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Hydrodynamic Characteristic of Three-Phase (Liquid-Liquid-Solid) Fluidized Beds

Thamer J. Mohammed[1]*, Abbas H. Sulaymon[2] and Amer A. Abdul-Rahmun[1]

[1]*Chemical Engineering Department, University of Technology, Iraq*
[2]*Environmental Engineering Department, College of Engineering, University of Baghdad, Iraq*

Abstract

This work aims to study the hydrodynamic characteristic of three-phase fluidized beds (water, kerosene, spherical plastic or cylindrical PVC particles). The hydrodynamic parameters were investigated (phase hold-up, droplet diameter, droplet rising velocity, particle types and sizes, holes diameter of the liquid distributor, continuous and dispersed phase velocities. The experimental work was carried out using QVF glass 0.106 m I.D. and 2 m height. Three distributors with whole diameters (2.5, 5, 7 mm) were used. Plastic particles with diameters (8, 15 mm) and PVC particles (3.34 mm) were used as a solid phase. Two methods (quick, closing valves and pressure drop method) were used to measure the continuous, dispersed and solid hold-up. A cine camera was used to measure the droplet diameter and rising velocity. It was found that the droplet diameter and rising velocity increase with increasing superficial continuous and dispersed velocities. A relationships between the dispersed phase hold-up, droplet diameter and rising velocity with continuous and dispersed superficial velocities respectively were obtained Dimensionless correlation for the prediction of the droplet rising velocity was developed.

Keywords: Water; Kerosene; Solid fluidization

Nomenclature

A: Cross sectional area (m^2)

d_p: Particle diameter (m)

D_d: Droplet diameter (m)

D_h: Holes diameter in the distributor (m)

g: Acceleration due to gravity (m/s^2)

H_f: Bed height after fluidization (m)

U: Superficial velocity (m)

M_s: Mass of solid (kg)

V_d: Droplet rising velocity (m/s)

V_{dr}: Droplet relative rising velocity (m/s)

Dimensionless group

Fr_{mix}: Mixture Froude number (U^2_{mix}/gdp)

Re_{mix}: Mixture Reynolds member ($\rho_{mix}\ dpU_c/\mu_{mix}$)

We_{mix}: Mixture Weber number ($V^2_c\rho_{mix}dp/\sigma_{mix}$)

Greek letters

ε: Porosity

ε_c: Continuous phase holdup

ε_d: Dispersed phase holdup

ε_s: Solid phase holdup

μ_{mix}: Mixture viscosity of continuous and dispersed phase (mPa.s)

ρ_{mix}: Mixture density of continuous and dispersed phase (Kg/m^3)

σ_{mix}: Mixture surface tension of continuous and dispersed phase N/m

Subscripts

c: Continues phase

d: Dispersed phase

s: Solid phase

Introduction

Fluidized beds have been widely used in many industries because of its desirable characteristics such as high heat and mass transfer rates between phases, temperature homogeneity, easy holding and rapid mixing of particles [1,2]. In three phase fluidization, the particles are fluidized by the co-current flow of liquid and gas [2-6] or immiscible liquid-liquid phase. The gas or immiscible liquid form discrete bubbles or droplets phase and the liquid a continuous phase containing the solid particles [7].

A liquid-liquid-solid three phase fluidized beds as liquid-liquid extractor has been studied by Roszak and Gawrouski [8] and Dakshinamurty et al. [9,10]. It has been found that a three phase fluidized beds extractor is superior in its performance, based on HTU values. A spray extraction column using kerosene/n-butyric acid/water and kerosene/propionic acid/water system has been investigated [9]. A higher mass transfer coefficient has been obtained in three phase fluidized beds extractor with the higher interfacial area due to the droplet breakage caused by the turbulent nature of the fluidized solid particles, using toluene/acetic acid/water and toluene benzoic acid water systems. Phase holdup and mass transfer coefficients in three phase liquid-liquid fluidized beds of three different particles with

***Corresponding author:** Thamer J. Mohammed, Chemical Engineering Department, University of Technology, Iraq, Tel: 9647901425828; E-mail: thamer_jasim58@yahoo.com

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Fluids	Density (kg/m³)	Viscosity (mPa.s)	Surface tension (mN/m)
Water	1000	1.5	72.0
Kerosene	780	2.5	28.0

Table 1: Physical properties of the fluids.

Distributors	Size of holes (mm)	No. of holes (water)	No. of holes (kerosene)	Free surface area (%)
Teflon	7	28	25	23.11
Teflon	5	56	29	18.46
Copper	2.5	52	29	4.5

Table 2: Properties of the distributors.

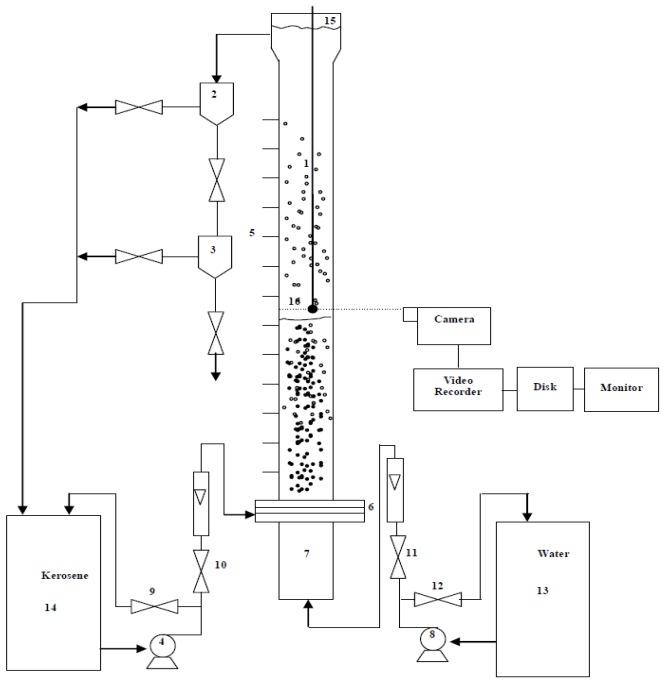

Figure 1: Schematic diagram of the experimental 1.Test column (QVF). 2. first separator. 3. second separator. 4. Kerosene Pump. 5. Pressure taps. 6. Distributor. 7. Distributor box. 8. Water Pump. 9,10,11,12. Globe valves. 13. Water storage (continuous phase).14. Kerosene storage (dispersed phase). 15. Liquid overflow header. 16. Spherical ball.

n-butyric acid/kerosene/water system have been determine by Kim et al. [11].

Various aspects of gas-liquid-solid three phase fluidized beds have been studied extensively as can be seen in previous review articles [12,13]. As a process operation, three phase fluidization is in competition industrially with other and perhaps more conventional contacting reactors, such as the agitated-slurry, trickle flow, and the bubble-column slurry [13]. Three-phase fluidized beds have been applied successfully to many industrial process such as in the H- oil process for hydrogenation and hydro-desulphurization of residual oil, and the bio-oxidation process for wastewater treatment [2].

