Hydrodynamic Parameters of Gas-Solid-Liquid Three Phase Fluidized Bed with Low Density Solids

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This paper presents the experimental investigation on the hydrodynamic behaviour of a counter current gas-solid-liquid three phase fluidized bed with low density solid. Based on the experimental results, the effect of gas velocity, liquid spray density and solid particle density on the pressure drop, expanded bed height, liquid holdup was determined. It was observed that the pressure drop increases with the increase of gas velocity, spray liquid density and solid particle density. The expanded bed height increases with the increase of gas velocity and decreases with solid particles density. The liquid holdup increases with the increase of the solid particles density and gas velocity at constant liquid spray density. Liquid holdup increases more pronounced with increasing liquid spray density at constant density of the solid particles.

Keywords: Gas-solid-liquid three phase fluidization, fluidized bed with low density solid, hydrodynamic parameters, pressure drop, expanded bed height, liquid holdup.

Gas absorption is one of the most important processes in chemical industry and flue gas purification processes. In the last few years this process was adopted on wider scale in ecological engineering for flue gas depolluation. In order to obtain maximum absorption efficiency it is necessary to use proper equipments to maximize gas-liquid contact, such as new columns with three phase fluidized beds [1, 2]. Three phase fluidized bed absorber is mass transfer equipment in which the bed of low density packing is fluidized by the counter current flow of gas and liquid. The gas flows as a continuous phase and the liquid flows as a dispersed phase in absorber. Low density inert solid of hallow spherical balls are used in column and fluidized by gas.

Packing solid, having no chemical effect on process, maximizes gas-liquid contact. The intensive mixing of the solid packing in the column determines high turbulence and therefore enhances the mass transfer comparative to conventional fixed packed beds. This contact mode between phases has many advantages for mass transfer systems: low pressure drop in the column, very high interfacial contact area per unit volume of the column, capability to process large volume of gases. In the same time, the solid packing is easily handled and can be removed from the column with pneumatic transport, the solid packing does not need special expertise and it can easily be made from chemical resistant plastics, depending on the chemicals involved in process, low capital cost of equipment.

These attributes make three phase fluidized bed absorbers technology favourable for pollution control. There are two operating modes for three phase fluidized bed [3]:

I -fluidization without flooding, when fluidization starts before flooding point in the column;

II -fluidization at incipient flooding, when fluidization starts after flooding point.

Important parameters to be considered in the design of three phase fluidized bed are minimum fluidization velocity, pressure drop, expanded bed height, liquid holdup and mass transfer coefficients.

Density of solid packing has a major contribution in deciding the mode of operation. For beds with relatively low-density particles (ρ_s <300 kg/m^3), fluidization occurs at a gas velocity lower than the flooding point of the equivalent counter current fixed bed. For beds with relatively high-density particles (ρ_s >300 kg/m^3), fluidization always occurs after the flooding point of the packed bed [3].

The hydrodynamic study plays an important role in the economical design and operation of a three phase fluidized bed. Based on the experimental work, the effect of fluid rates on the various parameters was studied and the observed data were reported. The objectives of the present study were: to determinate the pressure drop and fluidized bed expansion, and to estimate liquid holdup in gas-solid-liquid three phase fluidized bed column with low density solids.

Experimental part

The present work is an experimental investigation on the hydrodynamic behaviour of a counter current three phase fluidized bed with gas (air) as a continuous phase, low density inert packing, with size of 10 mm spherical balls and the liquid (water) flow as a dispersed phase.

The schematic diagram of the experimental setup is shown in figure 1. The column was made of glass, with 0.14 m diameter and 1.10 m height. The packing are hollow spheres of 0.01 m diameter and with the density of 170; 210 and 337 kg/m^3. The air as the continuous phase was fed at the bottom of the column and exit at the top. Water as dispersed phase was sprayed from the liquid distributor over the column cross section. For measuring the fluidized bed height, a scale arrangement was made. The height of the expanded bed was noted, when the steady state conditions were attained.

