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EXPERIMENTAL STUDY OF LIQUID DISPERSION IN BUBBLE COLUMN

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ABSTRACT - The main object of this study is to investigate the influence of column diameter and superficial gas velocity on liquid phase dispersion coefficients (axial and radial dispersion coefficients), mixing times, gas holdup, and bubble dynamics (bubble diameter and rise velocity). The liquid phase dispersion, gas holdup, and bubble dynamics (D_b and V_b^0) were measured for the air-water system in bubble columns of two different diameters,15 and 30 cm. The superficial gas velocity, U_g, was varied in the range 1-10 cm/s, spanning both the homogeneous and heterogeneous flow regimes. The height of liquid in the column was kept constant at 130 cm for the two column. Axial and radial dispersion coefficients and mixing times were measured at various axial and radial locations inside the columns (Z = 25, 75, 125 cm and r/R = 0, 0.45, 0.85), bubble dynamics were measured at three axial location (Z=25, 75, 125 cm). From the experimental data it was found that, the value of the radial dispersion coefficient ($D_{r,L}$) and axial dispersion coefficient ($D_{ax,L}$), gas holdup, bubble diameter and bubble rise velocity, increase with increasing superficial gas velocity. The results emphasise the significant influence of the column diameter on the hydrodynamics. Gas holdup showed a decrease with increasing column diameter, while the radial dispersion coefficient ($D_{r,L}$), axial dispersion coefficient ($D_{ax,L}$), bubble diameter and bubble rise velocity increased with increasing column diameter. A statistical analysis was performed to get a general correlations for the axial liquid dispersion coefficient as a function of the mixing time and dispersion height (H_d), this correlations are: $D_{ax,L}$=0.15 $H^2{}_d$ /$\theta_{0.3}$ for 30 cm column diameter and $D_{ax,L}$=0.11 $H^2{}_d$ /$\theta_{0.3}$ for 15 cm column diameter.

Keywords: Bubble column, axial dispersion, radial dispersion, gas hold-up, hydrodynamics.

1. INTRODUCTION

Bubble columns have emerged as the reactor of choice due to the advantages over other reactor types. In general the bubble column is an adaptable type of reactor which is reasonable in price and can be built in large size. Simple construction and the lack of any mechanically operated parts are two characteristic aspects of bubble column reactors. Hence, little maintenance and low operating costs are required. They have excellent heat and mass transfer characteristics, meaning high heat and mass transfer coefficients(1,2,3,4). Bubble columns are used as reactors, absorption columns, or strippers in a wide variety of processes(5).

Deckwer, 1992(6), reported that the liquid phase dispersion results from the entrainment of the surrounding liquid by the rising gas bubbles that carry the entrained liquid upwards. The liquid phase dispersion in bubble columns is characterized by using dispersion coefficients that are analogous to the diffusion coefficient of Fick's law of diffusion. Unlike diffusion, dispersion arises from convective motion of fluid caused by the following main factors: relative movement of the gas and liquid phases; bubble coalescence and break up; the carry forward of fluid in wakes behind the rising gas bubbles and the consequent return flow generated for maintaining mass balance; and turbulence generated by any superimposed flow of liquid(7). In a bubble column reactor the dispersion has the effect of reducing conversion in reactors, also influence reaction selectivity. Lievenspiel, 1992(8), considered the degree of backmixing as categorized in a measure of the deviation from the ideal plug flow system.

The liquid dispersion in bubble column is two types, axial and radial. Axial mixing, axial dispersion, and longitudinal dispersion are all terms used to describe a phenomena that causes a distribution of residence time for a reaction mixture. Fig. (1) shows the ideal velocity and non ideal velocity profile. Literature correlations for liquid phase axial dispersion vary considerably in complexity and details(9). All correlations anticipate a significant increase in $D_{ax;L}$ with increasing column diameter D_C, often correlated as a power-law dependence D^n_C. The value of the power law index n varies between 1 and 1.5(9). The dependence of $D_{ax,L}$ on the superficial gas velocity (U_g) usually ranges within ($U^{0.25}$– $U^{0.5}$)(9). Several correlations for liquid phase axial dispersion coefficient as Riquart, 1981(10) and Joshi and Sharma, 1979(11), express $D_{ax;L}$ as a product of the liquid circulation velocity and the column diameter D_C. This liquid re-circulation is the cause of liquid phase dispersion and backmixing.

Krishna et al., 2000(9), reported that, the liquid viscosity does not have any significant influence on $V_L(r)$ or $V_L(0)$. A simple linear proportionality is observed with a coefficient 0.31, i.e.

$$D_{ax,L} = 0.31 D_C V_L(0) \qquad (1)$$

The influence of the physical properties of the liquid on the dispersion coefficient has been investigated by several authors (Aoyama et al., 1968(12); Hikita and Hikukawa, 1974(13); Riquarts, 1981(14); Walter and Blanch, 1983(15); Bernemann, 1989(16)) but little effect had been observed by them. The degree of axial dispersion is also affected by vessel internals and surface-active agents that delay the coalescence. Konig et al., 1978(17), demonstrate the effects of surfactants and sparger type by experimenting with weak alcohol solutions using three different porous spargers. They clearly indicated that the interaction of surfactants and sparger can be very complex. Surfactants can produce either much more or much less backmixing than surfactant-free systems, depending on the bubble size, which, in turn, depends on the sparger used. While substantial information exists on axial dispersion of fluid in bubble columns, radial mixing in these reactors has been ignored almost completely(18,19,20,21,22). The few measurements of radial dispersion coefficients cited by Deckwer, 1992(6), suggest that the radial dispersion coefficient is approximately less than one-tenth of the value of the axial coefficient. When bubble columns are used as photobioreactors, the need arises for quantifying the radial mixing. This is because photosynthesizing microalgae and cyanobacteria suspended in column photobioreactors cause a radial decline in irradiance from a high value near the externally illuminated transparent wall to a low value in the center of the photobioreactor. Consequently, the volume of the photobioreactor can be demarcated into a dark interior core region and a relatively better illuminated peripheral region(23,24).

The frequency of radial motion of the cell-suspending fluid between the light and dark zones, or the “flashing-light effect”, influences the productivity of photobioreactors(25,26). Therefore, ways should be found for quantifying the extent of radial mixing and its dependence on the aeration rate, the main operational variable in a bubble column photobioreactor.

For otherwise equal conditions, the $D_{r,L}$ value was typically only about 1% of the $D_{ax,L}$ value. The relatively low $D_{r,L}$ values were apparently a reflection of the fact that the bubble column operated in the bubble flow regime that exists typically when the superficial aeration velocity below 0.05 m/s. Operation in the churn-turbulent flow regime that is characterized by the presence of many spherical cap bubbles, is likely to enhance the $D_{r,L}$ value relative to the values(7).

The aim of the present work is to study experimentally the characteristic of the hydrodynamics of the operating system (i.e., hold-up, bubble size, radial and axial liquid dispersion coefficients) and to develop a model which is capable to predict the axial and radial coefficient.

2. EXPERIMENTAL APPARATUS AND PROCFDURE

2-1 Experimental Apparatus

In present work, experiments were carried out using columns of 0.15 and 0.3 m diameter respectively. The first column is made of QVF glass while the second one is made of plexiglass. Air-water system was used in the experiments. In all the experiments the height of liquid in the column was kept constant at 130 cm. The system was working in batch mode, i.e., during each run, the liquid was neither fed nor discharged. Fig. (2) shows a photographic view of the experimental apparatus and Fig.(3) shows a schematic diagram. Each column had a gas distributor plate with perforated holes of 2 mm diameter as shown in Fig. (5). Detailed description of the experimental setup can be found in Hayder, 2007(27). The physical properties and operating conditions are listed in table (1).

2-2. Measuring System

2-2-1 Conductivity meter

The conductivity of the liquid inside the column was measured by a bench-type single-electrode conductivity meter manufactured by Philips Company of range 100 μs to 1000 ms which provide a reading in units of conductance and three conductivity probes were used as the controller circuit. It includes all the necessary circuitry to measure the liquid conductivity. The objective is to build an interface circuit that reads the conductivity from three conductivity electrodes distributed along the bubble column simultaneously by multiplexing each probe alone to the conductivity meter. The calibration of the conductivity meter was done by using standard NaCl solutions prepared for this process.

2-2-2 Conductivity probe

Three conductivity probes were used in this work. The probes 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, every one of them was connected to PC via interface circuit. Fig. (4) shows the locations of the probes along the column and the tracer injection point.

