



## Gas Holdup and Axial Liquid Dispersion in Three Phase Bubble Columns

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### Abstract

Axial liquid dispersion coefficient was studied in air-tap water systems (bubble columns) and air-tap water-PVC solid particles system for different column diameters, 10, 15 and 30 cm, Each column had a gas distributor plate with perforated holes of 2 mm diameter. The superficial gas velocity was varied in the range 0.87–7.54 cm/s. The overall gas holdup was measured experimentally by bed expansion technique and the liquid-phase axial dispersion coefficients were obtained by adjusting the experimental profiles of tracer concentration with the predictions of the model. The experimental results show the overall gas hold up decreases with increase in column diameter and solid concentration. The experimental results show that the one-dimensional axial dispersion coefficient,  $D_{ax,L}$ , reveal strong scale dependence. Liquid phase axial dispersion coefficient increases with the diameter of the reactor and with solid concentration. Correlations have been used for the estimation of the overall gas holdup and the liquid phase dispersion coefficient in gas-liquid-solid system bubble column.

**Keywords:** Bubble columns, overall gas holdup, liquid phase dispersion

### محتوى الغاز ومعامل التشتت المحوري في الاعمدة الفقاعية ذات الثلاث اطوار

#### الخلاصة

تمت دراسة معامل التشتت المحوري للطور السائل في أنظمة هواء- ماء (أعمدة فقاعة) ونظام هواء- ماء- حبيبات PVC صلبة بأعمدة فقاعية مختلفة الاقطار، 10 و 15 و 30 سم، كل من الاعمدة يحتوي على موزع للغاز قطر فتحاته 2 ملم. وقد اختلفت سرعة الغاز السطحية بين 0.87، 7.54، و 7.54 سم/ثا. تم قياس محتوى الغاز الكلي تجريبيا بتقنية تمدد الحشوة والسائل وتم قياس معامل التشتت المحوري عن طريق ضبط الاشكال المختبرية (experimental profile) لتركيز الدليل مع توقعات النموذج. أظهرت النتائج بانخفاض بمحتوى الغاز مع زيادة في قطر العمود وتركيز المادة الصلبة، كما أظهرت النتائج بأن معامل التشتت المحوري يزداد مع كل من قطر المفاعل و تركيز المادة الصلبة. وقد استخدمت إرتباطات لتقدير كل من محتوى الغاز وتشتت الطور السائل في نظام غاز سائل صلب للعمود الفقاعي.

**الكلمات الدالة:** الاعمدة الفقاعية، محتوى الغاز العام، معامل تشتت الطور السائل

**Notation**

- $C_s$  Concentration of solids, (g/l)
- $D_T$  Column diameter, m
- $D_{ax,L}$  Axial liquid dispersion coefficient ( $m^2/s$ )
- $g$  Acceleration due to gravity,  $m/s^2$
- $H_0$  Total liquid height in the column, m
- $H_d$  Dispersion height, m
- $L_1$  Distance from injection point to the measurement point probe 1, m
- $L_2$  Distance from injection point to the measurement point probe 2, m
- $L_3$  Distance from injection point to the measurement point probe 3, m
- $L_4$  Distance from injection point to the measurement point probe 4, m
- $U_g$  Superficial gas velocity, m/s
- $U_l$  Superficial liquid velocity (m/s)
- $U_{b\infty}$  Terminal velocity of air bubble (m/s)
- $V_c$  Circulation velocity (m/s)
- $V_s$  Hindered settling velocity (m/s)
- $x$  Longitudinal co-ordinate (m)
- $Z$  Distance (measuring point to feed point) (m)
- $\epsilon_g$  Fractional gas hold-up
- $\epsilon_l$  Fractional liquid hold-up
- $\epsilon_s$  Fractional hold-up of solids
- $\rho_s$  Solid density ( $kg/m^3$ )