The objective of the present work is to find experimentally the effect of superficial velocity continuous (water), dispersed (kerosene) phase, particle diameters particle, and holes of diameter of the distributor on continuous, dispersed and solid holdup , droplet diameter , and droplet rising velocity in three phase fluidized beds system.

Experimental work and materials

Materials

Two different sizes of low density solid spherical plastic particles with diameter (8 and 15 mm, density = 1044.7 and 1115.3 kg/m³ respectively) and cylindrical PVC particles with diameter (3.34 mm and density = 1025 kg/m³) were used. Water continuous and kerosene as dispersed phases were used. Their physical properties are listed in Table 1.

Experimental apparatus and measurements

A Q. V. F. glass column of 0.106 m I. D. and 2 m height was used. The column consisted of three sections. These are liquid distributor section, working section and liquid overflow header. The working section in which fluidization takes place is located between the liquid distributor and the liquid overflow header. It is provided with thirteen pressure taps located (0.13 m) apart. These taps are connected to an inverted U-tube mercury manometer to measure the pressure profile along the column.

Two liquid distributors were made of perforated teflon and the third one was made of copper. Distributors the properties of three water-kerosene are given in Table 2. Superficial velocities of dispersed and continuous phases are (0.00135-0.006 m/s) and (0.0359-0.04251 m/s) respectively. The liquids flow rate was measured by means of calibrated rotameter [14]. Schematic diagram of the experimental apparatus (Figure 1).

Measurement of phase holdup

The method of quick- closing values described by Ostegaard [15] has been adopted. The expanded bed height (Hp) was measured, and then the flow of kerosene and water were stopped simultaneously. The distance (Δh) between the upper edge of the fluidized beds and the upper edge of the fixed bed (solid) was measured. For the water-solid system, the expanded heights were measured by visual observation. The solid holdup was formed by using Equations (1).

$$\varepsilon_s = \frac{M_s}{\rho_s A H_f} \quad (1)$$

The other method used by Kim et al. [16,17], Shaikh and Al-Dahhan [18], and Begovich and Waston [19] to find the holdup was adopted. The theoretical basis of this method is that under the steady-state conditions, the total axial pressure gradients of any cross section in the column represents the total weight of the bed consisting of the three phases per unit volume Equation (2).

$$\Delta p = gHp(\rho_c\varepsilon_c + \rho_s\varepsilon_s + \rho_d\varepsilon_d) \quad (2)$$

$$\varepsilon_c + \varepsilon_s + \varepsilon_d = 1 \quad (3)$$

Droplet characteristics measurement

The drop let size and velocity were measured by means of cine camera (National, M-7, 25 frame/s) [14].

Results and Discussion

Phase holdup characteristics

Continuous phase holdup in liquid-solid fluidized beds: The effect of continuous phase velocity on the continuous holdup (Figures 2a, 2b, 2c). The continuous phase holdup increases with increasing water velocity and decreases with increasing particle diameter. These figures show also the continuous phase holdup increases with increasing the size of holes of the liquid distributor.

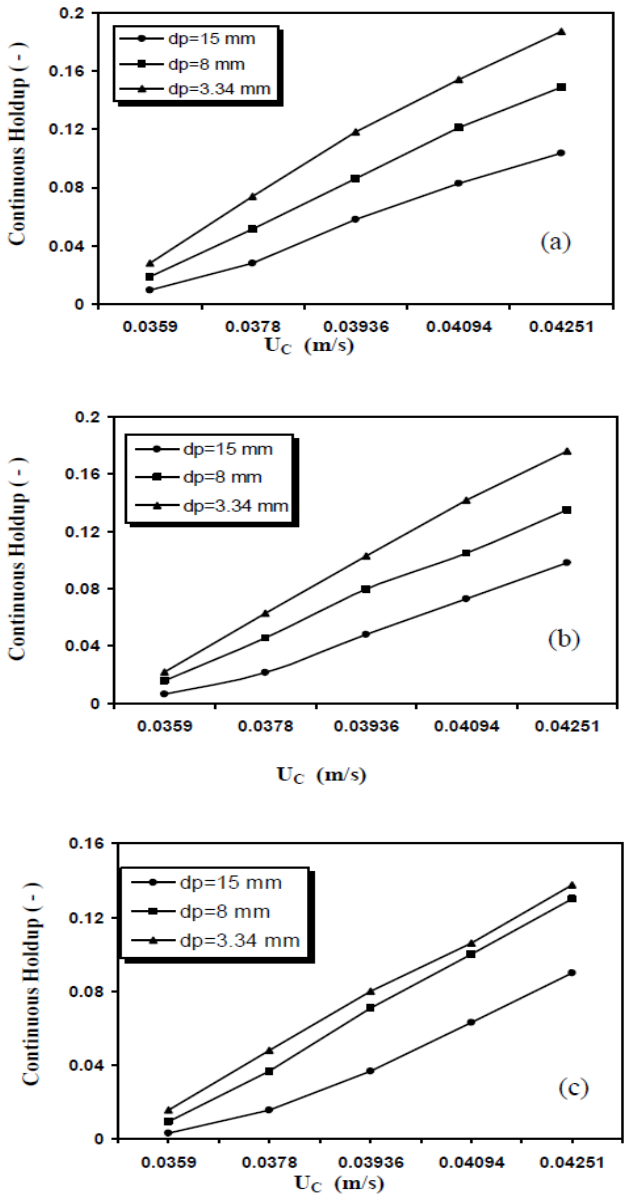

Figure 2: Continuous holdup vs continuous velocity for copper distributor (a) (Dh=7mm) , (b) =5mm, (c) (Dh=2.5mm).

Bed porosity: The variation of bed porosity with dispersed phase velocity for different continues phase velocity particle sizes (Figure 3a-3c). It can be seen that the bed porosity increases with increasing dispersed phase and increases with increasing particle size. The droplets are broken due to high momentum generated from the large particles so that small droplets are well dispersed in the beds of larger particles. Consequently wake velocity and its volume behind small droplets are much lower and smaller in these beds (expansion beds) than in beds of smaller particles (contraction beds). However in the present work, no bed concentration was observed in all fluidized beds studied. On the other hand, the bed porosity increases with continuous phase velocity the expanded this is due to bed height increases with the continuous phase velocity. This observation is agreement with Kim et al. [11].

