Tikrit Journal of Eng. Sciences/Vol.18/No.3/September 2011, (13-22)

Gas Holdup and Liquid-Phase Dispersion in Packed Bubble Columns

Dr. Burhan Sadeq Abdulrazzaq
Lecturer
Department of Chemical Engineering-University of Tikrit,

Abstract
The gas holdup and liquid phase axial dispersion coefficient are measured in two semi batch packed bubble columns, 10 and 15 cm diameter for an air–water system, at atmospheric conditions. The experiments were carried out using a transient method (the tracer response method). The dispersion coefficient was obtained by adjusting the experimental profiles of tracer concentration with the predictions of the model. Experiment results of packed bubble column, shows a considerable reduction of the backmixing. The investigations have been carried out using RTD measurements and the backmixing is usually characterized by the axial dispersion coefficient obtained from the one-dimensional axial dispersion model. Also, a decrease in superficial gas velocity reduces the liquid backmixing. It is observed that the liquid circulation comprises an upward flow in the column core and a downward flow along the wall. It also seen that the transition from the bubbly flow to the pulsation flow regime occurred at 5-6 cm/s superficial gas velocity.

Keywords: Packed Bubble columns, Gas holdup, Liquid dispersion, Liquid circulation

محتوى الغاز وتشتت الطور السائل في الاعمدة الفقاعية المحشوة

الخلاصة

تم قياس محتوى الغاز (Gas holdup) ومعامل التشتت المحوري (Axial dispersion coefficient) للطور السائل للاثنين من الاعمدة الفقاعية المحشوة ذات الاقطار (10 ، 15 سم) باستخدام نظام ماء – هواء . نفذت التجارب باستخدام طريقة الانتقال (Transient method) اي طريقة استجابة الدليل (tracer response experiments) . وتم الحصول على معامل التشتت عن طريق ضبط الاشكال المختبرية (experimental profile) لتركيز الدليل مع توقعات النموذج . بينت النتائج بان الخلط الرجوعي (backmixing) ومعامل التشتت المحوري للطور السائل ينخفضان بشكل كبير في الاعمدة المحشوة مقارنة بالاعمدة الفقاعية غير المحشوة وان هذه التحقيقات تمت بقياس دالة زمن البقاء (RTD). ومن النتائج وجد بان الخلط الرجوعي للطور السائل يقل بانخفاض سرعة الغاز (superficial gas velocity). كما لوحظ بان توزيع السائل (circulation velocity) يتضمن تدفق صاعدا في مركز العمود الفقاعي وتدفق تحتي على طول جدار العمود الفقاعي. كما لوحظ بان الانتقال من الجريان الفقاعي الى الجريان النبضي حدث بسرعة غاز بين 5–6 سم/ثا.

الكلمات الدالة : الاعمدة الفقاعيةالمحشوة ، محتوى الغاز ، تشتت السائل ، توزيع السائل

Notation

C_0 Final concentration achieved when $t = \infty$, kg/m^3
C_L Liquid concentration, kg/m^3
$D_{ax,L}$ Liquid phase axial dispersion coefficient , m^2/s
D_T Column diameter, m
H Total height of the column, m
H_0 Total liquid height in the column, m
Hd Dispersion height, m
L Total liquid height in the column, m
t Time, s
U_g Superficial gas velocity (m/s)
z Axial coordinate, m
ε_G Fractional gas hold-up

Introduction

Bubble columns are frequently used in the chemical industry to perform gas liquid reactions. Although this kind of equipment has been extensively investigated during the last decades, the number of published articles regarding the hydrodynamics in packed bubble columns is not as substantial. These works have considered, in particular, the influence of gas velocity, different packing, packing sizes, liquid flow, tower diameter and bed height.

There are some discrepancies in the effects of the gas velocity on the liquid dispersion coefficient. According to Shah et al.[1], it is generally believed that an increase in gas velocity increases the liquid dispersion coefficient.

In packed bubble columns, the gas holdup has been investigated by several authors for different kinds of packings (Stiegl and Shah[2]; Abraham and Sawant[3]; Niranjan and Pangarkar[4]). It is a well known fact that the gas holdup increases with increasing gas flow and that the packing size influences the gas holdup considerably. Several empirical correlations have been developed, but it is obvious that there is currently no universal correlation for predicting the gas holdup in packed bubble columns.

