Pressure Drop and Liquid Holdup in Co-current Gas-Liquid Downflow of Air-CMC Solutions through Packed Beds

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The present available experimental data on liquid holdup (735 data points) and pressure drop (863 data points) in co-current gas-liquid downflow through packed beds obtained using air-non-Newtonian (w = 0.2 %, 0.5 %, 0.8 % and 1 % CMC) solutions, were analyzed for their dependency on the system variables. Modified Reynolds number and Morton's number involving flow consistency index (k) and flow behavior index (n), along with other variables were used for the development of unified correlations to represent the available data.

Key words:
Packed bed contactors, two-phase downflow, liquid holdup, pressure drop

Introduction

Packed bed contactors are widely used in many chemical, biochemical, petroleum and petrochemical industries and in specific operations like distillation, absorption, humidification, desulfurisation and hydro-treating. Most biological processes are carried out in large scale by employing immobilized enzymes or cells where the system exhibits non-Newtonian properties. For optimum design and the scale up of these types of contactors, the knowledge of hydrodynamic parameters such as identification of flow regimes, pressure drop and phase holdup are inevitable using non-Newtonian liquid systems.

More than a few decades, there have been several reviews of literature on liquid holdup/saturation and pressure drop in co-current gas-liquid downflow through packed bed contactors using Newtonian fluids, but only a very few have used non-Newtonian liquid systems for their study. The important literature correlations for the estimation of liquid holdup and pressure drop, along with their range of applicability are compiled and discussed in detail (*Jegadeesh Babu* 2006).[11] Among the available correlations, most of them (*Larkins et al.,* 1961; *Sato et al.,* 1973;[23] *Charpentier* and *Favier*, 1975;[1] *Midoux et al.,* 1976;[17] *Ellman et al.,* 1990[4]) are based on the *Lockhart-Martinelli* (1949)[13] parameter 'χ', which requires *a priori* knowledge of single-phase pressure drop. The suggested correlations of *Turpin* and *Huntington* (1967),[26] *Hochman* and *Effron* (1969),[7] *Goto et al.* (1975),[5] *Speechia* and *Baldi* (1977),[25] *Matsuura et al.* (1979),[16] *Clements* and *Schmidt* (1980a and 1980b),[3] *Govardhana Rao et al.* (1983),[6] *Rao et al.* (1985),[19] *Sai* and *Varma* (1987 and 1988)[21] and *Iliuta et al.* (1997),[9] for the estimation of liquid holdup and pressure drop were developed using dimensionless groups to characterize the flow phenomena of both gas and liquid phases, properties of gas and liquids and the packing geometries. For the flow of non-Newtonian liquid systems through packed bed contactors, *Mohunta* and *Laddha* (1965),[18] and *Sai* and *Varma* (1987 and 1988)[21] have established semi-empirical correlations, whereas *Iliuta et al.* (1997 and 1999)[10] have made a comparative study of the flow behavior through upflow and downflow contactor. Even though a number of correlations are available, an accurate prediction of liquid holdup/saturation and two-phase pressure drop in a co-current downflow through packed bed contactors using a generalized approach is yet to be accomplished. Due to these constraints, it is imperative to develop unified correlations for the estimation of liquid holdup/saturation and two-phase pressure drop for air-CMC systems, covering a wide range of system variables.

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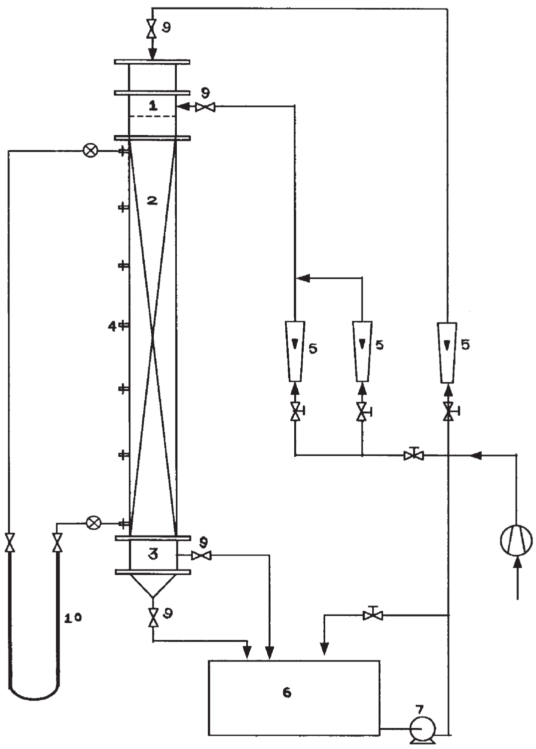
Experimental setup

Experiments were conducted using a cylindrical column of 0.092 m i.d. and a height of 1.86 m. Fig. 1 shows the schematic view of the experimental setup. The experimental column has a testing section of 1.84 m, mounted with a gas-liquid distributor to have a uniform distribution of the phases while entering into the testing section. Liquid was recirculated back to the storage tank using a gas-liquid separator provided at the bottom of the column. Liquid from a storage tank was pumped to the top of the column using a centrifugal pump and the compressed air was sent into the gas-liquid distributor through a regulating valve. Flow rates of gas and liquid phases were measured using individual calibrated rotameters. Pressure tapings were provided in the testing section to measure the pressure drop across the packed column using manometers. Solenoid valves were provided both at the inlet and outlet ends.