The pressure drop across the bed was measured using manometer method and the height of the expanded fluidized bed was read on the scale and noted. Using these values, the minimum fluidization velocity, expanded bed height and liquid holdup can be determined. The experimental conditions are presented in table 1.

In literature it has been reported that the column diameter to packing diameter ratio should be higher than 10, in order to eliminate wall effects. The static bed height to column diameter ration should be lower than one to

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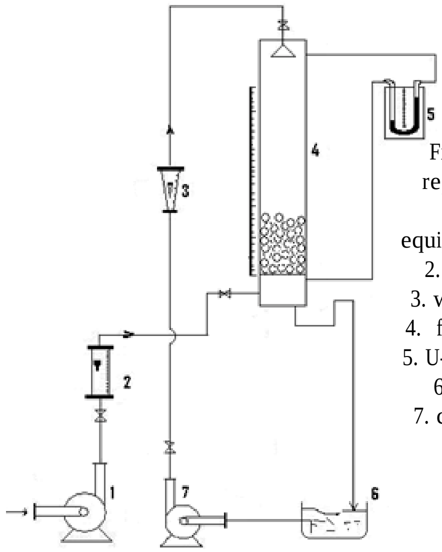

Fig.1. Schematic representation of experimental equipment: 1. blower, 2. air flow meter, 3. water flow meter, 4. fluidizing column, 5. U-tube manometer, 6. storage tank, 7. centrifugal pump

Table 1
VALUES OF EXPERIMENTAL CONDITIONS

Variable	Values
Column diameter D_c [m]	0.14
Diameter of solid particle d_p [m]	0.01
Solid particle density [Kg/m^3]	170; 210; 337.
Static solid bed height H_0 [m]	0.12
Grid free area [%]	78%
Liquid spray density q_l [m^3/m^2 h]	6.49; 9.74; 13; 16.24
Gas velocity v_g [m/s]	0-2.2

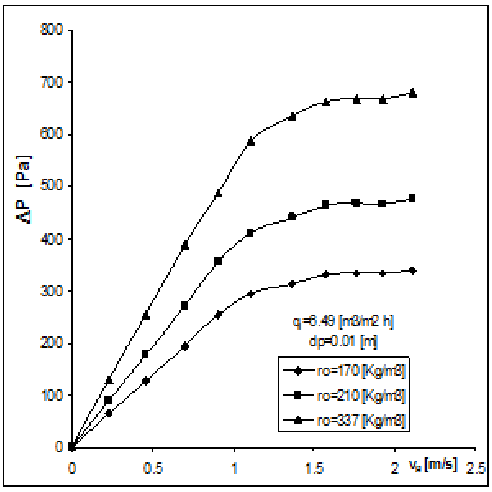

Fig.2. Influence of gas velocity and solid particles density on pressure drop at q_l of 6.49 [m^3/m^2h]

eliminate the pulsation and non-uniform fluidization. It has been reported in literature that the hydrodynamic and mass transfer effects of grid becomes negligible for grid free areas greater than 70% [4]. All these conditions were fulfilled in the present study.

Results and discussions

The hydrodynamic parameters important for mass transfer in fluidized equipments include: minimum fluidization velocity, pressure drop, expanded bed height and liquid holdup.

Pressure drop and expanded bed height

The pressure drop in any mass transfer device determines liquid holdup, interfacial contact area and its operating cost. The measurement results of the pressure drop in gas-solid-liquid three phase fluidized bed are presented in figures 2 - 8.

Figures 2-8 show the increase of the pressure drop with the increase of gas velocity, spray liquid density and solid particle density. For a constant spray liquid density (q_l=ct.) and increasing gas velocity, the pressure drop increases until minimum fluidization velocity is attained. This zone of operation column is the static bed region. With further increase in gas velocity, the entire solid bed is in fully fluidized state, thereafter there is a small change in pressure drop. This is the fluidized bed region. The column is preferably operated in the fully fluidized bed, the region in which the turbulence of the bed increases with the increase of gas velocity and spray liquid density, until the true flooding point is reached.