Turpin and Huntington[5] have identified three different flow regimes in packed bed reactors: the bubble flow, the pulse flow and the spray flow regimes. The bubble flow regime is characterized by individual gas bubbles flowing in an unbroken stream upwards in the bubble column. The pulse flow results in an increase in the gas flow to greater than 7–10 cm/s, and alternate portions of more dense and less dense phases pass through the column. The spray flow is where the gas is the continuous phase and the liquid acts as the dispersed phase.

Since the axial dispersion model characterises the backmixing by only a single parameter, its simplicity made it the most widely used representation of the non-ideal mixing behaviour for each phase in bubble column reactors. The time variation of the liquid phase concentration of a tracer is given by Fick's law:

$$\frac{\partial C_L}{\partial t} = D_{ax,L} \frac{\partial^2 C_L}{\partial z^2} \qquad \text{- - - - - (1)}$$

Where the axial dispersion coefficient ($D_{ax,L}$) is characterized as a unique parameter for the degree of backmixing during process. The term 'axial' is used in order to mark clearly the difference between the mixing in the direction of flow and the mixing in the lateral or radial direction.[6] In bubble columns these two quantities are quite different in magnitude, so that the axial dispersion coefficient is significantly exceeding the radial dispersion coefficient [7].

Siemes and Weiss (1957)[8] were the first using the pulse method for measuring the dispersion in bubble

columns. They were followed by Ohki et al., (1970)[9], Hikita et al., (1974)[10]. The partial differential equation based on one-dimensional model has been solved analytically by Siemes and Weiss (1957)[8]. The boundary condition for equation (1) are:

$$\frac{\partial C_L}{\partial z} = 0 \quad \text{at} \quad z = 0 \quad \text{and} \quad z = L$$

and the initial condition are

$C_L(z,0) = C_0$ for $0 \le z \le \lambda$
$C_L(z,0) = 0$ for $z \ge \lambda$

Where λ is the height at which tracer is injected. The solution of Eq.(1) under consideration of the corresponding boundary conditions gives[21]:

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

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

Several authors state that increasing the gas velocity increases the dispersion coefficient (Stiegel and Shah[2] and Hofman[11]). Stiegel and Shah[2] propose a correlation where the dispersion coefficient varies as $U_G^{0.16}$ for packing sizes of approximately 4 mm. For a bubble column packed with glass cylinders with a diameter of 3.8mm and a length of 4.8 mm, Gelder and Westerterp[12] propose the exponent of U_G to be 0.313. Niranjan and Pangarkar[4] concluded that for packings with nominal dimensions of 25mm and above, the influence of the gas velocity on the dispersion is considerable: the dispersion coefficient in this case varies approximately with $U_G^{0.3}$, while for smaller packings, the dispersion coefficient is almost independent of the gas velocity. Niranjan and Pangarkar[4] explained this by the fact that smaller packings suppress the bulk, and the probable cause of mixing in smaller packings is micro-turbulence. They suggest that for larger packings, mixing predominantly occurs by liquid circulation. Moreover, results obtained by Campos and Guedes de Carvalho[13] and Carleton et al.[14] give no rise in the dispersion when increasing the gas velocity.

The aim of the present study is to determine the gas holdup, the flow regimes and the liquid axial dispersion coefficient in a packed bubble column and unpacked bubble column experimentally.

Experimental work

The experiments were carried out in two batch type bubble columns with internal diameters of (10, 15) cm and (174, 160) 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). The columns were open at the top; hence the pressure corresponded with ambient conditions. Both of the bubble columns were packed with glass cylinders with a diameter of 1.0 cm and a length of 1.5 – 2 cm. Perforated plate spargers of identical design were used in columns to distribute the gas phase. The distributor plates were made of plastic plate with holes of 2 mm diameter. Air was used as the gas phase and tap water as liquid phase. The gas was introduced at the bottom of the columns. The experiments were carried out at various gas velocities, carefully adjusted and controlled using a 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.

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 was 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 meter (of Philips type), of range 100 μs to 1000 ms which provide a reading in units of conductance. The meters were connected with an interface to a personal computer.

Tracer was injected as a pulse input. Local changes in tracer concentration were displayed and saved continuously on PC. Three electric conductivity probes were inserted 2 cm away from the inside wall, located at different heights as shown in Fig. (2), each of them was connected to PC via interface circuit. The distance from the injection to the measuring points, L_1, L_2, L_3 and Hd are given in Table (1).

Time for each experiment was about 10 min to reach final concentration in the column.