After attaining steady-state conditions, the air and liquid flow rates were simultaneously stopped using solenoid valves and further the column was disconnected from the supply lines and weighed (m_1) using an electronic weighing balance having an accuracy of ± 5 g. After weighing the column the liquid was allowed to drain for 30 min and again the column was weighed (m_2). Hot air was sent into the column to have a complete drying and the dried column was weighed (m_3) again. The difference between the weight of the column with the liquid phase and the dry mass gave the total liquid holdup (m_1-m_3), and the difference between the dry mass of the column and the mass of the column measured after 30 min gave the static liquid holdup (m_2-m_3). Similar experimental methodology was applied for all the experimental conditions. In the present research, air-water and air-CMC solutions (w = 0.2 %, 0.5 %, 0.8 % and 1 %) were used. The CMC solution is a power law model liquid (pseudoplastic fluid) and exhibits thixotropic nature. Hence, the properties viz., flow consistency index 'k' and flow behavior index 'n' were evaluated using the power law model,

$$\tau = k\left(\frac{\mathrm{d}u}{\mathrm{d}y}\right)^n \qquad (1)$$

The Haake Viscometer was used to measure the shear rate '$\mathrm{d}u/\mathrm{d}y$' for various shear stress 'τ'. The experimental data on liquid holdup and two-phase pressure drop used for the present analysis are given in Tables 1 and 2.



1. Gas-liquid distributor
2. Testing section
3. Gas-liquid separator
4. Manometer tapping
5. Rotameter
6. Liquid storage tank
7. Centrifugal pump
8. Compressor
9. Solenoid valve
10. Manometer

Fig. 1 – *Schematic diagram of the experimental setup*

Results and Discussion

Development of correlation for two-phase pressure drop

Among the hydrodynamic quantities, pressure drop is one of the important quantities used in the design of the packed bed contactors and it is further essential for estimating the energy required for pumping the fluids through the packed bed contactor. Two-phase pressure drop is also used as a correlating variable for the prediction of mass transfer rates in packed bed contactors (*Govardhana Rao*e t al., 1983).[6] For the past several decades, two-phase pressure drop was correlated either with the *Lockhart-Martinelli* (1949)[15] parameter 'χ' (initially established for the two-phase co-current flow in hollow pipes, which basically requires single-phase pressure drop data) in modified form, or by the modified Ergun equations (*Hutton* and *Leung*, 1974)[8] with the estimated constants, for individual particles. Few authors (*Turpin* and *Huntington*, 1967;[26] *Specchia* and *Baldi*, 1977;[25] *Sai* and *Varma*, 1987)[21] have used the dimensionless frictional factor 'f' for correlating the pressure drop in co-current gas-liquid downflow through packed bed contactors.

Fig. 2 shows the dependency of the two-phase pressure drop on gas and liquid flow rates. It is evident that the two-phase pressure drop increases with increasing gas and liquid flow rates, but the variation differs for different conditions of column

Table 1 – *Physical properties of the liquids used for estimating the liquid holdup and two-phase pressure drop*

Authors	Non-Newtonian liquids	ρ / $kg \cdot m^{-3}$	k / $kg \cdot m \cdot s^{n-2}$	n / –	σ / $\times 10^3$ $kg \cdot s^{-2}$
Present work	0.2 % CMC	996	0.0270	0.820	70.0
	0.3 % CMC	998	0.2354	0.705	70.0
	0.8 % CMC	1004	0.4660	0.703	70.0
	1.0 % CMC	1009	1.0830	0.641	70.0
Sai and *Varma* (1988)	0.2 % CMC	1000	0.00397	0.856	72.0
	0.5 % CMC	1001	0.01027	0.789	70.0
	1.0 % CMC	1004	0.01834	0.735	66.0
	2.0 % CMC	1010	0.06243	0.785	55.0
Iliuta et.al. (1997)	0.1 % CMC	1000.30	0.00496	0.936	72.0
	0.5 % CMC	1001.40	0.01778	0.900	72.0
	1.0 % CMC	1004.67	0.05599	0.849	72.0

Table 2 – *Details of the packing used for estimating the liquid holdup and two-phase pressure drop*

Authors	Packing	d_p / 10^3 m	ε
Present Work	Spherical	11.72	0.373
	Cylindrical	8.09	0.337
	Raschig Ring	3.14	0.707
Sai and Varma (1988)	Spherical	11.72	0.39
		6.72	0.35
	Raschig Ring	3.14	0.72
Iliuta (1997)	Spherical	3.3	0.356