## 1. Introduction

The application of three-phase bubble column reactors (BCR) is well established in a wide variety of chemical processing operations such as, production of fine chemicals, hydro treating of petroleum residues, coal liquefaction, wastewater treatment and fermentation processes. The performance of the reactor is, however, strongly influenced by the complex interaction between gas and liquid hydrodynamics as well as the solids mixing pattern<sup>[1]</sup>. Although considerable attention has been given to the analysis of fluid dynamics and associated effects on mass transfer coefficients, phase hold-up and other physical processes<sup>[2]</sup>, there is a paucity of information on the solids mixing behavior<sup>[3]</sup>. In many situations, however, the solid particles are catalysts for the reaction, thus the role of solid flow dynamics is important due to the transient nature of catalyst activity and deactivation. Chen et al, 1994<sup>[4]</sup> and Reese and Fanl,1997<sup>[5]</sup>, investigated the dynamics of solid suspension in BCRs.

To design a column of this type as a slurry reactor, the behavior of the suspended solid particles, that is, the values of the critical gas velocity required for complete suspension of solid particles and the concentration distribution of the solid particles, should be known. However, only three researcher works have been done on the critical gas velocity (VGc) required for complete suspension of solid particles<sup>[6]</sup>.

The solid concentration is defined as the volume fraction, ( $\epsilon_s$ ), of solids in the gas-free slurry. The effect of solid concentration and particle size on gas hold up has been investigated by a number of researchers. Several of researchers concluded that an increase in solids concentrations generally reduced the gas hold up<sup>[7-9]</sup>. Smith et al<sup>[10]</sup> found a strong dependence of gas hold up on solids concentration at low solid concentration. Kato et al<sup>[11]</sup> reported that the effect of solid concentration on gas hold up becomes significant at high gas velocities (> 10-20 cm/s).

Axial mixing of the liquid phase in a bubble column (BC) has been extensively investigated. It is generally described by the axial liquid dispersion coefficient  $D_{ax,L}$ . Deckwer et al.<sup>[12]</sup>, Baird and Rice<sup>[13]</sup>, Joshi<sup>[14]</sup>, Rice et al.<sup>[15]</sup>, and Kawase and Moo-Young<sup>[16]</sup> have measured this liquid axial dispersion coefficient in BCs. All of their results indicate that  $D_{ax,L}$  depends strongly on the superficial gas velocity and the column diameter.

For the gas-liquid-solid (G-L-S) reactors, Hebrard et al. [17] investigated the effect of the gas sparger on two types of multiphase reactor: the classical three-phase fluidized bed (TPFB) and the slurry bubble column. Nacef et al. [18] investigated the effect of the gas sparger and of the scale of the column on the minimum fluidization velocity and the flow regime in the TPFB. Kato et al. [19], used glass beads ( $63 \mu\text{m} < d_p < 177 \mu\text{m}$ ) and characterized the axial liquid mixing in columns 6.6 and 12.2 cm in diameter. They reported that the axial liquid dispersion coefficient  $D_{ax,L}$  increases with the gas velocity at low liquid throughput and that the liquid mixing in the TPFB is very similar to that in a BC. Michelsen and Ostergaard [20] and Ostergaard [21], showed that with glass beads the intensity of axial liquid mixing depends upon the particle size and superficial gas and liquid velocities.

Joshi [14] proposed the following relations for  $D_{ax,L}$ , applicable to BCs and TPFBs respectively

$$D_{ax,l} = 0.33 (V_C + U_l) D_C \quad (1)$$

and

$$D_{ax,l} = 0.29 (V_C + U_l) D_C \quad (2)$$

where  $V_C$  is the circulation velocity calculated from the energy balance method on the basis of Whalley and Davidson's [22] approach, for BCs and TPFBs respectively

$$V_C = 1.31 \left\{ g D_C \left[ U_g - \varepsilon_g U_{b\infty} \right] \right\}^{(1/3)} \quad (3)$$

$$V_C = 1.31 \left\{ g D_C \left[ U_g + U_l - \frac{\rho_l U_l}{\varepsilon_s \rho_s + \varepsilon_l \rho_l} - \varepsilon_s \left( \frac{\rho_s}{\varepsilon_s \rho_s + \varepsilon_l \rho_l} - 1 \right) V_s - \varepsilon_g U_{b\infty} \right] \right\}^{(1/3)} \quad (4)$$

When the weight of solids in the column and/or  $V_s$  are low, Eq. (4) reverts to Eq. (3) available for gas-liquid (G-L) systems.