2-3 Experimental Procedure

Detailed of the procedure followed during the experiments can be found in Hayder, 2007(27). Table (2) shows a sample of experimental arrangements of the experiments.

3. THEORY

At each experimental run, the over all gas hold-up (ε_g) was measured using bed expansion method(28,29) according to:

$$\pounds_g = (H_d - H_o)/H_d \qquad (2)$$

3-1. Determination of Transition Point for the Flow Regime

The detection of regime transition from homogenous to churn turbulent flow and the investigation of the transition regime are quite important(30). In the present work, the drift flux model hypothesis which is introduced by Wallis, 1969(31), is utilized to detect the transition point. According to Mouza et al., 2005(32),

$$j_{GL} = U_g(1 - \pounds_g) \qquad (3)$$

A plot of j_{GL} vs. ε_g reveals the gas velocity at which transition occurs represented by a change in slope of the best fitting line.

3-2. Axial and Radial Dispersion

For a batch of liquid in a bubble column, the dimensionless differential equation which describes the axial and radial profile of an injected tracer through the column is:

$$\frac{\partial C_T}{\partial^{-}} = \frac{\partial^2 C_T}{\partial y^2} + \frac{1}{x}\frac{\partial C_T}{\partial x} + \frac{\partial^2 C_T}{\partial x^2} \qquad (4)$$

where:

$$\theta = \frac{D_{ax,L}t}{L^2} \qquad (5)$$

$$y = \frac{h}{L} \qquad (6)$$

$$x = \frac{r}{L}\sqrt{\frac{D_{ax,L}}{D_r}} \qquad (7)$$

$$C_T = \frac{C_L}{C_\infty} \tag{8}$$

with the following initial and boundary conditions:

1- x = 0 $\frac{\partial C_T}{\partial x} = 0$

2- $x = R/L\sqrt{Dax,L/Dr,L} = \beta$ (i.e., at the wall when r=R), $\frac{\partial C_T}{\partial x} = 0$

3- when $\theta > 0$ and $y = 0$ $\frac{\partial C_T}{\partial y} = 0$ (i.e. at the surface of the dispersion).

4- when $\theta > 0$ and $y = 1$ $\frac{\partial C_T}{\partial y} = 0$ (i.e. at the bottom of the reactor).

5- when $\theta=0$ $C_T = 0$

The analytic solution of equation (4) is found to be:

$$C_T = \sum_{n=1}^{\infty} \frac{J_0(\upsilon_n x)}{J_0^2(\upsilon_n \;)} e^{-(\upsilon_n^{2}{}^{-})}\left(1 + 2\sum_{m=1}^{\infty} \cos(m\text{®}y)\, e^{-(m^2\text{®}\,^{2-})}\right) \tag{9}$$

where J_0 is the zero-order Bessel function and υ_n is the *n*th root of the first order Bessel function, J_1(7). Detailed analytic solution is found elsewhere Hayder, 2007(27).

Wei et al., 1995(33) and Chen et al., 2001(34), found a similar two dimensional dispersion model which has been used to characterize mixing in liquid–solid fluidized beds.

It is noted that, when the C_T in Eq. (9) is radially invariant (i.e. $D_{r,L}=\infty$), υ_n,β and x become zero and $J_0(\upsilon_n\beta) = J_0(\upsilon_n x)=1$ (7). In this case Equation (9) reduces to:

$$C_T = 1 + 2\sum_{m=1}^{\infty} \cos(m\text{®}y)\, e^{-(m^2\text{®}\,^{2-})} \tag{10}$$

The conductivity data used for the calculation of dispersion coefficients are smoothed in order to remove the noise that already present due to occasional gas bubbles being in contact with the conductivity probes. A typical set of the pulse-response data and the best fit model curve generated using Eq. (9) are shown in Fig. (6) for 15 cm column diameter. The value of the radial dispersion coefficient influenced the height of the model generated peak, whereas the value of the axial dispersion coefficient influenced the width of the peak.

4. RESULTS AND DISCUSSION

4-1. Determination of Flow Regime Transition Point

Fig. (7) indicates the location of the flow regime transition point using drift flux model introduced by Wallis, 1969(31). Effect of the column diameter is obvious. The gas velocity at which transition occurs is proportional to the column diameter.

4-2. Effect of Superficial Gas Velocity and Column Diameter on Gas Holdup

The velocity-holdup relationship of the gas phase is the most important design parameter for gas-liquid bubble column reactors, providing the basis for the prediction of heat and mass transfer coefficients and information on hydrodynamic conditions. Fig. (8) shows this relationship, it also shows the effect of column diameter on gas holdup.

At low superficial gas velocity, a steep increasing relationship occurred with gas holdup. This behavior characterizes bubble columns operating in bubbly flow regime. At gas velocity above the transition point (i.e., 4 cm/s for 15 cm column diameter and 7 cm/s for 30 cm column diameter) , the rate of increasing becomes slower. Also the gas holdup is found to decrease slightly with increasing column diameter. This decrease in gas holdup evident in both homogenous and heterogeneous flows regime is due to increase in liquid circulation with increasing column diameter, due to these strong circulations, the bubble will be accelerate and reduction in gas holdup occurred. These results are in agreement with (Krishna et.al.,1999(35), Krishna et. al., 2001(36), Akita and Yoshida, 1973(37)).

4-3. Effect of Column Diameter, Superficial Gas Velocity, And Liquid Height On Bubble Size

Figs. (9) a, b and c show the effect of column diameter and liquid height on bubble diameter at varying superficial gas velocity. It indicates that a positive proportional relationship with both gas velocity an column diameter, but it inversely proportional with liquid height. These relations can be explained as follows:

- Relation of gas velocity with bubble diameter can be attributed to two reasons: an increase in the air flow rate produced a large number of bubble per unit volume, increasing frequency of collision. Furthermore higher air flow rate generate stronger convection and a more impact of the bubble occurred leading to coalescence of these bubble. These results are in agreement with (De Swart et al., 1996(38); Ueyama et al., 1980(39); Koide et al., 1985(40) and Yamashita, 1994(41)).

- Relation of column diameter with bubble size can be attributed to that, at smaller column diameter most of large bubble culled with the wall of column producing smaller bubbles. This is in agreement with (Onna Kramer, 2000(42); Krishna and van Baten, 2001(43) and Koide et al., 1979(44)).
- Relation of liquid height with bubble diameter can be attributed to that, when a bubble rises up through the liquid, it will undergo a phenomena of brake up which results bubble with smaller diameter. This is in agreement with (Lockett and Kirkportick, 1975(45); Kolbel et al., 1972(46) and Krishna 2000(47)).

4-4. Effect of Superficial Gas Velocity, Column Diameter, Radial Position, And Liquid Height on Liquid Axial Dispersion Coefficient

Figs. (10) a, b, and c show the effect of superficial gas velocity, column diameter and different axial position of liquid on axial dispersion coefficient. It can be seen that, there is nearly a linear proportional relationship of axial dispersion coefficient with gas velocity, a proportional relationship between column diameter and axial dispersion is also observed. The relationship between liquid height and axial dispersion coefficient is of proportionality in less extent:

- The relationship of U_G with $D_{ax,L}$ can be attributed to the effect of the average liquid circulation velocity (V_C) which increases with the increase of (U_g) leading to a decrease in mixing time and consequently to an increase in liquid dispersion coefficient. This result is in agreement with Pandit and Joshi, 1982(48), Whalley and Davidson, 1974(49), and Field and Davidson, 1980(50).
- The relationship of D_C with $D_{ax,L}$ can be attributed to that, the decrease in column diameter causes an increase in the gas hold-up which reduces the liquid circulation velocity and as result the mixing time increases leading to decrease in ($D_{ax,L}$). This result is in agreement with Pandit and Joshi, 1982(48).
- The decrease in ($D_{ax,L}$) with increasing (r/R) is due to that, the flow resistances to the gas phase in the direction of the flow increases so the gas gets redistributed in the radial direction. This uniform distribution of the dispersed phase minimizes the density gradient effects, which results in the reduction in the liquid recirculation. The reduction in the liquid

circulation velocities results in lower backmixing. This is in agreement with the findings of many researchers such as (Ueyama and Miyauchi, 1977(51), Riquart, 1981(10) and Kawase and Moo-Young,1989(52)). .

- The increase in ($D_{ax,L}$) with increasing (Z) is due to a decrease in bubble diameter which leads to a decrease in bubble rise velocity and consequently increasing the liquid circulation velocity (V_C) as shown in Equation (11), then increase ($D_{ax,L}$). These results are in agreement with Pandit and Joshi, 1982(48) and Krishn et al.,2000(9).