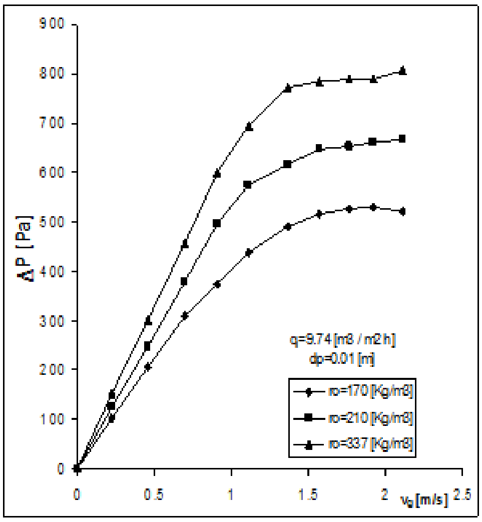

Fig.3. Influence of gas velocity and solid particles density on pressure drop at q_l of 9.74 [m^3/m^2h]

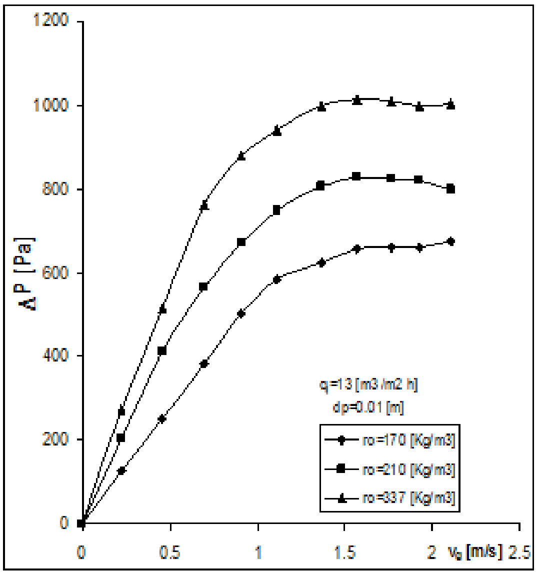

Fig.4. Influence of gas velocity and solid particles density on pressure drop at q_l of 13 [m^3/m^2h]

The static bed begins to expand once the superficial gas velocity reaches the minimum fluidization velocity. The expansion of the bed increases with the increase of gas velocity, due to the upward force of the gas which tends to lift the packing. During the experiments, it was observed that the bed expansion begins at lower gas velocities when the spray liquid density is increasing, at the same density of the solid particles. Variables which affect bed expansion are: gas flow rate, liquid flow rate, free area of the grid, density and diameter of the packing and static bed height.

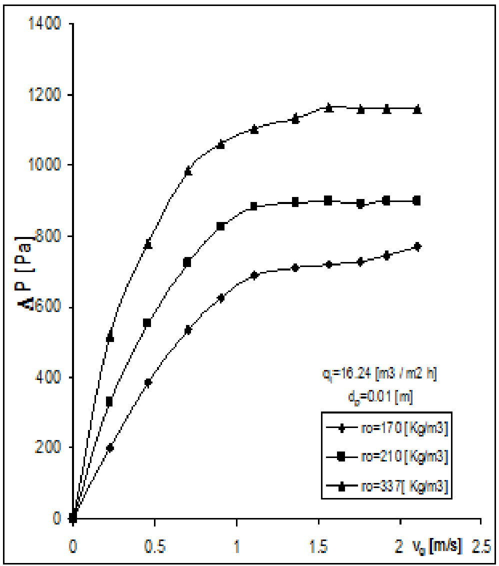

Fig.5. Influence of gas velocity and solid particles density on pressure drop at q_l of 16.24 [m^3/m^2h]