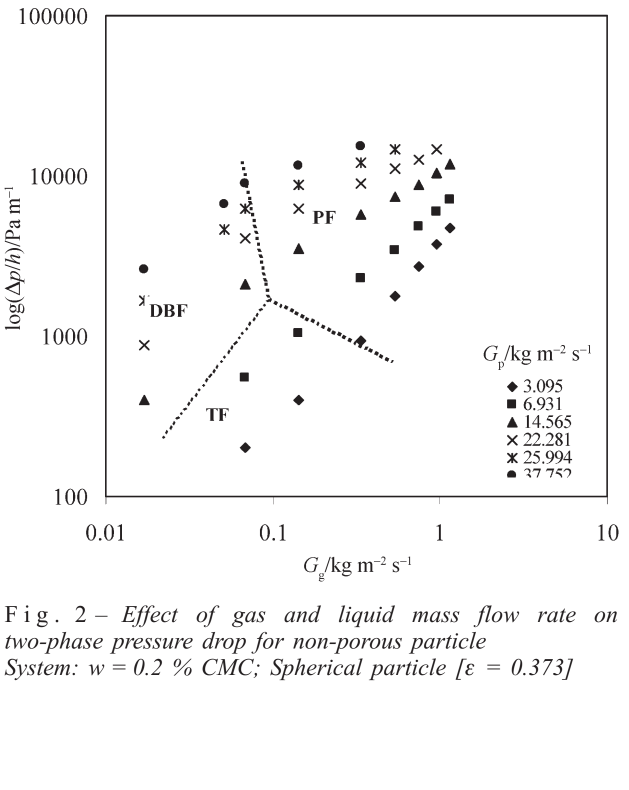
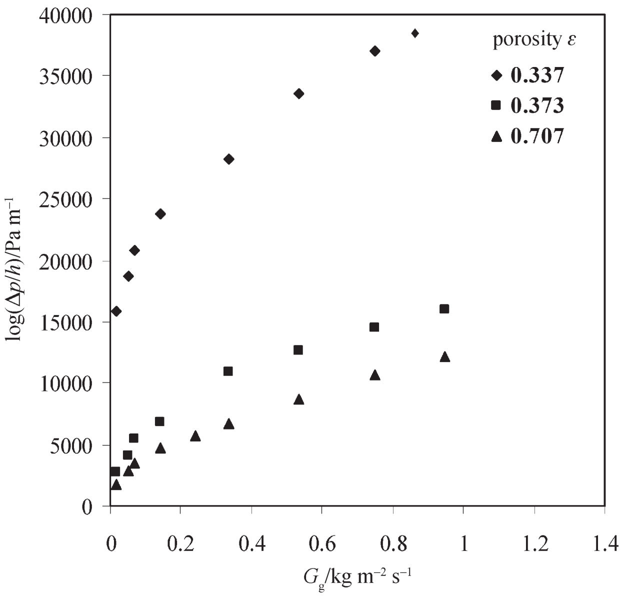
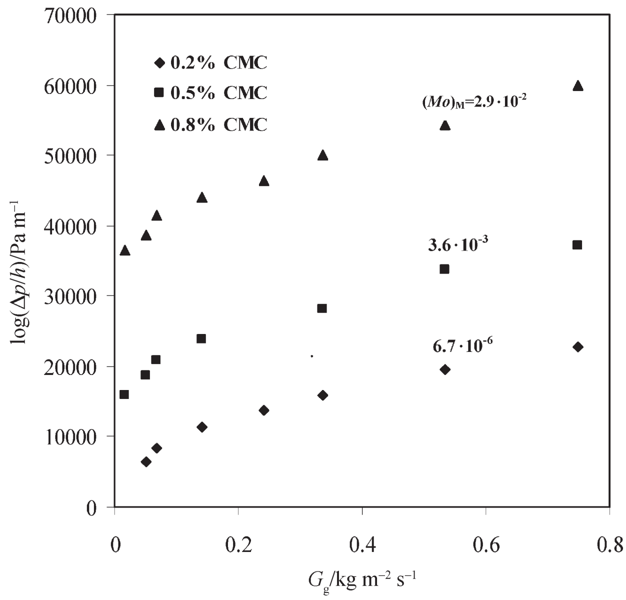


Fig. 2 – *Effect of gas and liquid mass flow rate on two-phase pressure drop for non-porous particle*
System: w = 0.2 % CMC; Spherical particle [ε = 0.373]

operation viz., i) at both low gas and liquid flow rates, ii) low gas flow rates with moderate increase in liquid flow rates, iii) at high gas and liquid flow rates, indicating the existence of different hydrodynamic flow regimes. The slopes and the intercepts of the graphs were found to vary with the flow regimes. By visual observations, the following flow patterns, namely trickle flow, disperse bubble flow and pulse flow were observed and the regime transitions were compared with the pressure drop measurements. Based on the analysis of the present data, it was found that the regime transitions vary with the liquid properties, packing characteristics and the column geometry, etc.. For the case of downflow mode of operation, the flow patterns are broadly classified as low interaction and high interaction regimes (*Midoux* et al., (1976)[17] and *Specchia* and *Baldi*, (1977)).[25] Low interaction regime consists of trickle flow, where the gas phase is continuous and the liquid is dispersed. Pulse and disperse bubble flow were considered as a high interaction regime where the liquid is continuous and the gas dispersed. Since the interactions between the phases are different in the individual hydrodynamic flow regime, it is always advantageous to have individual correlations. It is very difficult to generalize the delineation of flow regimes from the first principles, hence the flow regime transitions were made using a simulator developed by *Iliuta* et al. (1999)[10] based on a neural network model. The observed flow regime transitions based on visual observation as well as from the pressure drop measurements are favorably compared with those obtained from the neural network technique. The change in two-phase pressure drop is significantly more in the high interaction regime (disperse and pulse flow) when com-



F i g . 3 – *Effect of bed porosity on two-phase pressure drop for non-porous particle*
System: w = 0.5 % CMC; G_p = 10.16 kg m^{-2} s^{-1}



F i g . 4 – *Effect of physical properties on two-phase pressure drop for non-porous particle*
System: Cylindrical particle [ε = 0.337]; G_p = 10.726 kg m^{-2} s^{-1}

pared with the low interaction regime (trickle flow). It is also observed that the two-phase pressure drop decreases with an increase in bed porosity (Fig. 3) and it is found that the pressure drop is more for spherical (ε = 0.373) and cylindrical particles (ε = 0.337) when compared with raschig rings (ε = 0.707), which may be attributed to the fact that the available void volume for the flow of gas and liquid phases through the packed section were less. Further, it is also observed that the two-phase pressure drop increases with an increase in CMC concentrations for a constant gas and liquid mass flow rate (Fig. 4). For higher concentrations, the liquid side shear stress was greater, which leads to an increase in the two-phase pressure drop. The suggested correlation of *Sai* and *Varma* (1987),[21] established for the estimation of the friction factor in non-Newtonian liquid systems, when analyzed using the present experimental data, gave high deviations (> 35 %). The simulator proposed by Iliuta *et al.* (1999) also over-predicts (> 40 %) the two-phase pressure drop for non-Newtonian systems.