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

Fig. (3) shows typical transient tracer concentrations from the 10 cm unpacked bubble column, operated at 4.68 cm/s superficial gas velocity. These signals were fitted using the analytic solution to the diffusion equation presented in [15]. In this way, for a given experiment, only one variable was adjusted, i.e. the axial dispersion coefficient, $D_{ax,L}$.

Results and Discussion

The gas holdup was measured using an overflow technique. The bubble column was filled with water to the same level as one of the sample points. The gas was introduced and, by measuring the volume of the entrained liquid, the gas holdup was calculated. Fig. (4) shows the gas holdup versus the superficial gas velocity in two different packed bubble columns.

As expected, the gas velocity increases the gas holdup. It is well known that packing prevents coalescence, and that there is no formation of larger bubbles as there is in an empty column operating in the churn-turbulent regime. The figure also shows that there is no significant difference in gas holdups due to the different bubble column diameter.

In a packed bubble column, the maximal bubble size is determined primarily by the packing size. Therefore, it could be suspected that the diameter of the column is of little importance. It is apparent that no correlation in the literature takes the column diameter into consideration. Niranjan and Pangarkar[4] reported identical gas holdups for two different columns of diameter 0.2 and 0.38 m. Sahay and Sharma[16] also used two different column diameters, 0.1 and 0.2 m for a variety of packing. These results are in agreement with the results obtained in this work.

Two different flow regimes were observed visually at atmospheric conditions: namely, bubble and pulsation flow. The pulses, which started at the lower part of the bed and moved upwards, were easily detected. These pulses form the pulsation flow region. In Fig. (5), the pulse frequencies have been plotted against the superficial gas velocity. The values of the pulse frequencies were determined by calculating the number of pulses during 1 min.

It can be seen that the flow regime transition point occurs at approximately 5–6 cm/s. It is interesting to note that the transition point in a packed bubble column occurs at a higher gas velocity compared to an unpacked bubble column, where the churn turbulent flow regimes starts at approximately 4 cm/s.[6] This phenomenon can be explained by the fact that the packing effectively prevents the gas bubbles from coalescing.

The dispersion coefficients are obtained by fitting the experimental data with theoretical predictions of Eq. (2). One example for fitting the measured tracer response is given in Fig. (3). clearly, the model simulates the measured data closely.

The measured axial dispersion coefficients of air-water system are compared in Fig. (6) with available literatures data at same operating conditions for the both unpacked columns.[17-26]

In Fig. (7), the results of the dispersion measurements in packed bubble column are shown. It is evident that the dispersion coefficient depends on the gas velocity.

Figure (8) shows the electrical conductivity probe responses in bubble column and packed bubble column at superficial gas velocity equal to 4.68 cm/s. it seem that the addition of packing to the bubble column reduces the backmixing (reduces the axial dispersion coefficient). In case of bubble column, intense liquid flow is developed which is centrally upward and downwardly near the wall. As a result, bubble rise faster and this leads to increase axial dispersion coefficients. In the presence of packing, the liquid circulation is substantially reduced, which inhibits the bubble rise velocity very much that consequently increase the residence time for gas bubbles, higher values of effective interfacial area and lower axial dispersion coefficients in packed bubble column.

In order to investigate the validity of the dispersion model in a packed bubble column, further dispersion experiments were conducted. Conductivity probes are fixed at approximately half the bed height at three different radial positions inside the bed: 5, 25 and 50 mm from the column wall (r/R = 0.9, 0.5, 0). The results obtained from these trials are shown in Figs. 9 and 10. In Fig. 9, the superficial gas velocity was approximately 0.87 cm/s and the bubble column was operated in the bubble flow regime. The figure illustrates that the distribution of the tracer is dependent on the radial position. The tracer near the column wall is distributed faster than the tracer in the middle of the column; the tracer at r/R = 0.5 is distributed with a velocity between these two. Such trends can only be explained by the fact that the liquid flows upwards in the centre and downwards near the wall region. This situation can also be compared to the conditions existing in empty bubble columns where the time averaged liquid velocity profile shows a comparable rising liquid flow in the centre and a descending flow at the periphery.

It is apparent from the figures that the concentration gradients across the column cross-section decrease gradually when the

gas velocity is increased. At a superficial gas velocity of approximately 5.36 cm/s, there is a transition from bubble flow to pulsation flow. When the pulses rise upwards along the bubble column axis, each pulse increases the exchange of liquid elements in the radial direction.

