decreased with the rise in liquid velocity whereas liquid holdup followed just the opposite trend. However, the bed porosity increased linearly with increase in gas and liquid velocities. These observations largely comformed to the general trend reported in the literature. The particle size distribution was found to have significant influence on hydrodynamic parameters. The empirical correlations (from literature) for predicting bed porosity and liquid holdup were found to lie within ±10 to ±20 percent deviations from our experimental results. One such correlation was modified to closely fit our experimental data. The investigation validates the existing data for scaleup to large scale, as the bed diameter studied was closer to pilot scale than most of the works reported in the literature.

INTRODUCTION

In the past two decades, there have been growing applications of three phase (gas-liquid) fluidized bed reactors in numerous industrial chemical, petrochemical and biochemical processess and particularly in the petroleum industry. Such reactors have several advantages over conventional fixed bed reactors such as intimate and efficient contact between the phases, conducive for efficient heat and mass transfer, reduced risk of blockage, small pressure drop and easy handling of solids. To name a few, three phase-fluidized bed reactors have been employed in processes like hydrotreatment of petroleum feed stock and flue gases, hydrometallurgy and the production of antibiotics. Three phase-fluidized beds are now being used for bio-oxidation/anaerobic treatment of wastewater. Here fluidized particles serve to support microorganism like those found in trickling bed filters, thus obviating the danger of blockage of the bed due to microbial flocs and also providing large gas-liquid interfacial area. Recently, fluidized beds have found increasing applications in biotechnology as bioreactors due to the advantage of higher biomass retention in such systems and thus a better efficiency (Buffiere et al. 1998; Buffiere & Moletta 1999). Especially, inverse three-phase fluidized bed contacting where gas and liquid flows counter-currently through a bed of solid lighter than the liquid has received growing interest over the last year as biological reactors (Buffiere & Moletta 1999). Muroyama and Fan (1985), Wild et al. (1984) and Godia and Sola (1995) have presented a comprehensive review on the application of three-phase fluidized beds.

In spite of wide applications and numerous interesting features of the gas-liquid fluidized bed reactors, especially with a continuous liquid phase in upward flow, literature dealing with this type of contacting particularly in large dimension columns is still inadequate and the behavior of such systems is still only poorly understood. Nevertheless, in the last two decades, considerable work has been done on the hydrodynamics of gas-liquid co-current fluidized bed. The phase holdup characteristics, the expansion/contraction behavior of the bed, bubble and wake characteristics, and distribution of flow regimes in the bed have especially attracted attention of investigators (Darton & Harisson 1975; Begovich & Watson 1978; Dhanuka and Stepanek (1978), Epstien (1981), Catros et al. (1985), Saberian-Broudjenni et al. (1987), Chen et al. 1995. Majority of reported studies were performed on small ($\leq$ 15 cm) diameter columns. However, hydrodynamics characteristics in large columns (of gas-liquid or gas-liquid- solid system) are quite different from those in small ones. Therefore, studies performed in large diameter columns are more reliable for scale up to industrial scale than those of small size columns. Catros et al. (1985) have used 17.15 cm internal diameter (ID) column and a nozzle perforated plate gas-liquid distributor in their experimental study of air, water and 3 mm glass bead system with liquid velocity in the range of 1.35-6.75 cm/s and gas velocity ranging from 0.0-3.0 cm/s. Saberian-Broudjenni et al. (1987) have studied the hydrodynamics of gas-liquid-solid fluidization at low liquid superficial velocities (0.0-3.0 cm/s) in 15.0 cm ID and 9.0 m high column using different liquids, gases, and solids. Based upon their experimental data, Saberian-Broudjenni et al. (1987) have proposed empirical correlations for calculating bed porosity and gas liquid holdups. Recently, Chen et al. (1995) have investigated the flow regime distribution as a function of axial position and gas velocity in a plexiglass column of 28.5 cm inside diameter and 4.1 m height operated with air, tap water and glass beads. They also studied gas and solid holdups in different flow regimes as function of operating parameters.