The analysis of the present experimental data on air-CMC systems in packed bed downflow contactors, confirms the dependency of the two-phase pressure drop on all the variables viz., phase flow rates, physical properties of liquid systems (density, viscosity and interfacial tension) and geometrical quantities (particle dimension and shape, bed porosity) of the packed bed. Therefore, it is proposed to analyze the effect of all the variables through different combinations of dimensionless groups. The effect of individual flow rates (gas and liquid), viscosity and density have been accounted by the corresponding Reynolds number ($(Re)_l$ and $(Re)_g$), whereas the combined effect of the liquid properties in the form of Morton number have also been considered. In order to consider the power law quantities (flow consistency index and flow behavior index), the modified Reynolds number $(Re)_{lM}$ and modified Morton number $(Mo)_M$ were used to form a generalized correlation of the following form,

$$\left(\frac{\Delta p}{\rho g h}\right)_{lg} = K(Re)_{lM}^{a}(Re)_{g}^{b}\left(\frac{\varepsilon}{1-\varepsilon}\right)^{c}\left(\frac{D_e}{D_c}\right)^{d}(Mo)_M^{e} \quad (2)$$

A collection of 863 data on two-phase pressure drop (low interaction regime (128) and high interaction regime (735)), were used for regression analysis, to establish the following correlations for representing the pressure drop data.

For low interaction regime

$$\left(\frac{\Delta p}{\rho g h}\right)_{lg} = 2.4\cdot 10^{-3}(Re)_{lM}^{0.6}(Re)_{g}^{0.35}\cdot \cdot\left(\frac{\varepsilon}{1-\varepsilon}\right)^{-0.83}\left(\frac{D_e}{D_c}\right)^{-2.38}(Mo)_M^{0.55} \quad (3)$$

For high interaction regime

$$\left(\frac{\Delta p}{\rho g h}\right)_{lg} = 4.4\cdot 10^{-4}(Re)_{lM}^{0.6}(Re)_{g}^{0.49}\cdot \cdot\left(\frac{\varepsilon}{1-\varepsilon}\right)^{-0.72}\left(\frac{D_e}{D_c}\right)^{-2.22}(Mo)_M^{0.26} \quad (4)$$

The estimated RMS errors for eq. 3 and eq. 4 were found to be ±18.3 % and ±19.2 % respectively and the parity plots for the comparison of the predicted and experimental pressure drop data are shown in Fig. 5 and 6. For validation of the present proposed correlations to the limiting conditions, they were tested with the present data obtained using air-water systems ($k = 0.001(\mathrm{Pa \cdot s}^n)$, $n = 1$) with spherical particle (d_p = 1.172 mm and ε = 0.373) and the estimated RMS error was ±14.5 %.

Development of correlation for dynamic liquid holdup

In general, the liquid holdup is defined as the fractional amount of liquid retained in the packed section for any specific flow rate of gas and liquid phases through the packed bed. Normally, for porous particles the liquid holdup has two parts i.e. external and internal holdup, whereas for the case of non-porous particles only the external holdup plays a vital role, which includes the dynamic (or free flowing) and the static liquid holdup, and thus total liquid holdup can be explained as,

$$\varepsilon_{\mathrm{t}} = \varepsilon_{\mathrm{d}} + \varepsilon_{\mathrm{s}} \tag{5}$$

The static liquid holdup, for all the experimental conditions have been measured. It is observed from the literature that the static liquid holdup is a function of bed porosity and liquid properties, in particular interfacial tension of the liquid. The present data (12 points) on static liquid holdup were tested using the available correlations of *Saez* and *Carbonell* (1985)[20] and *Sivakumar* and *Murugesan* (2003)[24] suggested for Newtonian systems, and the corresponding errors were found to be ± 9.87 % and ± 11.3 %. The static holdup forms around 5 to 8 % of the total holdup for the range covered in this work. Hence, the contribution of static holdup alone, towards the pressure drop may be neglected. Also, the quantities affecting the static holdup (bed porosity, interfacial tension, etc.) have been considered for the development of pressure drop correlation (in the previous part), which in turn will take care of the effect of static holdup on the pressure drop, if any.

Several authors (Table 1) have analyzed liquid saturation instead of dynamic liquid holdup in gas-liquid downflow through packed bed contactors. Liquid saturation is defined as the ratio of the volume of liquid phase to the void volume and hence the liquid saturation and liquid holdup are interrelated as,

$$\varepsilon_{\mathrm{t}} = \beta_{\mathrm{t}}\varepsilon \tag{6}$$

Based on the experimental dynamic liquid holdup data obtained using air-CMC (0.2%, 0.5%, 0.8% and 1.0% CMC) systems, the dependency of liquid holdup on the variables viz., flow rates of individual phases, physical properties, the column and the bed geometries have been evaluated. Fig. 7 shows the effect of mass flow rates of the individual phases on dynamic liquid holdup (1.0% CMC) and it is observed that, both the gas and liquid mass flow rates have significant effects on dynamic liq-