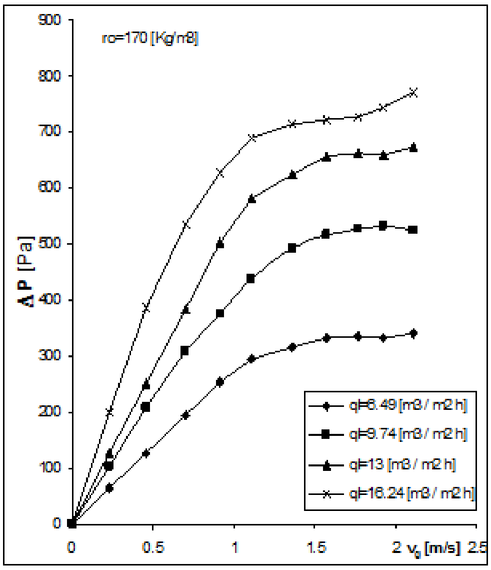

Fig.6. Influence of liquid spray density and gas velocity on pressure drop for packing with density of 170 [kg/m^3]

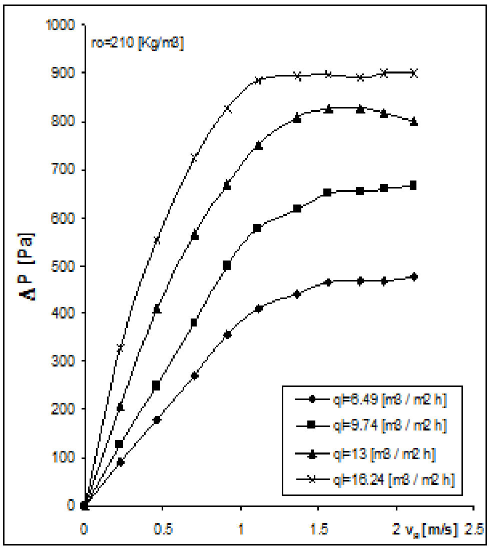

Fig.7. Influence of liquid spray density and gas velocity on pressure drop for packing with density of 210 [kg/m^3]

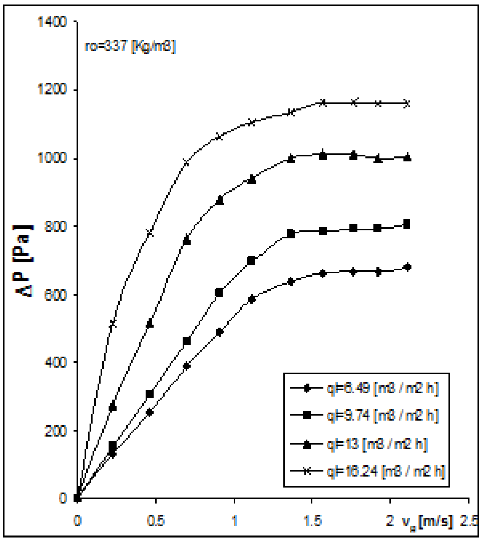

Fig.8. Influence of liquid spray density and gas velocity on pressure drop for packing with density of 337 [kg/m^3]

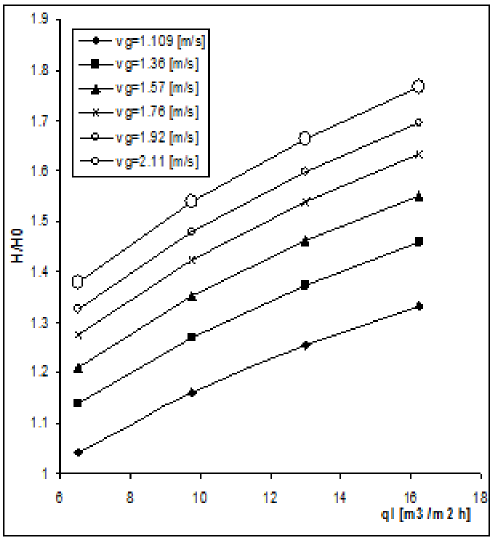

Fig.9. Influence of liquid spray density and gas velocity on expanded bed height for packing with the density of 170 [kg/m^3]

Expanded bed height was determined by visual observations. In case of bed pulsation, maximum and minimum heights of the bed were noted and average height was selected.

Bed expansion data obtained in present study is represented in dimensionless form as ratio of the expanded bed height to the static bed height (H/H_0). The effect of liquid spray density and gas velocity on expanded bed height (H/H_0) is shown in figure 9.