4-5. *Effect of Superficial Gas Velocity, Column Diameter, Radial Position, And Liquid Height on Liquid Radial Dispersion Coefficient*

Figs. (11) a, b, and c show the effects of gas velocity, column diameter, height of liquid (Z), and different radial position on radial liquid dispersion coefficient. Considering symmetric behavior around the axial center line, it can be deduced that, the radial dispersion coefficient increases with the increase in superficial gas velocity (U_g). This can be attributed to the effect of the average liquid circulation velocity (V_C) (the axial and radial components of velocity) which increases with the increase of (U_g) leading to an increase in liquid radial dispersion coefficient. This result is in agreement with (Pndit and joshi, 1982(48), Field and Davidson, 1980(50).

From Fig. (11) a, it can be seen that, the radial liquid dispersion coefficient ($D_{r,L}$) increases with increasing axial distance of probe's location from distributor (Z) . This increase in ($D_{r,L}$) is due to decrease in bubble diameter leading to a decrease in bubble rise velocity and consequently increasing the liquid circulation velocity (V_C) then increase ($D_{r,L}$). These results are in agreement with (Pandit and Joshi, 1982(48) and Krishn et al., 2000(9)).

Figure (11) b, shows the effect of radial probe's position (r/R) on the radial liquid dispersion coefficient ($D_{r,L}$), it can be seen that, the ($D_{r,L}$) values measured at $r/R = 0$ were in-between the values measured at $r/R = 0.45$ (i.e. in the upflow core zone of the reactor) and $r/R = 0.85$ (i.e. in the downflow region close to the wall). In contrast to the $D_{ax,L}$ measurements, the error in the $D_{r,L}$ values was much greater specially for the measurements at $r/R > 0$. The large deviations from the average in $D_{r,L}$ values were probably a reflection of the real random fluctuations in the $D_{r,L}$ caused by passage of large bubbles and continual fluctuations of the fuzzy interface between the upflow and the downflow zones. These results are in agreement with (Camacho et al., 2004(7)). It is evident that, the value of the radial dispersion coefficient of the liquid seem to depend on the radial component of the liquid velocity, which has higher value at the radial distance in between $r/R=0$ and $r/R=0.85$.

From Fig.(11) c, It can be seen that, $D_{r,L}$ increases with the increase of column diameter, the explanation for this increasing in ($D_{r,L}$)is that, the decrease in column diameter causes an increase in the gas hold-up which reduces the liquid circulation velocity and as result the mixing time increases leading to a decrease in ($D_{r,L}$). This result is in agreement with (Pandit and Joshi, 1982(48)).

4-6. Effect of Dr,L on Dax,L

Figs. (12) a and b show the effect of radial liquid dispersion coefficients on axial liquid dispersion coefficients. It can be seen that, the axial liquid dispersion coefficient increases considerably with the radial liquid dispersion coefficient. This increase in the axial liquid dispersion can be attributed to that, considering the radial liquid dispersion coefficient means that, the liquid circulation velocity will increase hence increase axial dispersion coefficient. These results are in agreement with (Camacho et. al., 2004(7)).

4-7. Effect of Superficial Gas Velocity, Column Diameter And Liquid Height on Mixing Time

Mixing time, which is a direct indicator of the mixing capacity of a reactor, was defined as the time required to attain a 5% deviation from complete homogeneity from the instance of tracer addition (7).

The mixing time can be calculated from fitting conductivity data as shown in Fig. (6)

From Figs. (13) a and b, it is clearly seen that the mixing time decreases with increase in (U_g). This can be attributed to the average liquid circulation velocity (V_C), which increases with increase in (U_g). These results are in agreement with (Pandit and Joshi, 1982(48)).

Beside (U_g), there might be an effect of the quantum of energy that is dissipated in liquid motion. Joshi (1980)(55) proposed the following equation to estimate liquid circulation velocity:

$$V_c = 1.31\left\{g\, D_C \left(U_g - £_g\, V^0{}_b\right)\right\}^{1/3} \tag{11}$$

this equation contains the term (U_g - ε_g $V^0{}_b$) which is equivalent to the quantum of energy that is supplied or available for the liquid motion. This quantity alter due to the variation in ε and V_b, relative to the variation in (U_g). It is simply concluded that when (V_C) is lower, the mixing time is higher. In other words, one can say that due to the relatively lower (V_C) values than a lower energy is available for the liquid motion, that causes an increase in mixing time.

Figs. (14) a and b show the effect of column diameter on mixing time. It can be seen that, the mixing time decreases with the increase in column diameter. This decrease in mixing time is due to the increase in liquid circulation velocity with the increase in column diameter.

From Figs. (13) a, b, and c, it is clearly seen that the mixing time increases with the increase in (r/R). This was because of the local liquid velocity had its maximum value at the center of the column while it decreases with the increase in (r/R). This is in agreement with the findings of many researchers such as (Riquart, 1981(10), Ueyama and Miyauchi, 1977(51), and Kawase and Moo-Young,1989(52)).

5. MATHEMATICAL CORRELATION

Whalley and Davidson, 1974(49), reported the relationship between the mixing time and dispersion coefficient. For relatively low values of U_L
the following equation is appropriate.

$$D_{ax,L}=0.5\ H^2_d/\theta_{0.3} \qquad (12)$$

Field and Davidson (1980)(50) reported similar equation for longitudinal dispersion coefficient with different constants (0.1 instead of 0.5). To correlate the experimental data of this study similarly using a nonlinear regression technique via a Statistica software, the value was equal to 0.1115 with $R^2 = 0.93$ for 15 cm column diameter and 0.1499 with $R^2 = 0.97$ for 30 cm column diameter as shown below:

For 15 cm column diameter

$$D_{ax,L} = 0.11 \times H_d^2 \big/ \theta_{0.3} \qquad (13)$$

For 30 cm column diameter

$$D_{ax,L} = 0.15 \times H_d^2 \big/ \theta_{0.3} \qquad (14)$$

6. COMPARISON BETWEEN MEASURED $D_{AX.L}$ AND AVAILABLE LITERATURE CORRELATIONS

The $D_{ax,L}$ values measured in the present work were compared with the predictions of the literature correlations shown in Table (3). This comparison is shown in Fig. (15).

7. CCNCLUSIONS

From the present work, the following conclusions are extracted:

1. In a bubble column the over all gas holdup ascending with increasing superficial gas velocity and descending with increasing column diameter.
2. The bubble size increases with increasing superficial gas velocity and column diameter but it decreases as it rises up through the column.
3. The axial and radial liquid dispersion coefficients increase with increasing both superficial gas velocity and column diameter.
4. The axial and radial liquid dispersion coefficients increase when the bubble rise up through the liquid.
5. Based on the present work, a mathematical correlation is formulated to predict the axial dispersion coefficient.

$$D_{ax,L} = 0.15 \times H_d^2 \Big/ {}_{0.3}$$ for 30 cm column diameter and

U_g=1 – 10 cm/s with R^2=0.97

$$D_{ax,L} = 0.11 \times H_d^2 \Big/ {}_{0.3}$$ for 15 cm column diameter and

U_g=1 – 10 cm/s with R^2=0.93

6. The value of the radial dispersion coefficient was typically about 1% of the value of the axial dispersion coefficient under the given conditions of operation.
7. It has been observed that mixing time decreases with an increase in both superficial gas velocity and column diameter.
8. The mixing time increases with increase in dimensionless radial distance r/R.
9. The axial liquid dispersion coefficient ($D_{ax,L}$) increases with decrease mixing time (t_m).

NOMENCLATURE

Symbol	***Definition***
Ac	Column cross section area (m^2)
$D_{ax,L}$	Axial liquid phase dispersion coefficient (cm^2/s)
$D_{r,L}$	Radial liquid phase dispersion coefficient (cm^2/s)
D_C	Column diameter (cm)
d_b	Bubble diameter (m)
d	Hole diameter of gas distributor (m)
$d_\circ$	Effective hole diameter in gas distributor (m)

g	Acceleration due to gravity (cm/s^2)
H_d	Final liquid height with gas (cm)
H_o	Initial liquid height without gas (cm)
h	Distance between tracer point injection and probe (cm)
j_{GL}	Drift flux velocity (cm/s)
N	Number of holes in distributor
r	Radial position inside the column (cm)
R	Radius of the column (cm)
t_m	Mixing Time (s)
U_g	Superficial gas velocity (cm/s)
U_{trans}	Superficial gas velocity at transition regime (cm/s)
V_b	Bubble rise velocity (cm/s)
V_{small}	Small bubble rise velocity (cm/s)
VL(0)	Center line Liquid circulation velocity (cm/s)
VL(r)	Radial liquid circulation velocity (cm/s)
V_C	Average Liquid circulation velocity (cm/s)
Z	Axial position (cm)

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Table (1). Liquid physical properties and selected operating conditions

Ug (Air) , cm/s	1-10
Pressure = atm	1
Liquid mode	Batch
Distilled water	$\rho_\ell = 1$ g/cm^3, $\mu_\ell = 10^{-4}$ g/cm.s , $\sigma_\ell =$ 72 dyn/cm
Temperature, °C	20
Initial Liquid height , H_o cm	130
Volume of tracer for each run, ml	400 for 30 cm column diameter, and 93 for 15 cm column diameter.

