Journal of Advanced Engineering Research
ISSN: 2393-8447
Volume 1, Issue 1, 2014, pp.8-15

Holdup Studies in IFBR for Wastewater Treatment

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ABSTRACT

In the present investigation, bed porosity and phase holdups in a three-phase inverse fluidized bed reactor are determined for Newtonian (aqueous solutions of glycerol) and non-Newtonian (aqueous solutions of carboxy methyl cellulose) systems using low-density polyethylene (LDPE) and polypropylene (PP) particles of different diameters (4, 6 and 8 mm). The gas holdup (ε_g) and porosity (ε) in the three-phase inverse fluidized bed reactor was observed to increase with an increase in U_g. The solid holdup (ε_s) and liquid holdup (ε_l) decreased as the U_g increased for a fixed U_l. The liquid holdup increased with increase in U_l and the gas holdup increased with increase in particle diameter. The effect of liquid concentration on gas holdup and liquid holdup are also studied. Empirical correlations are developed to determine the phase hodups. Modified drift flux model have been presented for the prediction of the phase holdups (solid, liquid and gas) for Newtonian and non-Newtonian systems.

Keywords - Three-phase inverse fluidization, phase-holdup, modified drift flux model, Newtonian fluids and non-Newtonian fluids.

1. INTRODUCTION

Fluidization is an operation, which involves the flow of solids in contact with gas, liquid or gas and liquid. Fluidization has been the preferred method of operation for contacting solid and fluid phases for physical and chemical processes due to its many advantages viz. small pressure drop, good solid mixing and high heat and mass transfer rates. Fluidization technology has been applied significantly in chemical, petrochemical, metallurgical, mineral and biochemical operations. In the conventional fluidized bed, the solid particles have a higher density than that of the fluid. When the solids have a density lower than that of the fluid, it can be fluidized by a down flow of this liquid and this multiphase system is called inverse fluidized bed.

During recent years considerable effort has been expanded in exploring and understanding the hydrodynamics of fluid flow and heat and mass transfer in two- and three-phase fluidized beds [1]. Fluidized bed systems have proved their versatility for carrying out aerobic fermentation processes, bio-treatment of wastewater, refineries, and hydro metallic operations, biochemical engineering and polymeric industries [2-5]. Fan *et al.* [1] studied the hydrodynamic aspects of an inverse gas-liquid-solid fluidized bed reactor. Chern *et al.* [7] modified the Wallis drift flux model to describe the gas holdup in a constrained inverse fluidized bed. Buffiere and Moletta [8] investigated the hydrodynamic characteristics of inverse three-phase fluidized beds using two types of low-density particles. Miura et al., 1997 carried out the gas holdup and bed expansion measurements for a bed of glass beads fluidized in Newtonian liquids and Non-Newtonian liquids with gas.

1.1 Non-Newtonian Fluids

The wide occurrence of non-Newtonian fluids has recently motivated the investigation of the flow behavior of these fluids in multi-particle systems. Examples of such flows include the flow of oil through porous rock, the movement of aqueous polymer solutions through sand and sand-stone in tertiary oil recovery, the filtration of polymer solutions and slurries, the flow of non-Newtonian liquids through the ion-exchange beds, catalytic polymerization in hydroxydation processes, etc. Experimental studies of the flow of non-Newtonian fluids through fixed and fluidized beds were carried out by several investigators. In biochemical industries a number of fluids represent non-Newtonian behaviors.

However, little attention has been focused on three-phase inverse fluidized beds with non-Newtonian fluids. The objective of the present study is to determine the holdups in three-phase inverse fluidized beds and to modify the drift flux model for non-Newtonian fluids.

2. EXPERIMENTAL SET UP

The experimental set-up of the three-phase inverse fluidized bed reactor is shown in Fig 1. The column (100 mm) was made up of Perspex with a maximum height of 1800 mm and a wall thickness of 3 mm. The column consisted of three sections, namely liquid distribution section, test section and the liquid discharge section. The liquid distribution section comprises an inverted conical shape liquid distributor located at the top of the column in such a way that uniform distribution of the liquid throughout the column is ensured and an overflow arrangement to maintain a constant liquid level inside the column. An air vent is also provided at the top of the column. The test section consists of a wire mesh provided both at the top and the bottom to prevent the elutriation of the particles. Above the liquid discharge section, the gas sparger is provided for airflow. The airline is connected to a compressor through a calibrated flow meter. The liquid discharge section connects a pipe to transfer the liquid to the tank so that it is recirculated. A control valve is also provided in the discharge line to adjust the flow-rate. All runs were made at room temperature.