Continuous phase holdup in three-phase fluidized beds

The effect of the continuous and dispersed phase's velocities on continuous phase holdup for different particles size and densities (Figure 4a, 4b). The continuous phase holdup increases with increasing continuous phase velocity and decreases with increasing dispersed phase velocity. The continuous phase decreases with increasing size and particle density. This figure shows the agreement between the two methods (the quick-closing and the pressure drop) with error equal to 0.3%. The continues-phase holdup in the three phase fluidized beds of

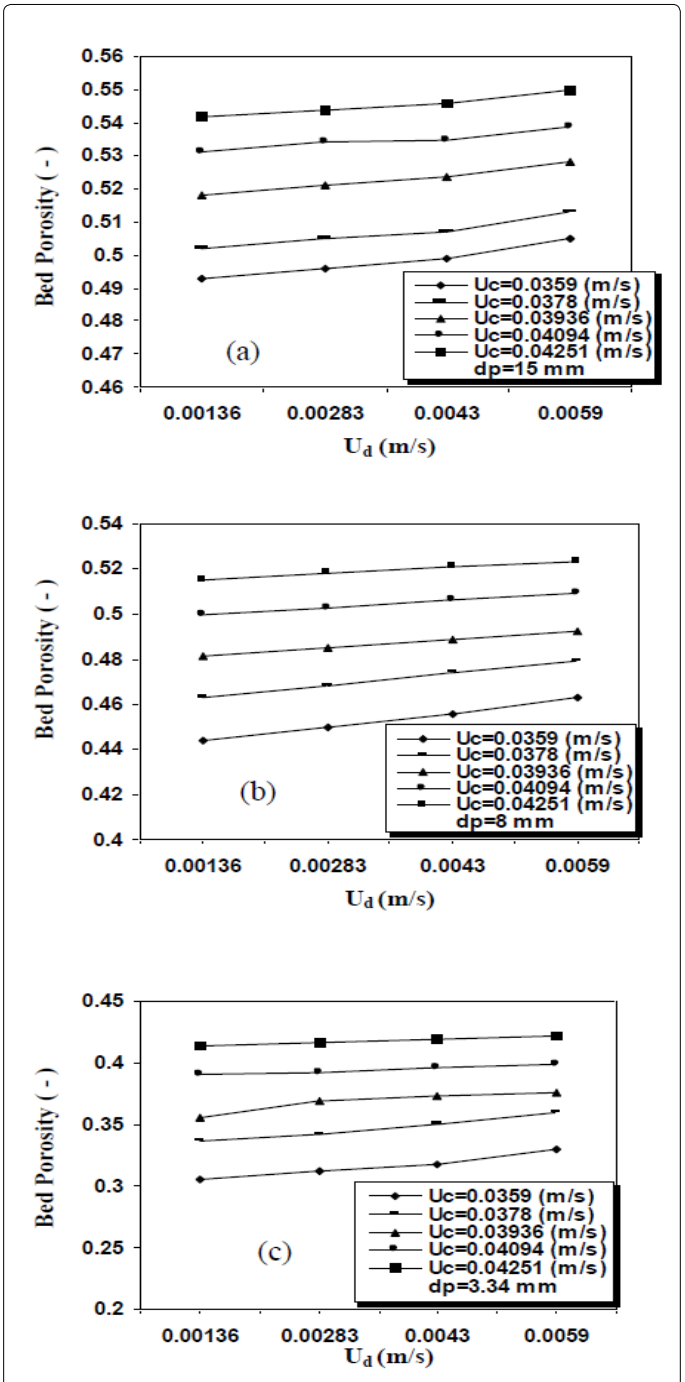

Figure 3: Effect of dispersed velocity on bed porosity for teflon distributor (Dh=7mm) , (a) dp= 15 mm,(b) dp=8mm, (c) dp= 3.34mm.

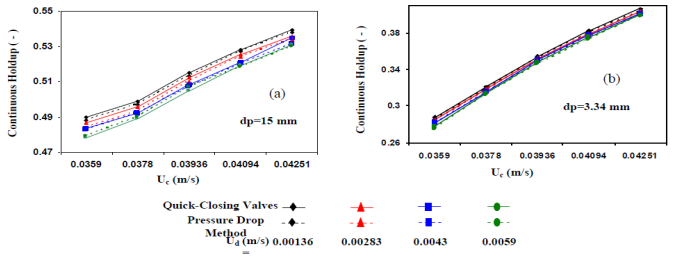

Figure 4: Effect of continuous and dispersed velocities continuous holdup teflon distributor (Dh = 7mm) in liquid-liquid-solid fluidized bed (a) dp = 15 mm, (b) dp = 3.34 mm.

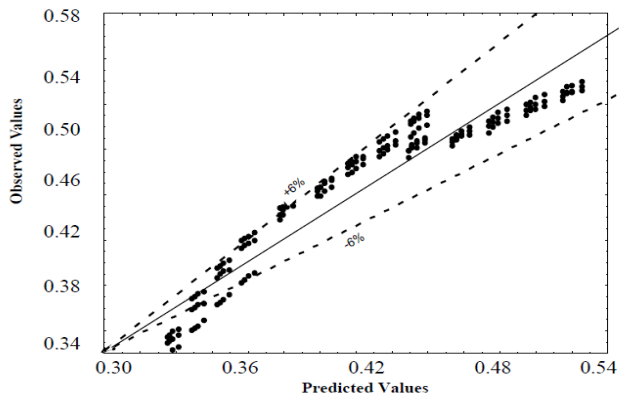

Figure 5: Comparison between experimental and predicted values of continuous phase holdup.

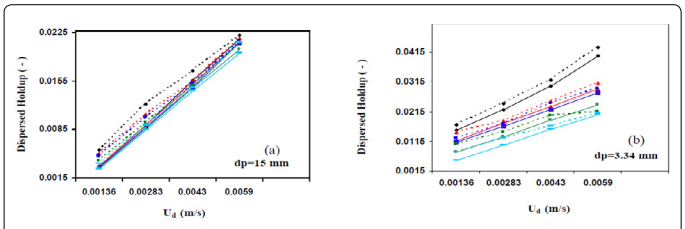

Figure 6: Effect of dispersed velocity on dispersed holdup in teflon distributor (Dh=5mm) in liquid-liquid-solid fluidized beds, (a) = 15mm (b) = 3.34 mm.

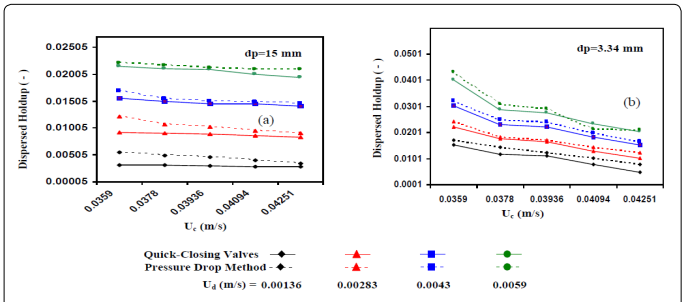

Figure 7: Effect of continuous and dispersed velocity on dispersed holdup for teflon distributor (Dh=5mm) in liquid-liquid-solid fluidized beds (a) = 15mm (b) = 3.34mm.

different particle sizes can be predicted by the following equations (in different flow regimes)

$$\varepsilon_c = 68.72 U_c^{1.161} U_d^{-0.06} d_p^{0.267} \quad (4)$$

Correlations coefficient = 0.948 for D_h = 2.5-7 mm of distributor,

Figure 5, shows the relationship between the experimental and predictable results of continues phase with error 6%.