Table (3). Correlations of $D_{ax,L}$ cited in literature

Correlation	Reference
$D_{ax,L} = 0.31\ D_C\ V_L(0)$	9
$D_{ax,L} = 0.1 \times H_d^2 \big/ \theta_{0.3}$	49
$D_{ax,L} = 0.5 \times H_d^2 \big/ \theta_{0.3}$	50
$D_{ax,L} = 0.343\ (D_C)^{4/3} (g\ U_g)^{1/3}$	54

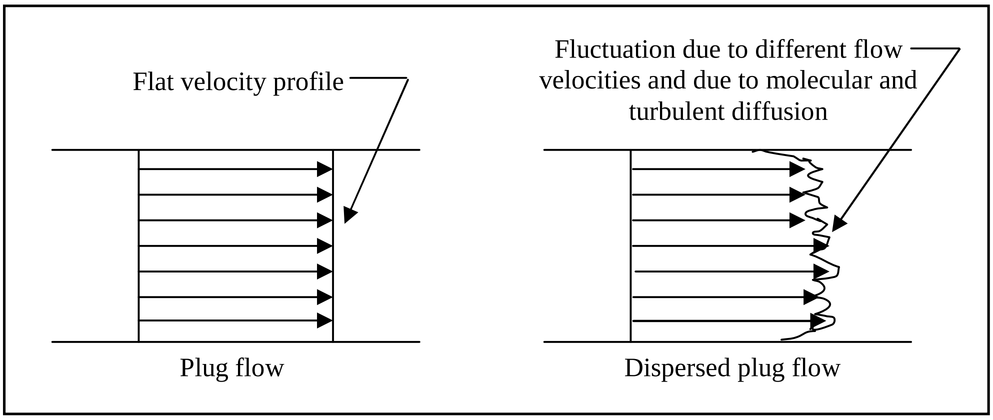


Fig.1. Representation of the dispersion (dispersed plug flow) model$^{(8)}$.

Fig. (2). General View of Experimental Apparatus.

1	**Air-compressor**
2	**Check valve**
3	**Air-filter**
4	**Needle valve**
5	**Air-flometers**
6	**on-off valve**
7	**Discharge valve**
8	**Gas distributors**
9	**Conductivity Probes**
10	**Interface**
11	**Pc (PIII)**
12	**Digital Camera**
13	**Conductivity meter**

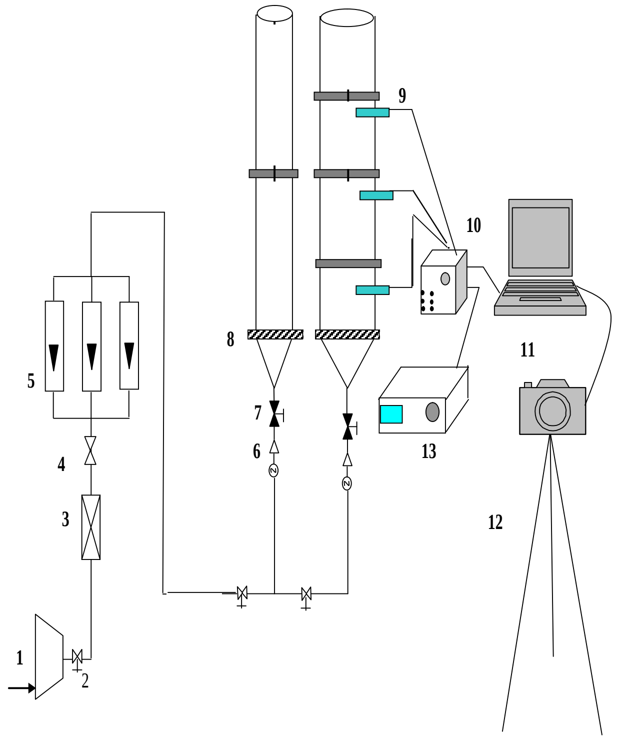


Fig. (3). Schematic diagram of the experimental apparatus.

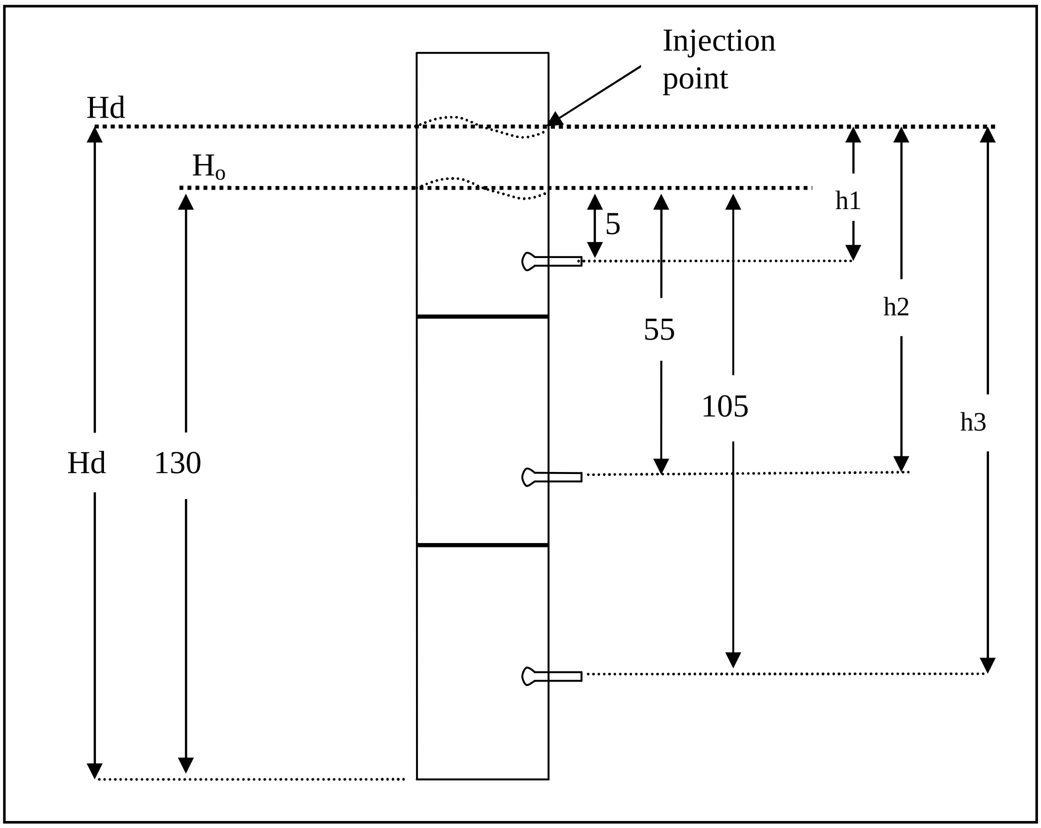
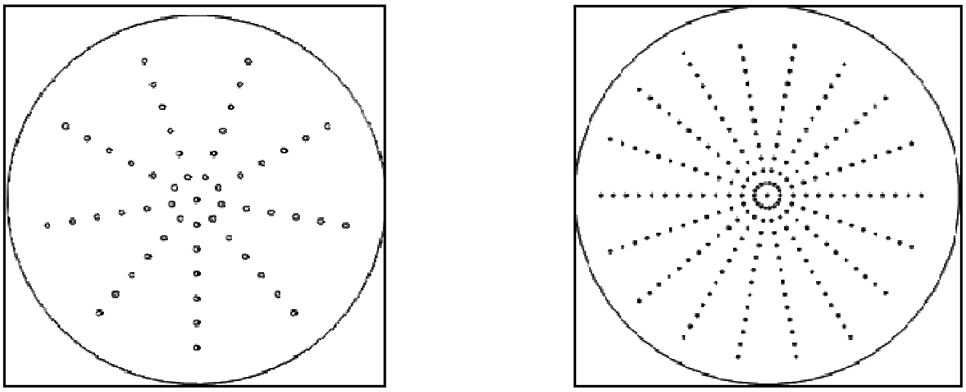


Fig. (4) Details of experimental setup used during tracer experiments (all dimensions in cm)



DC = 15 cm No. of holes = 5٢
Distance between two holes = 1 cm

DC =30 cm No of holes = 207
Distance between two holes = 1 cm

Fig. (5) Dimensions of the gas distributors used in the two bubble columns.