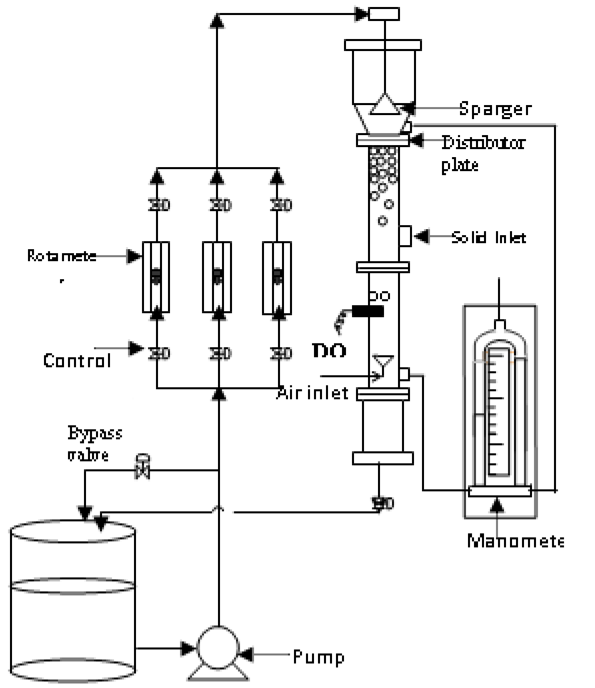

Fig. 1 The schematic diagram of Inverse Fluidized Bed Reactor

2.1 Measurement of phase holdup in a three-phase inverse fluidized bed reactor

The bed porosity (ε) and the holdups of the solid particles (ε_s), liquid (ε_l) and the gas (ε_g) are interrelated through the following relationships.

$$\in_s + \in_l + \in_g = 1 \quad (1)$$

$$\in = \in_l + \in_g = 1 - \in_s \quad (2)$$

The mean holdup of solids in the bed (ε_s) was calculated on the basis of the weight of dry particles, M_s using the equation:

$$\in_s = \frac{M_s}{HA\,\rho_S} \quad (3)$$

The gas holdup was determined using volume expansion technique using the following equation.

$$\in_g = \frac{H - H_0}{H} \quad (4)$$

Substituting ε_s and ε_g in equation (1), the liquid holdup, ε_l is calculated.

The properties of solid particles are presented in Table 1. The rheological properties of glycerol and CMC were measured using concentric cylinder viscometer (Haake VT 181) and are given in Table 2 and 3 respectively.

Table 1 Properties of solid particles

Particle	Diameter, d_p (mm)	Initial Porosity, ε_0	Density, ρ_s (kg/m^3)
LDPE	4	0.441	940
	6	0.421	
	8	0.431	
PP	4	0.431	830
	6	0.418	
	8	0.433	

Table 2 Rheological properties of aqueous solutions of glycerol

Concentration of glycerol	ρ_l (kg/m^3)	μ_l (mPa.s)
30%	1056	1.94
40%	1088	2.56
50%	1107	3.74
60%	1119	6.11
70%	1132	9.35

Table 3 Rheological properties of CMC at 30 ℃

CMC concentration (Wt %)	ρ (kg/m^3)	K ($Pa.s^n x10^2$)	n
0.1	1033	1.42	0.86
0.2	1041	2.38	0.84
0.3	1046	3.59	0.81

