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Dedicated to prof. dr. I. C. Popescu
on the occasion of his 70th anniversary

HYDRODYNAMIC CHARACTERIZATION OF THREE PHASE FLUIDIZED BED

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ABSTRACT. Hydrodynamics study plays an important role in the economical design and operation of three phase fluidized bed contactor. This paper presents the experimental investigation on the hydrodynamic behavior of a counter current gas-solid-liquid three phase fluidized bed. The results have shown 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.

Keywords: *three phase fluidized bed contactor, pressure drop, expanded bed height*

INTRODUCTION

Intensification of the mass transfer between a liquid and a gaseous phase needs as high contact area as possible between the two phases. So, is very important to construct such equipments which could generate as high interfacial area as possible. Fluidized beds offer very high interfacial area per unit volume of the active column. A specific equipment which uses relatively large size solid inert packing as the fluidizing media is the turbulent contact absorber [1].

Three phase fluidized bed column 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. The intensive mixing of the solid packing in the column determines high turbulence and therefore enhances the mass

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transfer comparative to conventional fixed packed beds. This contact mode between phases determine low pressure drop in the column, very high interfacial contact area per unit volume of the column, and capability to process large volume of gases. These advantages make three phase fluidized bed absorbers technology favourable for pollution control.

There are two operating types for three phase fluidized bed [2]. In type I, the solid fluidization starts before flooding in the column, whereas in type II, fluidization starts after flooding in the column. 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 [2, 3].

Vunjak-Novakovic et al. [4] developed a chart for the demarcation of type I and type II modes of three phase fluidized bed operation, as shown in Figure 1.

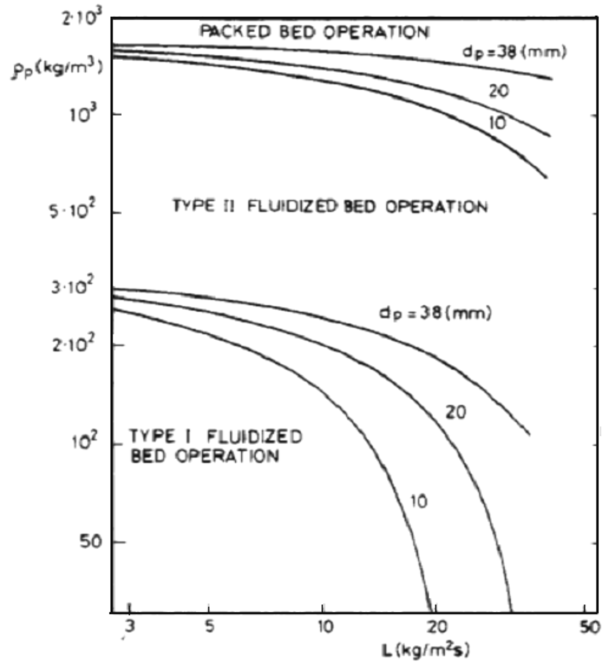

Figure. 1. Operating flooding zones in turbulent contact absorber [4].

Above classification of type I and type II regimes of turbulent contact absorber operation is very broad. The increase in liquid flow rate and decrease in solid packing diameter also shifts the regime from type I into type II.

Pressure drop, expanded bed height and liquid holdup are regarded as key hydrodynamic parameters of any mass transfer equipment The objectives of the present study were to determinate the pressure drop and fluidized bed expansion in gas-solid-liquid three phase fluidized bed column with low density solids.

RESULTS AND DISCUSSION

The hydrodynamic parameters important for mass transfer in fluidized equipments include: minimum fluidization velocity, pressure drop, expanded bed height and liquid holdup. For determination of these hydrodynamic characteristics were used experimental methods.

Pressure drop in any mass transfer device determines liquid holdup, interfacial contact area and its operating cost. In literature, the dependence of pressure drop on liquid and gas velocities, diameter and density of packing, static bed height and support grid free area have been studied. It is consensus in literature that pressure drop in fluidized bed contactor increases with increasing liquid flow rate at a fixed gas flow rate.