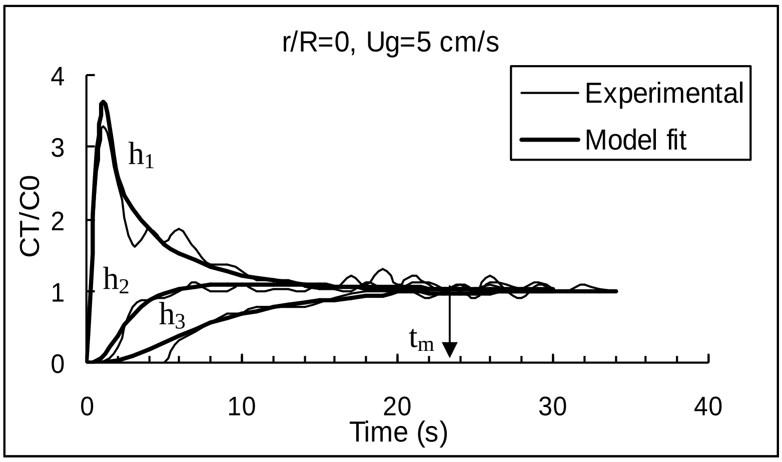


Fig. (6) Typical conductivity responses of differe probes for 15 cm column diameter.

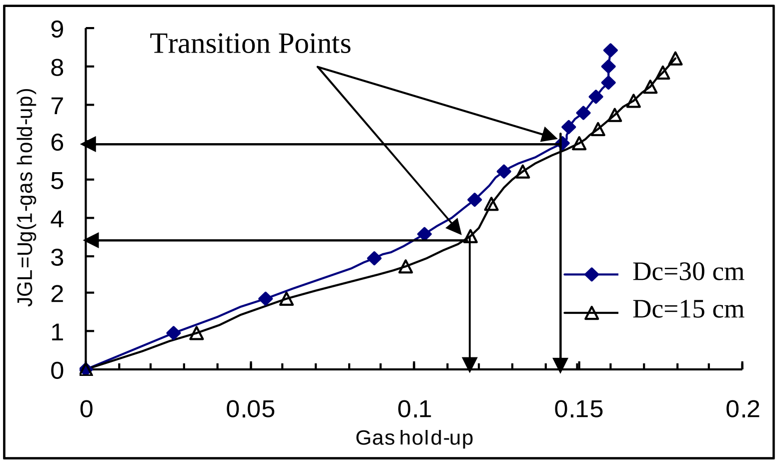


Fig. (7). Determination of flow regime transition point using drift flux model concept.

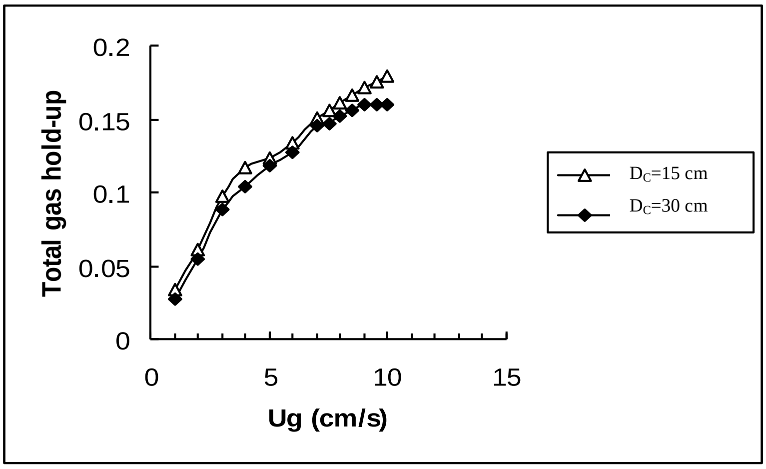


Fig. (8). Effect of superficial gas velocity and column diameter on gas hold-up.

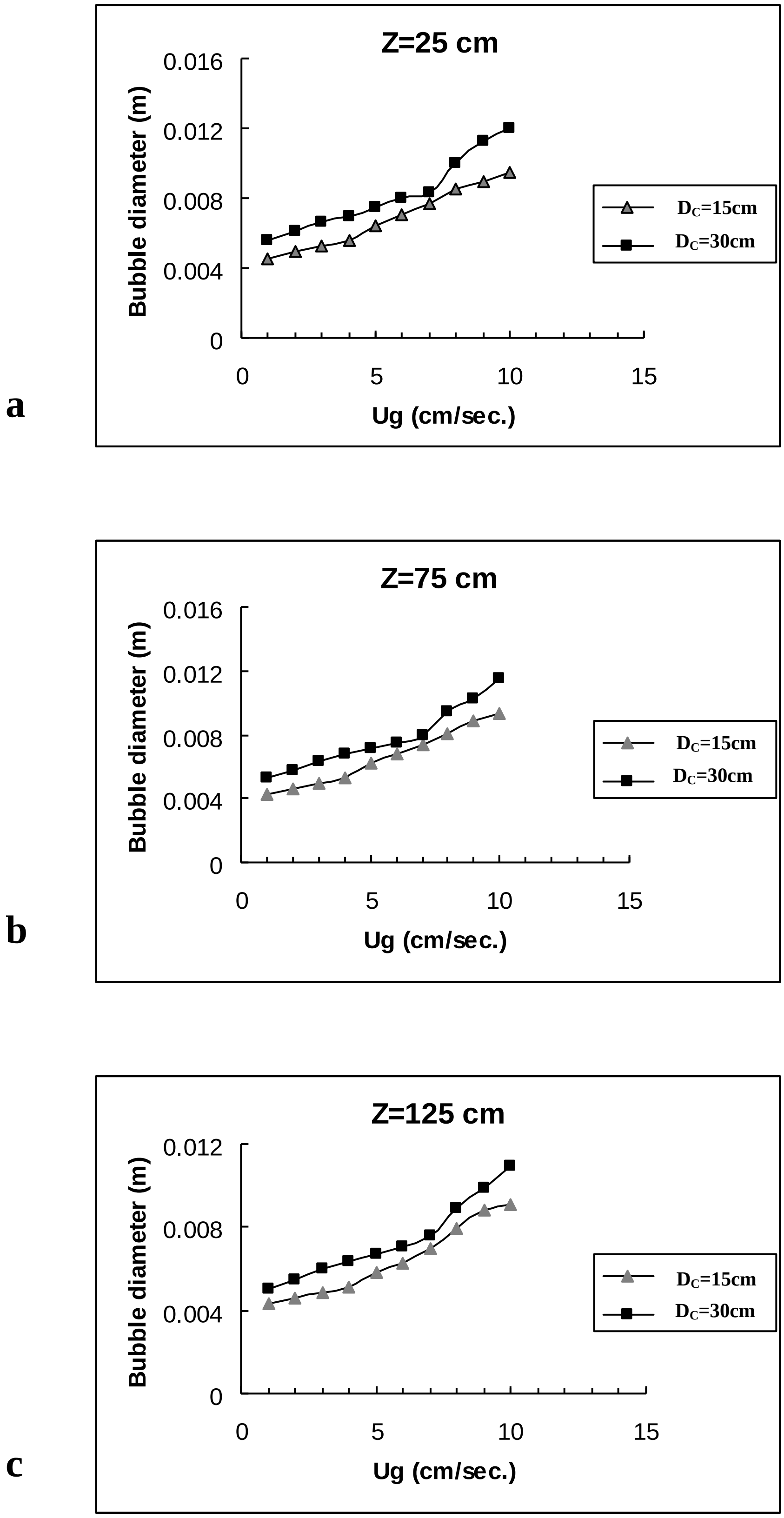


Fig. (9). Effect of gas velocity, height of liquid, and column diameter on bubble diameter.

