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Hydrodynamic Studies on Three-Phase Combined (Internal & External) Loop Airlift Fluidized Bed Reactor Using Newtonian and non-Newtonian Liquids: Minimum Fluidization Velocity and Liquid Holdup

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Hydrodynamic Studies on Three-Phase Combined (Internal & External) Loop Airlift Fluidized Bed Reactor Using Newtonian and non-Newtonian Liquids: Minimum Fluidization Velocity and Liquid Holdup*

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Abstract

A novel combined airlift loop fluidized bed reactor was proposed in this work. The internal and external loops were combined and the hydrodynamic parameters like minimum fluidization velocity and liquid holdup were studied for Newtonian and non-Newtonian fluids. Studies were conducted using Newtonian fluids of water, n-butanol, 60% and 80% glycerol and non - Newtonian fluids such as 0.25%, 0.6% and 1.0% Carboxy Methyl Cellulose (CMC) aqueous solutions were employed in the liquid phases. Spheres, Bearl saddle and Raschig ring with different sizes were used as solid phase. The experimental results indicated that the increase in particle size and sphericity increased minimum fluidization velocity whereas increase in superficial gas velocity decreased minimum fluidization velocity. In addition, the liquid holdup increased with increase in particle size and superficial liquid velocity. Furthermore, an increase in superficial gas velocity decreased the liquid holdup for Newtonian and non-Newtonian systems. Two separate correlations were developed to predict the minimum fluidization velocity and liquid holdup based on the experimental results for both Newtonian and non-Newtonian liquids for a wide range of operating conditions. The capability of the proposed correlation for minimum fluidization velocity and liquid holdup

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was examined and reasonable agreement between predicted and experimental results of Newtonian and non-Newtonian liquids suggested the applicability of the proposed correlations.

KEYWORDS: combined loop airlift fluidized bed reactor, minimum fluidization velocity, liquid holdup

INTRODUCTION

In general, the reactors used in industries are broadly classified into stirred vessels or column reactors. In stirred vessels, agitation is accomplished by rotating impellers and it does provide excellent mixing, high mass and heat transfer coefficients. Drawbacks of the stirred vessels are maintenance cost and high energy consumption. Apart from this, these vessels are found to be not suitable for biological application because, some level of contamination due to mechanical moving parts such as shafts and impellers. Column reactors are widely used in fermentation and other biological process applications because of good mixing. In general, the column reactors are classified into bubble column and airlift reactors. The airlift differs from the bubble column by the introduction of inner draft tubes or outer tubes, which improves liquid circulation and enhances the liquid mixing without addition of external energy. The bubble column and airlift reactors are pneumatically agitated where gas-liquid contact is important. In these two pneumatically agitated column reactors, bubble column is a simple tower gas sparged at the bottom of the column with no recirculation of liquids whereas airlift reactors are divided into two sections, the riser and down comer, depending on the arrangement of down comer, classified into the internal-loop or external loop airlift reactors. In the internal loop reactors, the riser and the down comer are present in the same column whereas in the external-loop reactors riser and down comer are separate tubes put up side by side and connected at the top and bottom. In airlift reactor, gas and liquid are passed at the bottom of the riser, which results in a difference in static pressure in the two sections, which leads to the circulation of the liquid.

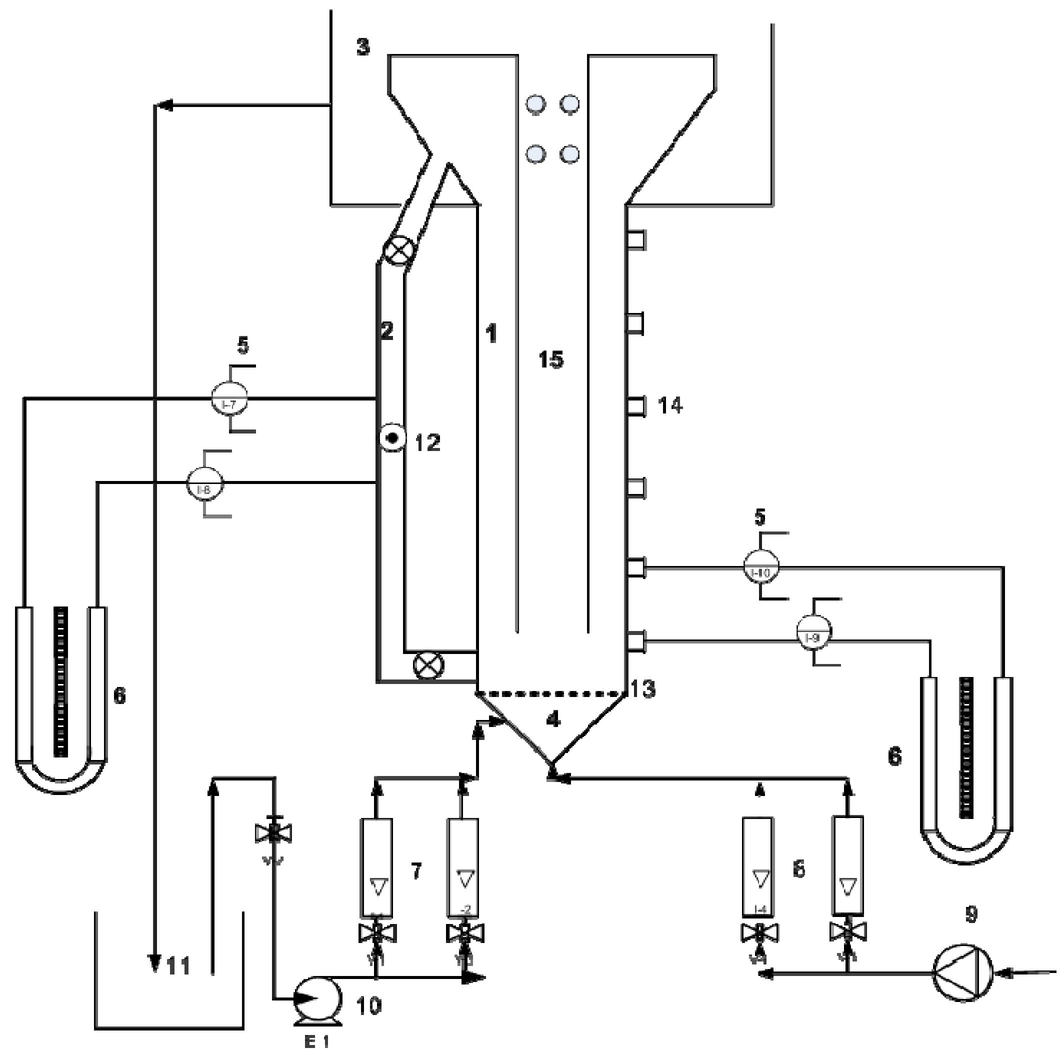
In three-phase operation, solid particles are added as catalyst or inert material to increase the mixing characteristics of liquid and gas phase reactants. The complexity increases when a solid is added to the system either as inert or catalyst. Recently, live cells have been immobilized on inert packing materials for wastewater treatment and in other biochemical applications, where mass transfer phenomena were accompanied with biochemical reactions. In three-phase fluidization, a bed of solid particles present in the column is suspended by an upward co-current flow of both gas and liquid phases whereas in three phase airlift fluidized bed reactor, a bed of solid particles present in the riser is suspended by an upward co-current flow of both gas and liquid phase. Recently three phase airlift fluidized bed reactors have been widely used in chemical industries, environmental remediation technologies, biochemical industries, etc., because of their significant advantages such as simple structure, no moving parts, lower investment, lower gas requirement for complete solid fluidization, fine gas dispersion, high mixing and mass transfer performance, rapid and uniform distribution of the reaction components, etc (Korpijarvi et al.1999, Tobajas et al.1999, Wen et al.2005, Jin et al.2006). Numerous hydrodynamic studies have been carried out using either by using internal loop or by using external loop separately. To our knowledge no work has been carried out to study the influence of fundamental and operating variable on hydrodynamic parameters of airlift fluidized bed using combined loop configuration. Since combined loop reactor has more circulation rate and mixing characteristics, in this paper an attempt has been made to study the influence of superficial gas and liquid velocities, particle diameter, sphericity, physical and rheological properties of Newtonian and non-Newtonian liquids on minimum fluidization velocity and liquid holdup. Apart from this, unified correlations have been developed to determine the minimum fluidization velocity and liquid holdup from the fundamental and operating variable for both Newtonian and non-Newtonian liquids.