3. RESULTS AND DISCUSSION

3.1 Effect of superficial gas velocity (Ug) on phase holdup

The variation of phase holdups (solid holdup (ε_s), liquid holdup (ε_l) and gas holdup (ε_g) with U_g is shown in Fig. 2 for 8 mm LDPE particle. The ε_s and ε_l decreased as U_g increased whereas the ε_g increased as U_g increased. When U_g increased the bed expansion and the total volume of fluidized bed increased, but the volume fraction of the solids decreased. With an increase in airflow rate, the bed porosity and the volume fraction of gas, ε_g also increased in the fluidized bed and hence ε_l decreased. A similar trend was observed for 6, 4 mm LDPE and 8, 6, 4 mm PP particles. In the present study, no hysteresis effects were observed in the gas holdup with respect to gas velocity under the given conditions. Gas holdup in three-phase inverse fluidized beds showed trends similar to that found in a bubble column or fluidized beds. Essentially, at a given gas velocity, the same gas holdup was obtained independent of whether the preceding gas velocity used was lower or higher [1].

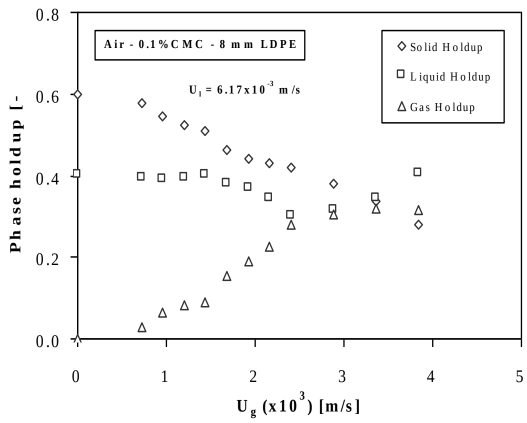

Fig. 2 Effect of gas velocity (U_g) on phase holdup in air-0.1% CMC–8 mm LDPE system for fluidized bed (H_f)

The typical porosity-velocity relationship for non-Newtonian fluid (CMC) is shown in Fig. 3. It was observed that there was a linear relationship between porosity and superficial gas velocity and the bed expansion was a function of rheological properties of liquid.

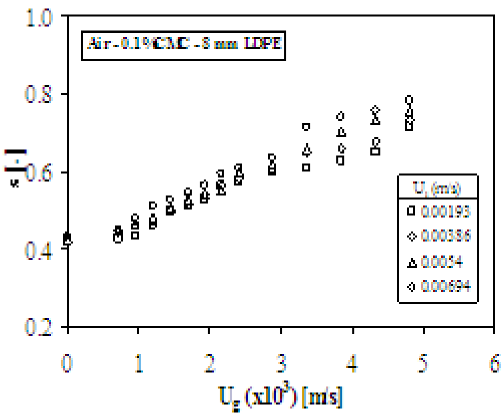

Fig. 3 Effect of gas velocity (U_g) on bed porosity in air - 0.1% CMC - 8 mm LDPE system for fluidized bed (H_f)

3.2 Effect of superficial liquid velocity (U_l) on gas holdup (ε_g)

The effect of U_l on gas holdup for air - water - 8 mm LDPE system is shown in Fig. 4. As U_l increased the bed expansion and the total volume of the fluidized bed increased. Increase in liquid flow rate increased the bed porosity and the volume fraction of liquid in the fluidized bed and the gas holdup decreased for a fixed U_g. A similar trend was observed for glycerol and CMC systems.

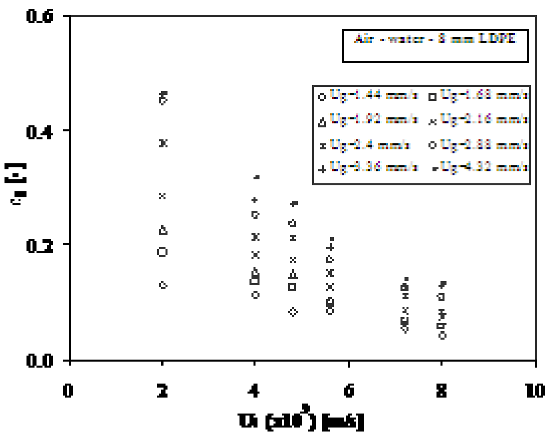

Fig. 4 Effect of liquid velocity (U_l) on gas holdup, ε_g in air-water-8 mm LDPE system