In the present work, some of the important hydrodynamic parameters like bed expansion and contraction, bed porosity and gas and liquid holdups have been investigated in a 19.0 cm ID column operated with cocurrent upward flow of air and water, with water as continuous phase through a bed of glass beads. Nearly spherical glass beads having narrow granulometric distribution with average diameters 2.50, 2.56 and 3.7 mm and density 2.41 gm/cm^3 were studied.

EXPERIMENTAL

EQUIPMENT

The experiments were carried out in a 19.0 cm diameter and 2.7 m high perspex column equipped with flow and pressure measuring devices (Figure

1). The test column was comprised of three main sections: the gas-liquid distribution section, the gas-liquid fluidized bed section and the gas-liquid disengagement section. The distribution section consisted of 20.0 cm high calming section followed by 25.0 cm high packed section filled with 10.0 mm rasching rings, sandwiched between two aluminium perforated plates having 220 holes oof 2.0 mm diameter on 8.0 mm squre pitch. A fine wire-mesh screen was placed over the upper grid to prevent fine particles from choking the perforations.Air was introduced in the calming section through six nozzles on 8.0 mm ID facing downward (Figure 2) in order to reduce the risk of channeling and bypassing by uniformly distributing air through the cross section. The 1.9 m high test section above the distribution section comprised of 40.0 cm of packing followed by two-phase gas-liquid section that served to settle down the particles carried into the dilute phase by bubble wakes.

Thirty three pressure taps were installed on the wall of the column along the axial distance above the upper grid. Due to anticipated large pressure gradient in the three-phase region, the pressure taps in this section of the bed were more closely spaced than in the two-phase section. The pressure tappings connected to open tube manometer gave static pressure in terms of the head of liquid (water) flowing in the test column.

Adjustable pinchcock was used to grip each flexible tubing connecting themanometers with the pressure taps (in order to narrowdown the cross sectional area), to dampen the fluctuations on the liquid level in mannometers caused by the pulsating bed. Just above the upper grid, a 38.0 mm opening in the column wall of three phase section (Figure 1) was provided for emptying the solids from the bed. Gas-liquid disengagement was achieved in a 40.0 cm high and 30.0 cm ID concentric cylinder installed at the top of the column. A wire-mesh screen restrainer was used to check the particles from entertainment.