EXPERIMENTAL SETUP AND PROCEDURE

Combined loop airlift fluidized bed reactor is constructed by mounting a draft tube inside a bubble column, and an external down comer connected at the top and bottom of the column. This reactor is divided into three zones; one is gas sparged riser, an unsparged down comer inside the column and external down comer, a degassing zone at the top of the column, which removes the gas bubbles from the riser. The compressed gas is sparged into the riser zone creates a density difference between the riser and down comers induces liquid circulation. All the experiments have been carried out in a Perspex column, which has a diameter of 0.15 m, 1.63 m height, with a flat bottom and draft tube of 1.54 m in height with 0.084 m diameter. The details of experimental setup are shown in Fig.1.The vertical clearance between draft tube and gas distributor was 0.09 m. Air was sparged into the draft tube through sparger which was 0.08 m in diameter with holes of 0.0008 m each located slightly below the perforated plate. The gas flow rate was measured by calibrated rotameters with an accuracy of ± 2 %. The density of the liquids was measured with a specific gravity method and the rheological properties of non-Newtonian liquids were measured by using

Brookfiled Rheometer. Superficial liquid velocities were varied from 0.001 m/s to 0.12 m/s. Superficial gas velocities were varied from 0.0001 – 0.0056 m/s. Minimum fluidization velocity was determined by visual measurement (Koide et al.1983 and Zhang et al.1995). In visual observation method, the velocity at which the first particle move upwards is taken as minimum fluidization velocity. The minimum fluidization velocity results obtained by visual observation of spherical, Raschig rings and Bearl saddles were compared with the results obtained from pressure drop method. Almost 80 % of results obtained by visual observation method were found to be good fit with the results obtained by pressure drop method with less than ± 3 % error and maximum deviation was found to be ± 5 % for air -water system using Raschig ring particles. The detail of measurement of minimum fluidization velocity by using pressure drop method is given elsewhere (Sivakumar and Senthilkumar, 2010). Liquid holdup was measured by displacement method (Nacef et al.1992 and Miura et al .2001) in which the gas and liquid flow attained a steady state condition and both flow rates were suddenly stopped by closing the valves simultaneously and then the column was filled with liquid. The difference between the void volume and volume of liquid used to fill the column gave the liquid holdup. The details of measuring technique are given in our earlier article (Sivakumar et al. 2010).

In the present work, water, n- butanol and various concentrations of glycerol were used as Newtonian fluids and different concentrations of Carboxy Methyl Cellulose (CMC) were used as non-Newtonian fluids. Different diameters of spheres, Bearl saddle and Raschig ring were used as solid phases. All the experiments were carried out in an atmospheric temperature with oil free compressed air as gas phase. A minimum of 3-5 readings were taken and the average value was used for calculations and the error was found to be less than ± 3 %. The properties of solid particles and liquids used in the present study are given in Tables 1 and 2.



1. Riser. 2. Down comer, 3. Gas liquid Separator, 4. Calming section, 5.Gas liquid Separator for Manometer. 6. Manometer. 7. Rota meter for Liquid flow rate. 8.Reta meter for Gas flow rate. 9. Gas inlet valve. 10. Liquid circulating pump. 11. Liquid Storage tank, 12. Orifice, 13. Supporting Screen, 14.Manometer Tapings, 15.Draft tube.

Fig.1 Schematic diagram of experimental setup

Table-1 Properties of solids used for the present study

Particle Description	Size, dp, m	Density,kg/m3	Particle Sphericity
Particle 1 (Sphere-Porcelian)	0.001	2478	1
Particle 2 (Sphere-Porcelian)	0.002	2478	1
Particle 3 (Sphere-Porcelian)	0.003	2478	1
Particle 4 (Sphere-Porcelian)	0.004	2478	1
Particle 5 (Sphere-Porcelian)	0.005	2478	1
Particle 6 (Sphere-Porcelian)	0.006	2478	1
Particle 7 (Sphere-Porcelian)	0.01036	2478	1
Particle 8 (Bearl saddles)	0.0115	2456	0.33
Particle 9 (Raschig rings)	0.01366	2083	0.58

Table-2 Properties of liquids used for the present study

Type of liquids	Density of liquids,(ρ_L) kg/m3	Surface tension (σ_L) N/m	Viscosity	
			K kg $m^{-1}s^{n-2}$	n
Water	1000	0.0700	0.00083	1
n-Butanol	1008	0.0350	0.00098	1
80 % Glycerol	1180	0.0650	0.030	1
60 % Glycerol	1155	0.0660	0.0185	1
0.25 % CMC	1026	0.0730	0.0197	0.87
0.6 % CMC	1020	0.0735	0.0308	0.86
1.0 % CMC	1017	0.0740	0.0565	0.85

RESULTS AND DISCUSSIONS

EFFECT OF GAS VELOCITY AND PARTICLE SIZE ON MINIMUM FLUIDIZATION VELOCITY

The results of influence of superficial gas velocity and particle diameter on minimum fluidization velocity are shown in Fig.2. They show that an increase in superficial gas velocity decrease the minimum fluidization velocity because increasing gas flow creates high density difference between riser and down comer which leads to more liquid circulation and hence decrease minimum fluidization velocity (Sivakumar et al. 2010). It is also observed that minimum fluidization velocity increases with increasing particle diameter (Petrovic et al.1992).