The static bed begins to expand once the superficial gas velocity reaches the minimum fluidization velocity. The expansion of the bed with the increase of gas velocity is due to the upward force of the gas which tends to lift the solid packing. During the experiments, it was observed that the bed expansion begins at lower gas velocities when the liquid spray density is increased. This means that the bed expands with the increase of gas velocity.

The effect of solid packing density on the reduced bed height (H/H_0) is shown in figure 10.

As figure 10 shows the low density packing is easier to fluidize. The solid packing with same diameter but with higher density fluidizes to a smaller height, at the constant liquid spray density and constant gas velocity.

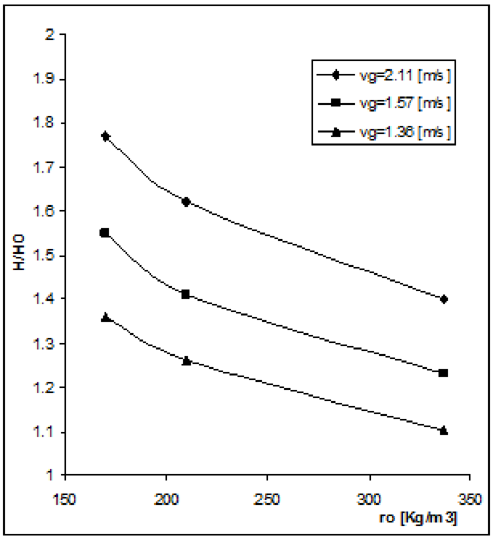

Fig.10. Influence of density of solid packing and gas velocity on bed expansion at q_l of 16.24 [m^3/m^2h]

Liquid holdup

Three methods have been reported in the literature for the determination of liquid holdup: shut off valve, tracer technique and pressure drop measurements. In this study liquid holdup was determined from hydrodynamic model of pressured drop [5] and experimental measurements.

The geometrical model representation of the hydrodynamic behaviour in gas-solid-liquid three phase fluidized bed is shown in figure 11.

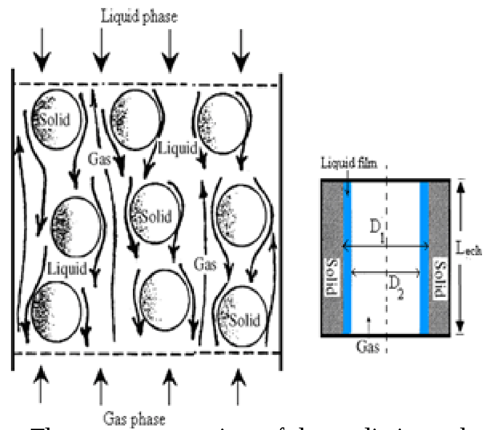

Fig.11. Geometrical model of gas-solid-liquid three phase fluidized bed

The system consists of three distinct phases: solid, gas and liquid and it assumes that the liquid phase wets completely the solid particles. There is not direct contact between the gas phase and the solid particles. The boundary between the liquid and the solid phases is defined in the model by diameter D_1, while between the gas and liquid phases is defined by diameter D_2.

Volumes of liquid (e_l) eand gas (e_g) in the system, and the liquid-solid contact area are given by equation (1) – (3):

$$\varepsilon_l=\frac{\pi}{4}\left(D_1^2-D_2^2\right)L_{ech} \text{ and } \varepsilon_g=\frac{\pi}{4}D_2^2L_{ech} \tag{1}$$

$$\frac{\varepsilon_l}{\varepsilon_g}=\left(\frac{D_1}{D_2}\right)^2-1 \text{ or } \frac{D_1}{D_2}=\sqrt{1+\frac{\varepsilon_l}{\varepsilon_g}} \tag{2}$$

$$A_c=\pi D_1L_{ech}=\frac{V_s}{d_{ech}}=\frac{\frac{\pi}{4}D_c^2H\varepsilon_s}{d_{ech}}=\frac{\pi}{4}D_c^2H\varepsilon_s\cdot\frac{6}{d_p}$$

$$d_{ech}=\frac{V_p}{A_p}=\frac{d_p}{6} \tag{3}$$

For liquid phase the equivalent diameter is given by equation (4):

$$D_{ech}=D_1-D_2=\frac{2\varepsilon_g d_p}{3\varepsilon_s}\left[\left(1+\frac{\varepsilon_l}{\varepsilon_g}\right)-\sqrt{1+\frac{\varepsilon_l}{\varepsilon_g}}\right] \tag{4}$$