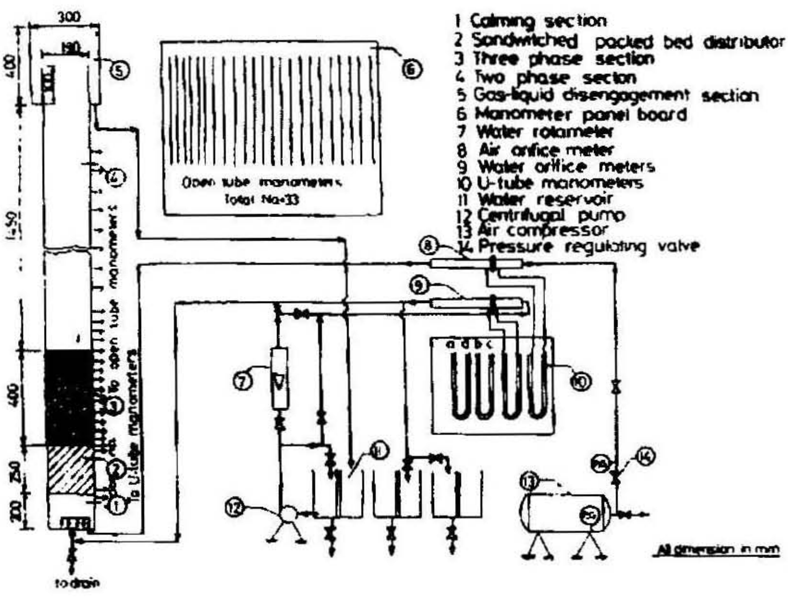

FIGURE 1. Schematic diagram of experimental setup

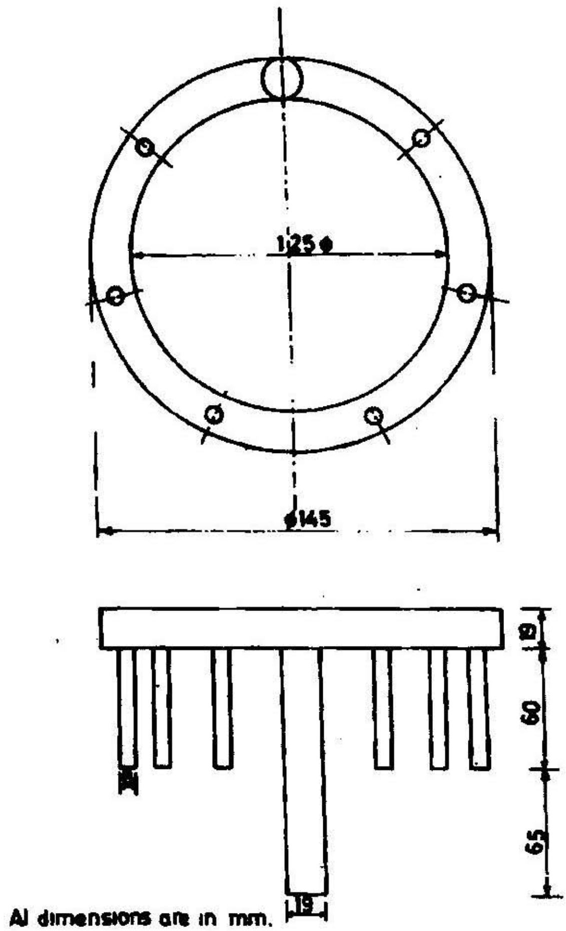

FIGURE 2. Details of gas spartger

PROCEDURE

The column was loaded with glass beads and fed with water at the desired flow rate. Air was then injected into the column and adjusted to desired flow rates. At steady state, i.e., when the liquid level in the open tube manometers became steady, the mean value of the manometer readings was recorded to alleviate the error due to small fluctuations in the manometer level. The experiments were carried out for the following range of parameters.

Gas velocity, cm/s	: 0.0-1.1
Liquid velocity, cm/s	: 0.0-3.64
Initial bed heigh, cm	: 32.5-40.0
Average particle diameter, cm	: 0.250, 0.256, and 0.371
Particle density, gm/cm^3	: 2.41
Pressure and temperature	: 15 psi, 28-33 °C

HOLDCUP CALCULATION

A Fortran program was employed to implement the following pressure profile technique (widely reported in literature) for the complete analysis of the experimental data. First the axial pressure profiles for the two-phase and three-phase regions were separately calculated using linear regression of the raw data (Jameel 1989). The intersection point of the two regression lines gave bed height, which was subsequently used along with the slope of the pressure profiles of the three-phase region, i.e., pressure gradient in the three phase region, to compute phase holdups in the bed using following equations (Ostergaard 1971).

$$1-\varepsilon=\frac{M_s}{\rho_s AH}$$

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

$$\varepsilon_l+\varepsilon_g+\varepsilon_s=1 \qquad (2)$$

Where e is the bed porosity, e_g, e_l, e_s are gas, liquid and solid holdups (3) respectively and ρ_g, ρ_l, ρ_s are densities of the gas, liquid and solid phases respectively. M_s is the mass of the solid bed, A is the cross sectional area of the fluidized bed column, and H is the height of the solid bed. $\Delta P/\Delta H$ is the pressure drop in the bed, which is balanced by the weight per unit volume of the bed, and g is the acceleration due to gravity.

RESULTS AND DISCUSSIONS

QUALITATIVE BEHAVIOR OF FLUID PARTICLE BED

Which fluid velocities approaching minimum fluidization velocity in the gas- liquid-solid bed, the particles at the top of the bed started to vibrate with subsequent interchange of particles throughout the bed ultimately leading to complete fluidization of the bed. This phenomenon could be attributed to the increase in the size of gas bubbles as they rise from the bottom to the top of the bed, coupled with migration of smaller size particles (for a size distribution of particles) towards the top of the bed. There was a reduction in the settled bed height after initial fluidization as also reported by Saberian-Broudjenni et al. (1987). A minimum of 1.3 % bed contraction was observed in the present study, whereas Saberian-Broudjenni et al. (1987) have reported bed contraction after initial fluidization as high as 30%. The discrepancy may be due to the different particle size distribution of the bed and owing to comparatively low gas and liquid velocities in the present study.