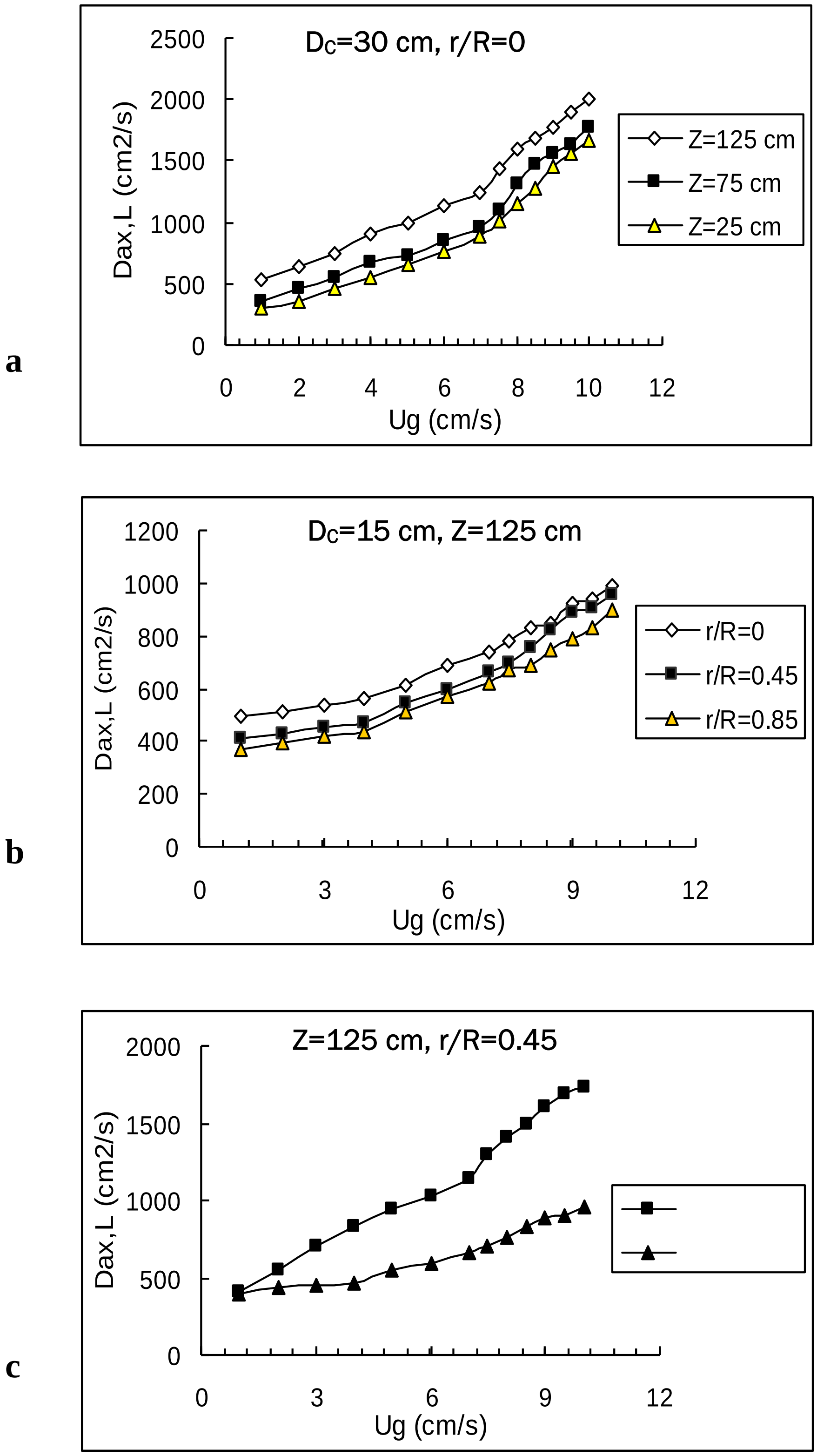


Fig. (10). Effect of gas velocity, column diameter, radial position, and height of liquid on axial dispersion coefficient

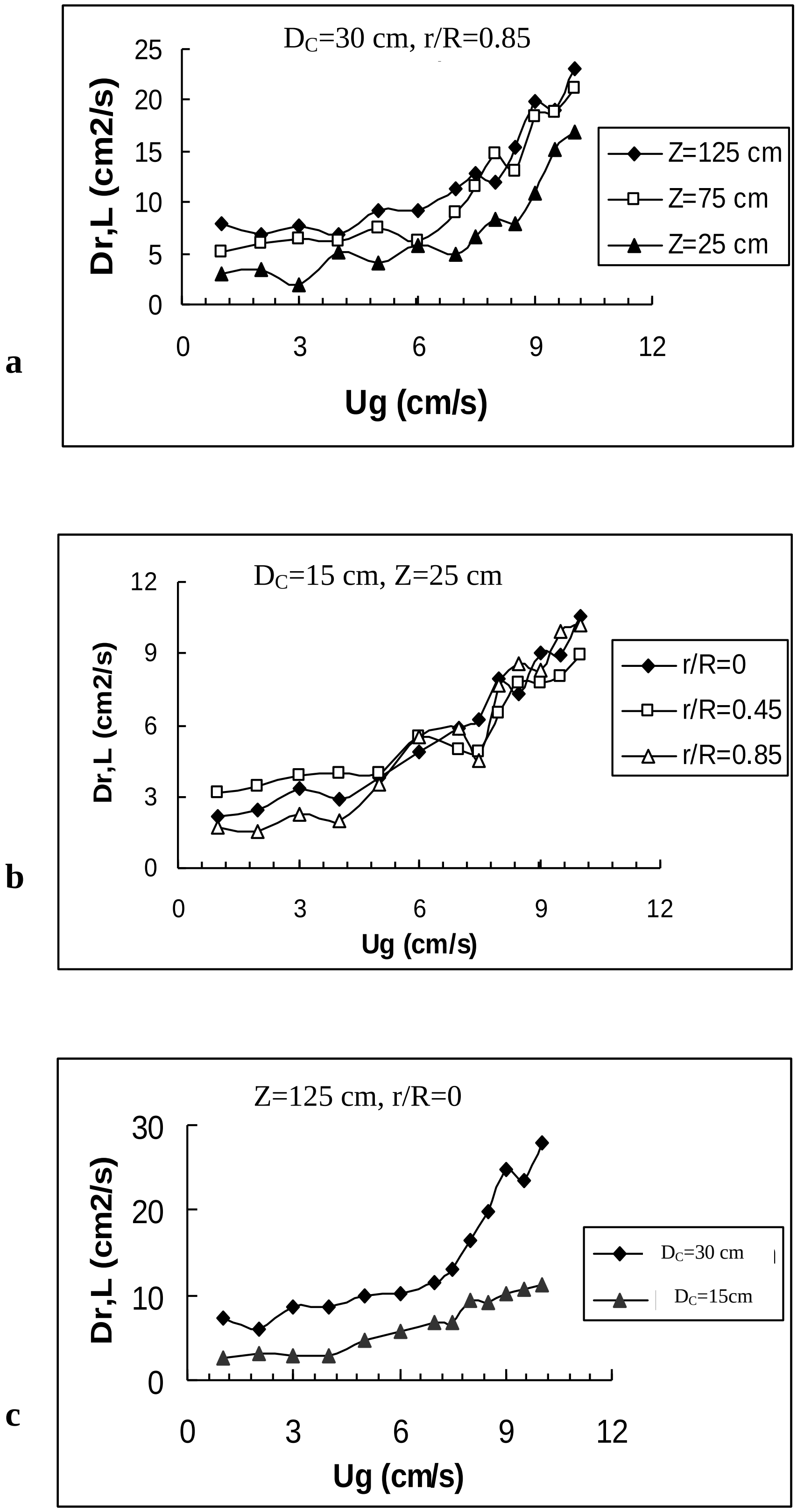
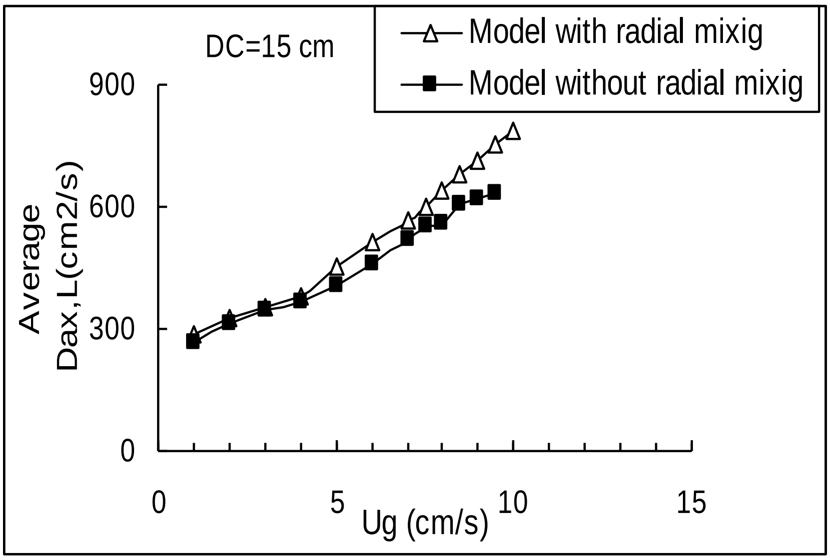
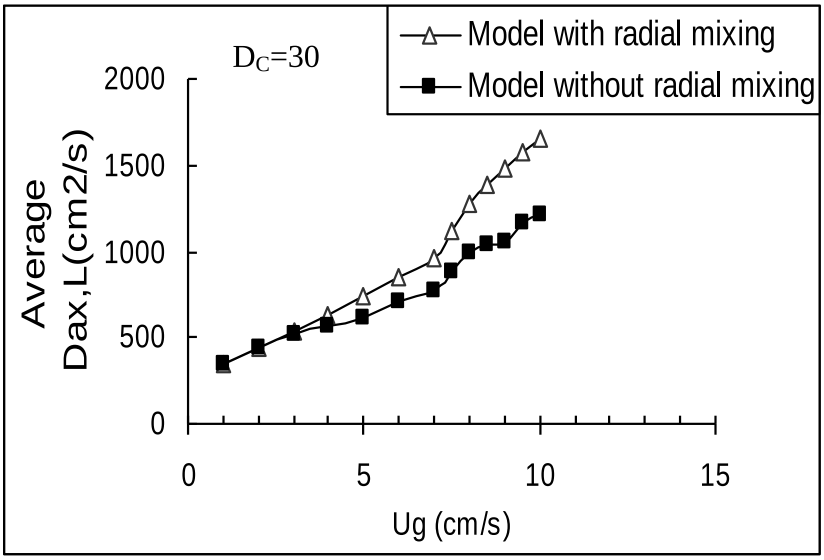