Assuming momentum transfer due to acceleration of each phase as negligible, the momentum balances for gas and liquid phases in the bed under steady state condition can be expressed by [5]:

$$\begin{aligned}\left(-\frac{dp}{dz}\right)^{bed}&=\rho_g g+\left(-\frac{dp}{dz}\right)^{gas}_{friction}\\ \left(-\frac{dp}{dz}\right)^{bed}&=\rho_l g+\left(-\frac{dp}{dz}\right)^{liquid}_{friction}\end{aligned} \tag{5}$$

Pressure drop due to friction in the gas phase is gas-liquid friction:

$$\left(-\frac{dp}{dz}\right)^{gas}_{friction}=\left(-\frac{dp}{dz}\right)^{g-l}_{f} \tag{6}$$

and pressure drop due to friction in the liquid phase consists of two component, one is contributed by the solid phase and the other is contributed by the gas phase:

$$\left(-\frac{dp}{dz}\right)^{liquid}_{friction}=\left(-\frac{dp}{dz}\right)^{l-s}_{f}+\left(-\frac{dp}{dz}\right)^{l-g}_{f} \tag{7}$$

Since the mutual forces at the interface are cancelled by each other, we have [5]:

$$\begin{aligned}&\frac{\varepsilon_l}{1-\varepsilon_s}\left(-\frac{dp}{dz}\right)^{l-g}_{f}+\frac{\varepsilon_g}{1-\varepsilon_s}\left(-\frac{dp}{dz}\right)^{g-l}_{f}=0 \text{ or}\\ &\left(-\frac{dp}{dz}\right)^{l-g}_{f}=-\frac{\varepsilon_g}{\varepsilon_l}\left(-\frac{dp}{dz}\right)^{g}_{f}\end{aligned} \tag{8}$$

From equations (7) and (8) we obtain:

$$\left(-\frac{dp}{dz}\right)^{bed}=\rho_l g+\left(-\frac{dp}{dz}\right)^{l-s}_{f}-\frac{\varepsilon_g}{\varepsilon_l}\left(-\frac{dp}{dz}\right)^{g}_{f} \tag{9}$$

Expressing liquid holdup ε_l from equations (5) and (9) we obtain:

$$\varepsilon_l=(1-\varepsilon_s)\frac{\left(-\frac{dp}{dz}\right)^{g}_{f}}{(\rho_l-\rho g)+\left(-\frac{dp}{dz}\right)^{l-s}_{f}} \tag{10}$$

where $[-dP/dZ]^g_f$ can be written as [6]:

$$\left(-\frac{dp}{dz}\right)^{g}_{f}=\frac{4\tau_i}{D_2} \tag{11}$$

τ_i, the interfacial shear stress at the gas-liquid interface, is given by equation (12):

$$\tau_i=\frac{0.5f_g\rho_g\left[\frac{v_g^2}{(1-\varepsilon_s)^2}\right]}{(D_2/D_1)^4} \tag{12}$$

For the friction factor in counter current gas-liquid flow is proposed the following equation [6]:

$$f_g=\left[0.005+1.5(D_2/D_1)\right] \tag{13}$$

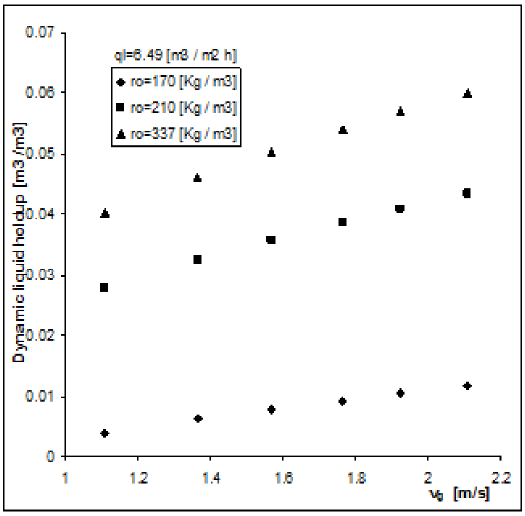

Fig.12. Influence of gas velocity and solid particles density on liquid holdup at q_l of 6.24 [m^3/m^2h]