Due to undefined level of fluidized bed especially at high velocities, the error involved in the calculated and visually observed bed height was in the range of ±12%.

BED POROSITY AND BED CONTRACTION

Bed porosity of the expanded bed increased with the increasing liquid and gas velocities except during bed contraction period, which is discussed in the next paragraph. An almost linear variation of bed porosity with gas and liquid velocities has been consistently observed in the present investigation as shown in Figure 3. Earlier investigators (Dikshinamurthy et al. 1971); Dhanuka & Stepanek 1978; Saberian-Broudjenni et al. 1987 have also reported similar trend. However, some investigators have shown non-linear variation of bed porosity especially near incipient fluidization conditions (Lee & deLasa 1987). On the other hand, a fall in bed porosity with an increase in average particle size was observed which is evident from parity diagrams (Figure 6, 7 and 9).

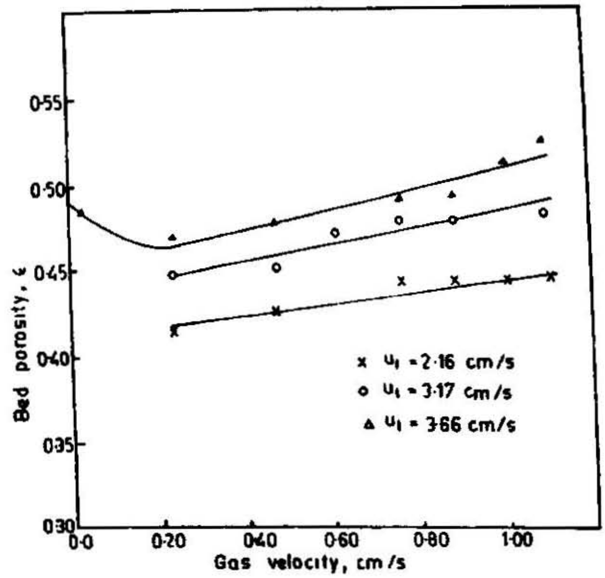

FIGURE 3. Dependence of mean bed porosity on superficial gas and liquid velocities for air-water-glass bead system and d_p = 2.56 mm

The phenomenon of initial bed contraction on introduction of gas into liquid-solid fluidized bed was observed in some cases as revealed by Figure 3 for a liquid velocity of 3.66 cm/s and for particle size of 2.56 mm. For this case, the bed porosity of liquid-solid fluidized bed (before the introduction of the gas) was 0.49. The porosity decreased to a minimum of approximately 0.46 at a gas velocity of 0.2 cm/s. For further increase in gas velocity, started to increase. The maximum bed contraction observed in the present study was 10.4%. Many earlier investigators have also reported the phenomenon. Steward and Davidson (1946) and later, Epstein (1981) have proposed qualitative elucidation of this phenomenon. Later, Nikov et al. (1990) have reported an initial contraction in a 2.0 mm particle bed an a throughout expansion in the bed of 6.0 mm particles. They also reported that for the 3.9 mm particle bed, contraction or expansion was observed depending on the superficial liquid velocity and on the liquid viscosity. They have attributed these contradicting behaviors to different flow regimes. Thus the initial contraction is a characteristic of the coalesced bubble regime and the expansion that of the bubble disintegration regime. The possibility of contraction or expansion depending upon liquid velocity is a characteristic of the transition region between the disintegrating and coalescing bubble regimes (Nikov et al. 1990).

In our case, the bed contraction was observed at a liquid velocity, u_l of 3.66 cm/s (maximum of u_l studied), which suggest that the transition between the disintegrating and coalescing bubble regime (according to Nikov et al. classificaton) occured somewhere around this velocity. For liquid velocities lower than 3.66 cm/s (i.e. at u_l = 2.16 and 3.17 cm/s), the flow still happened to be in disintegrating bubble regime and therefore, the bed contraction did not occur on the introduction of the gas and the bed kept on expanding with increasing gas velocities. Thus the results from the present study and those from Nikov et al. (1990) supplement each other.