Conclusions

The main results presented in this work are:

1- It is observed that the gas holdup increases at increased superficial gas velocity. The diameter of the packed bubble column does not influence the gas holdup.
2- Two flow regimes are observed: bubble and pulsation flow. The pulsation flow regimes start at Ug approximately 5-6 cm/s.
3- In both regions, the one dimensional axial dispersion coefficient increases with increasing gas velocity.
4- There is a non-uniform liquid velocity distribution in the radial direction in a packed bubble column. This liquid circulation comprises an upward flow in the column core and a downward flow along the wall.

References

1- Shah, Y.T., Stiegel, G.J., Sharma, M.M., "Backmixing in Gas-Liquid Reactors", AIChE J. 24 (1978) 369-400.
2- Stiegel, G.J. and Shah, Y.T., "Backmixing and Liquid Holdup in a Gas Liquid Co-Current Up flow Packed Column", Ind. Eng. Chem. Proc. Des. Dev. 16 (1977) 37.
3- Abraham, M. and Sawant, S.B., "Hydrodynamic and Mass Transfer Characteristics of Packed Bubble Columns", Chem. Eng. J. 43 (1990) 95-105.
4- Niranjan, K. and Pangarkar, V.G., "Gas Holdup and Mixing Characteristics of Packed Bubble Columns", Chem. Eng. J. 29 (1984) 101-111.
5- Turpin, J.L. and Huntington, R.L., "Prediction of Pressure Drop for Two-Phase, Two Component Co-current flow in Packed Beds", AIChE J. 13 (1967) 1196-1202.
6- Krishna, R., Urseanu, M.I., van Baten, J.M., Ellenberger J., "Liquid Phase Dispersion in Bubble Columns Operating in the Churn-Turbulent Flow Regime", Chem. Eng. J., 78, 43–51, (2000).
7- Rubio, F. C. , miron, A. S. , Garcia, M. C. , Camacho, F. G. , Grima, E. M. , Chisti, Y. "Mixing in Bubble Columns: A New Approach for Characterizing Dispersion Coefficients", Chem. Eng. Sci., 59, 4369-4376, (2004).
8- Siemes, W. and Weiss, W., 1957, Flfissigkeitsdurchmischung in engen Blasens~ulen. Chem.-lngr.-Tech. 29, 727-732, (1957). (sited in Ohki et al., 1970)
9- Ohki, Y. and Inoue H., "Longitudinal Mixing of The Liquid Phase in Bubble Columns", Chem. Eng. Sci., 25, 1-16, (1970).
10- Hikita, H. and Kikukawa, H.,"Liquid-Phase Mixing in Bubble Columns: Effect of Liquid Properties", Chem. Eng. J., 8, 191-197, (1974).
11- Hofmann, H., "Der derzeitige stand bei der vorausberechnung der verweilzeitverteilung in technischen reaktoren", Chem. Eng. Sci. 14 (1961) 193-208.
12- Gelder, K.B., and Westerterp, K.R., "Residence Time Distribution and Hold-up in a Co-current Upflow Packed Bed Reactor at Elevated Pressure", Chem. Eng. & Technology 13 (1990) 27-40.