Fig. (11). Effect of gas velocity, column diameter, radial position, and height of liquid on radial dispersion coefficient



a



b

Fig. (12). Effect of radial liquid dispersion coefficient ($D_{r,L}$) on axial liquid dispersion coefficient ($D_{ax,L}$) for different column diameter.

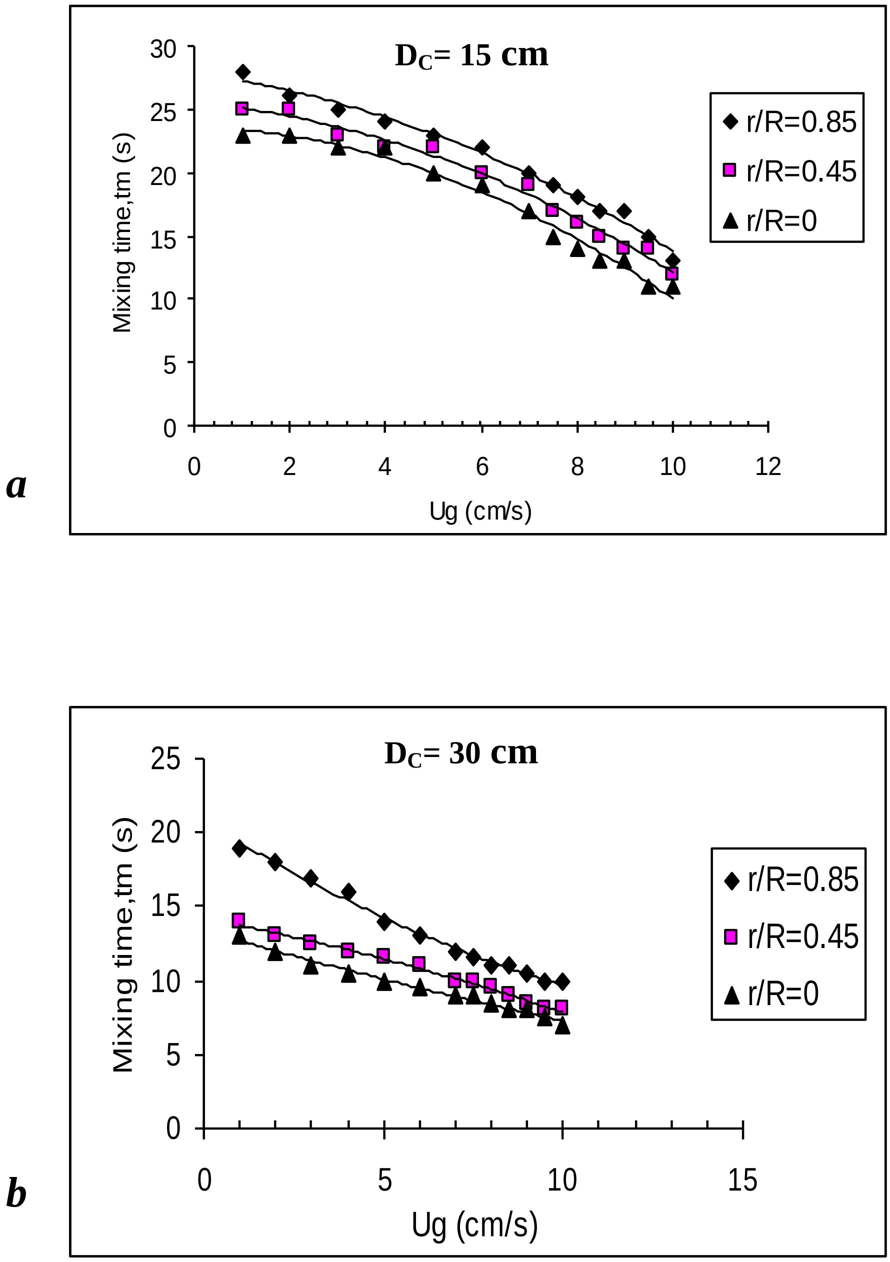


Fig.(13). Effect of gas velocity and radial position on mixing time for different column diameter

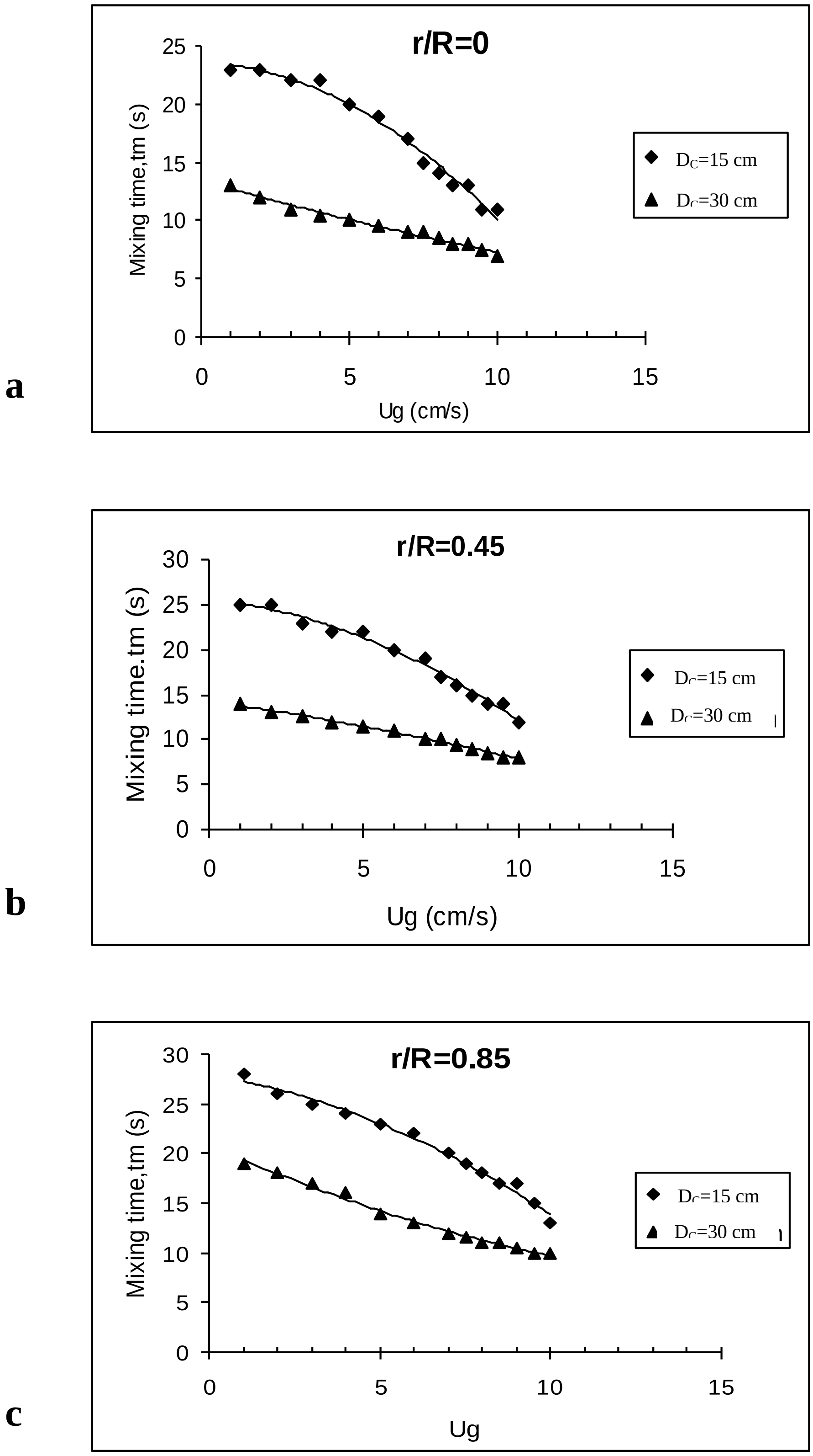


Fig. (14). Effect of column diameter on radial dispersion coefficient for different radial position.