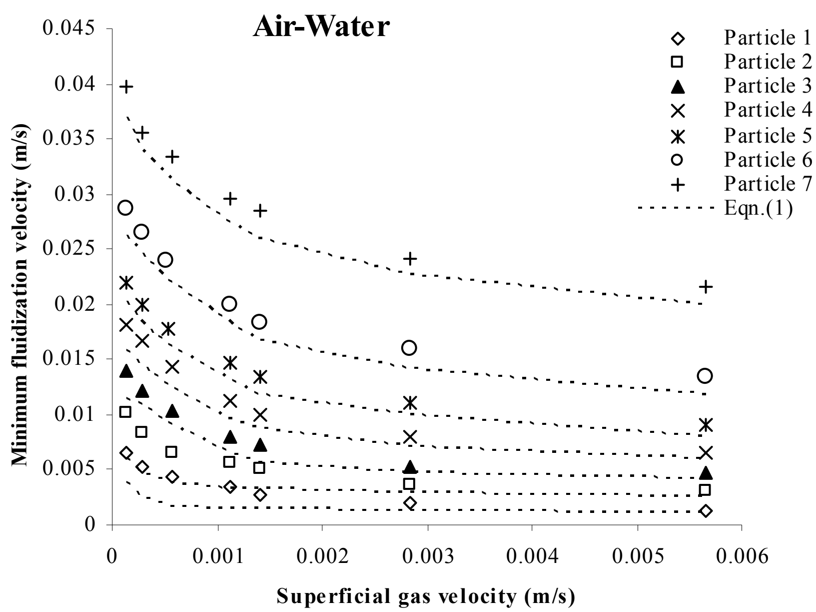


Fig.2 Effect of particle diameter on Minimum fluidization velocity

EFFECT OF SPHERICITY OF PARTICLE ON MINIMUM FLUIDIZATION VELOCITY

Different sphericity of particles was used to find out the effect of particle shape on minimum fluidization velocity. From Fig.3 it is observed that when the particle sphericity increases minimum fluidization velocity will also increased. This is due to the fact that increase in sphericity decreases bed voidage and hence the minimum fluidization velocity increases. The similar trend of result was also observed for air - 60 % glycerol in Fig.4 and air - 0.6 % CMC shown in Fig.5.