Since the bubble rising velocity increased as the liquid velocity increased and the residence time of bubbles decreased, the increase of liquid velocity resulted in the decrease of the gas holdup. The gas holdup on U_g did not change with liquid velocity. A similar trend was reported by Fan et al [2], Song *et al.* [9] and Miura *et al.* [10]. *Song et al.* [9] also found a monotonous decrease in gas holdup with liquid velocity at low gas flow rates. They found, however, that at high gas flow rates, an increase in liquid velocity beyond a critical value of U_l caused an increase in gas holdup due to the enhancement of bubble break-up and simultaneous reduction of bubble coalescence.

3.3 Effect of superficial liquid velocity (U_l) on liquid holdup (ε_l)

The effect of U_l on liquid holdup for air - water - 8 mm LDPE system is shown in Fig 5. With an increase in U_l, the bed expansion increased and the total volume of the fluidized bed also increased. A rise in liquid flow rate increased the bed porosity and the volume fraction of liquid (ε_l) in the fluidized bed. The same trend was observed for glycerol and CMC systems.

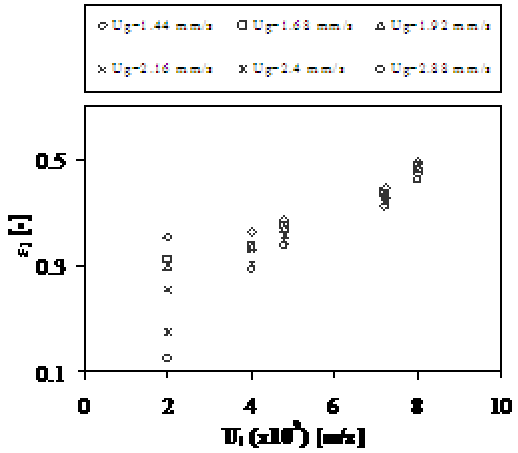

Fig. 5 Effect of liquid velocity (U_l) on liquid holdup, ε_l in air - water- 8 mm LDPE system

3.4 Effect of diameter of particle (d_p) on gas holdup (ε_g)

The effect of U_g on gas holdup for 8, 6 mm PP and 8, 6 mm LDPE is given in Fig. 6. As the particle diameter increased, the gas holdup also increased. This was due to the increased porosity in the bed, which increased the space for the gas. The gas holdup was high for 8 mm PP particle than for 6 mm PP as the porosity is high for the 8 mm PP bed. The Figure 4.27 also shows that the gas holdup for LDPE (density 940 kg/m^3) is lower when compared to PP (density 830 kg/m^3). Fan et al [1] and Saberian-Broudjenni *et al.* [12] reported that the gas holdup always depended on the nature of the liquid and on the size and the density of the particles. Miura et al. (2001) reported that for a given liquid velocity, ε_g in the three-phase fluidized bed (with water and CMC) as liquids increased with increasing U_g and d_p

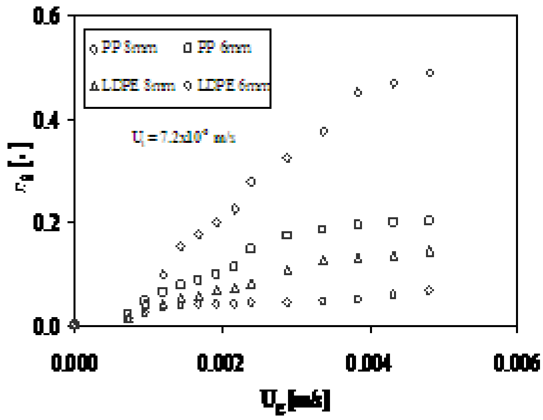

Fig. 6 Effect of particle diameter (d_p) on gas holdup (ε_g) in air - water - solid system

3.5 Effects of glycerol and CMC concentration on gas holdup (ε_g)

The effects of glycerol and CMC concentrations on gas holdup with U_g are shown in Fig. 7 and 8 respectively. It was observed from the Fig. 7 that the gas holdup increased as the glycerol concentration increased. From the Fig. 8, it was observed that the gas holdup increased as the CMC concentration increased from 0.1% to 0.3% gradually. The gas holdup was only slightly affected by the rheological properties of the liquids. A similar trend was reported by Miura *et al.* [11].