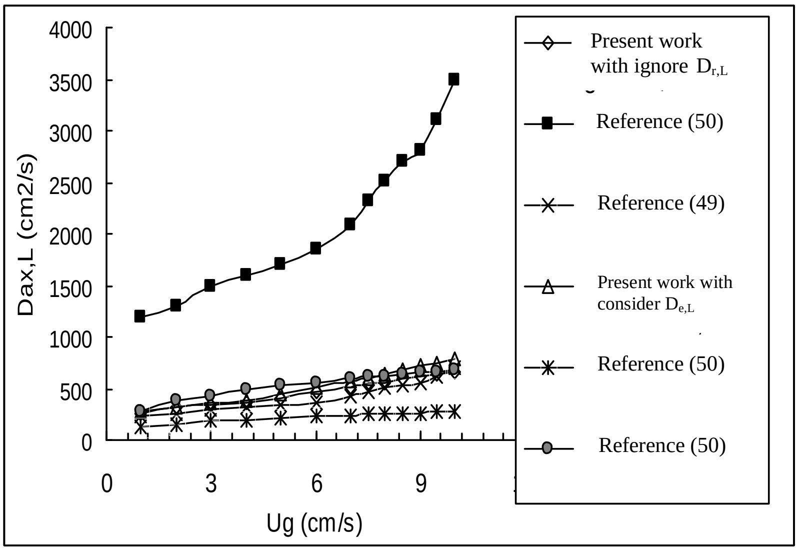


Fig.(15, a). Comparison between measured ($D_{ax,L}$) with available literature correlations for 15 cm column diameter.

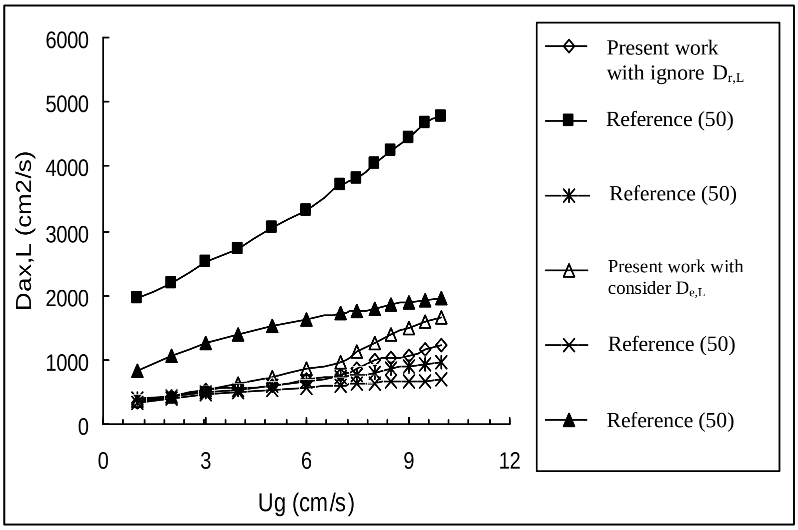


Fig.(15, b). Comparison between measured ($D_{ax,L}$) with available literature correlations for 30 cm column diameter.

دراسة تجريبية للسائل المشتت في العمود الفقاعي

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الخلاصة

إنّ الهدف الرئيسيَ لهذه الدراسةِ هوَ تحرّي تأثيرَ قطرِ العمودِ وسرعةِ الغازِ السطحيةِ على معاملاتِ تشتت السائلِ (المعاملات المحوريةِ والقطريةِ) و زمن الخلط و المحتوى الحجمي للغار و سلوكية الفقاعةِ (قطر الفقاعةِ وسرعة إرتفاعها). تشتت السائلِ و المحتوى الحجمي للغار و سلوكية الفقاعةِ (قطر الفقاعةِ وسرعة إرتفاعها) قِيستْ لنظامِ (ماءِ _ هواء) في عمودين لقطرين مختلفينِ (١٥ و ٣٠ سنتيمتر) و بمدى سرعة غاز سطحيةَ يتراوح بين(١-١٠ سنتيمتر/ثا) يغطّي كلتا أنظمة التدفقِ المتجانسةِ والمضطربة. إرتفاع السائلِ كان ١٣٠ سنتيمتراً للعمودين. معاملات تشتت السائلِ (المعاملات المحوريةِ والقطريةِ) و زمن الخلط قِيستْ في مواقعِ محورية وقطرية مُخْتَلِفة داخل الأعمدةِ (Z = ٢٥، ٧٥، ١٢٥ سنتيمتراً و r/ R= ٠، ٠.٤٥، ٠.٨٥) بينما سلوكية الفقاعةِ (قطر الفقاعةِ وسرعة إرتفاعها) قِيستْ في ثلاثة مواقع محورية (Z=٢٥، ٧٥، ١٢٥ سنتيمتراً).

هذا العملِ يُقدّمُ تحليلا نظريا يأْخذُ في الحسبان الإختِلاط في أعمدةِ الفقاعةِ المحوريةِ والشعاعيةِ بشكل آني. هذا التحليلِ أستعمل لحِساب معاملاتِ التشتت المحوريةِ والشعاعيةِ مِنْ بياناتِ إستجابة الراسم المدروسةِ بشكل تجريبي. النموذج الكامل لتشتت السائلِ أستَعمل لمعرفة طبيعة الخلط المحوريِ والشعاعيِ في عمودِ الفقاعةِ بشكل آني.

التقنيات التجريبية ذات العلاقة لقياس تشتت السائلِ نوقشَ بشكل شامل مِن قِبل Deckwer (1992) [6]. معاملات التشتت تَحْسبُ بواسطة قياس استجابة (تركيزِ _ زمن) المدروس و ذلك باستعمال راسم غير ماصّ خامد غير تفاعلي في المفاعلِ. نموذج التشتت استعمل لمُلائَمَة بياناتِ الرَدِّ المدروسةِ. إنّ قيمة معاملَ التشتت القطري كانت تساوي ١% من قيمة معاملِ التشتت المحوري.

مِنْ البياناتِ التجريبيةِ وجد بأن قيمة معاملِ التشتت القطري ومعامل التشتت المحوري و محتوى الغاز الحجمي وقطر الفقاعة وسرعة الفقاعة تزداد بزيَاْدَة سرعةِ الغازِ السطحيةِ. تؤكّد النتائِج التأثير الهامَّ لقطرِ العمودَ على محتوى الغاز

الحجمي حيث وجد أن محتوى الغاز الحجمي يقل بزيَادة قطرِ العمودِ، بينما معامل التشتت القطري و معامل التشتت المحوري و قطر الفقاعةِ وسرعة الفقاعة تزداد بزيادَة قطرِ العمودِ.

وقد تم إجراء تحليلات إحصائية للحصول على معادلة عامة لمعامل تشتت السائل المحوري ($D_{ax,L}$) كدالة لزمن الخلط ($\theta_{0.3}$) و إرتفاع التشتت (H_d) التي تم دراستها عملياً، و كانت المعادله بالصيغة ألصيغة التالية:

للعمود ذي القطر ٣٠ سم $$D_{ax,L} = \left(\frac{0.15 \times H_d^{2}}{\underset{0.3}{-}} \right)$$

للعمود ذي القطر ١٥ سم $$D_{ax,L} = \left(\frac{0.11 \times H_d^{2}}{\underset{0.3}{-}} \right)$$