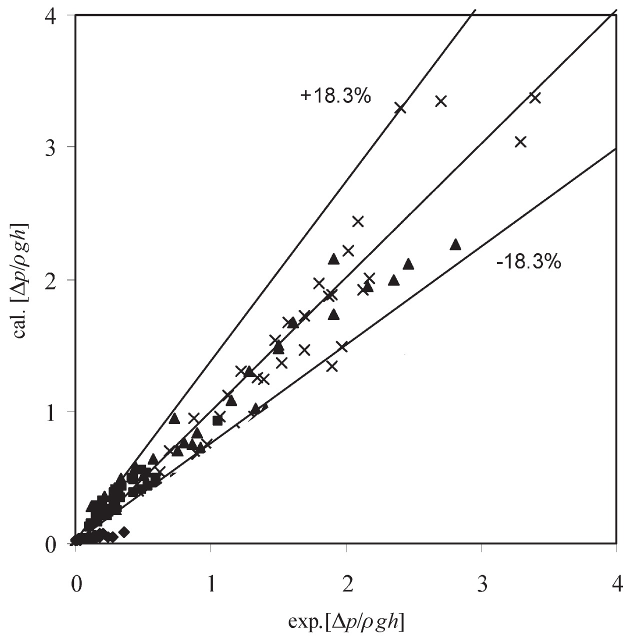


Fig. 5 – *Comparison of experimental and predicted two-phase pressure drop for low interaction regime*
Mass fraction (w): 0.2 % CMC, 0.5 % CMC, 0.8 % CMC, × 1 % CMC.

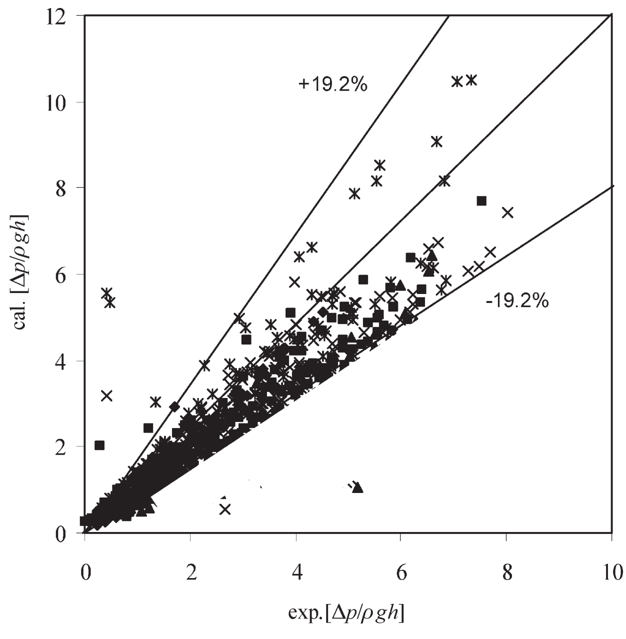


Fig. 6 – *Comparison of experimental and predicted two-phase pressure drop for high interaction regime*
*Mass fraction (w): 0.2 % CMC, 0.5 % CMC, 0.8 % CMC, × 1.0 % CMC, * 2.0 % CMC*

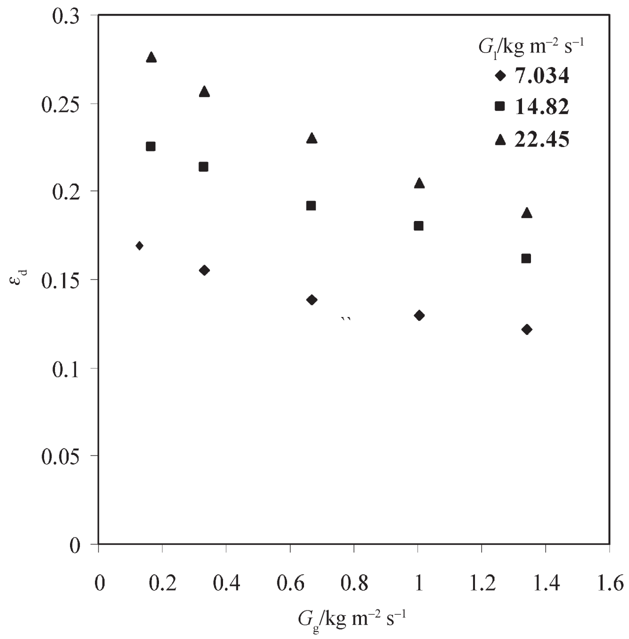


Fig. 7 – *Effect of gas and liquid mass flow rate on dynamic liquid holdup for non-porous particle System: w = 1 % CMC; Raschig Ring [ε = 0.72]*

uid holdup. The dynamic liquid holdup decreases with increasing gas flow rate for a constant liquid flow rate and it increases with increasing liquid flow rate for a constant gas flow rate, as observed and reported by Sai and Varma (1988). The change in the flow pattern of gas and liquid phases has a significant effect in the variation of the dynamic liquid holdup. It is also observed that the changes in the dynamic liquid holdup are minimal in the trickling flow regime when compared with the pulse and disperse bubble flow regime. In continuous and low gas flow rates (trickle flow), the effect of gas flow rate on the dynamic liquid holdup is negligible since the effect of drag force exerted by the gas on liquid is insignificant. For higher liquid flow rates, where the liquid is continuous, the effect of drag force by the gas phase on the liquid phase increases sharply with increasing gas flow rates (pulse and disperse bubble flow). The increase in the drag forces lead to a reduction in liquid residence time in the packed bed, and hence a decrease in dynamic liquid holdup is observed. In packed bed contactors, an alternative gas-rich and liquid-rich mixture forms the pulse flow, whereas in disperse bubble flow the gas phase disperses as bubbles in the liquid phase. In the disperse bubble flow regime, changes in the liquid holdup are due to the fact that the gas side shear stress appears to be important in balancing the driving force and the buoyancy force exerted by the gas bubbles, thereby a gradual increase in the gas flow rate shows a sharp decrease in the dynamic liquid holdup.