The axial liquid dispersion coefficient  $D_{ax,L}$  was calculated by fitting the Inoue and Ohki [23] and Kafarov et al. [24] model to the experimental curve generated for the pulse response. Inoue and Ohki [23] and Kafarov et al. [19] have considered that the propagation process is described in analogy to Fick's Second Law

$$\frac{\partial C}{\partial t} = D_{ax,L} \frac{\partial^2 C}{\partial x^2} \quad (5)$$

solution of this equation gives[1]

$$\frac{C(t,Z)}{C_0} = 1 + 2 \sum_{n=1}^{\infty} \left[ \cos \frac{n\pi}{L} L_z \right] \exp \left( - \frac{n^2 \pi^2}{L^2} D_{ax,L} t \right) \quad (6)$$

Where  $C_0$  is the final concentration when  $t = \infty$  and  $L_z$  is the distance to the measuring point ( $L_1, L_2, L_3$  and  $L_4$ ).

The axial dispersion coefficient  $D_{ax,L}$  was obtained by adjusting the experimental profiles with the solution of Eq. (6) model. A number of  $n = 20$  terms were found to be sufficient. As it can be seen, the evaluation of  $D_{ax,L}$  from the curves produced by the pulse method is fast and reasonable good.

However, although extensive work has been performed on these hydrodynamic parameters in G-L systems and classical TPFBs using heavy particles, there is little information on TPFBs using large r light particles. The objectives of this work is to investigate the effect of solid concentration on the overall gas holdup and to characterize the axial liquid mixing generated in multiphase reactors in columns with varying diameters. The results of our study can be expected to be useful for scale up purposes.

## 2. Experimental

The experiments were carried out in three batch type bubble columns with internal diameters of (10, 15, 30) cm and (174, 160, 150) cm in height respectively. The 10 cm column was made of PVC incorporated with glass window for the purpose of visual inspection, the 15 cm column was made of glass type (QVF), and the third column was made of Perspex. The columns were open at the top; hence the pressure corresponded with ambient conditions. Each column had a (6) mm thick plastic plate gas distributor with perforated holes of (2) mm diameter. In the 10 cm diameter column a total of 19 holes were drilled; for the 15 cm diameter column the distributor plate had 80 holes and for the 30 cm diameter column 334 holes were drilled.

Water and air were used as the liquid and gas phases. The solid phase consisted of (PVC) particles of diameter  $d_p = 4$  mm, density  $\rho_s = 0.0139$  Kg/m<sup>3</sup>. Two concentrations of these particles (10 and 25) kg/m<sup>3</sup> were used in the experiments.

The columns were operated batch wise with respect to the liquid and solid phases. The gas was introduced at the bottom of the columns. The experiments were carried out at various gas velocities, carefully adjusted and controlled using calibrated rotameters. Before starting

acquisition of data for a given gas flow rate, the system was given time to achieve steady state. A typical experimental set-up is shown in Fig. 1 for the 10 cm column.