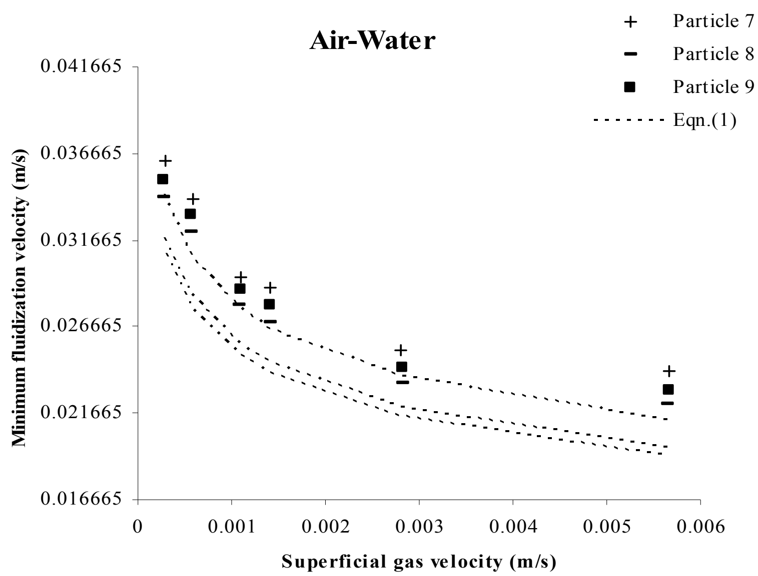


Fig.3 Effect of particle sphericity on Minimum fluidization velocity

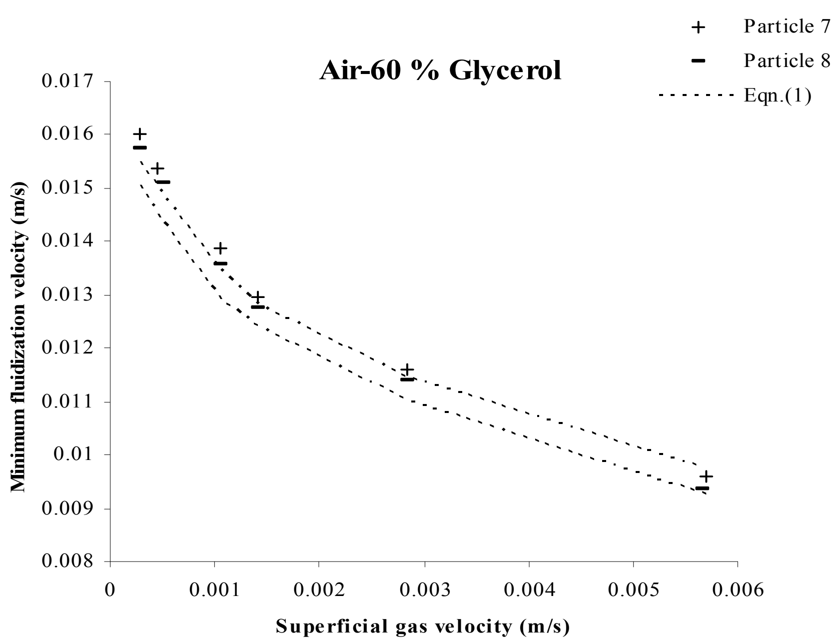


Fig.4 Effect of particle sphericity on Minimum fluidization velocity

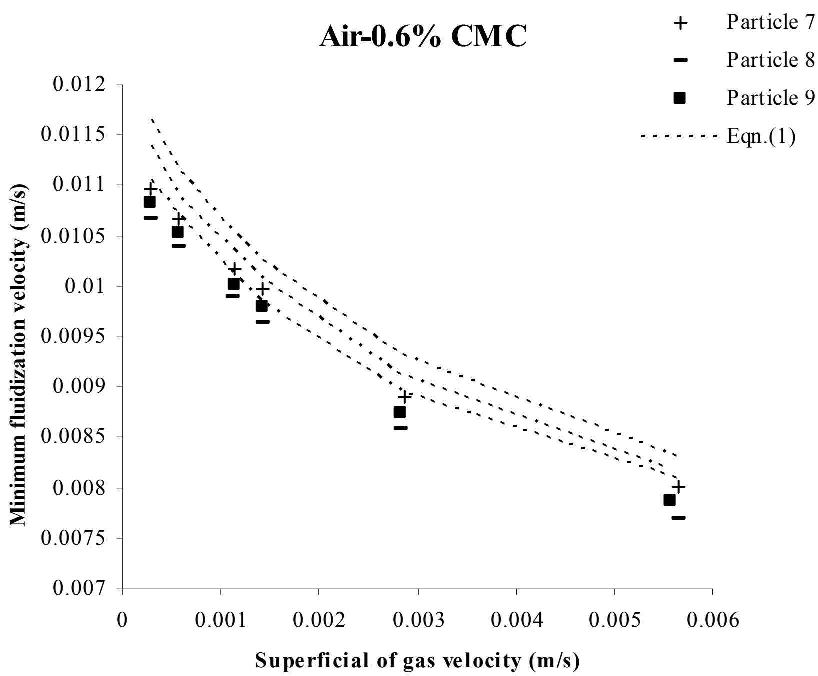


Fig.5 Effect of particle sphericity on Minimum fluidization velocity

EFFECT OF PHYSICAL PROPERTIES OF LIQUIDS ON MINIMUM FLUIDIZATION VELOCITY

Fig.6 shows the effect of physical properties of Newtonian liquids on minimum fluidization velocity. From the Figure 6, it is observed that when the viscosity of liquids increases the minimum fluidization velocity will decrease and it is due to high shear forces extended by more viscous liquids on the particles. From the result, it is observed that the surface tension of liquids does not have significant influence on the minimum fluidization velocity. Fig.7 shows the effect of rheological properties of non-Newtonian liquids on minimum fluidization velocity. Three concentrations 0.25 %, 0.6 % and 1.0 % of CMC were used as liquid phase. It is observed that an increase in fluid consistency index will decrease the minimum fluidization velocity (Sivakumar et al. 2010).

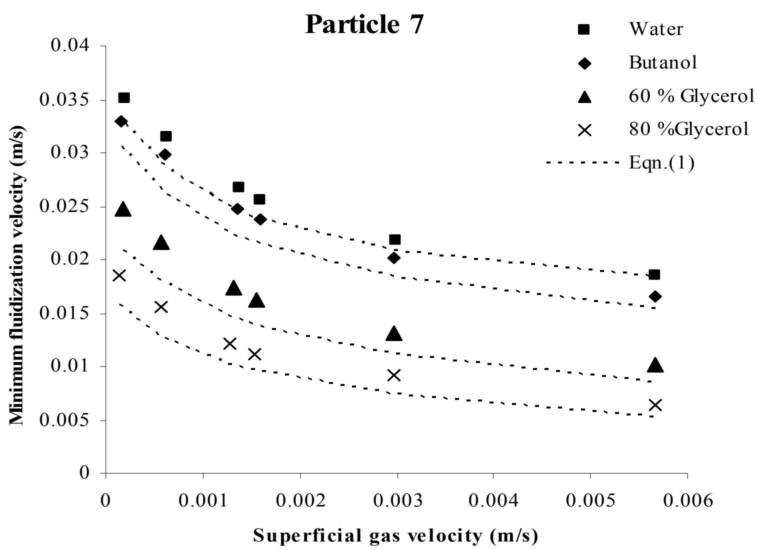


Fig.6 Effect of physical properties on Minimum fluidization velocity

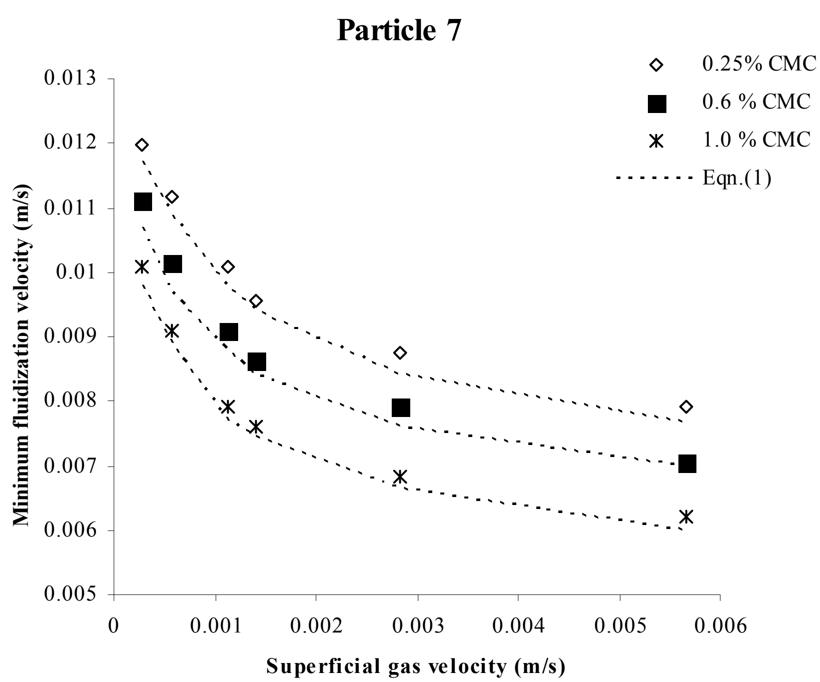


Fig.7 Effect of rheological properties on Minimum fluidization velocity

EFFECT OF PARTICLE DIAMETER AND SPHERICITY ON LIQUID HOLDUP

Effect of particle diameter on liquid holdup for Air-Water system is shown in Fig.8 which is drawn between superficial liquid velocity and liquid holdup. From this graph it is observed that an increase in particle diameter retains more liquid in the column leading to increase in liquid holdup. Fig.9 shows the effect of sphericity of particles on liquid holdup for the superficial gas velocity 0.000283 m/s. It is observed that increase in particle sphericity does not have any significant influence on liquid holdup.The same trend was also observed for air - 0.6 % CMC system, which is shown in Fig.10.