For the low interaction regime, the effects of bed geometry and porosity have little influence on the hydrodynamic quantities, whereas in the high interaction regime (pulse and disperse bubble flow) the effect of bed porosity is comparatively greater. Since the particle (diameter and shape) characteristics and the bed characteristics (bed porosity) are interrelated, it is difficult to identify the effect of the individual relationship with the liquid phase holdup. It is obvious from literature, even though for a same diameter of packing materials, i.e. sphere, cylinders, raschig ring, Berl saddle etc., the bed porosity for the raschig ring was reported (*Ellman* et al. (1989))[4] to be higher when compared with other particles, which is in good agreement with the present observations. Hence, for the further analysis of data, the effect of bed porosities was considered. The dynamic liquid holdup increases with an increase in the bed porosity (Fig. 8). For the development of correlations for liquid holdup 'ε_d', the bed porosity 'ε' and 'D_e' the equivalent diameter (which incorporate the effect of 'φ' the shape factor) were considered apart from other variables. From the experimental results it is observed that the changes in the physical properties of the liquid systems have considerable effect on the dynamic liquid holdup. It is found (Fig. 9) that the dynamic liquid holdup increases with an increase in liquid properties irrespective of the packing particles used. For 0.2 % CMC solution, dynamic liquid holdup is low when compared with 1.0 % CMC solution, for a constant gas and liquid flow rate. At low gas flow rates, the tendency of variation of dynamic liquid

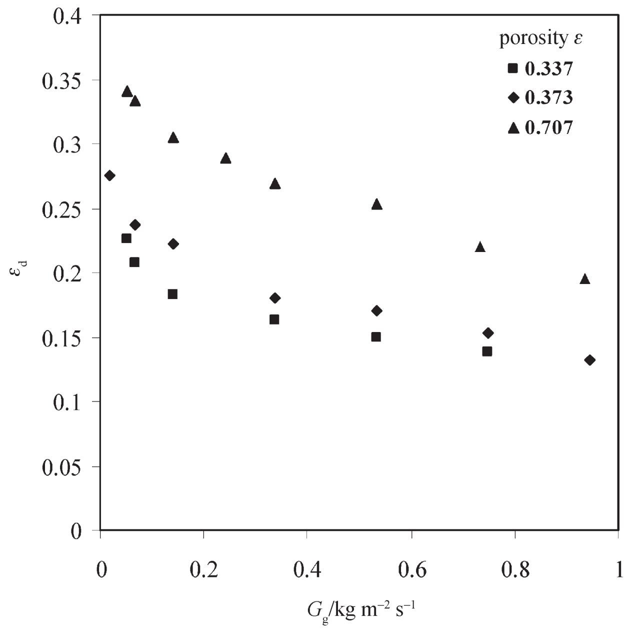
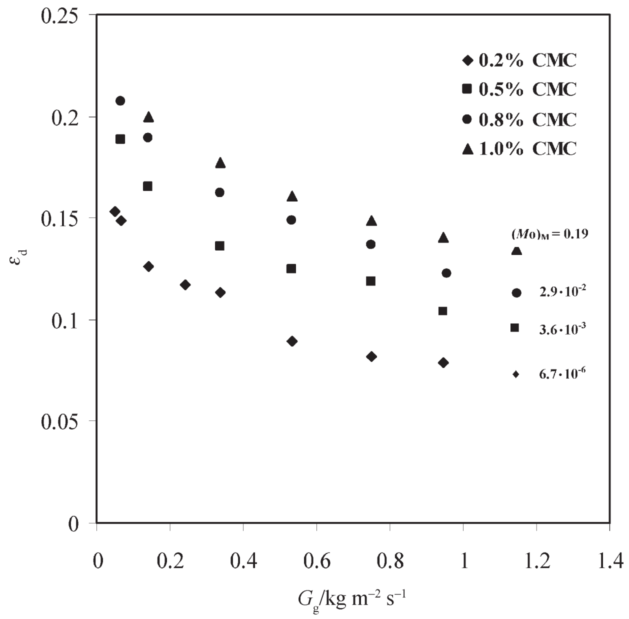
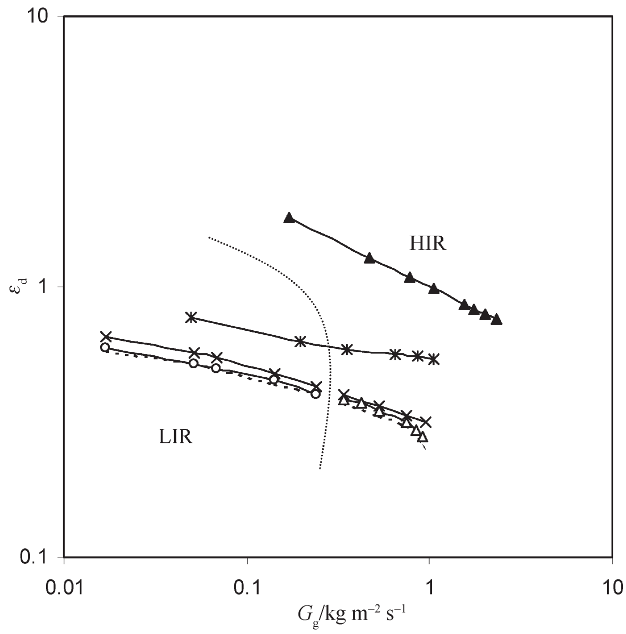


Fig. 8 – *Effect of bed porosity on dynamic liquid holdup for non-porous particle System: w = 0.5 % CMC; G_l = 17.239 kg m^{-2} s^{-1}*



F i g . 9 – *Effect of physical properties on dynamic liquid holdup for non-porous particle*
System: Cylindrical particle [ε = 0.337]; G_p = 6.931 kg m^{-2} s^{-1}