GAS AND LIQUID HOLDUPS

Figure 4 and 5 show the dependence of gas and liquid holdups respectively, of the gas and liquid velocities. It is observed that:

1. The gas holdup increases with increasing gas velocity. With and increase of liquid velocity, gas hold up decreases. The experimental data show somewhat linear variation with gas velocity, however, deviation from this trend has been found significant in some cases. Chean et al. (1995) have reported a steep increase at first and then a gradual increase in gas holdup with increasing gas velocity. However, our experimental data do not show a definite trend of this type. Probably, more extensive experimentation is required to resolve this discrepancy.
2. Liquid holdup decreases with an increase in gas velocity and increases with increasing liquid velocity. Figure 5 shows almost linear variation of l(liquid holdup with gas and liquid velocities. Begovich and Watson (1978), Briens et al. (1997), Epstein (1981), Kim et al. (1972), Lee and deLasa

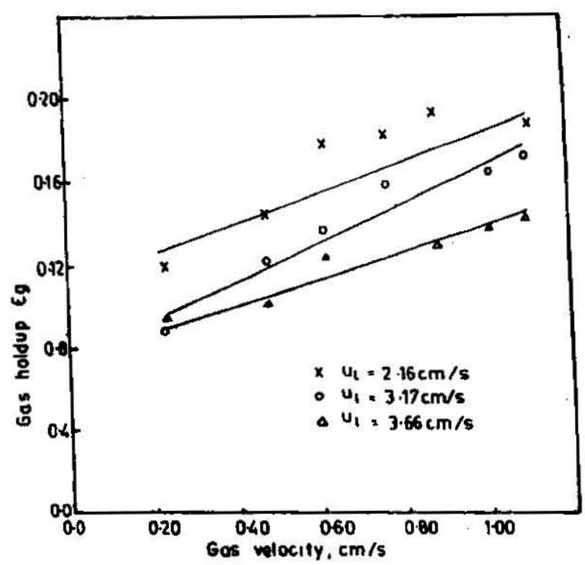

FIGURE 4. Dependence of gas holdup on superficial gas and liquid velocities for air-water-glass bead system and d_p = 2.56 mm

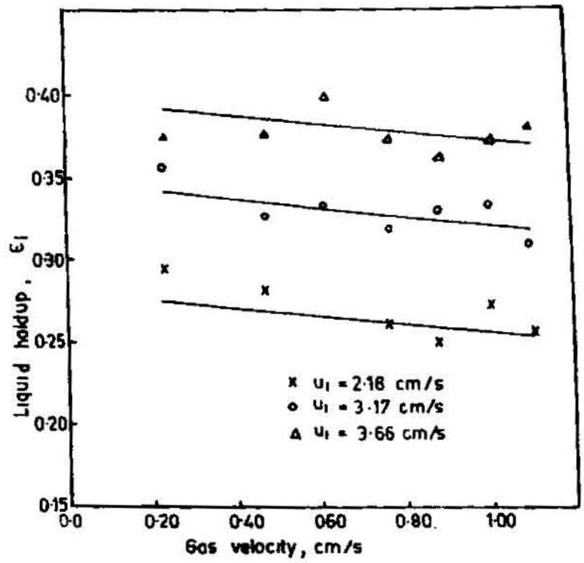

FIGURE 5. Dependence of liquid holdup on superficial gas and liquid velocities for air-water-glass bead system and dp = 2,56 mm

(1987), and Saberian-Broujenni et al. (1987) have also reported similar type of dependence.

TEST OF CORRELATIONS FROM THE LITERATURE

A number of correlations for calculating bed porosity and gas and liquid holdups have been proposed (Begovich and Watson 1978; Dakshinamuethy et al. 1971; Saberian-Broudjenni et al. 1987, however, their validity lies only within the range of experimental parameters investigated. So far, no correlation, valid in a wider range of parameter values could be found. Some of the correlations from literature, which were obtained under similar experimental conditions as those in the present study, are shown in the Table 1. The predictions from these correlations were compared with our experimental results (Figure 6-9) and the following conclusions were drawn.