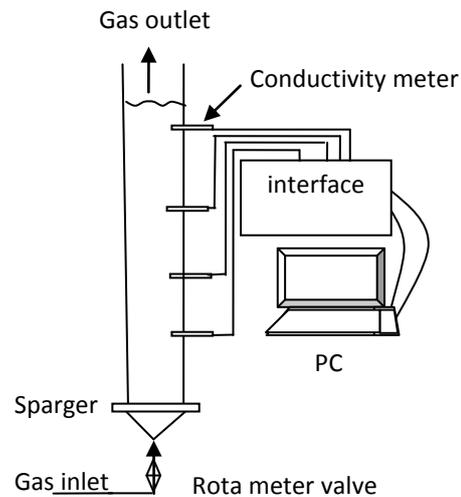


Figure (1) Typical experimental set-up for the 10 cm diameter column.

Residence time distribution (RTD) of the liquid phase was measured using different amounts of saturated solution of NaCl as a tracer. Different volumes of tracer were used to obtain the optimal amount of tracer that corresponds to optimal signal within the operating range of conductivity cell. This optimal amount of a saturated solution of NaCl was found equal to 3.38 wt %.

The conductivity probes used in this work were manufactured by Philips Company, dimensions 1cm in diameter and 15 cm long. They simply consist of two electrodes, approximately 3 mm apart, and encapsulated in plastic tubing. The probes were properly calibrated by measuring their responses to solutions of different known tracer concentrations.

The signals from the electrodes were transmitted to conductance meters of range 100  $\mu$ s to 1000 ms which provides a reading in units of conductance. The meters were connected to a personal computer via interface.

A tracer was injected as a pulse input. The local changes in tracer concentration were displayed and saved continuously on PC. Four electric conductivity probes were inserted 2 cm away from the inside wall, located at different heights as shown in Fig. (2), The distance from the injection to the measuring points,  $L_1$ ,  $L_2$ ,  $L_3$ ,  $L_4$  and  $H_0$  are given in Table (1).

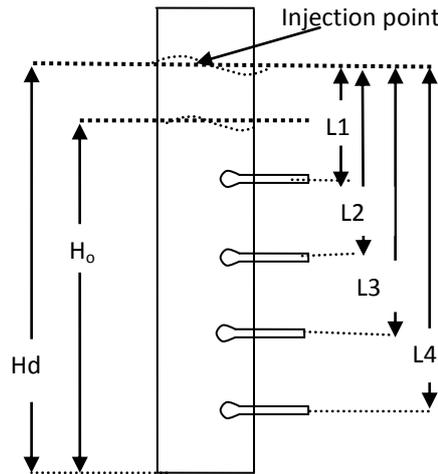


Figure (2) Distances to the measuring points in the column.

Table (1) Constructional detail about the backmixing experiments

Operating condition	Column diameter		
	$D_T = 10$ cm	$D_T = 15$ cm	$D_T = 30$ cm
Liquid height $H_0$ /cm	131	135	100
Distance to the measuring point / cm	$L_1 = 2.8$	$L_1 = 2.5$	$L_1 = 3$
	$L_2 = 44$	$L_2 = 45$	$L_2 = 54$
	$L_3 = 82$	$L_3 = 100$	$L_3 = 88$
	$L_4 = 112.8$	$L_4 = 115$	$L_4 = 115$

The constructive details of the backmixing experiments in the three bubble columns are specified in Table (1). The operating conditions used in the experiments are given in Table (2).

Table (2) Operating conditions used for measuring the overall gas holdup

Column diameter $D_T$ (cm)	Superficial gas velocity range, $U_g$ (cm/s)	Gas holdup range $\epsilon_G$
10	0.87-7.54	0.061-0.566
15	0.87-7.54	0.058-0.445
<b>30</b>	0.87-7.54	0.053-0.320

Fig. (3) shows typical transient tracer concentrations from the 10 cm column, operated at 4.68 cm/s superficial gas velocity. These signals were fitted using the analytic solution to the diffusion equation presented in <sup>[1]</sup>. In this way, for a given experiment, only one variable was adjusted, i.e. the axial dispersion coefficient,  $D_{ax,L}$ .