13- Campos, J.B.L.M. and Guedes de Carvalho, J.R.F.A., "A new Experimental Technique to Study Backmixing in Packed Bubble Columns", Chem. Eng. Sci. 47 (1992) 4063-4067.
14- Carleton, A.J., Flain, R.J., Rennie, J., Valentin, F.H.H., "Some Properties of a Packed Bubble Column", Chem. Eng. Sci. 22 (1967) 1839-1845.
15- Deckwer, W. D. "Bubble column reactors". New York : Wiley(1992).
16- Sahay, B.N., and Sharma, M.M., "Absorption in Packed Bubble Columns", Chem. Eng. Sci. 28 (1973) 2245-2255.
17- Yang, Y. B., Devanathan, N. and Dudukovik, M. P., "Liquid Backmixing in Bubble Columns", Chem. Eng. Sci., 47, 2859, (1992).
18- Miyauchi, T., Furusaki, S., Morooka, S. and Akida, Y., "Transport Phenomena and Reaction in Fluidized Catalyst Beds", Advances in Chemical Engineering, Vol. 11, p. 275-448 (1981). (Sited in Degaleesan et al., 1998)
19- VanBaten, J. M., and Krishna, R., "Scale Up Studies on Partitioned Bubble Column Reactors with the Aid of CFD Simulations", Catalysis Today, 79-80, 219-227, (2003).
20- Degaleesan, S., and Dudukovik, M. P., "Liquid Backmixing in Bubble Columns and the Axial Dispersion Coefficient", AIChE J., 44, 11, 2369-2378, (1998).
21- Wilkinson, P. M. , Spek, A. P. , and van Dierendonck, L. L., "Design Parameters Estimation for Scale-Up of High-Pressure Bubble Columns", AIChE J., Vol. 38, 544-554, (1992). (Sited in Krishna, et al., 1996)
22- Kantak, M. V., Shetty, S. A. and Kelkar, B. G., "Liquid Phase Backmixing in Bubble Column Reactors – A New Correlation", Chem. Eng. Commun., 127, 23-34, (1994). (sited in http://ct=cr4.chem .uva.nl/partition.)
23- Towell, G. D. and Ackerman, G. H., "Axial Mixing of Liquids and Gas in Large Bubble Reactor", Proceedings of 2nd International Symposium Chem. React. Eng., Amsterdam, The Netherlands, B3.1-B3.13, (1972). (sited in http://ct=cr4.chem.uva.nl /partition.)
24- Schugerl, K., Todt, J., Lucke, J., and Renken, A., "Gas Holdup and Longitudinal Dispersion in Different Types of Multiphase Reactors and their Possible Application for Microbial Processes", Chem. Eng. Sci., 32, 369-375, (1977).
25- Baird, M. H. I. and Rice, R. G., "Axial Dispersion in Large Unbaffled Columns", Chem. Eng. J., 9, 171-174, (1975).
26- Krishna, R., Urseanu, M.I., van Baten, J.M., Ellenberger J., "Liquid Phase Dispersion in Bubble Columns Operating in the Churn-Turbulent Flow Regime", Chem. Eng. J., 78, 43–51, (2000).

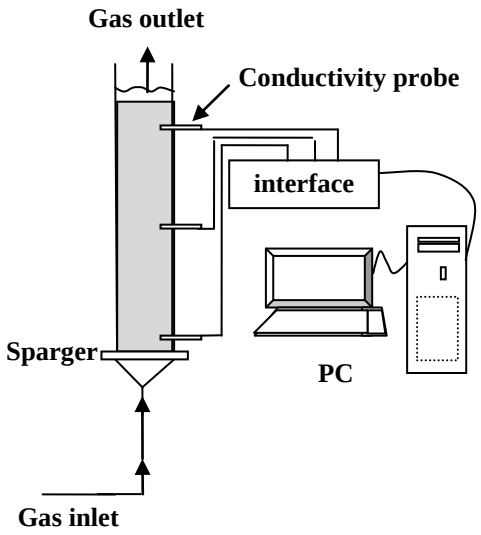


Figure (1) Typical experimental set-up for the 10 cm diameter column.

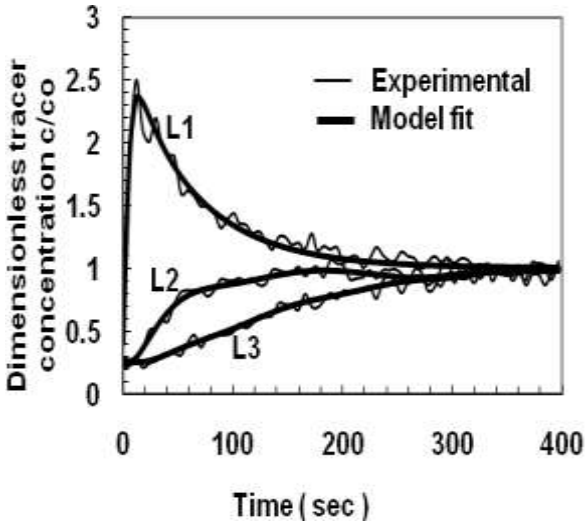


Figure (3) Normalized liquid-phase tracer concentration measured at three different locations along the height of the column in response to pulse tracer injection. The smooth curves represent the fits to the curves from fitting a diffusion model presented in [15].