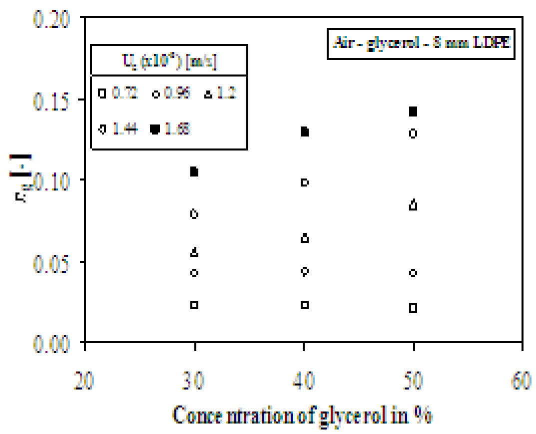

Fig. 7 Effect of glycerol concentration on gas holdup (ε_g) at U_l = 7.2x10^{-3} m/s

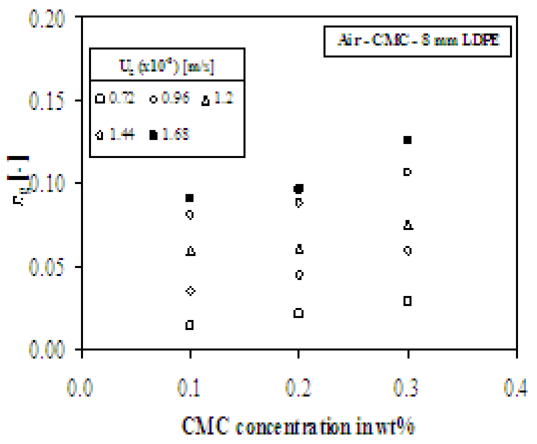

Fig. 8 Effect of CMC concentration on gas holdup (ε_g) at $U_l = 7.2x10^{-3}$ m/s

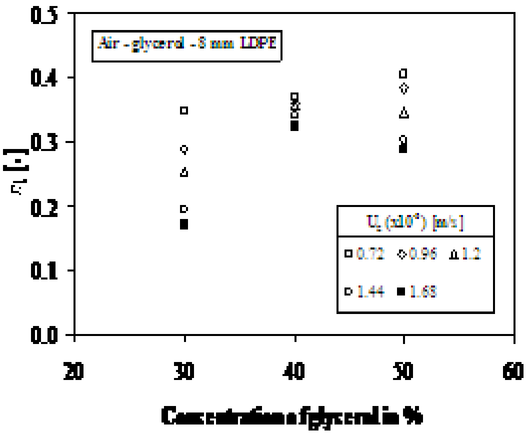

Fig. 9 Effect of glycerol concentration on liquid holdup (ε_l) at $U_l = 7.2x10^{-3}$ m/s

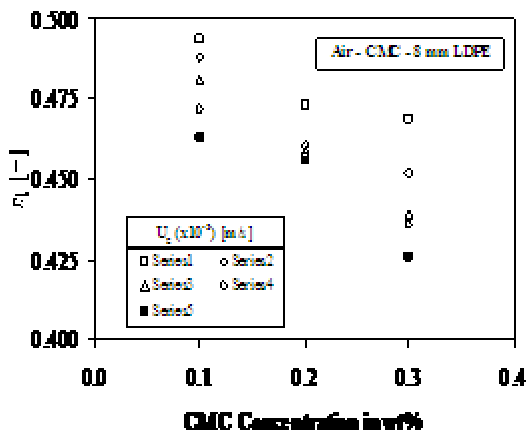

Fig. 10 Effect of CMC concentration on liquid holdup (ε_l) at $U_l = 7.2x10^{-3}$ m/s

3.6 *Effects of glycerol and CMC concentration on liquid holdup (ε_l)*

The effects of glycerol and CMC concentration on liquid holdup with U_g are shown in Fig. 9 and 10 respectively. It was observed from the Fig. 9 that the liquid holdup increased as the glycerol concentration increased from 30% to 40% and then decreased as the concentration increased to 50%. From the Fig. 10, it was observed that the liquid holdup decreased as the CMC concentration increased from 0.1 to 0.3%.