Dispersed phase holdup in three-phase fluidized beds

The effect of dispersed phase velocity on dispersed phase holdup in beds of plastic (15 mm diameter) and PVC (3.34 mm diameter) particles respectively (Figure 6a, 6b). It can be seen that dispersed phase holdup using two different methods are in good agreement so that the local and mean values of the dispersed phase holdup are nearly the same, since variation of droplet size in marginal throughout the bed. The error between the two methods is 0.9%. Thus the droplet size distribution is fairly uniform so that the liquid-liquid-solid fluidized bed gives better performance compared with spray column for liquid-liquid extraction [17].

The dispersed holdup increases with increasing dispersed phase velocity and decreases with increasing continuous phase velocity, because at high continuous phase velocity provides a high droplet rising velocity [11,16,17]. In the bed of small particles, droplet coalescence increases and the rate of increase in the dispersed phase holdup decreases with increasing the dispersed phase velocity.

The effect of the continuous phase velocity on the dispersed phase holdup in beds of plastic (15 mm diameter) and PVC (3.34 mm) particles (Figure 7a and 7b), where the dispersed phase holdup decreases with increasing the continuous phase velocity due to the reduction of the viscosity and increases in the droplet rise velocity that make the result in the reduction of the residence time of the droplet in the bed. As can be seen the dispersed phase holdup in the bed of 15mm plastic is smaller than 3.34 mm PVC particles because droplet may break down in the former bed while coalesce in the latter bed so the drag force acting on the large droplet is large and the residence time of the larger droplet is small in the bed of large particle [11,16]. The dispersed phase holdup in three phase fluidized beds is correlated with the experimental variables as:

$$\varepsilon_d = 0.0141 U_c^{-1.462} U_d^{1.087} d_p^{0.303} \quad (5)$$

Absolute average error = 7% Figure 8, shows the relationship between the experimental and predictable results for disposed holdup for D_h = 2.5-7 mm liquid distributor.

Solid holdup in three-phase fluidized beds

Two types of flow regime maps in three-phase fluidized beds are

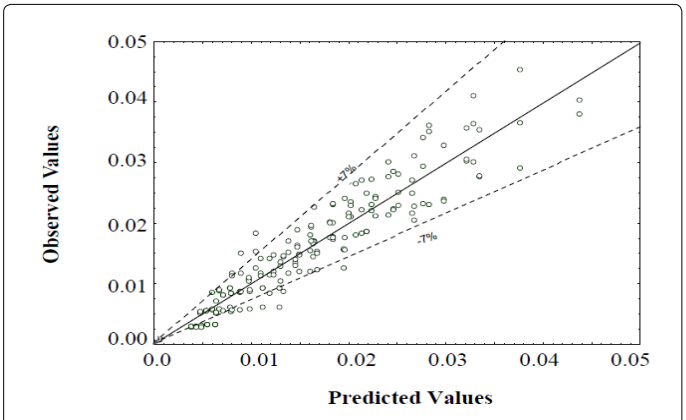

Figure 8: Comparison between exp. and predicted values of dispersed phase holdup.

Holes diameter in the distributor (m×10³)	Particle diameter (m×10³)	Regime
7	15	Droplet disintegrating regime.
7	8	From Transition to coalescing regime.
7	3.34	Droplet coalescing regime.
5	15	Droplet disintegrating regime.
5	8	From disintegrating to transition regime.
5	3.34	Droplet coalescing regime.
2.5	15	Droplet disintegrating regime.
2.5	8	From disintegrating to transition regime.
2.5	3.34	From transition to coalescing regime.

Table 3: Flow Regime Results.

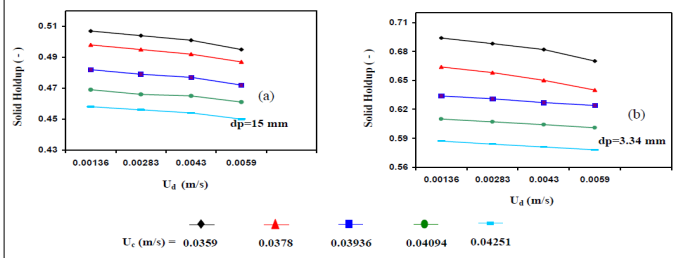

Figure 9: Effect of dispersed and continuous and dispersed velocity on slid holdup in teflon distributor (Dh=7mm) in liquid-liquid-solid fluidized beds (a) dp=15 mm, (b) dp= 3.34.

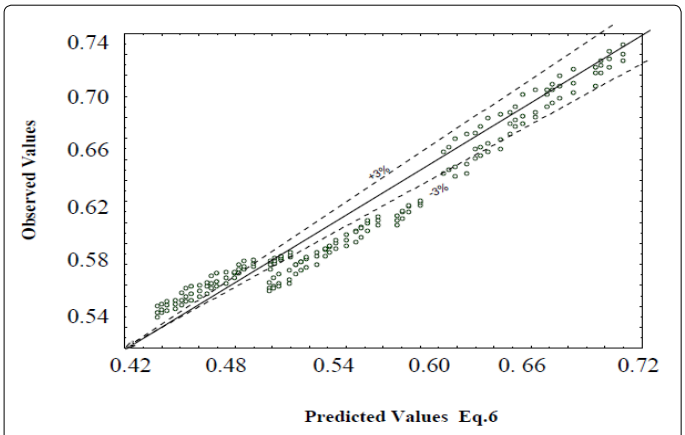

Figure 10: Comparison between experimental and predicted values by Equation (6) of solid holdup.

available, this is based on the drift flux theory, and the other one is the dimensional coordinates of superficial dispersed and continuous phase velocities to improve the existing flow regime maps. A new flow-regime map based on the drift flux theory in three-phase fluidized beds has been proposed by Han and Kim [20]. The flow regimes that are expected in the present work (Table 3).

The effect of continuous and dispersed phase velocities on the solid holdup at a given particle size (Figures 9a and 9b) respectively.

These figures show that the solid holdup decreases with increasing continuous and dispersed velocities. The solid holdup decreases with increasing the particles density. The experimental variable was correlated an:

$$\varepsilon_s = 0.0348 U_c^{-0.518} U_d^{-0.0146} d_p^{-0.199} \quad (6)$$

Correlation coefficient = 0.971 and absolute average error = 3%. Figure 10, shows the relationship between the experimental and predictable results of solid holdup for D_h = 2.5-7 mm liquid distributor.