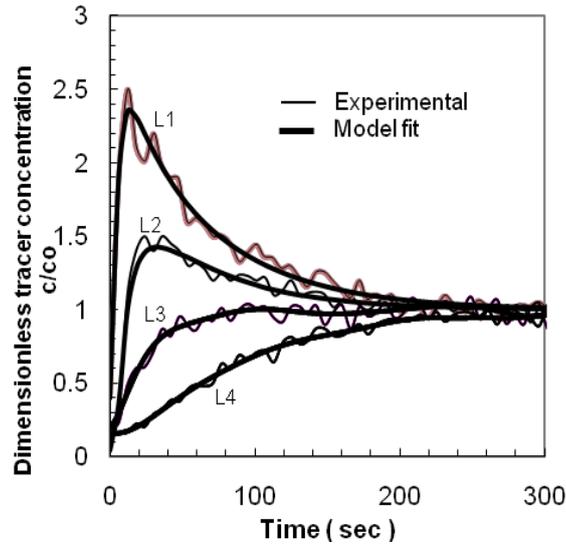


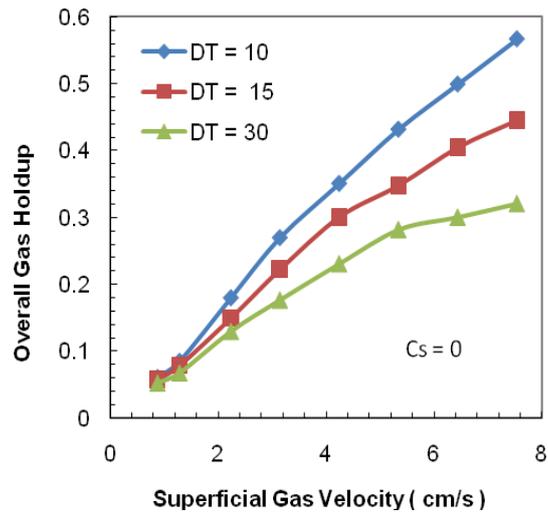
Figure (3) Normalised liquid-phase tracer concentration measured at four different locations along the height of the column in response to pulse tracer injection. The smooth curves represent the fits to the curves from fitting a diffusion model presented in <sup>[1]</sup>.

### 3. Results and Discussion

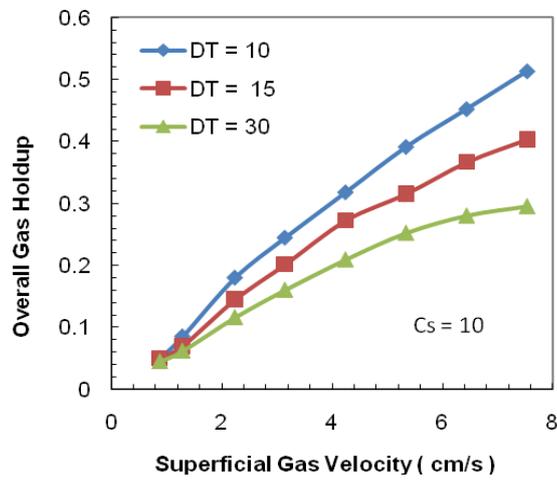
For the estimation of the overall gas holdup, according to bed expansion technique, the overall gas holdup is determined by measuring the heights of the dispersed phase at 161-183 cm that corresponds to initial and dynamic liquid heights respectively for 10 cm bubble column diameter. According to these two heights, the overall gas holdup is calculated by

using  $\left( \varepsilon_g = \frac{H_d - H_o}{H_d} \right)$ . Figure (4) shows the overall gas holdup versus the superficial gas