The correlation of Saberian-Broudjenni et al. (1987) represented our experimental data on bed porosity almost within ±10 % error with some deviations at lower bed porosity (Figure 6). The constants in the equation were modified to completely fit our experimental data within ±10 % error (Figure 7). The modified equation of Saberian-Broudjenni was obtained as:

$$\varepsilon = \varepsilon_0 (u_l / u_{l_{mf}})^{0.23} (0.95 + 0.09\,\mathrm{Re}_{\mathrm{lg}}^{0.41}) \tag{4}$$

where ε_0 is the bed porosity of the static bed, u_l, the liquid velocity and $u_{l_{mf}}$ is the minimum fluidization velocity in the liquid-solid fluidized bed. $\mathrm{Re}_{\mathrm{lg}}$ is the Reynolds number defined as, $\rho_l u_g d_p / \mu_l$, where ρ_l, u_l, μ_g, and d_p, are respectively liquid density, liquid viscosity, gas velocity and particle diameter.

There was some discrepancy in the method of calculating ε_0 and $u_{l_{mf}}$ which can be seen elsewhere (Jameel 1989). Two equations from literature

TABLE 1. Correlations (from literture) tested with our experimental data

Author	Correlation Proposed (SI Units)	Experimental Basis
Begovich and Watson (1978)	$\varepsilon = a u_l^b u_g^c (\rho_s - p_l)^d d_p^e \mu_l^f D_c^h$ a = 3.474, b = 0.282, c = 0.046 d = –305, e = –208, f = 0.063 h = –0.020	Air-water-alumni beads $D_c \leq 15.2$ cm, $d_p = 0.62$ cm $u_g = 0 - 17.3$ cm/s $u_l = 0 - 12/0$ cm/s
Saberian-Broudjenni et. al. (1987)	$\varepsilon = \varepsilon_0 (u_l / u_{l_{mf}})^a (1 + b \mathrm{Re}_{ig}^c)$ a = 0.27, b = 0.07, c = 0.34	Solids: Alumina and glass beads Liquids: Aqueous and non-foaming organic liquids $D_c \leq 15$ cm, $u_g = 0 - 3$ cm/s $d_p = 0.14 - 0.26$ cm
Ostergaard and Michelsen (1968)	$\varepsilon_l = a u_l^b 10^{c^{u_g}}$ a = 1.414, b = 0.415, c = –1.1	Air-water-glass beads $D_c \leq 21.6$ cm, $d_p = 0.6$ cm $0 < u_g < 2.2$ cm/s

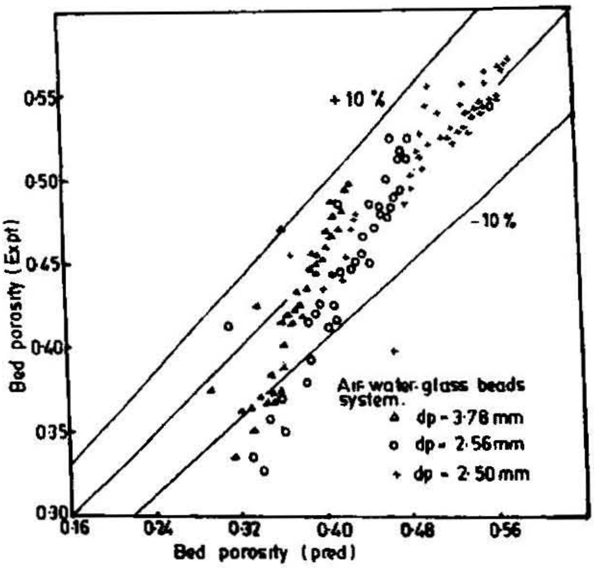

FIGURE 6. Parity diagram for the corelation of Saberian-Broudjenni et al. (1987). In all subsequent parity diagrams, experimental values of the parameter are plotted as a function of the corresponding value calculated from the correlation

(i.e., correlations of Begovich & Watson 1978 and Saberian-Broudjenni 1987, whose experimental basis resembles those in the present study), and the above equation (Equation 4) for the calculation of bed porosity are compared with our experimental data in Figure 8. Results presented in Figure 7 and 8, can be summarized as:

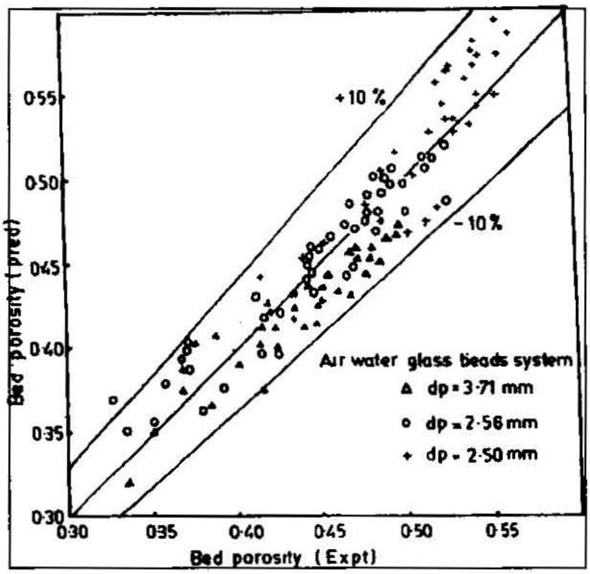

FIGURE 7. Parity diagram for the modified equation of Saberian Broudjenni et al. (1989)

1. Modified equation fits the experimental data with a tolerence of (10%, for the entire range of gas and liquid velocities investigated (Figure 7).
2. The bed porosity predicted by modified equation is somewhat intermediate between those predicted by correlations of Begovich and Watson (1978) and Saberian-Brudjenni et al. (1987) (Figure 8).
3. Correlation of Begovich and Watson (1978) shows some deviation from our experimental data at low gas velocities whereas the correlation of Saberian-Broudjenni et al. (1987) shows deviation at high gas velocities (Figure 8).

This leads to the conclusion that the modified equation of Saberian-Broudjenni (Equation 4) can be used to satisfactorily predict (within ±10% error) the bed porosity at low and high gas velocities in large diameter columns; hence more reliable for commercial scaleup (within the range of parameters investigated).

Correlation of Ostergaad and Michelsen (1968) has shown a good agreement with our experimental results as evident from Figure 9. The deviation being in the range of +30% to -15%.

Lastly, it is interesting to note that the two mixtures of particles with average diameter 2.50 mm and 2.56 mm respectively showed significant discrepancies in the hydrodynamic parameters studied as illustrated by Figure 6, 7 and 8. The anomaly, albeit the two diameter values close to each other, can be attributed to the difference in the granulometric distribution of the two mixtures of particles.

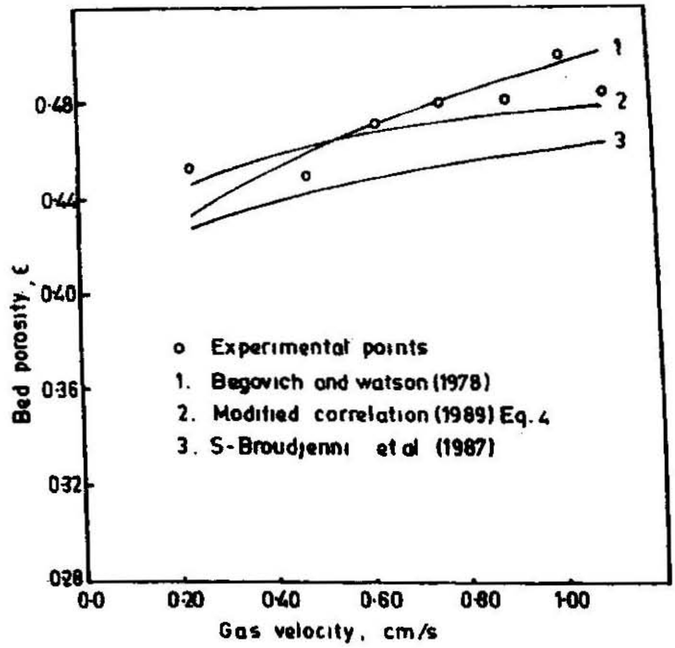

FIGURE 8. Comparison of different correlation for calcula-ting bed porosity (Table 1) for parameter values d_p = 2.56 mm, and u^l = 3.17 cm/s