Droplet diameter and droplet rising velocity

The effect of the dispersed velocity on the droplet diameter with different sizes of particles and different types of distributor (Figures 11a-11c) for Teflon and (Figures 12a-12c) for copper distributors. In the present of PVC 3.34 mm particles, the droplet diameter increases with, increasing the dispersed phase velocity, whereas the droplet diameter in the beds of 8 and 15 mm particles does not change appreciably with the dispersed phase velocity [17] because large particles have

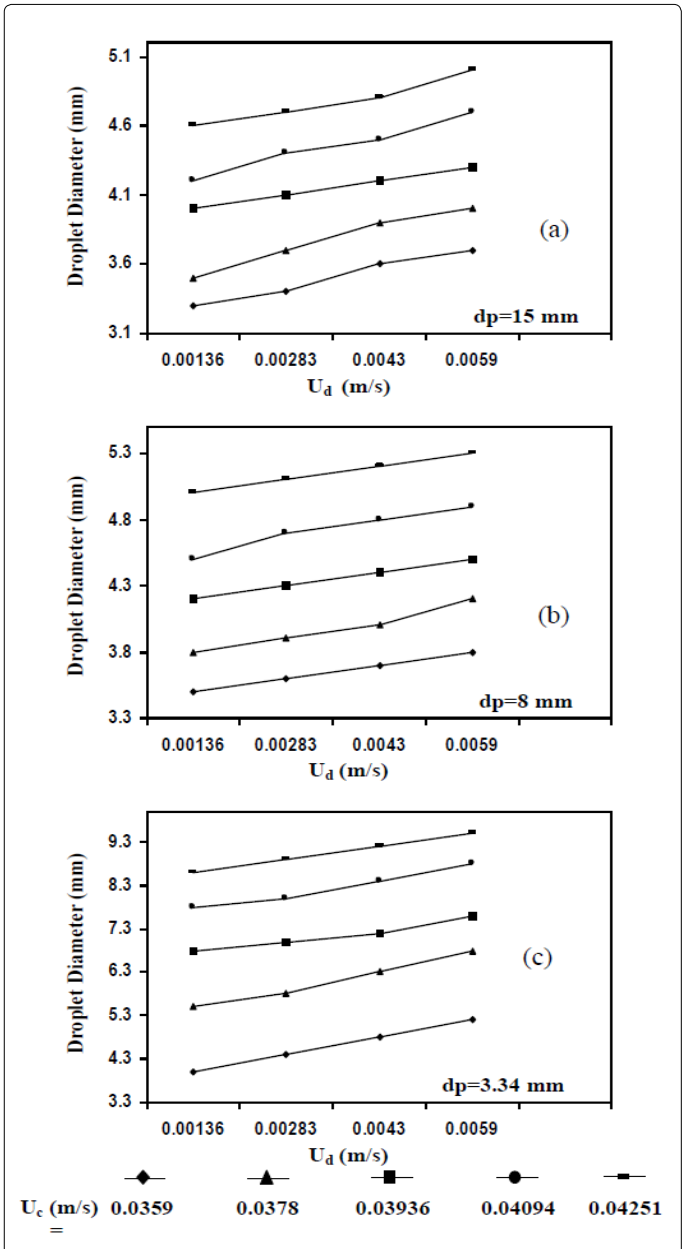

Figure 11: Effect of dispersed velocity on droplet diameter for teflon distributor (Dh=7mm) in liquid-liquid-solid fluidized beds (a) 15mm , (b) 8mm , (c) 3.34mm.

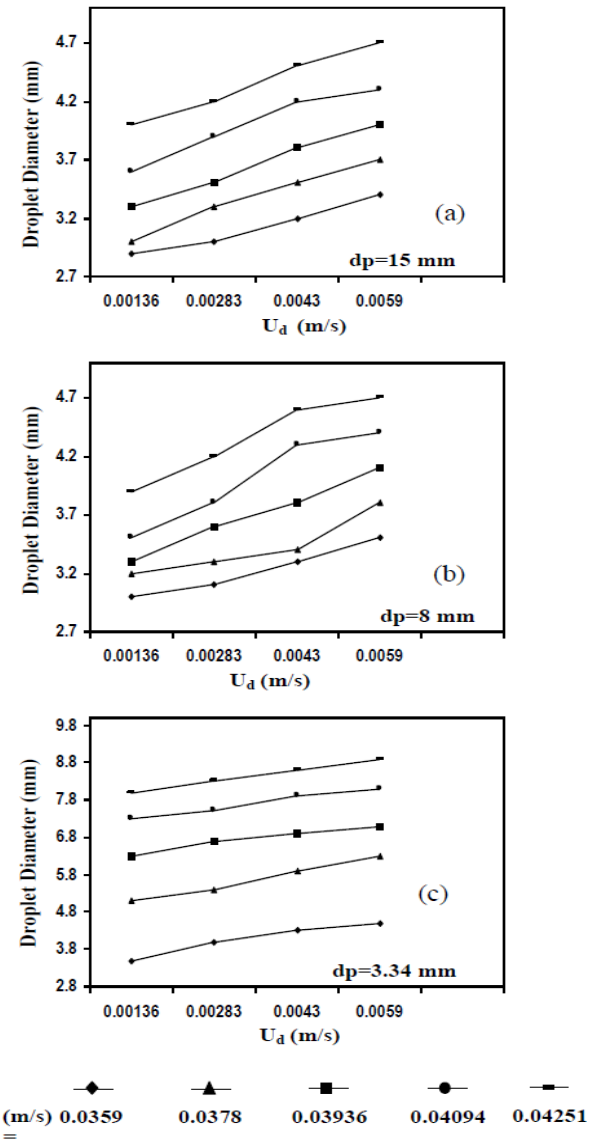

Figure 12: Effect of dispersed and continues velocities on droplet diameter in copper distributor (Dh=2.5mm) in liquid-liquid-solid fluidized beds dp= 15mm, (b) dp= 8mm, dp=3.34mm.

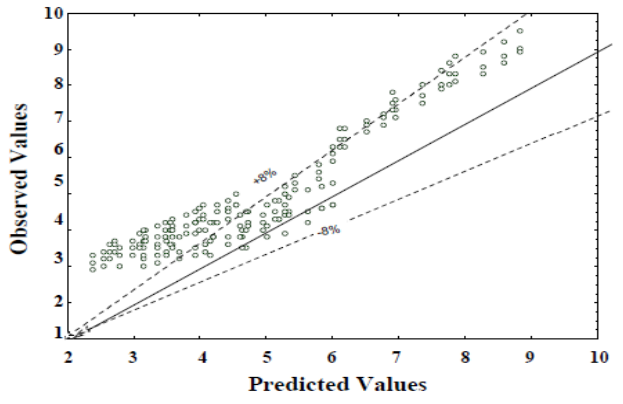

Figure 13: Comparison between experimental. and predicted values of droplet diameter. Equation 7.

droplet breaking potential. Due to the viscosity and small surface tension of the dispersed phase are comparative. Those of the continues phase, dispersed phase can contact the solid surface more easily than continuous phase, which can promote the droplet coalescence as in the case of solvent extraction from slurries [21]. On the other hand, in the beds of large particles, the particle inertia force is larger than the viscous and surface tension forces of the fluids so that the droplet breaks up regardless of the dispersed phase velocity.

The effect of the continuous phase velocity on the droplet diameter (Figures 11a-11c) for Teflon and (Figure 12a-12c) copper distributors. These figures show the droplet diameter increases with increasing continuous phase velocity. The effect of the continuous phase velocity on the bed of (3.34 mm) is more than on the beds of 8 and 15 mm particles. The experimental variable was correlated an:

$$D_d = 16.98 U_c^{3.047} U_d^{0.0882} d_p^{-0.441} \quad (7)$$

For D_h = 2.5-7 mm Figure 13, shows the relation between the predicted and observed values of droplet diameter with correlation coefficient = 0.9373 and absolute average error = 8%.