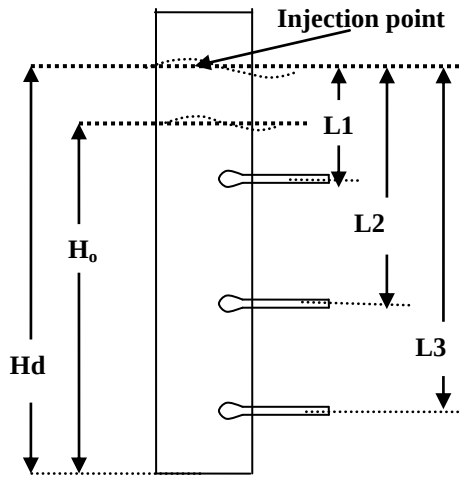


Figure (2) Distances to the measuring points in the column.

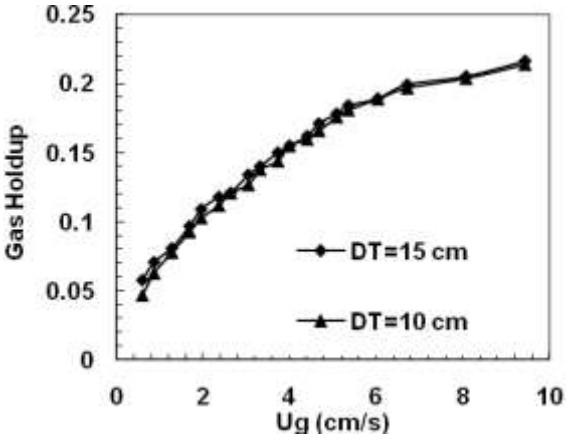


Figure (4) Gas holdup as a function of superficial gas velocity in two different packed bubble columns

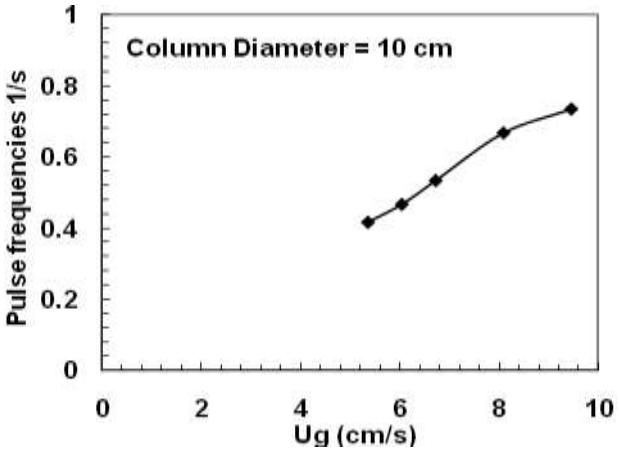


Figure (5) The pulse frequencies as a function of superficial gas velocity

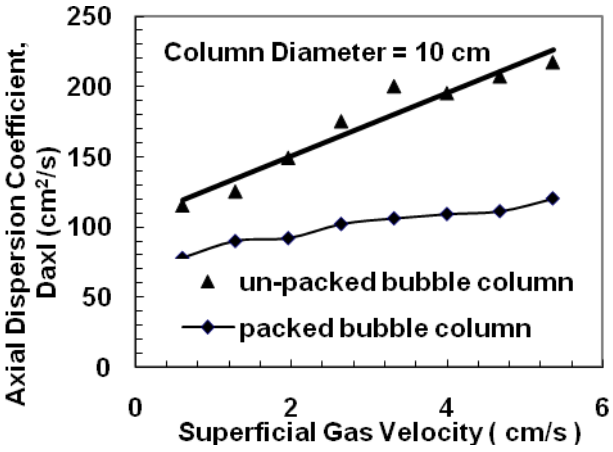
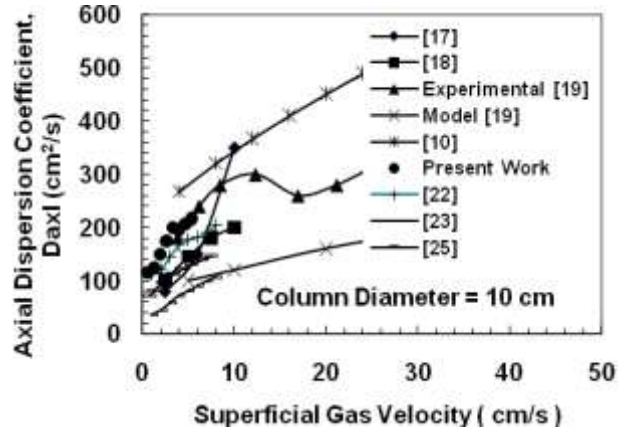
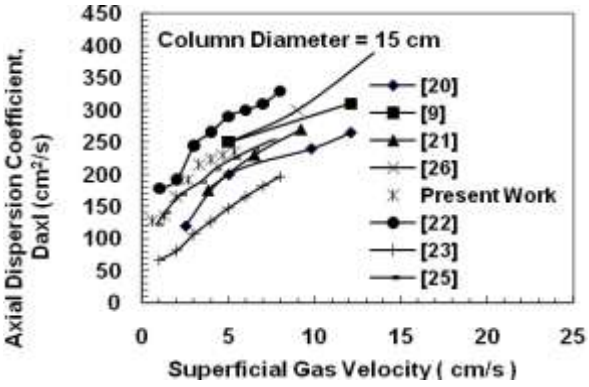