F i g . 1 0 – *Comparison of literature and present correlations of dynamic liquid holdup with experimental data*
System: Air-0.2 % CMC, particle cylinder, G_p = 18.433 kg m^{-2} s^{-1}
○ low interaction (equ. 4), Δ high interaction (equ. 5), ····· experimental data, × Sai and Varma (1988), ✱ Venkataratnam (1990), ▲ Iliuta et al. (1999)

holdup (Fig. 9) 'ε_d' for $Mo_M = 2.9 \cdot 10^{-2}$ (0.8 % CMC) and $6.7 \cdot 10^{-6}$ (0.2 % CMC) is found to be slightly different for the other two cases (Mo_M = 0.19 and $3.6 \cdot 10^{-3}$). The reason could be attributed to the difference in the operating hydrodynamic regimes, i.e. for the earlier case (0.2 % and 0.8 % CMC), the flow regime corresponds to trickle flow for low gas flow rates and shifts to pulse flow with increasing gas flow rates, whereas for the latter case, at low gas rates the observed flow regime was disperse bubble flow which then shifts to pulse flow regime, with increasing gas flow rates. The increase in CMC concentration leads to an increase in liquid side shear stress at gas-liquid and liquid-solid interfaces, and hence an increase in dynamic liquid holdup was observed.

A comparison plot of the important available literature correlations for non-Newtonian solutions for estimating liquid phase holdup is shown in Fig. 10. From the graph, it is observed that the simulated values of *Iliuta* et al. (1999) show higher deviations when compared with the present experimental data and other literature correlations. Though the correlations of *Venkataratnam* (1990)[27] show less deviation, the complete effect of physical properties of the liquid phase, particularly the interfacial tension, which is the major factor affecting the static holdup, has not been considered in his correlation. Further, the proposed equation is a dimensional correlation and care should be taken while using the constants. The correlations of *Sai* and *Varma* (1988)[21] and *Mohunta* and *Laddha* (1965)[18] over-predict the dynamic liquid holdup than the experimental value due to the variation in physical properties, in particular the flow consistency index '*k*' of the liquid systems used, and the suggested constants in the correlation are dimensional. Analysis of the present experimental data, show a strong dependency of dynamic liquid holdup on the variables viz. phase flow rates (gas and liquid), physical properties liquid phase and geometrical quantities of the packed bed (bed porosity and shape factor). The inconsistencies coupled with the limitations of the available literature correlations, made it necessary to establish a generalized dynamic liquid holdup correlation for air-CMC solution in co-current downflow through packed bed contactors. As discussed in the previous section, the modified Reynolds number $(Re)_{IM}$ and the modified Morton number $(Mo)_M$, along with other variables, i.e. bed porosity, particle diameter etc., were used for correlating the dynamic liquid holdup data,

$$\varepsilon_d = K(Re)_{IM}^a (Re)_g^b \left(\frac{\varepsilon}{1-\varepsilon}\right)^c \left(\frac{D_e}{D_c}\right)^d (Mo)_M^e \quad (7)$$

The suggested simulator of *Iliuta* et al. (1999)[9] was used for delineation of hydrodynamic (high and low interaction) regimes. In low interaction regime (gas continuous) liquid trickles over the packing and the gas passes through the voids of the packing, leading to a minimum gas-liquid interaction, whereas in the high interaction regime (liquid continuous) which includes disperse and pulse flow, the gas and the liquid interaction is expected to be greater (*Turpin* and *Huntington*, 1967;[26] *Charpentier* and *Favier*, 1975).[1] Regression analysis of the available experimental data on both low and high interaction regimes yielded the following constants and indices of eq. (7),

For low interaction regime

$$\varepsilon_{\mathrm{d}} = 0.182\,(Re)_{\mathrm{lM}}^{0.14}\,(Re)_{\mathrm{g}}^{-0.12} \cdot \cdot \left(\frac{\varepsilon}{1-\varepsilon}\right)^{0.01} \left(\frac{D_{\mathrm{e}}}{D_{\mathrm{c}}}\right)^{-0.34} (Mo)_{\mathrm{M}}^{0.1} \tag{8}$$

For high interaction regime

$$\varepsilon_{\mathrm{d}} = 0.764\,(Re)_{\mathrm{lM}}^{0.17}\,(Re)_{\mathrm{g}}^{-0.18} \cdot \cdot \left(\frac{\varepsilon}{1-\varepsilon}\right)^{0.24} \left(\frac{D_{\mathrm{e}}}{D_{\mathrm{c}}}\right)^{-0.04} (Mo)_{\mathrm{M}}^{0.1} \tag{9}$$

Statistical error analysis of the proposed correlations (Eq. 8 and 9) showed an RMS error of ±11.4 % for the low interaction regime and ±10.12 % for the high interaction regime, indicating a satisfactory representation of the available data on air-CMC systems. The parity plots are shown in Fig. 11 and 12. The ranges of variables considered for the development of correlations are,

flow consistency index k = 0.00397 – 1.083 Pa · s^n
flow behavior index n = 0.641 – 0.936
density ρ = 996 – 1009 kg · m^{-3}
liquid mass flux G_{l} = 1.51 – 37.75 kg · m^{-2} · s^{-1}
gas mass flux G_{g} = 0.017 – 1.34 kg · m^{-2} · s^{-1}
porosity ε = 0.337 – 0.720

