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Modelling of Gas-Liquid Hydrodynamics through Packed Beds

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ABSTRACT

Two-phase pressure drop and liquid holdup were measured in a acrylic tube of 76.2 mm diameter and 1000 mm height using 6.75 mm glass spheres as packing. The gas rates were varied from 0.12 to 0.75 kg/m .s and liquid rates were varied from 4.225 to 32.15 kg/m .s to realise gas-continuous trickle flow, pulse flow and dispersed bubble flow. The two-phase pressure drop in gas-continuous flow agreed satisfactorily with the earlier theoretical models. The pressure drops in pulse flow and dispersed bubble flow were predicted by modifying the earlier models. The liquid holdup data agreed well with earlier correlations.

INTRODUCTION

Pressure drop and liquid holdup are the important hydrodynamic parameters from the point of system design and operation of gas-liquid downflow packed columns. Extensive literature reviews are available on this topic (Satterfield, 1975, Gianetto et al, 1978, Hofmann, 1978). Depending on the extent of interaction between the phases, the flow regimes are classified as poor interaction regime (gas-continuous flow) and high interaction regime (Spray flow, pulse flow and dispersed bubble flow). Owing to the complexity of two-phase flow in packed beds, only a few attempts were made to predict pressure drop and liquid holdup through simplified theoretical models. The objective of this work is to look into the usefulness of the theoretical models proposed earlier with the help of additional experimental data and to improve the existing models if necessary.

THEORY

Gas-Continuous Flow :

The interaction between the flowing phases in gas-continuous flow is geometric in the sense that the flowing phases affect each other to the extent of reduction of cross-sectional area for the flow of each phase. Specchia and Baldi (1977) proposed a model that the pressure drop for a gas and liquid system in packed bed is only a function of gas velocity and liquid holdup for a particular packing as given by Eq.(1).

$$\delta_{LG} = \frac{K_1[1 - \varepsilon (1 - \beta_s - \beta_d)]^2}{d_e^2 \, \varepsilon^3 (1 - \beta_s - \beta_d)^3} \mu_G u_G + \frac{K_2[1 - \varepsilon(1 - \beta_s - \beta_d)]}{d_e \, \varepsilon^3 (1 - \beta_s - \beta_d)^3} \rho_G u_G^2 \quad .. (1)$$

The constants K1 and K2 were estimated for each packing by measuring single phase pressure drop on wet packing. Rao et al (1983) derived a geometric interaction model by a simple momentum balance for each of the phases to give the following equation :

$$\delta_{LG} = \frac{\alpha_L \overset{o}{\delta}_L}{\beta^3} - \rho_L g = \frac{\alpha_G \overset{o}{\delta}_G}{(1 - \beta)^3} - \rho_G g \quad ..(2)$$

α_L and α_G are the wetting parameters and were taken unity in gas-continuous flow. Rao and Drinkenburg (1985) modified the gas phase part of the eq.(2) by substituting single phase pressure drop of gas on wet packing. The modified geometric interaction model is given by

$$\delta_{LG} = \overset{o}{\delta}_{GW} / (1 - \beta_d)^3 - \rho_G g \quad .. (3)$$

eq.(3). Comparison of Eqs.(2) and (3) with experimental values showed an improvement with modified geometric interaction model (Rao and Drinkenburg, 1985). Sweeney (1967) developed a set of equations similar to Eq.(2) based on a different approach. The various geometric interaction models as given by Eqs.(1),(2) and (3) are found to be inadequate in high interaction regime because of friction at the interface, bubble formation, internal mixing and droplet formation.

Pulse Flow :

The pulse flow regime is characterized by low density and high density parts of gas and liquid mixtures. The flow mechanism in low density part is similar to gas-continuous flow. Rao and Drinkenburg (1985) estimated the pressure drop in the pulse (high density part) as being contributed by i) pressure drop due to bubble dispersion, ΔP_B, ii) pressure drop due to acceleration of liquid in the pulse, ΔP_A,

iii) pressure drop due to liquid mixing in the pulse ΔP_M, iv) pressure drop due to geometric interaction in the pulse, ΔP_L.

The pressure drop due to liquid mixing per unit height is given by

$$\delta_{LG,M} = n_M \Delta P_M \quad ..(4)$$

ΔP_M is the pressure drop due to liquid mixing in the pulse and is given by

$$\Delta P_M = \mu^3 H_p a_s^4 / \rho^2 V_p \varepsilon^4 (\beta_p - \beta_b)^4 \quad ..(5)$$

n_M is the total number of encounters of a liquid element (f_p / v_L).