Figure (7) The dispersion Coefficient as a Function of Superficial Gas Velocity

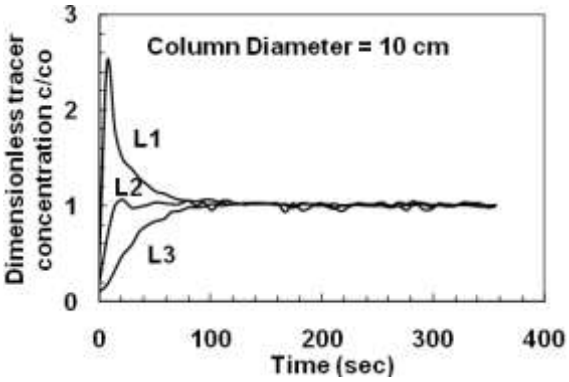


(a)

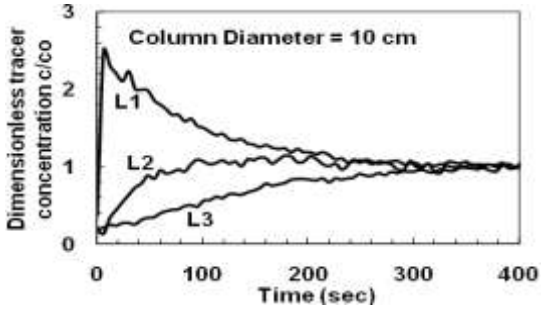


(b)

Figure (6) Comparison between the measured axial dispersion coefficient $D_{ax,L}$ (our data and from the literature)



(a)



(b)

Figure (8) Typical conductivity responses of different probes in (a) bubble column (b) packed bubble column

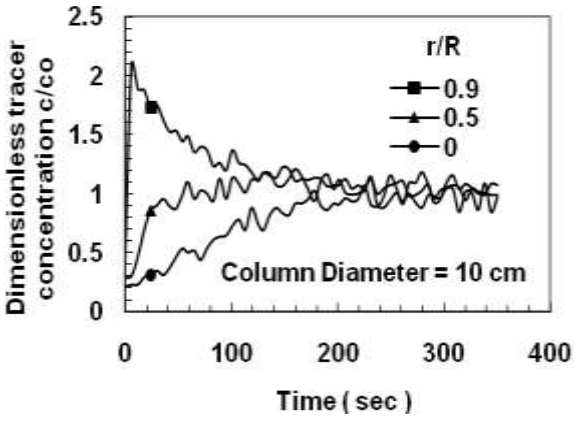


Figure (9) Normalized liquid-phase tracer concentration measured at different radial position plotted verses time, Ug=0.87 cm/s

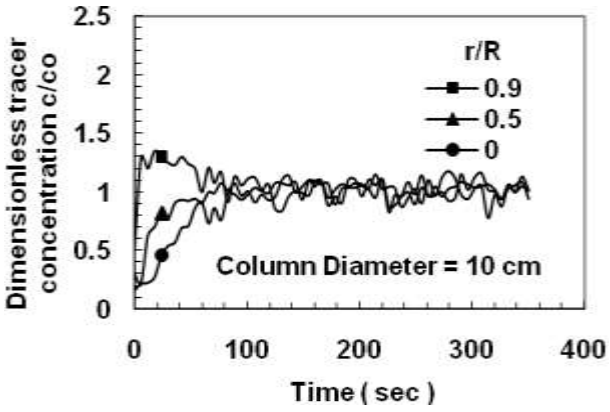


Figure (10) Normalized liquid-phase tracer concentration measured at different radial position plotted verses time, Ug=5.36 cm/s