The momentum balances for gas and liquid phases in the bed under steady state condition can be expressed by next equations [5]:

$$\left(-\frac{dp}{dz}\right)^{bed} = \rho_g g + \left(-\frac{dp}{dz}\right)^{gas}_{friction}$$
$$\left(-\frac{dp}{dz}\right)^{bed} = \rho_l g + \left(-\frac{dp}{dz}\right)^{liquid}_{friction} \quad (1)$$

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

$$\left(-\frac{dp}{dz}\right)^{gas}_{friction} = \left(-\frac{dp}{dz}\right)^{g-l}_{f} \quad (2)$$

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:

$$\left(-\frac{dp}{dz}\right)^{liquid}_{friction} = \left(-\frac{dp}{dz}\right)^{l-s}_{f} + \left(-\frac{dp}{dz}\right)^{l-g}_{f} \quad (3)$$

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

$$\frac{\varepsilon_l}{1-\varepsilon_s}\left(-\frac{dp}{dz}\right)_f^{l-g}+\frac{\varepsilon_g}{1-\varepsilon_s}\left(-\frac{dp}{dz}\right)_f^{g-l}=0 \text{ or}$$
$$\left(-\frac{dp}{dz}\right)_f^{l-g}=-\frac{\varepsilon_g}{\varepsilon_l}\left(-\frac{dp}{dz}\right)_f^{g} \tag{4}$$

From equations (3) and (4) results:

$$\left(-\frac{dp}{dz}\right)^{bed}=\rho_l g+\left(-\frac{dp}{dz}\right)_f^{l-s}-\frac{\varepsilon_g}{\varepsilon_l}\left(-\frac{dp}{dz}\right)_f^{g} \tag{5}$$

where $\varepsilon_g+\varepsilon_l+\varepsilon_s=1$, are gas, liquid and solid holdup.

The measurements of the bed pressure drop in gas-solid-liquid three phase fluidized bed are presented in Figure 2 to Figure 4.

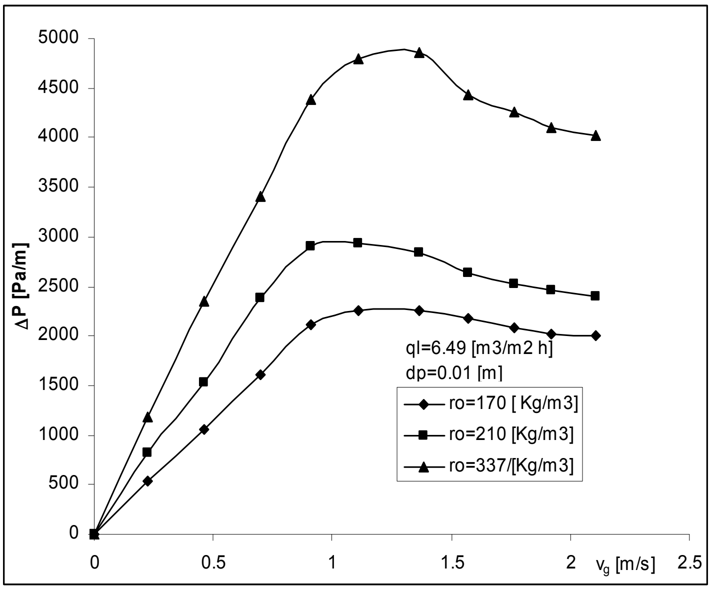

Figure 2. Influence of gas velocities and particle density on pressure drop at q_l = 6.49 m^3/m^2h.

Figures 2 and 3 show the increase of the pressure drop with the increase of gas velocities, solid particle density, spray liquid density and height solid packing.