The effects of the continuous and dispersed phase velocities on the droplet rising velocity (Figures 14a and 14b) for teflon distributed while Figure 15a and 15b for copper distributor (the droplet coalescing bed 3.34 mm particles). The droplet rising velocity increases with

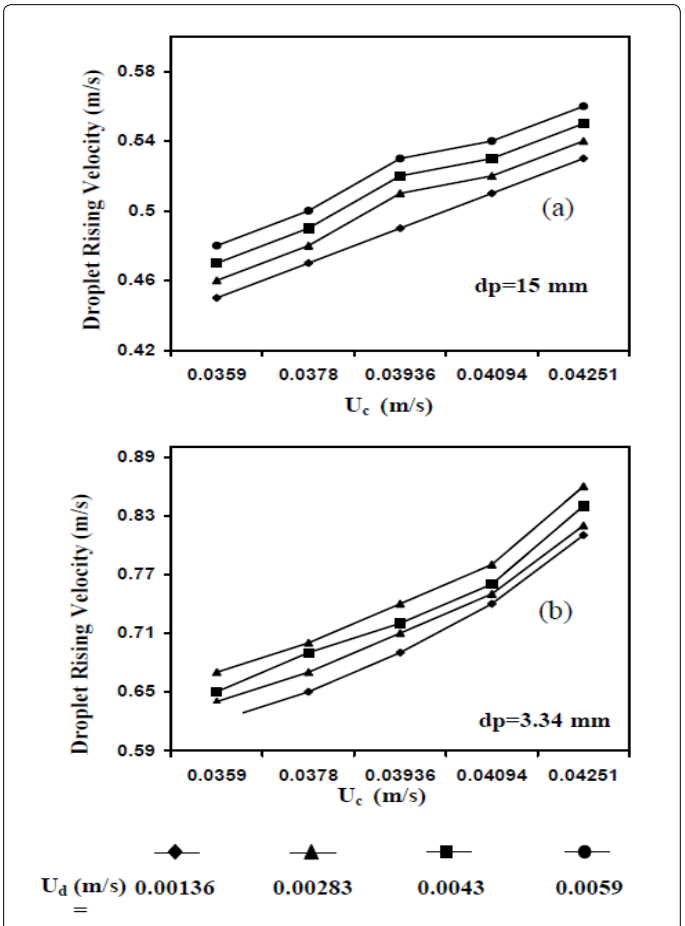

Figure 14: Effect of continues and dispersed velocities on droplet rising velocities for Teflon distributor (Dh=7mm) in liquid-liquid-solid fluidized beds dp= 15mm, (b) dp= 8mm , dp=3.34mm.

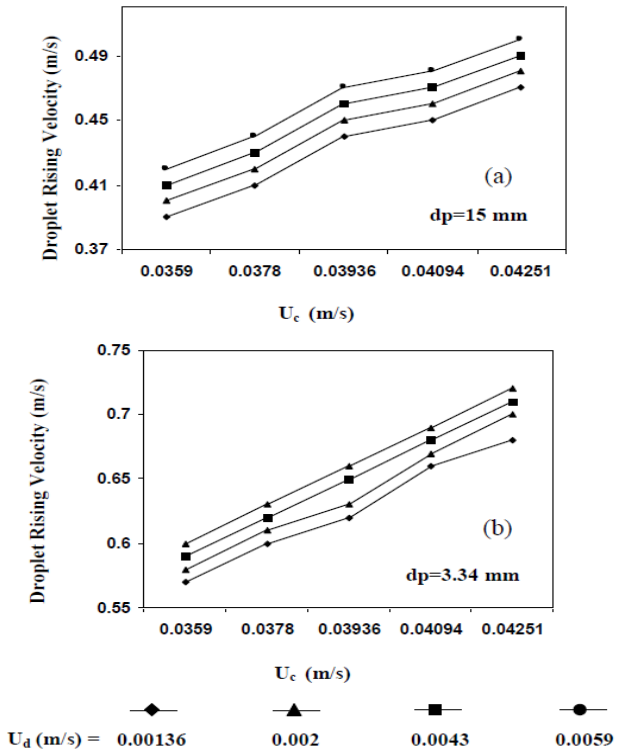

Figure 15: Effect of ntinues and dispersed velocities on droplet rising velocities for copper distributor (Dh=7mm) in liquid-liquid-solid fluidized beds dp= 15mm, (b) dp= 8mm , dp=3.34mm.

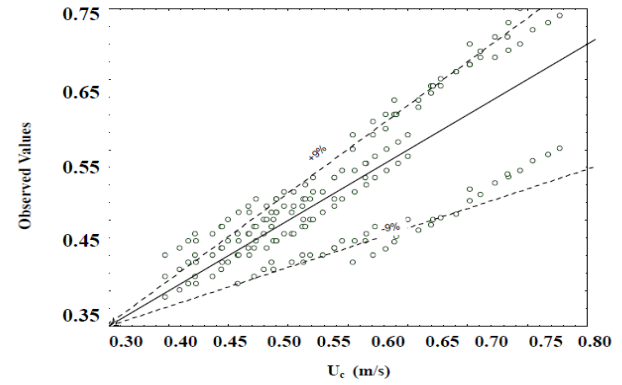

Figure 16: Comparison between experimental and predicted values of solid phase holdup.

increasing dispersed and continuous phase velocities, whereas the rate of increase in the dispersed velocity in the droplet disintegrating beds (15mm) particles is less than in the droplet-coalescing bed, because the dispersed phase velocity increases with droplet size. The dispersed phase velocities and the variable are correlated an:

$$V_d = 12.92 U_c^{1.242} U_d^{0.0417} d_p^{-0.2213} \quad (8)$$

Figure 16 shows the relation between the experimental and predictable values of droplet rising velocity for D_h = 2.5-7 mm with absolute average error 9%.

The relative droplet rising velocity ($V_{dr} = V_d - \frac{U_c}{\varepsilon_c}$) with variation of the continuous and dispersed velocities for Teflon distributor (Figures 17a and 17b), while Figure (18a and 18b) for copper distributor. As can be expected (V_{dr}) increases in the droplet-coalescing bed (3.34 mm particles) with increasing in the continuous and dispersed phase velocities. On the other hand, the rate of increase of (V_{dr}) in the droplet disintegrating bed (15 mm particles) is less than in the droplet-coalescing bed.

The relation between the droplet diameter and droplet rising velocity (Figure 19). The droplet rising velocity increases with increasing droplet diameter as shown in the following correlation:

$$V_d = 246.5\, D_d^{0.507} \quad (9)$$

The correlation coefficient is 0.85 and the error is 8% (Figure 20).