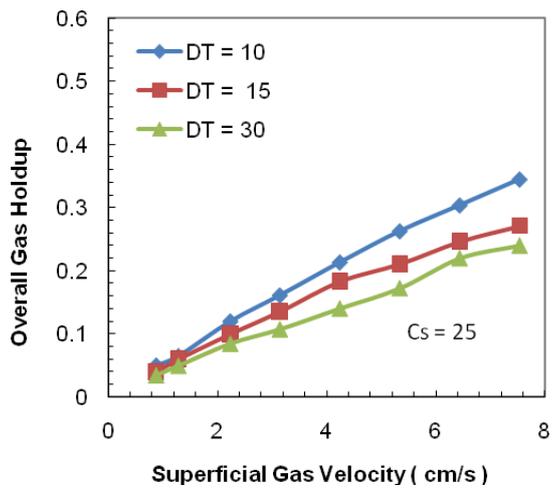
velocity, for 10, 15 and 30 cm bubble column diameters respectively, for the G-L and for G-L-S bubble columns at different solid concentrations, for the G-L bubble columns the overall gas holdup always increases as the superficial gas velocity (or gas flow rate) increases. This is observed for the different solid concentrations studied. The variation of gas holdup due to the addition of solid particles is caused by their influence on bubble size and bubble movement. The direct interaction between bubbles and solid particles may also favor coalescence of bubbles, increasing their diameter and thus their rise velocity, leading in this case to a decrease of overall gas holdup.



(a)



(b)



(c)

Figure (4) Overall gas holdup with solid concentration equal to, (a) 0 (b) 10 (c) 25

The effects of slurry concentration on gas holdup are presented in Fig. 5, where gas holdup is plotted as a function of slurry concentration for given gas velocities. The rate of decrease of gas holdup with increasing slurry concentration seems to depend on the operating gas velocity and the gradient increases with increasing gas velocity. The decrease in gas holdup can be attributed to either an increase in bubble coalescence rate or a reduction in bubble break-up rate. The increase in gas holdup obtained at high gas velocities in air–water system can be attributed to the higher rate of bubble break-up caused by interaction of turbulent eddies with bubbles [25]. The addition of solid particles can cause a dampening effect on the bubble break-up rate due to higher suspension viscosity. For low gas velocities < 4.24 cm/s., the bubble break-up rate is expected to be low leading to a smaller effect of slurry concentrations on gas holdup. The decrease in average gas holdup with increasing slurry concentration has been observed by other researchers [7,8,9].

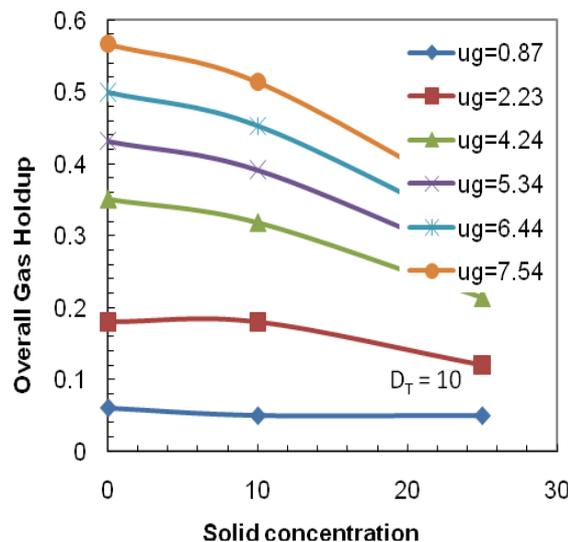


Figure (5) Effect of solid on overall gas holdup

The overall gas holdup is represented as a function of the variables studied in this work  $[\varepsilon_g = f(U_g, D_c, C_s)]$  that can be expressed in the form  $[\varepsilon_g = k U_g^a D_c^b + k1 C_s^c]$