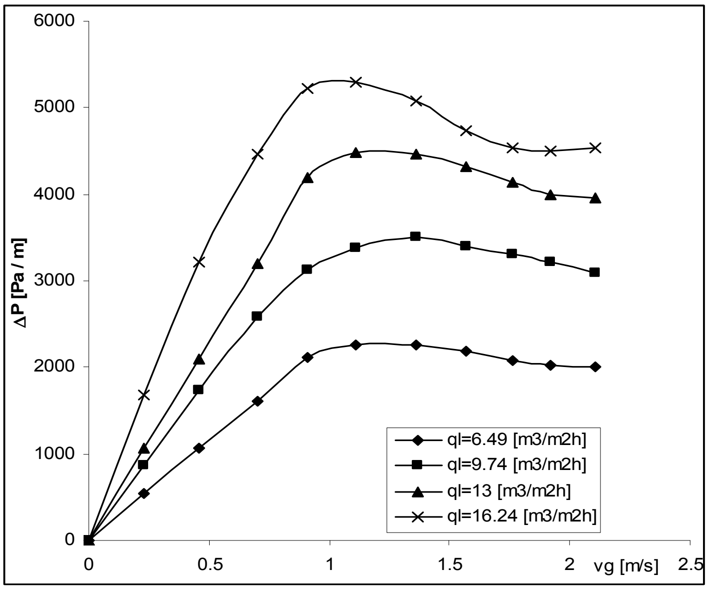

Figure 3. Influence of liquid spray density and gas velocities on pressure drop for packing with ρ = 170 kg/m^3.

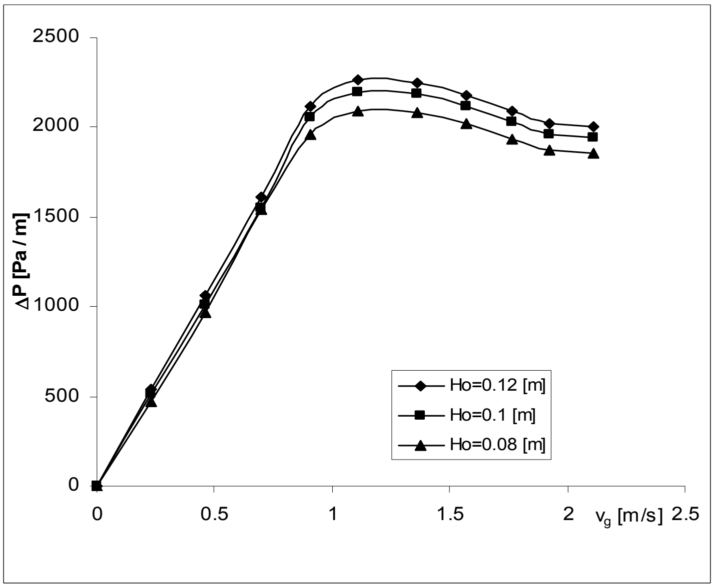

Figure 4. Effect of height of solid packing on pressure drop at q_l = 6.49 m^3/m^{2}h, for packing with ρ = 170 kg/m^3.

For a constant spray liquid density (q_l = ct.) and the increase 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 a fully fluidized state, thereafter it is a small change in pressure drop. This is the fluidized bed region. The column is preferably operated in the fully fluidized bed 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.

The effect of increasing static bed height on pressure drop is given in Figure 4. There is a slight decreasing effect of static bed height on the reduced bed height for type I fluidization. This initial reduced pressure drop in bed height can be explained by considering that as the static bed height increases with the increase in weight of the bed, which is more than the force generated by the upward moving gas.

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. Expanded bed height was determined by visual observations. In the 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 Figures 5 and 6.

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 solid packing with the same diameter but with higher density fluidizes to a smaller height, at the constant liquid spray density and constant gas velocity.