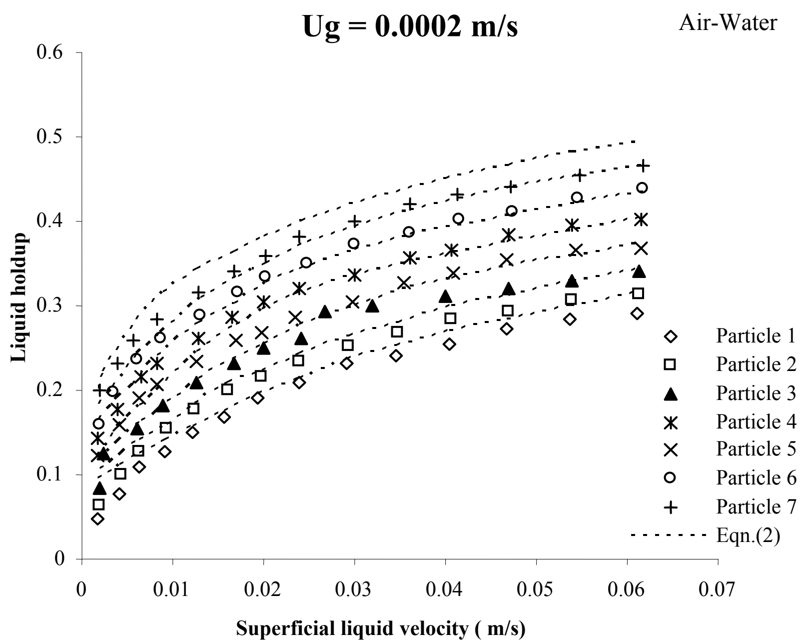


Fig.8 Effect of particle diameter on liquid holdup

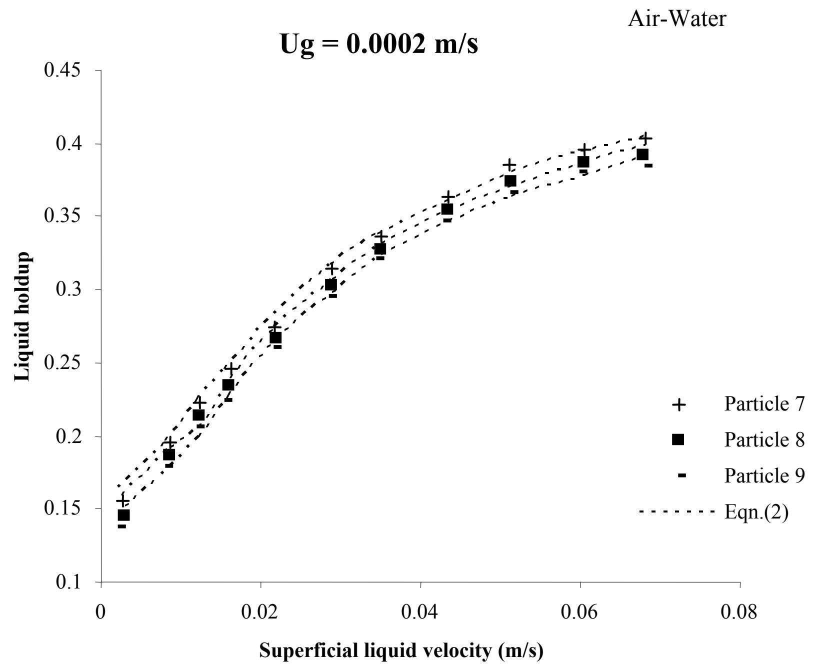


Fig.9 Effect of particle sphericity on liquid holdup

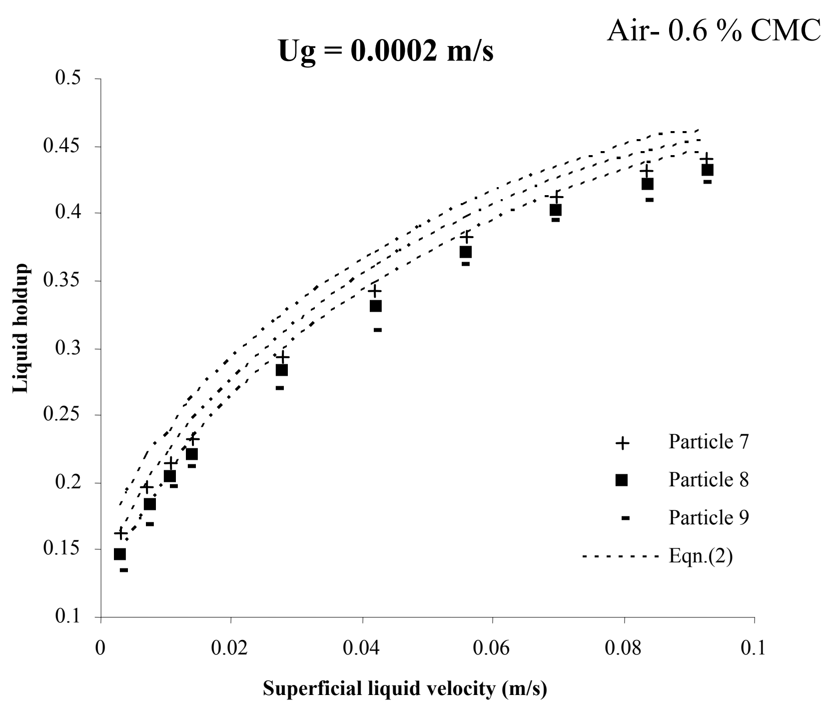


Fig.10 Effect of particle sphericity on liquid holdup

Effect of phase flow rates on liquid holdup

Figs.11 & 12 shows the effect of superficial gas and liquid velocity on liquid holdup for Air - Water system. It is observed that an increase in superficial liquid velocity increases the velocity of the gas bubble and therefore the liquid holdup increases. Increase in superficial gas velocity increases the fraction of gas bubbles in the column and hence liquid holdup decreases (Sivakumar et al.2010). The same trend was also observed for air- n- butanol, air - 80 % glycerol systems and air -1.0 % CMC which is shown in Figs. 13, 14 and 15.