The pressure drop due to acceleration of the liquid in the pulse is given by

$$\Delta P_A = \rho_L \varepsilon \beta_P (v_{LP} - V_P)^2 - \rho_L \varepsilon \beta_b (v_{Lb} - V_P)^2 \quad ..(6)$$

and $$\delta_{LG,A} = n_M \Delta P_A \quad ..(7)$$

Rao and Drinkenburg (1985), in their bubble dispersion model assumed that the size of the spherical bubble is equal to packing size, which may be approximately true when the porosity of the packing is of the order of 0.5. When the porosity of the packing is very much less than 0.5, then the bubble diameter may be approximately taken as

$$d_B = 2 \varepsilon d_p \quad ..(8)$$

Using Eq.(8), their equation is suitably modified to estimate the bubble dispersion pressure drop as,

$$\Delta P_B = 6 H_p \sigma_L (2d_B/3d_h - 1)/d_p d_B (1 - V_p/v_G) \quad ..(9)$$

The pressure drop due to bubble dispersion per unit height is then given by

$$\delta_{LG,B} = n_B \Delta P_B \quad ..(10)$$

n_B is the number of encounters of a gas bubble with the pulses which is given by

$$n_B = (f_p / V_p)(1 - V_p / v_G) \quad ..(11)$$

Dispersed Bubble Flow :

In this flow regime, the total pressure drop is due to geometric interaction and dispersion of gas. The bubble dispersion model as represented by Fig.1 assumes that when a spherical bubble of size d_B encounters a layer of packing, it elongates to a cylindrical bubble of length ℓ and diameter d_h which is equal to the hydraulic diameter of the packing.

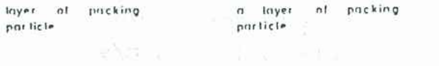

FIG 1: THE MODEL FOR BUBBLE DISPERSION

The difference in surface energy for a bubble in these two states is considered to be dissipated as heat during each cycle of surface enlargement and reduction process.

This process is repeated whenever the gas bubble encounters a layer of packing. The number of particle encounters of gas bubble is given by H/d_p where H is packing height. Hence from the total surface energy dissipated as heat the pressure drop due to bubble dispersion per unit height can be calculated to give

$$\delta_{LG,B} = 6 \sigma_L (2d_B/3d_h - 1)/d_p d_B \quad ..(12)$$

EXPERIMENTAL

A schematic diagram of the experimental set-up is given in Fig.2. It consists of a

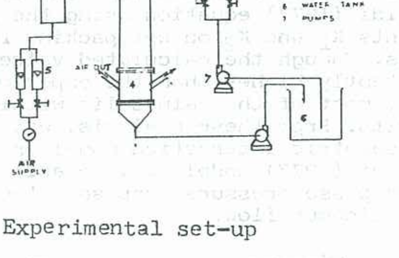

Fig.2 Experimental set-up

perspex column of 76.2 mm internal diameter with a packing height of 1000 mm. Air and water were distributed at the top of the column through a distributor consisting of two sets of copper tubes arranged alternately on a square pitch covering the entire cross-section of the packed column. Two solenoid valves provided at the top and bottom of the column facilitated the simultaneous shut-off of liquid for the measurement of dynamic liquid holdup. The gas was saturated by bubbling through water before it was allowed to enter the column. Glass beads of 6.75 mm diameter were used as packing. The porosity of the bed was 0.362. The static holdup β_s was found to be 0.064. Air rate was varied from 0.12 to 0.76 kg/m^2.s and liquid rate was varied from 4.225 to 32.15 kg/m^2.s to realize gas-continuous trickle flow, pulse flow and dispersed bubble flow. The pressure drops on dry and wet packing were measured at different gas velocities. Dynamic liquid holdup was measured by draining out the liquid for about 20 minutes which was trapped in the column using solenoid valves. Correction for the liquid present in the sections outside the packed section was made by conducting experiments with and without packed section. Pressure drops were measured using CCl /mercury manometers.

RESULTS AND DISCUSSION

Single Phase Pressure Drop :

The presssure drops measured at lower liquid rates when extrapolated to zero liquid rates at desired gas rates corresponded reasonably well with the measured pressure drops at zero liquid rate.

The measured values at zero liquid rate were then taken as the wet packing pressure drops of gas. The single phase pressure drops of gas on dry packing (β_s= 0) and on wet packing were fitted to Ergun-type equation as given by Eq.(1). The values of the constants K_1 and K_2 were then evaluated by linear least squares method. On dry packing (β_s= 0) K_1 and K_2 were found to be 300 and 2.18 respectively and on wet packing they were found to be 310 and 2.63 respectively. An estimate of the single phase pressure drop on wet packing was made from dry packing pressure drop data by geometric interaction model as given by Eq.(2). with $\alpha_G = 1$ and β replaced by β_s. The calculated values are consistantly lower than the experimental values but agreed within 20% error limit.

Two-Phase Pressure Drop

Experimental two-phase pressure drop data measured for 6.75 mm glass spheres is presented in Fig.3 that the variation of pressure drop with gas rate shows functional dependency in different flow regimes. The two-phase pressure drop δ_{LG} in gas-continuous flow was calculated using modified geometric interaction model as given by Eq.(3) and are presented in Fig.4 along with experimental values. The calculated and experimental values agreed within 25% error limit. δ_{LG} values on gas-continuous flow were also estimated by Speechia and Baldi (1977) equation using the estimated constants K_1 and K_2 on wet packing for 6.75 mm spheres. Though the calculated values are consistantly higher than the experimental values, most of the values lie within 25% error limits. From these analysis, either modified geometric interaction model or Speechia and Baldi (1977) model will be able to predict the two phase pressure drop satisfactorily in gas-continuous flow.