3.7 Empirical correlation of phase holdups for the fluidized bed

The various parameters affecting phase holdups are grouped and empirical correlations are developed as follows.

3.7.1 Newtonian system

Gas holdup

$$\in_g = 0.777\, d_p^{0.007} U_g^{1.256} U_l^{-0.796} R^{0.264} D_c^{2.249} \sigma_l^{-2.783}$$

R^2=0.89 (5)

Porosity

$$\in = 1.196\, d_p^{-0.293} U_g^{0.188} U_l^{0.02} R^{-0.176} D_c^{0.218} \sigma_l^{0.348}$$

R^2=0.92 (6)

3.7.2 Non-Newtonian System

Gas holdup

$$\in_g = 1.439\, d_p^{0.096} U_g^{1.233} U_l^{-0.149} R^{-0.39} \sigma_l^{-0.702}$$

R^2=0.8 (7)

Porosity

$$\in = 0..829\, d_p^{-0.373} U_g^{0.284} U_l^{0.115} R^{-0.248} \sigma_l^{0.155}$$

R^2=0.9 (8)

The liquid holdup, ε_l can be calculated by substituting the values of ε_g and ε in the equation 4.32.

$$\in_l = 1 - \in_s - \in_g \quad (9)$$

The comparison between experimental and predicted values of gas holdup and bed porosity for (6 mm PP particle) Newtonian and non-Newtonian systems using empirical correlation is presented in Figures 11 & 12 and in Fig. 13 and 14 respectively. The proposed equations predicted the gas holdup and bed porosity with an average RMS error of 23 and 3.5% for Newtonian system and 21 and 3.2% for non-Newtonian system respectively

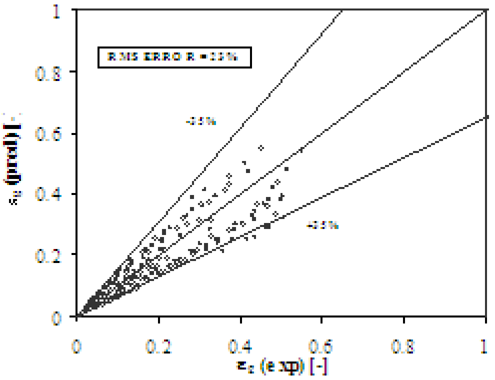

Fig. 11 Parity diagram for ε_g estimated from empirical equation 4.28 for Newtonian system

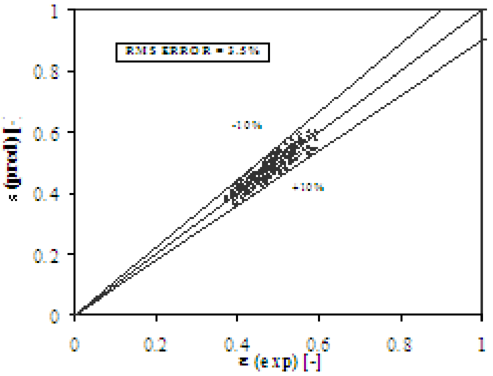

Fig. 12 Parity diagram for ε estimated from empirical equation 4.29 for Newtonian system

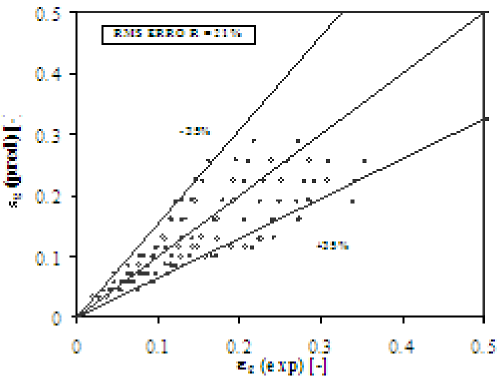

Fig. 13 Parity diagram for ε_g estimated from empirical equation 4.30 for non-Newtonian system