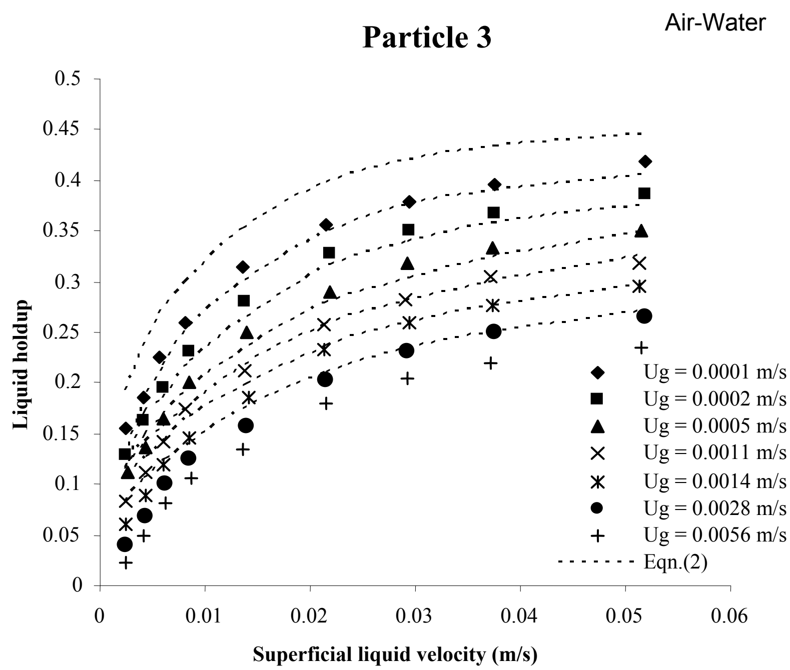


Fig.11 Effect of superficial gas velocity on liquid holdup

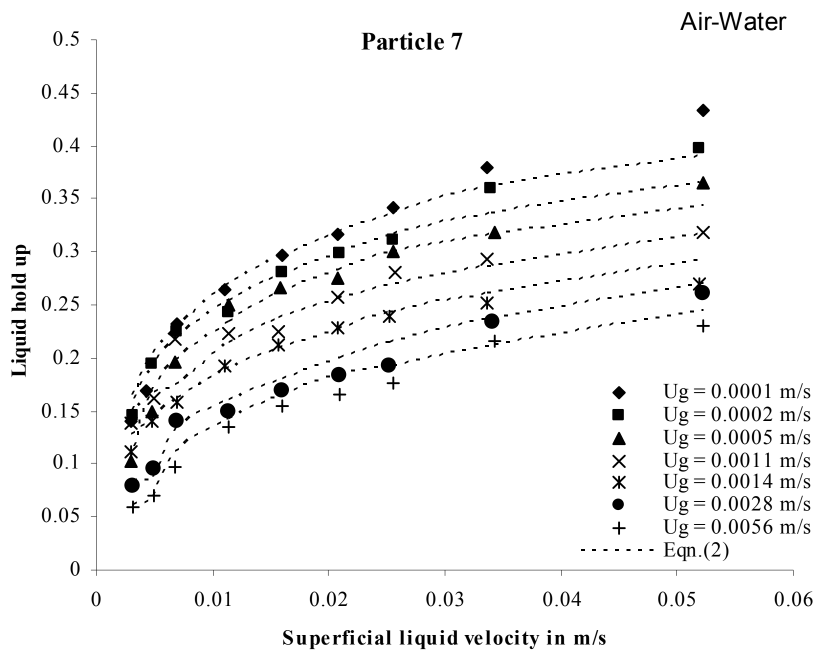


Fig.12 Effect of superficial gas velocity on liquid holdup

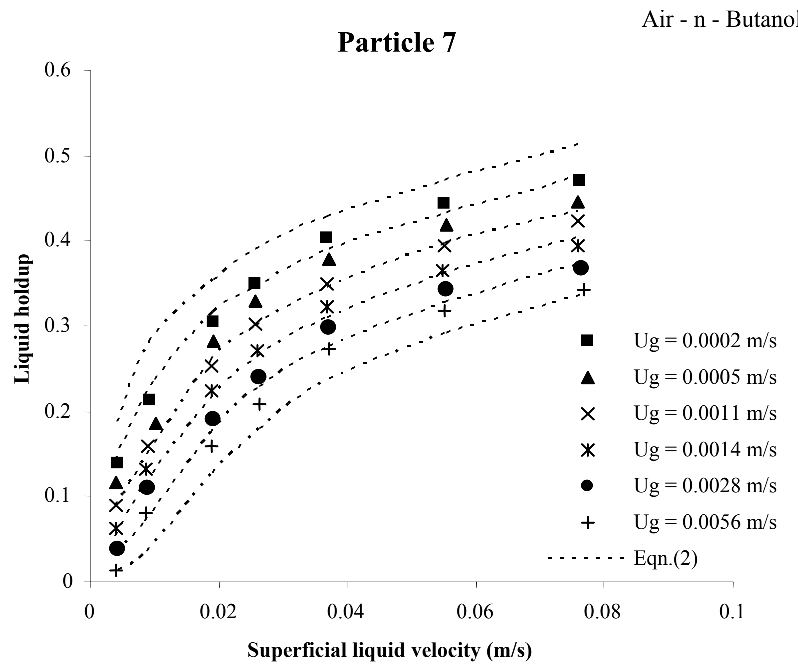


Fig.13 Effect of superficial gas velocity on liquid holdup

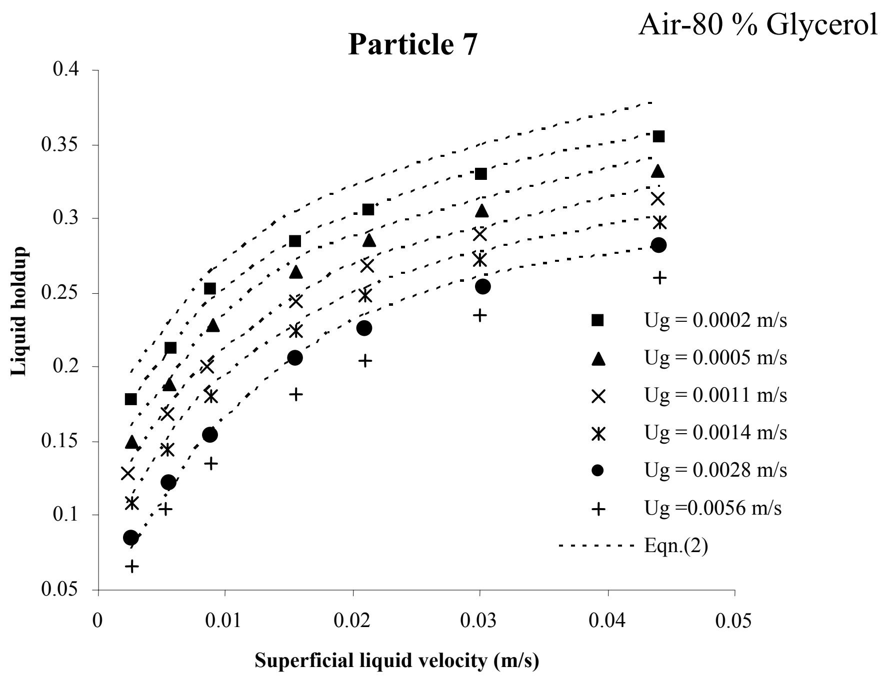


Fig.14 Effect of superficial gas velocity on liquid holdup

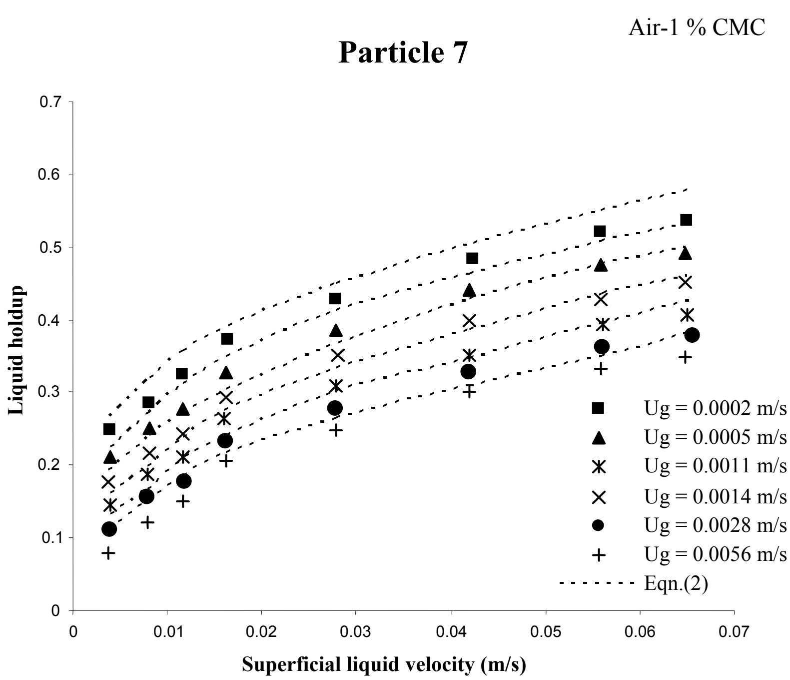


Fig.15 Effect of superficial gas velocity on liquid holdup

EFFECT OF PHYSICAL PROPERTIES OF LIQUIDS ON LIQUID HOLDUP

Fig.16 shows the effect of physical properties of air-water, air-n - butanol, Air - 60 % glycerol and air – 80 % glycerol systems on liquid holdup for the superficial gas velocity 0.0014 m/s. The figure shows that an increase in viscosity of liquid increases the liquid holdup and it was also observed that decreasing surface tension of liquid increases liquid holdup. Fig.17 shows the effect of fluid consistency index for 0.25, 0.6 and 1.0 % CMC liquids. From the experimental results it is observed that an increase in fluid consistency index of liquid increases the liquid holdup.