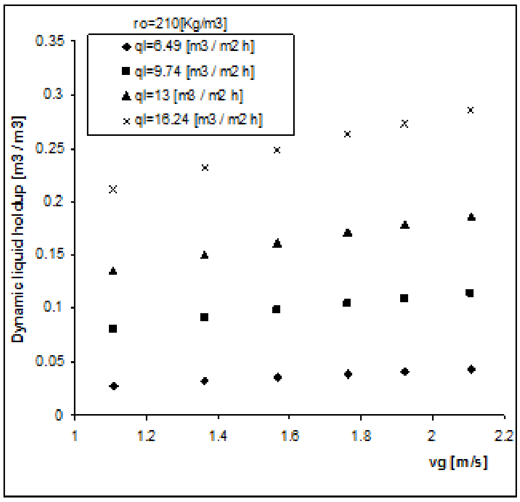

Fig.13. Influence of gas velocity and liquid spray density on liquid holdup at constant density of the solid

With these specifications we can written:

$$\left(-\frac{dp}{dz}\right)_f^g = 2\left[0.005+1.5(D_2/D_1)\right]\rho_g \frac{\dfrac{v_g^2}{(1-\varepsilon_s)^2}}{D_2(D_2/D_1)^4} \quad (14)$$

For $\left(-\frac{dp}{dz}\right)_f^{l-s}$ it can be written in terms of friction factor [5]: (15)

$$\left(-\frac{dp}{dz}\right)_f^{l-s} = 4 f_l (1/D_{ech})\left[\frac{\rho_l (v_l/\varepsilon_l)^2}{2}\right] \quad (16)$$

$$f_l = \frac{16}{\mathrm{Re}_l} \quad \text{and} \quad \mathrm{Re} = \frac{\rho_l v_l D_{ech}}{\eta_l \varepsilon_l} \quad (17)$$

$$\left(-\frac{dp}{dz}\right)^{bed} = (\varepsilon_g\rho_g + \varepsilon_s\rho_s + \varepsilon_l\rho_l)g \text{ and } \varepsilon_g + \varepsilon_l + \varepsilon_s = 1$$

Considering this hydrodynamic model and the experimental measurements for pressure drop, the liquid holdup was determined from equation (18):

$$\Delta p^b = \left[(1-\varepsilon_0)\rho_s + \frac{H}{H_0}\varepsilon_l\rho_l\right] g H_0 \quad (18)$$

Figure 12 shows the increase of the liquid holdup with increasing of the solid particles density and gas velocity at constant liquid spray density. Figure 13 shows the increase of liquid holdup with increasing liquid spray density at constant density of the solid particles. So, at low liquid spray density, the increase in liquid holdup is rather smaller. At high liquid spray density the increase of liquid holdup is quite substantial.

There is a contradiction in literature as far as effect of gas velocity on liquid holdup. Thus, some bibliographic references [1, 7, 8] reported no significant effect of gas velocity on liquid holdup, while others [9-11] reported the increase in liquid holdup with gas velocity. Very small changes in liquid holdup were observed with an increase in gas velocity, once the solid bed is completely fluidized.

Conclusions

The influence of solid packing density, gas velocity and liquid spray density on pressure drop, on expanded bed height and on liquid holdup, for gas-solid-liquid three phase fluidized bed with low density solid was determined.

The determined pressure drop increases with the increase of gas velocity, spray liquid density and solid particles density. The expanded bed height increases with the increase of gas velocity and decreases with the increase of solid particles density.

The liquid holdup values were correlated with density of the solid particles, gas velocity and liquid spray density. The liquid holdup increases with increasing of the solid particles density, gas velocity and spray liquid density.

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