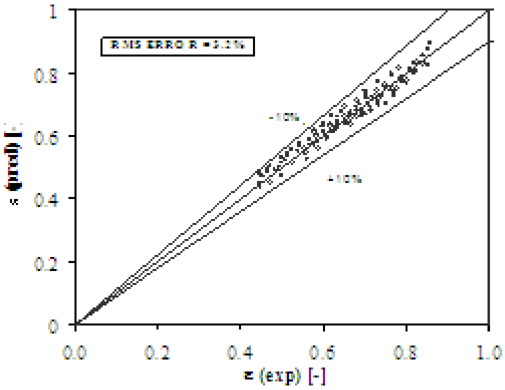

Fig. 14 Parity diagram for ε estimated from empirical equation 4.31 for non-Newtonian system

3.8 Modified drift flux model for holdups and bed porosity

Saberian-Broudjenni *et al.* [12] introduced the concept of drift flux to predict the liquid holdup and the bed porosity in a three-phase fluidized bed reactor. Buffiere and Moletta [8] used the modified gas drift flux to predict the liquid holdup and the bed porosity in a three-phase inverse fluidized bed reactor. The gas drift flux is defined as the difference between the interstitial gas velocity U_g/ε_g and the average interstitial velocity of the gas-liquid mixture $(U_g\text{-}U_l)/(\varepsilon_g+\varepsilon_l)$. In the case of a countercurrent of gas and liquid, the gas drift flux is expressed as follows:

$$j_{gl}' = \left(\frac{1}{\in_l + \in_g}\right)(U_g \in_l + U_l \in_g) \tag{10}$$

Buffiere and Moletta [8] noted that this expression allowed to make j_{gl}' almost independent from the type of particles used and proposed the following expression for a three phase up flow fluidized bed:

$$j_{gl}' = 0.017\,(\rho_l U_g^{2})^{0.45}$$

$$(= 0.38\,U_g^{0.9} \text{ for water}) \tag{11}$$

The experimental values of j_{gl}' were independent from the particles and the type of reactor and could be very well correlated with the gas velocity U_g as a power law (7).

$$j_{gl}' = 0.783\,U_g^{0.96} \tag{12}$$

In the present study, the experimental values of j_{gl}' were also independent from the particles and the type of reactor and could be very well correlated with the gas velocity U_g as a power law (8) as shown in Fig 15 for non-Newtonian system.

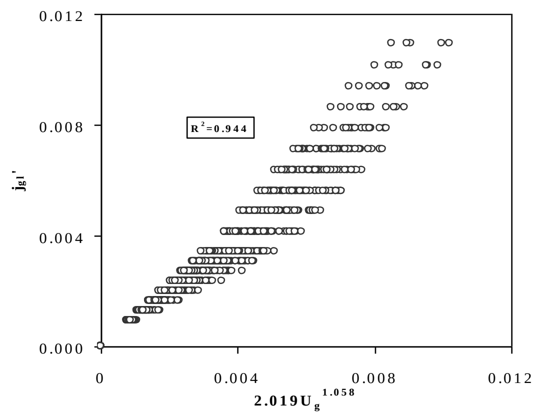

Fig. 15 Modification on gas drift flux, j_{gl}' - non-Newtonian system

$$j_{gl}' = 2.019\,U_g^{1.058} \qquad R^2=0.94 \qquad (13)$$

$$\epsilon_g = \frac{H - H_0}{H} \qquad (14)$$

From relation 14, the solid holdup in the bed can be expressed as a function of the solid fraction in the static bed ε_{so}. The ratio of static bed height to static column height H_0/H_c and the gas holdup are as follows:

$$\epsilon_s = \epsilon_{s0}\frac{H_0}{H} = \epsilon_{s0}\frac{H_0}{H_C}\frac{H_C}{H} = \epsilon_{s0}\frac{H_0}{H_C}(1-\epsilon_g) \qquad (15)$$