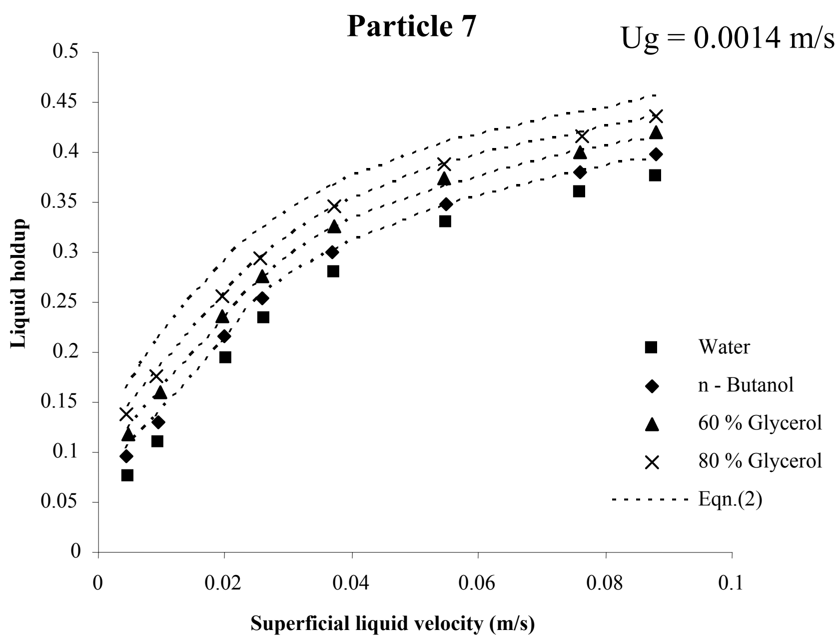


Fig.16 Effect of physical properties on liquid holdup

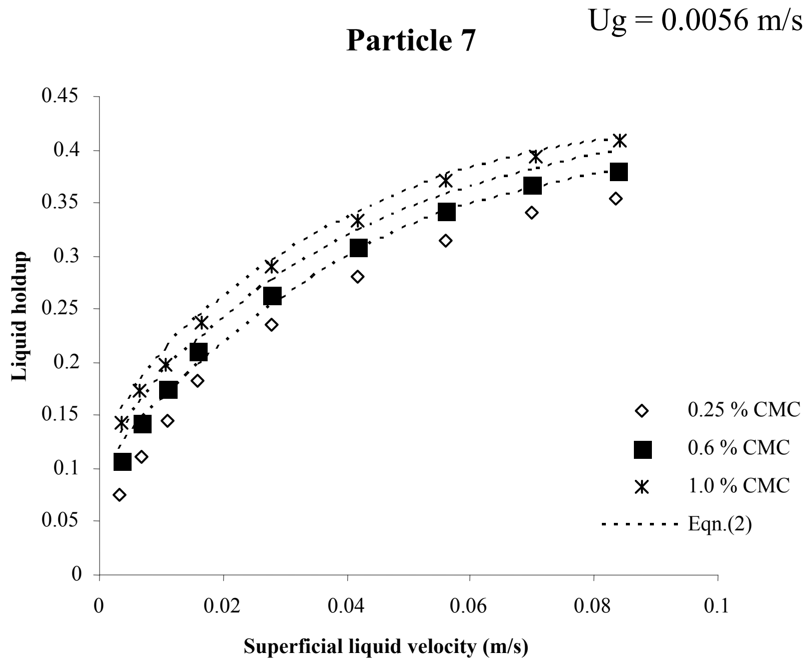


Fig.17 Effect of rheological properties on liquid holdup

CORRELATION

There is no literature evidence available for the combined loop reactor. From the analysis of literature f it is found that none of the authors developed correlation to predict minimum fluidization velocity and liquid holdup for a wide range of operating variables using Newtonian and non-Newtonian liquids. From the experimental data (172 for minimum fluidization velocity and 1580 for liquid holdup), two separate dimensionless correlations were developed to predict the minimum fluidization velocity and liquid holdup,

$$N_{\mathrm{Re}mf} = 0.98(Fr_G)^{-0.08}\left(\frac{\rho_P-\rho_L}{\rho_L}\right)^{0.39}\left(\frac{d_P}{d_C}\right)^{0.4}\left(\frac{\sigma_L}{\sigma_W}\right)^{0.15}(\phi)^{0.01}\left(A_r\right)_m^{\ 0.58} \quad \text{-- (1)}$$

$$\varepsilon_L = 0.98\left(Fr_G\right)^{-0.03}\left(Fr_L\right)^{0.17}\left(\frac{d_P}{d_C}\right)^{0.3}\left(\frac{\sigma_L}{\sigma_W}\right)^{-0.15}\left(\frac{\rho_P-\rho_L}{\rho_L}\right)^{0.73} Mo^{\,0.005} \quad \text{-- (2)}$$

The comparisons of our experimental and calculated values of the minimum fluidization velocity for Newtonian and non-Newtonian liquids are shown in Figs. 18 and 19. They show good agreement between the experimental and calculated minimum fluidization velocity with the average deviation of 15 % for 174 data. Figs. 20 and 21 show the comparison between the experimental and calculated values of liquid holdup for Newtonian and non-Newtonian liquids. Good agreement with experimental data was obtained with average deviation of 12%.

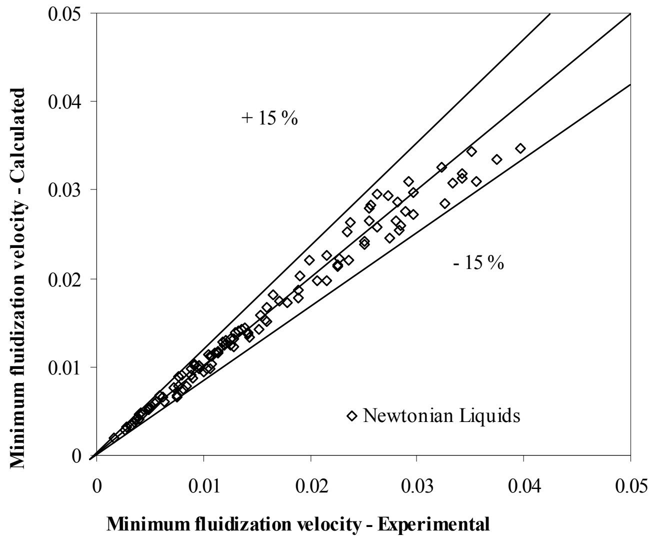


Fig.18 Comparison between the Experimental and Calculated Values of Minimum Fluidization Velocity for Newtonian Liquids