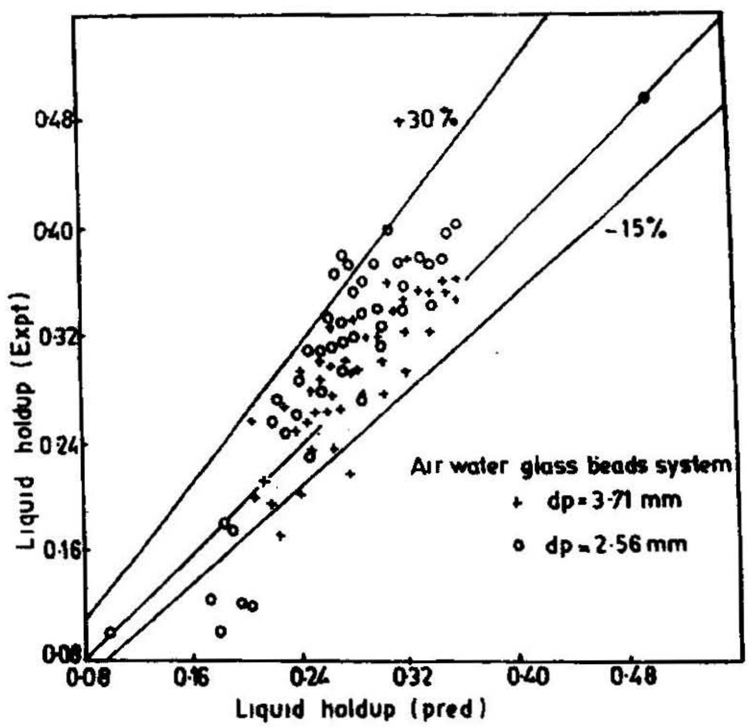

FIGURE 9. Parity diagram for the correlation of osterfaad and Michelsen (1968)

CONCLUSION

1. Over the range of gas and liquid velocities explored in the present study, pressure profile technique used for calculating hydrodynamic parameters proved sufficiently reliable and accurate yet simple compared to more expensive sophisticated techniques such as using pressure transducer employed by other investigator.
2. Results obtained on fluid-particle behavior of the system, and the influence of gas and liquid velocities on bed porosity, gas and liquid holdups are in close conformity with the trend reported in literature for similar experimental conditions.
3. The bed porosity for air-water-glass bead system within the range of parameters investigated can be satisfactorily represented by

   $\varepsilon = \varepsilon_0 (u_l / u_{l_{mf}})^{0.23} (0.95 + 0.09\, Re_{lg}^{0.41})$
4. The particle size distribution has significant influence on hydrodynamics parameters.
5. The present study shows that the results reported in literature on the hydrodynamic aspects of bed of diameter 15.0-22.0 cm are quite satisfactory as they show close conformity to each other and hence a high degree of reproducibility can be obtained.
6. However, future studies should be directed towards investigating hydrodynamics, heat and mass transfer charastirestic of a three phase fluidized bed using industrial gas-liquid system in large dimension column in order to ensure greater reliability for scaleup to commercial scale.

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Nomenclature

A	cross sectional area of the column, m^2
Dc	column diameter, m
dp	particle diameter, m
g	gravitational acceleration, m/s^2
H	height of the solid bed, m
Ms	mass of solid particles, kg
$\Delta P/\Delta H$	static pressure gradient in the bed, kPa/m
Re_{lg}	Reynolds number defined as $\rho_l u_g d_p/\mu_l$
u_l	liquid superficial velocity, m/s
u_g	gas superficial velocity, m/s
$u_{l_{mf}}$	minimum fluidization velocity of the two-phase liquid-solid fluidized bed, m/s
ε	bed porosity
ε_0	bed porosity of the static bed
ε_g	gas holdup
ε_l	solid holdup

ρ_l liquid density, kg/m^3
ρ_s solid density, kg/m^3
ρ_l liquid viscosity, kg/m.s

Subscripts
l liquid
g gas
s solid

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