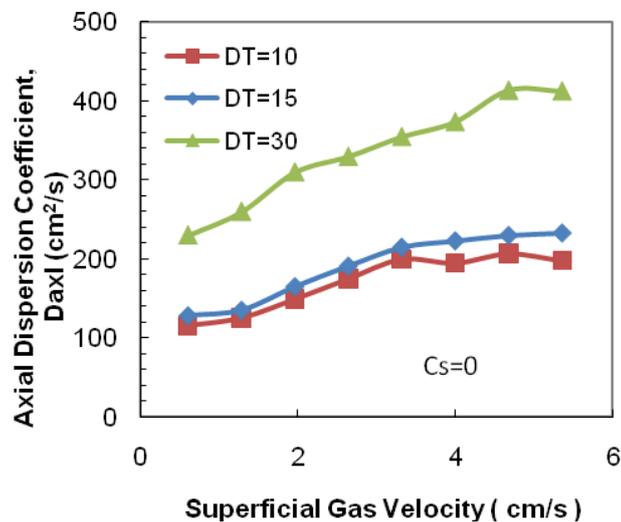
In order to find the coefficients k, a, b, k1, and c a nonlinear regression technique via Statistica software is used. The experimental data for a bubble column are regressed and the following relationship is determined with correlation coefficient of  $R^2 = 0.93$ :

$$[\varepsilon_g = 0.296 * U_g^{0.702} D_c^{-0.284} - 0.093 * C_s^{0.0005}] \quad (7)$$

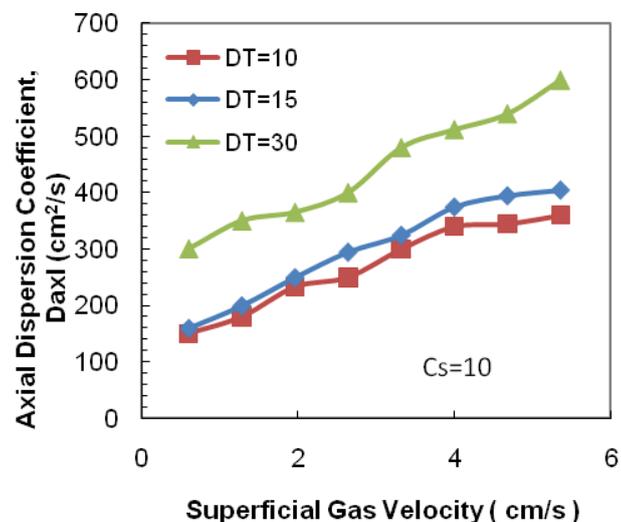
Typical fits of the RTD curves measured at four locations are shown in Fig. (3) for the 10 cm diameter column. Similar excellent fits were obtained for the whole range of gas velocities and for all the three columns studied.

The experimental measurements of the axial dispersion coefficients of the liquid phase show a strong function of the column diameter; Fig. (6-a). Liquid phase turbulence, induced

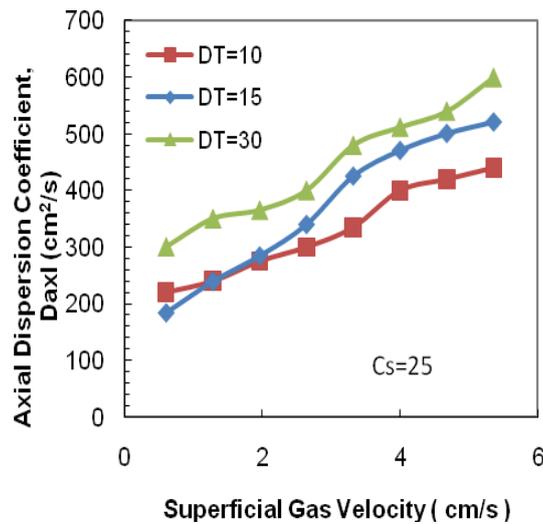
mainly by the movement of bubbles and the existence of large-scale liquid internal circulation, are the main causes of liquid mixing in bubble columns. Joshi, (1980)<sup>[14]</sup>, and Degaleesan et. al., (1997)<sup>[26]</sup> indicated in their studies, the presence of a large-scale liquid circulation cell in bubble columns, with liquid ascending at the central region and descending at the wall region. This liquid internal circulation is mainly driven by non-uniform radial gas distribution in the column. In homogeneous bubbly flow regime, there is no pronounced large-scale liquid circulation in the column and the liquid phase turbulence induced by rising bubbles is the main reason for liquid mixing. The scale of turbulence in homogeneous bubbly flow regime depends on the bubble size. As the gas velocity increases, the bubble size increases thus the bubble-induced turbulence increases which result in a rapid increase in the axial dispersion coefficient, as shown in Fig. (6-a). In churn-turbulent flow regime, both the convective liquid circulation and the liquid turbulent fluctuations play important roles in determining the mixing behavior of the liquid phase which causes liquid phase dispersion and backmixing.