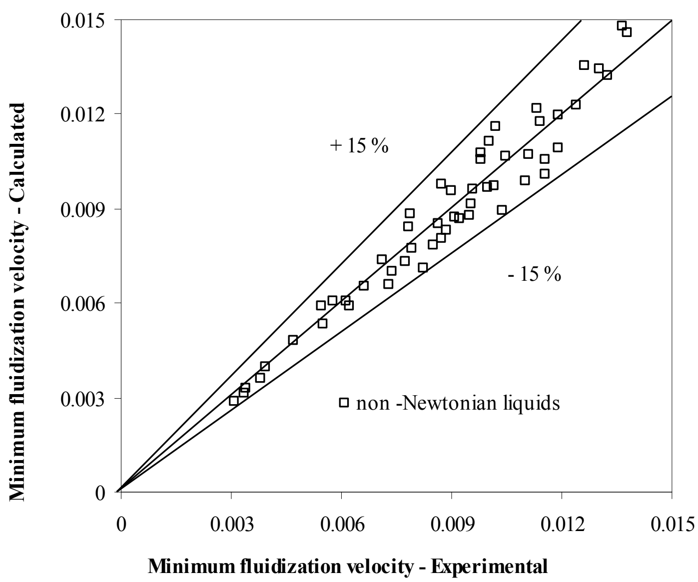


Fig.19 Comparison between the Experimental and Calculated Values of Minimum Fluidization Velocity for non-Newtonian Liquids

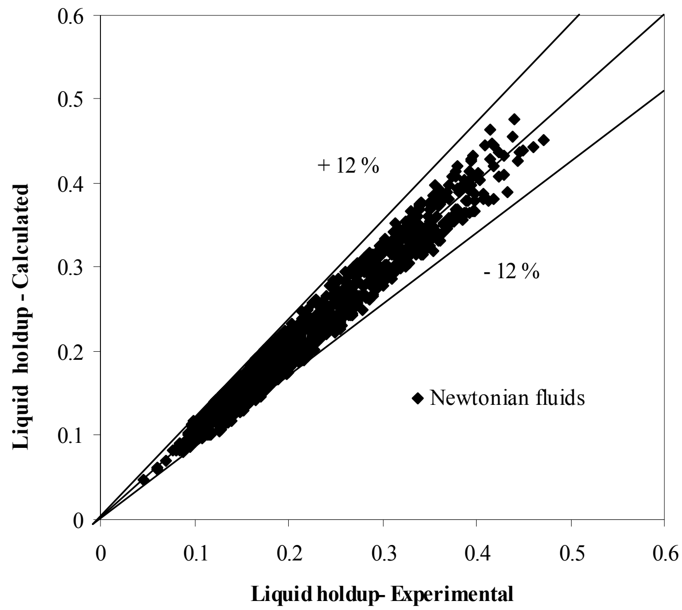


Fig.20 Comparison between the Experimental and Calculated Values of Liquid holdup for Newtonian Liquids

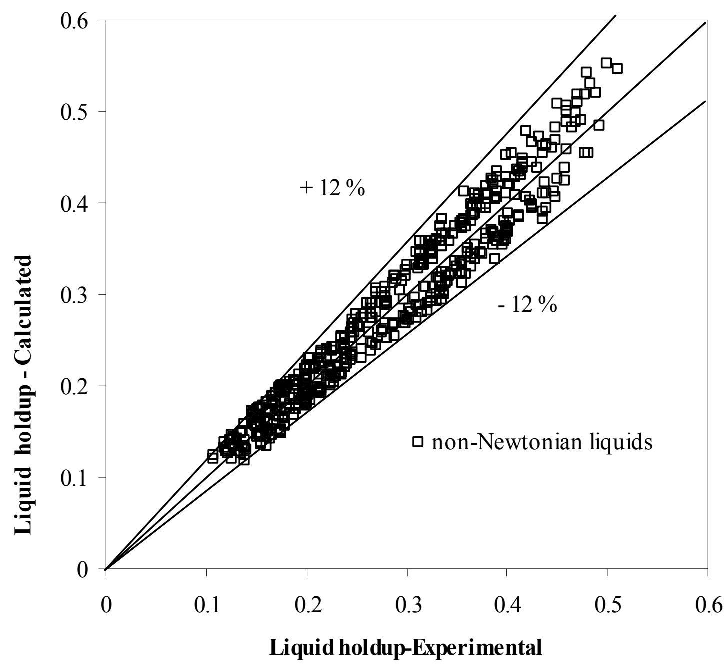


Fig.21 Comparison between the Experimental and Calculated Values of Liquid holdup for non-Newtonian Liquids

CONCLUSION

A three phase combined loop (Internal and External) airlift fluidized bed reactor was proposed by combining advantages of Internal loop airlift reactors and external loop airlift fluidized beds. Experiments were conducted to investigate the hydrodynamic parameters such as minimum fluidization velocity and liquid holdup. The experimental results show that the increase in particle diameter increases the minimum fluidization velocity and decreases with increase in superficial gas velocity. The increase in viscosity of liquids and superficial gas velocity decreases minimum fluidization velocity. The increase of flow consistency index of non-Newtonian fluid decreases the minimum fluidization velocity and decreases with increasing superficial gas velocity. The increase in superficial gas and liquid velocities increases the liquid holdup for the systems air-water, air - n - butanol, air - 80 % glycerol and air - 0.25 % CMC .The same trend is observed for the different particle diameters for air-water system. The increase in viscosity of liquids and superficial liquid velocity increases the liquid holdup. The increase in fluid consistency index increases the liquid holdup. The proposed correlations can be useful to design a commercial reactor.

NOTATION

$(A_r)_m$- Modified Archimedes number $= \dfrac{g.d_P^{\left(\frac{2+n}{2-n}\right)} \rho_L^{\left(\frac{n}{2-n}\right)} (\rho_P - \rho_L)}{K^{\left(\frac{2}{2-n}\right)}}$

d_P- Diameter of the particle, m

d_C- Diameter of the column, m

Fr_G- Froude number for gas- $Fr_G = \dfrac{U_G^2}{g.d_P}$

Fr_L- Froude number for liquid- $Fr_L = \dfrac{U_L^2}{g.d_P}$

g - Acceleration due to gravity, m/s^2

K – Fluid consistency index, kg.m $^{-1}$ s^{n-2}

M_o-Morton number- $Mo = \dfrac{\left[K\left(\dfrac{U_L}{d_P}\right)^{n-1}\right]^4}{\rho_L \sigma_L^3} g$

n- Fluid behaviour index

N_{Remf}- Modified Reynolds number, $N_{\mathrm{Re}mf} = \dfrac{d_P^n U_{mf}^{2-n} \rho_L}{K}$

U_G- Superficial gas velocity, m/s
U_L- Superficial liquid velocity, m/s
U_{mf}-Minimum Fluidization velocity, m/s

Greek letters

ρ_P- Density of the solid, kg/m^3
ρ_L- Density of the liquid, kg/m^3
ϕ - Particle sphericity
ε_L- Liquid holdup
σ_L – Surface tension of liquid, N/m
σ_W – Surface tension of water, N/m

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