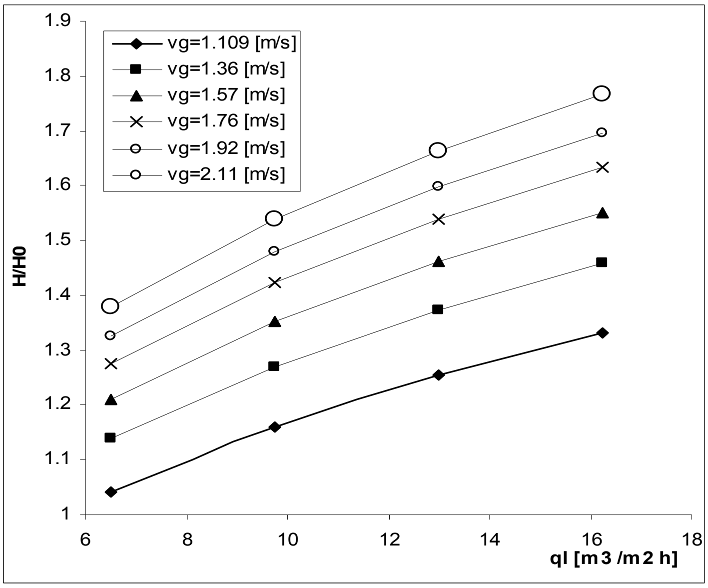

Figure 5. Influence of liquid spray density and gas velocities on expanded bed height, for packing with ρ = 170 kg/m^3.

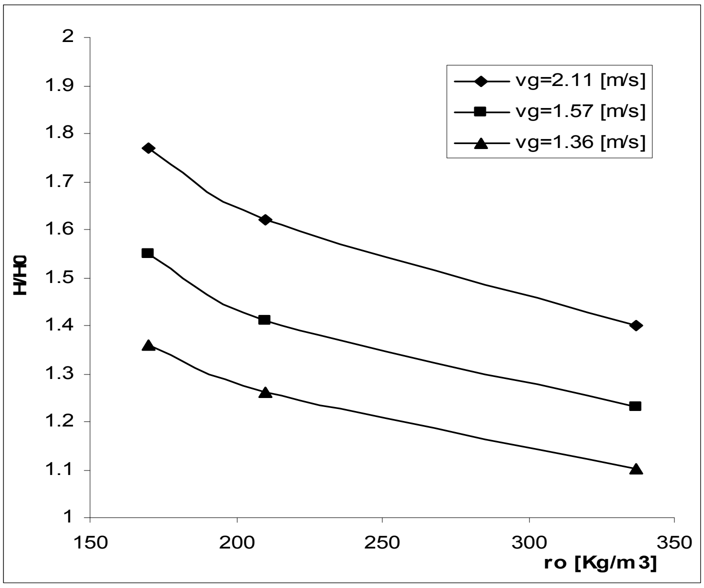

Figure 6. Influence of density of packing on bed expansion at q_l = 16.24 m^3/m^{2}h.

CONCLUSIONS

The pressure drop, expanded bed height and liquid holdup are regarded as the key hydrodynamic parameters of this mass transfer equipment.

Influence of gas velocities, solid density, liquid spray density and solid static bed height on pressure drop, expanded bed height for gas-solid-liquid three phase fluidized bed with low density solid was determined.

The pressure drop increases with the increase of gas velocities, solid particle density, spray liquid density and height solid packing. Pressure drop per unit bed height decreases lower when gas velocity increases over 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.

EXPERIMENTAL

The schematic diagram of the experimental setup is shown in Figure 7.

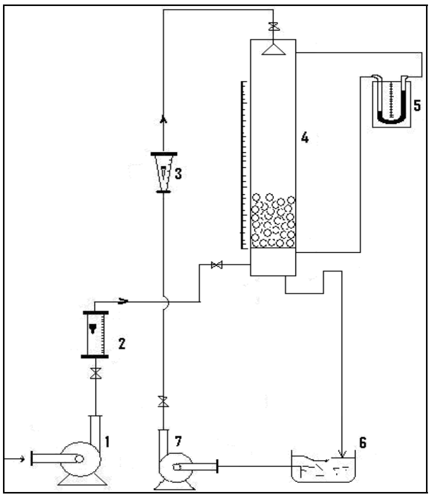

Figure 7. Schematic representation of experimental equipment.
Where: 1 - blower; 2 - air flow meter; 3 - water flow meter; 4 - fluidizing column; 5 - U-tube manometer; 6 - storage tank; 7 - centrifugal pump.

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 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 Table1.

Table 1. Range of the experimental conditions

Variable	Range
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; 0.1; 0.08
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

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 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 % [6]. All these conditions were fulfilled in the present study.

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