(a)



(b)



(c)

Figure (6) Experimental Axial Dispersion Coefficient in Different Column Diameter with solid concentration equal to, (a) 0 (b) 10 (c) 25

Figure (6) shows the effect of solid concentration on axial dispersion coefficient. It can be seen from the figure that increasing in the solid concentration increases the axial dispersion coefficient due to the increasing in the coalescence characteristics as mentioned before. From this figure the strong influence of the column diameter on the axial dispersion is quite evident.

The axial dispersion coefficient is represented as a function of the variables studied in this work  $[D_{ax,L} = f(U_g, D_T, C_s, \rho_s)]$  that can be expressed in the dimensionless form

$$\left[ \frac{D_{ax,L}}{U_g D_C} = k + k1 \left( \frac{g C_s D_C}{\rho_s U_g^2} \right)^a \right]$$

In order to find the coefficients k, k1, and a a nonlinear regression technique via Statistica software is used. The experimental data for a bubble column are regressed and the following relationship is determined with correlation coefficient of  $R^2 = 0.90$ :

$$\left[ \frac{D_{ax,L}}{U_g D_C} = 3.69 + 0.0002 \left( \frac{g C_s D_C}{\rho_s U_g^2} \right)^{0.58} \right] \quad (8)$$

An agreement between the experimental axial dispersion coefficients and the estimated values from the empirical expressions has been obtained Fig. (7).

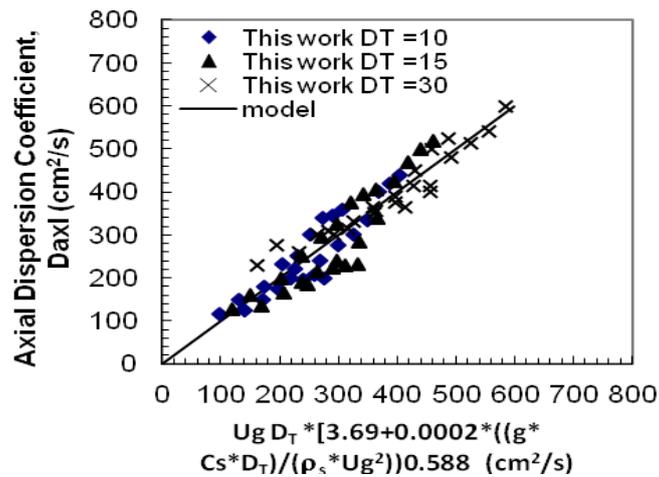


Figure (7) Comparison between the experimental and prediction correlation data of this work

#### 4. Conclusions

The main results derived from this work are:

- Average gas holdups generally decreased with increasing slurry concentration in the column. The gradient of this decrease was higher for gas velocities above 4.24 cm/s.
- The column diameter significantly influences the backmixing of the liquid phase. The axial dispersion coefficient of the liquid phase increases with the diameter of the reactor and with superficial gas velocity for the G-L and for G-L-S bubble columns.
- Increasing the solid concentration increases the axial dispersion coefficient.
- The empirical expression account for the effect of the studied parameters on the overall gas holdup Eq. (7) and axial dispersion in bubble column Eq. (8).

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