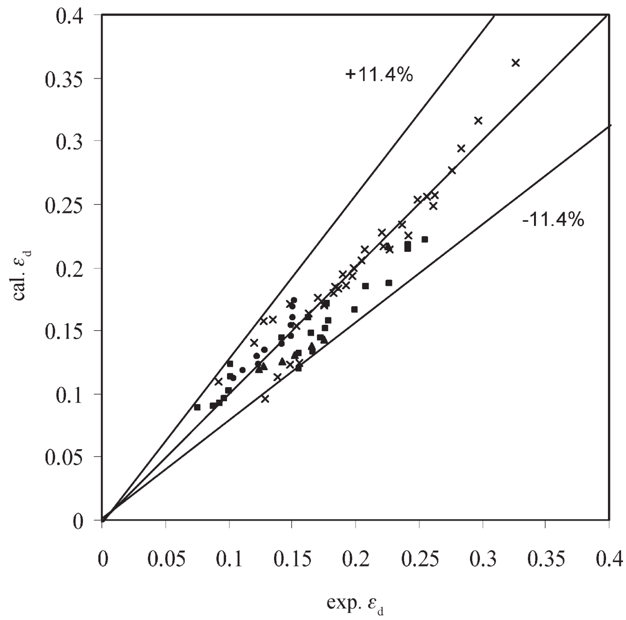


Fig. 11 – *Comparison of experimental and predicted dynamic liquid holdup for low interaction regime*
Mass fraction (w): 0.2 % CMC, 0.5 % CMC, 0.8 % CMC, × 1 % CMC

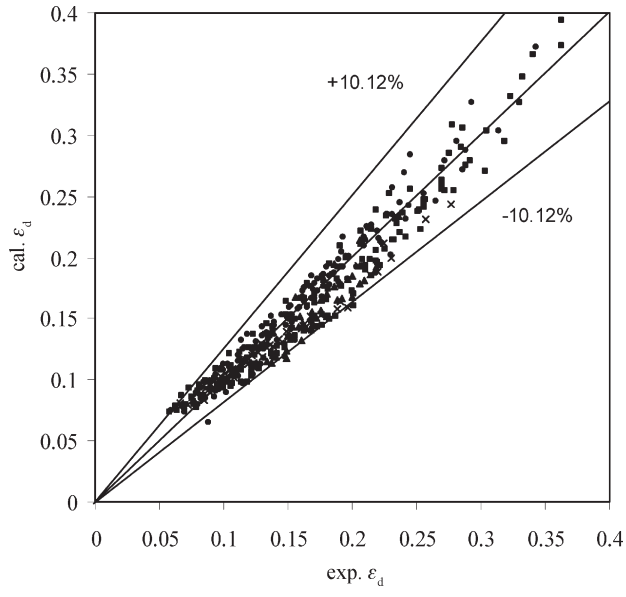


Fig. 12 – *Comparison of experimental and predicted dynamic liquid holdup for high interaction regime*
Mass fraction (w): 0.2 % CMC, 0.5 % CMC, 0.8 % CMC, × 1 % CMC

Conclusion

Detailed study on the dependency of liquid holdup and the two-phase pressure drop in co-current gas-liquid downflow through packed beds on the system variables such as flow rates of individual phases, physical properties and the bed characteristics, etc., were made to develop generalized correlations (eqs. (3), (4), (8) and (9)). For the development of the correlations, even though different combinations of dimensionless groups have been attempted, normal definition of Reynolds number and Morton number with the modification of the viscosity (in the dimensionless group) in terms of flow consistency index (k) and flow behavior index (n), apart from the bed characteristics, are found to be sufficient to represent the available data on air-CMC solutions. The range of the dimensionless groups used in the present analysis are, $(Re)_{\mathrm{lM}}$ = 0.0064 – 81.0273; $(Re)_{\mathrm{g}}$ = 2.884 – 853.52; $(\varepsilon/1-\varepsilon)$ = 0.5083 – 2.5714; $D_{\mathrm{e}}/D_{\mathrm{c}}$ = 0.0219 – 0.0545; $(Mo)_{\mathrm{M}}$ = 8.89 · 10^{-9} – 0.1933. However, the applicability of the present proposed correlations for all kinds of non-Newtonian liquids requires further verification using future experiments.

Nomenclature

D_c – column diameter, m

D_e – equivalent diameter $\left(\frac{2}{3}\frac{\varphi_s d_p \varepsilon}{(1-\varepsilon)}\right)$, m

d_p – particle diameter, m

g – acceleration due to gravity, m s^{-2}

G_g – gas mass velocity, kg m^{-2} s^{-1}

G_l – liquid mass flux, kg m^{-2} s^{-1}

h – height of testing section for pressure drop, m

k – flow consistency index, kg m^{-1} s^{n-2}

$(Mo)_M$ – modified Morton number $\left(\frac{g^{3n-2}k^4}{\rho^{2-n}\sigma^{n+2}}\right)$, –

m – mass, kg

n – flow behavior index, –

$(\Delta p/h)_{lg}$ – two-phase pressure drop, Pa m^{-1}

$(Re)_g$ – gas phase Reynolds number $\left(\frac{d_p u_g \rho_g}{\mu_g}\right)$, –

$(Re)_{lM}$ – modified liquid Reynolds number $\left(\frac{d_p^n u_l^{2-n}\rho_l}{k}\right)$, –

RMS – root mean square, –

u_g – gas velocity, m s^{-1}

u_l – liquid velocity, m s^{-1}

β_d – dynamic liquid saturation, –

β_s – static liquid saturation, –

β_t – total liquid saturation, –

ε – porosity, –

ε_d – dynamic liquid holdup, –

ε_s – static liquid holdup, –

ε_t – total liquid holdup, –

ρ_g – gas density, kg m^{-3}

ρ_l – liquid density, kg m^{-3}

μ_g – gas viscosity, Pa s^{-1}

μ_l – liquid viscosity, Pa s^{-1}

σ_l – interface tension, N m^{-1}

φ_s – sphericity, –

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