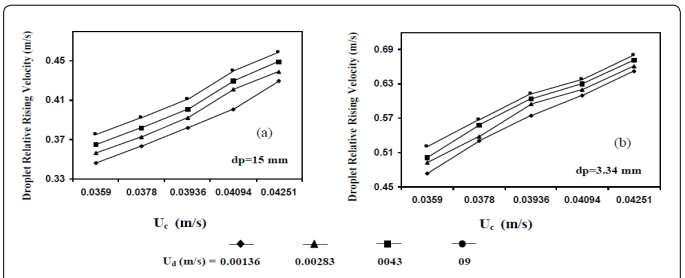

Figure 17: Effect of continuous and dispersed velocities on droplet relative rising velocity in teflon distributor (Dh=5mm) in liquid liquid-solid fluidized beds (a) dp= 15 mm , (b) dp = 3.34 mm.

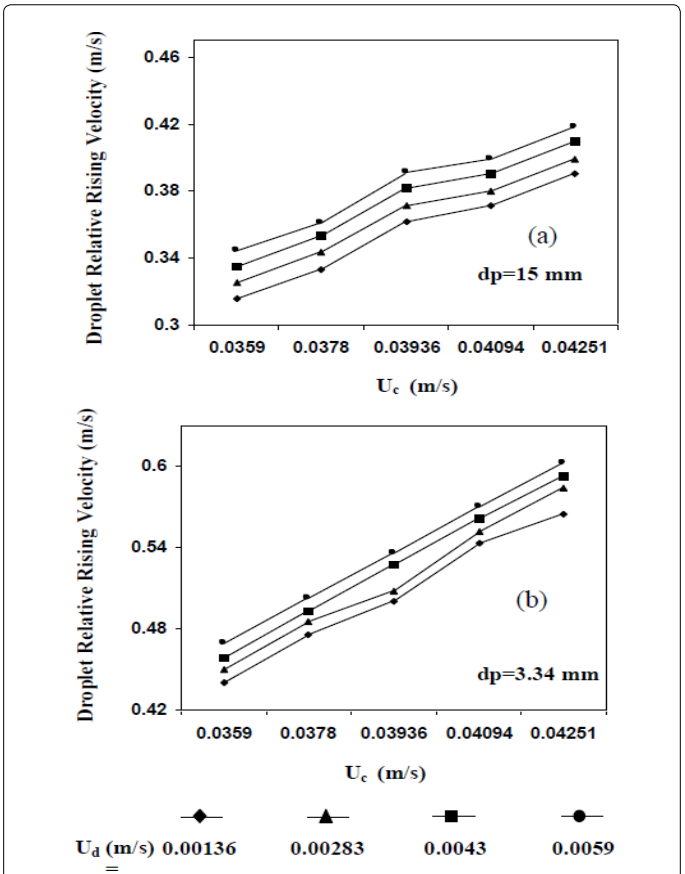

Figure 18: Effect of continuous and dispersed velocities on droplet relative rising velocity for copper distributor (Dh=2.5mm) in liquid-liquid-solid fluidized beds (a) dp= 15 mm , (b) dp = 3.34 mm.

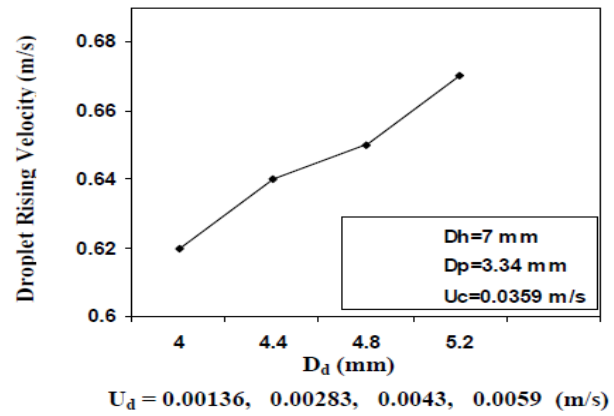

Figure 19: Effect of droplet diameter on the droplet rising velocity in three-phase fluidized beds.

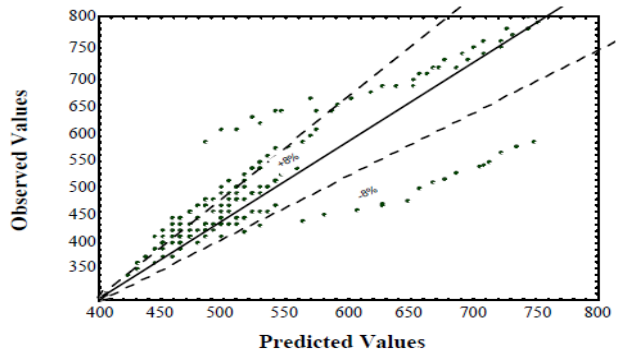

Figure 20: Comparison between exp. and predicted values of droplet rising velocity for equation (8).

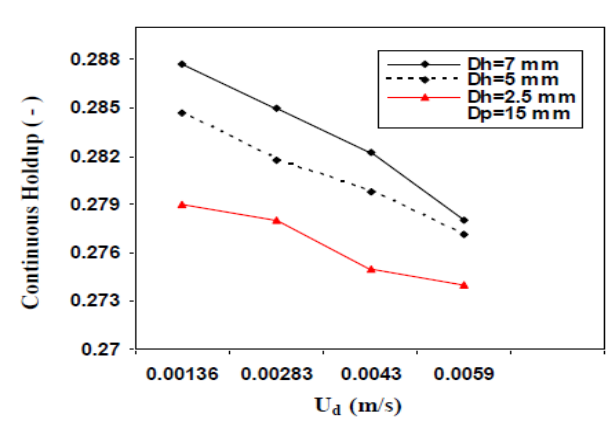

Figure 21: Effect the diameter of holes on the continuous holdup in three-phase fluidized beds.

Effect of Holes Diameter in the Distributor

Effect of Holes Diameter on the Holdup: The effect of holes diameter in the distributors on continuous, dispersed and solid holdup in three phase fluidized beds (Figures 21-23) respectively. The continuous phase holdup and dispersed holdup increase with increasing diameter of holes in the distributor, therefore the solid phase holdup increases with decreasing the diameter of the holes.

Effect of Holes Diameter on the Droplet Diameter and Droplet Rising Velocity: The effect of hole diameters in the distributor on the droplet diameter and droplet rising velocity (Figures 24 and 25) respectively. These figures show that droplet diameter increases with increasing holes diameters, therefore the droplet rising velocity increases due to the dispersed velocity increases with droplet size.