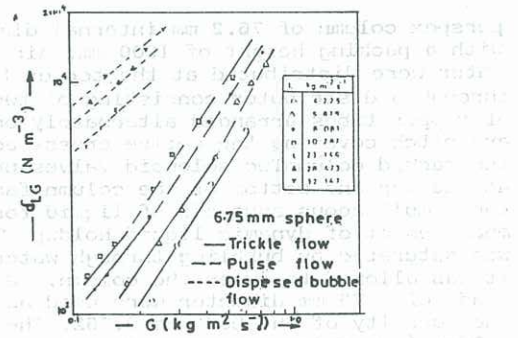

Fig.3 Experimental two phase pressure drop versus mass flow rate of gas and liquid

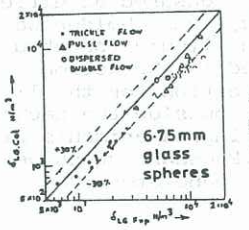

Fig.4 Experimental and calculated two phase pressure drop

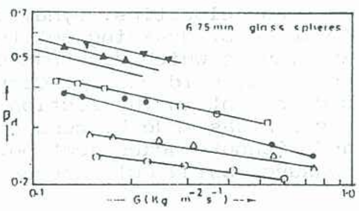

Fig.5 Experimental dynamic liquid hold-up versus mass flow rate of air and liquid (For key, Fig.3)

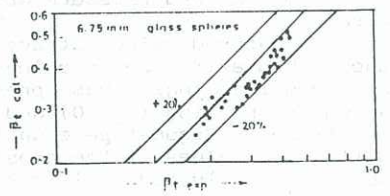

Fig.6 Experimental total liquid hold up with the values predicted by Rao et al (1983) correlation

The two-phase pressure drop in dispersed buble flow was calculated by adding the pressure drop contribution due to geometric interaction of the phases [Eq.(3)] and the pressure drop contribution by bubble dispersion [Eq.(12)]. The calculated and experimental values are compared in Fig.4. The agreement between the calculated and experimental values is better at lower levels of liquid rates. At higher levels of liquid rates, the difference between experimental and calculated values became larger especialy at higher gas rates. The pressure drop contribution due to bubble dispersion varies from 45 to 70% depending on the gas and liquid rates.

An estimate of the pressure drop in pulse flow was made using various pulse properties. Pulse frequency was measured by counting the pulses for a longer duration of time. The liquid holdups inside the pulse and outside the pulse were calculated by the correlations as proposed by Blok and Drinkenburg (1982). A typical average pulse height of 5 cm and pulse velocity of 1 m/s (Blok and Drinkenburg, 1982 and Rao and Drinkenburg, 1983) were assumed. The pressure drop in pulse flow was calculated by adding the pressure drop contributions due to geometric interaction in gas-continuous part and pulsing part, pressure drop contributions due to bubble dispersion, mixing and accleration of liquid in the pulsing part. The pressure drop due to the calculated values and are compared with the experimental values in Fig.4. As can be seen from the figure, almost all values lie within ± 25% error limits.

Liquid holdup :

Fig.(5) presents the dynamic liquid hold up data measured in gas-continuous flow, pulse flow and dispersed bubble flow at different gas and liquid rates. The experimental liquid holdup data agreed with the values calculated by the correlations proposed by Speechia and Baldi (1977) for poor interaction regime and for high interaction regime within 20% error limit. Total liquid holdup data in all the flow regimes agreed well with the values calculated by Rao et al (1983) correlation with a relative root mean square deviation of about 12% (Fig.6).

Nomenclature :

a_s	= specific surface area, m
d_B	= bubble diameter, m
d_p	= particle diameter, m
d_e	= effective particle diameter, m
d_h	= hydraulic diameter, m
f_p	= pulse frequency, s^{-1}
G	= gas mass flow rate, kg/m^2.s
H	= height, m
K_1, K_2	= constants
ℓ	= length, m
L	= liquid mass flow rate, kg/m^2.s
ΔP	= pressure drop, N/m^2
U	= superficial velocity, m/s
v	= real velocity, m/s
V_p	= pulse velocity, m/s

Greek letters

α	= wetting parameter
β	= liquid holdup, fraction of void volume
δ	= pressure drop, N/m^3

μ = viscosity, Ns/m^2

σ = surface tension, N/m

ε = bed porosity

ρ = density, kg/m^3

Subscripts

A = acceleration

B = bubble

b = base (gas-continuous part)

d = dynamic

G = gas

L = liquid

LG = liquid-gas (Two-phase)

M = mixing

P = pulse

s = static

Superscripts

o = single phase

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