The liquid holdup and porosity are:

$$\epsilon_l = \frac{1}{(1/1-\epsilon_{s0}(H_0/H_C)) - (U_g - j_{gl}')/(U_l - j_{gl}')} \qquad (16)$$

$$\epsilon = \epsilon_l + \epsilon_g$$

$$= \frac{(U_l - U_g)/(U_l - j_{gl}')}{(1/1-\epsilon_{s0}(H_0/H_C)) - (U_g - j_{gl}')/(U_l - j_{gl}')} \qquad (17)$$

The proposed equation for non Newtonian system (6 mm PP particle) using modified drift flux model predicted the bed porosity with an average RMS error of 0.8% as shown in Fig 16. This method cannot be used to predict the gas holdup (as the difference between ε and ε_l are of the order of values of ε_g). No globally satisfactory correlation was found to reflect the experimental data. So direct correlation for the gas holdup is derived under the form

$$\epsilon_g = aU_g^{\,b}R^{\,c} \qquad (18)$$

Non-Newtonian system

$$\epsilon_g = 153.555\,U_g^{1.665}R^{-0.437}$$

R^2=0.85 (19)

The proposed equations predicted the gas holdup with an average RMS error of 20% for non-Newtonian system (6 mm PP particle) as shown in Fig. 17.

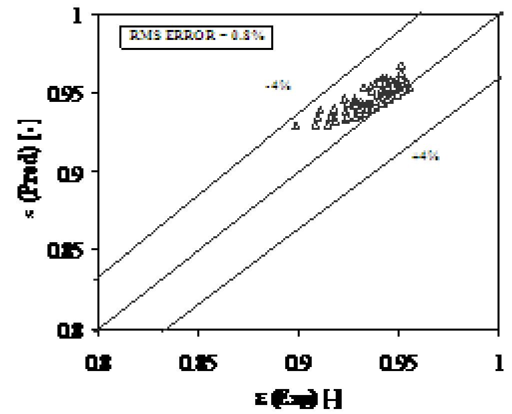

Fig 16 Parity diagram for ε estimated from modified drift flux model eqn. 12 for non-Newtonian system

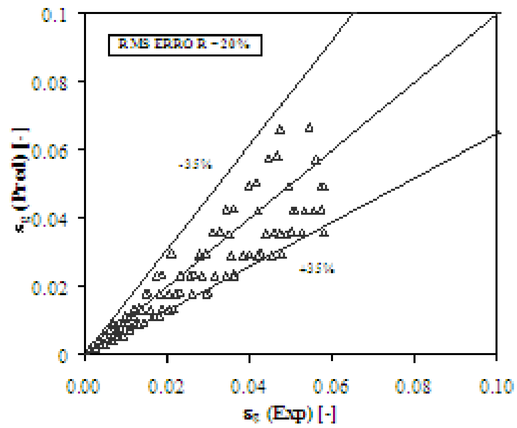

Fig. 17 Parity diagram for ε_g estimated from eqn. 14 for non-Newtonian system

5. CONCLUSION

The phase holdups of three-phase inverse fluidized bed reactor were experimentally investigated. The solid holdup and liquid holdup decreased with an increase in U_g. The gas holdup and porosity increased as U_g increased. The liquid holdup increased with increase in U_l and the gas holdup increased with increase in particle diameter. With the variation in concentration of glycerol or CMC, it was found that the gas holdup increased and a decreasing trend was observed for liquid holdup for the range of U_g used. The proposed empirical equations and modified drift flux model for bed porosity and phase holdups predicted the experimental data well for Newtonian and non-Newtonian systems.

LIST OF SYMBOLS

a,b,c	constants
j_{gl}'	modified drift flux, (ms^{-1})
H	total bed height, (m)
H_c	column height, (m)
H_0	static height, (m)
K	flow consistency index in power law model, ($Pa.s^n$)
M_s	weight of dry particles, (kg)
n	flow index in power law model
R	relative density difference, $(\rho_l-\rho_s)/\rho_l$
U_g	superficial gas velocity, (ms^{-1})
U_l	superficial liquid velocity, (ms^{-1})
$\in$	porosity
ε_s	solid holdup
ε_{so}	solid holdup in the static bed
ε_l	liquid holdup
ε_g	gas holdup
ρ_s	density of solid particles, (kgm^{-3})
ρ_l	density of liquid, (kgm^{-3})

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