Correlation of Droplet Rising Velocity

Buckingham's π theorem was used to find the dimensionless group to correlate droplet diameter, holes diameter in the distributor, superficial continuous and dispersed phase velocities, bed heights, particle diameter, density of particles and the physical properties of liquid phase mixture which the effect of the droplet rising velocity in the three phase fluidized beds, as shown in the following correlation:

$$\frac{U_c}{V_d} = 0.019\left(\frac{H_0}{D_d}\right)^{0.065}\left(\frac{H_0}{D_h}\right)^{0.144} \mathrm{Re}_{mix}^{0.039},$$

$$We_{mix}^{0.144}\left(\frac{\rho_{mix}}{\Delta P}\right)^{0.236}\left(\frac{d_p}{H_0}\right)^{0.236} Fr_{mix}^{0.152},$$

$$\left(\frac{U_c}{U_d}\right)^{0.045}\left(\frac{\rho_{mix}}{\rho_p}\right)^{0.179} \quad (10)$$

The Ranges of applicability of the correlation (10)

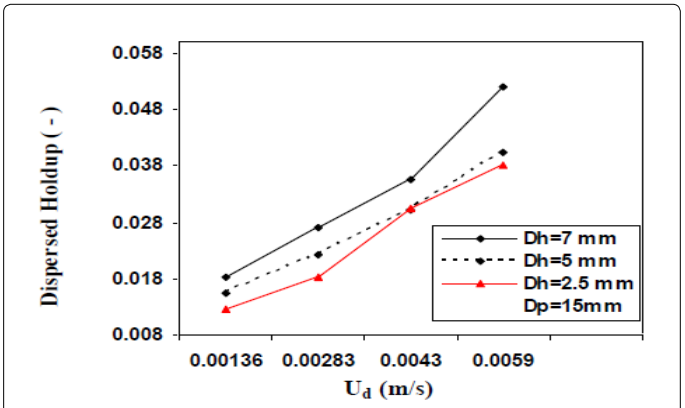

Figure 22: Effect the diameter of holes on the dispersed holdup in three-phase fluidized beds.

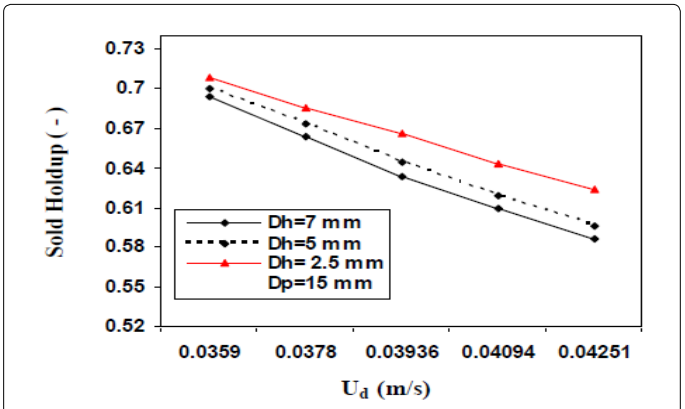

Figure 23: Effect the diameter of holes on the solid holdup in three-phase fluidized beds.

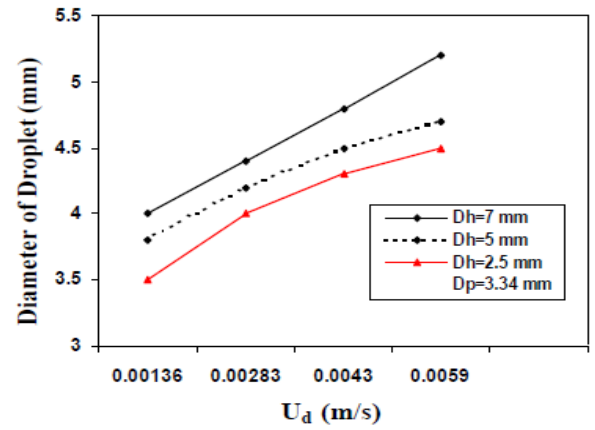

Figure 24: Effect of holes diameter on the diameter of droplet in three-phase fluidized beds.

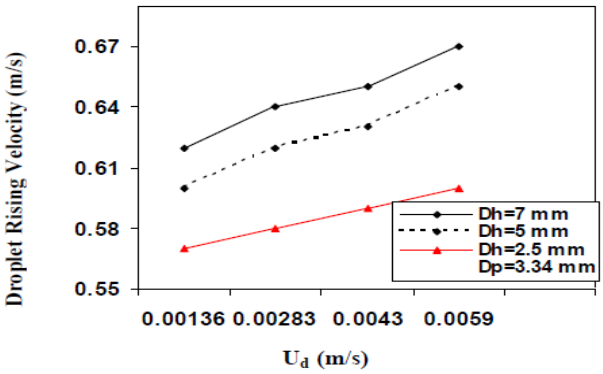

Figure 25: Effect the diameter of holes on the droplet rising velocity in three-phase fluidized beds.

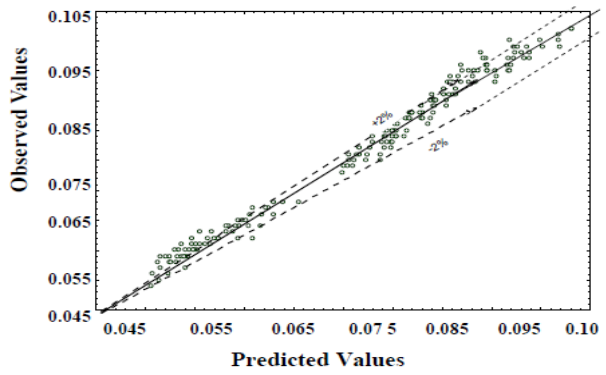

Figure 26: Comparison between exp. and predicted values of continuous phase holdup Equation. (10).

$$0.049 \le \frac{U_c}{V_d} \le 0.092 \text{ , } 33 \le \frac{H_o}{D_d} \le 108 \text{ , } 44.7 \le \frac{H_o}{D_h} \le 125.2 \text{ , } 120 \le \mathrm{Re}_{mix} \le 637.6 \text{ ,}$$

$$0.148 \le We_{mix} \le 0.934 \text{ , } 22.37 \le \frac{\rho_{mix}}{\Delta\rho} \le 40 \text{ , } 0.01 \le \frac{dp}{H_o} \le 0.0479 \text{ , } 8.7\times10^{-3} \le Fr_{mix} \le 0.0551 \text{ ,}$$

$$7.2 \le \frac{U_c}{U_d} \le 26.39 \text{ , } 0.896 \le \frac{\rho_{mix}}{\rho_p} \le 0.975$$

The correlation coefficient = 0.992

The ranges of applicability of the correlation are Figure 26, shows the relation between the experimental and predictable values of the ratio of superficial continuous velocity to the droplet rising velocity.

Conclusions

Bed porosity increases with increasing continuous and dispersed phase's velocities for all particle sizes studied.

- Continuous phase's holdup increases with increasing continuous phases velocity, particle diameter and decreases with increasing dispersed phases velocity.
- Dispersed phase holdup increases with increasing dispersed phase velocity, particle diameter and decreases with increasing continuous phase velocity.
- The solid holdup decreases with increasing continuous and dispersed phase velocities and size of particles.
- The droplet diameter of kerosene increases with increasing continuous and dispersed velocities.
- The droplet rising velocity and relative droplet rising velocity increase with increasing continuous and dispersed phase velocities.
- The continuous phase holdup and dispersed phases holdup increase with increasing holes diameter in the liquid distributor.
- The diameter of droplet increases with increasing holes diameter in the distributor, therefore the droplet rising velocity increases due to droplet rising velocity increases with droplet size.

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