Effects of Height to Diameter Ratio and Aeration rate on Liquid Mixing and Hydrodynamic Properties in a Bubble Column

M. Pirdashti[1], R. Kompany[2]*

1- Department of Chemical Engineering, Faculty of Eng., Shomal University, Amol, Iran.
2- Department of Chemical Engineering, Faculty of Eng., Razi University, Kermanshah, Iran.

Abstract
Experimental measurements of overall gas holdup (ε_g), residence time distribution (RTD) and liquid mixing time (t_m) have been carried out in an air-water system in a 17 lit bubble column with an 11 cm diameter, over a wide range of superficial gas velocity (0.14-1.46 ms^{-1}) and height to diameter ratio (1.36-8.84). The bed expansion method was used to obtain holdup values and the change of slope related to ε_g versus superficial gas velocity (U_{sg}) indicated the transition point from homogeneous to heterogeneous regime at U_{sg}= 0.7-0.9 ms^{-1}. Besides, the experiments illustrated that H/D ratio had no effect on holdup values. The axial dispersion model was used with semi-closed boundary conditions for prediction of RTD, and hence, the axial dispersion coefficients (D_{ax}). Moreover, fitting results of the model and experimental RTD curves achieved from tracer injection method attained the model parameter, D_{ax} , in three H/D ratios of 4.73, 6.36 and 8.84. Results showed that an elevation in H/D ratio, caused a rise in D_{ax}. On the other hand, the mixing time data declined with an increase in U_{sg}; however enhancement of H/D ratio caused an increase in t_m.

Keywords: *Bubble column, Overall gas holdup, Residence time distribution, Axial dispersion coefficient, Mixing time*

1- Introduction

Bubble columns are widely used as gas-liquid contactors. They have applications in chemical, petrochemical and biological industries [1]. Their advantages over other contacting devices are the simplicity of their construction and maintenance, low energy consumption, and minimal space requirements due to their vertical design [2]. The knowledge of the flow pattern and associated liquid circulation velocities help in the determination of the concentration gradient deciding local and overall rates of the reaction. Therefore, Hydrodynamics parameters and phase mixing are strongly dependent on the flow structure and the

---

* Corresponding author: Pirdashti@yahoo.com

corresponding pattern [3]. It is well established that knowledge of the degree of mixing within any reaction vessel is important for predicting its performance. For tubular reactors, axial mixing is usually several times higher than radial mixing. Thus for most practical purposes attention is focused only on the former state of mixing, not on the latter [4]. It is found that the axial dispersion coefficient is rather insensitive to changes in liquid velocity and liquid properties, i.e., viscosity and surface tension. Mixing in the axial direction in a bubble column is characterized by an axial dispersion coefficient, D_{ax}, which is mainly a function of aeration rate, the geometry of the column and the properties of the fluid. An axial dispersion coefficient in a gas–liquid system characterizes the axial mixing produced by rising gas bubbles that carry elements of circulating fluid in bubble wakes. Because bubble rises faster than the liquid, a certain amount of liquid is carried forward faster than the bulk flow of the liquid. This produces mixing in the axial direction [5]. Another indicator of liquid mixing is liquid phase mixing time (t_{mix}) which is an important performance parameter when the bubble column operates in a batch mode. Knowledge of the mixing time gives information regarding the liquid phase back mixing characteristics and the liquid phase flow pattern. An experimental study has been carried out by Doshi, et al. [3], in which the sparger with a percent free area (% FA) of 0.136 and 0.6% has been used and the superficial gas velocity (U_{sg}), liquid height to column diameter (H/D), and percent free area of the sectionalizing plate was varied from 0.06 to $0.295 ms^{-1}$, 3 to 4 and 4 to 23%,

respectively. It was found that there is no significant effect of the sparger design on the mixing time, but it does strongly depend on U_{sg}, H/D and % FA of the sectionalizing plate. The increment in the mixing time was about 100% for the low free area of the sectionalizing plate and the gas holdup increased 9% with an increase in H/D ratio from 3 to 4. In addition, for low range U_{sg}, the increase in the gas holdup was around 25%, however this growth fell down with an increase in the U_{sg} [3].While Vinaya found no dependency of liquid back mixing on the superficial gas velocity, Ichikawa et al. found significant effects of superficial gas velocity (U_{sg}) on axial dispersion coefficient (D_{ax}) when U_{sg} > 7 cm/s, where D_{ax} reaches a maximum at U_{sg} = 18 cm/s [6]. Thaker and Rao [7] obtained data on gas holdup, ε_g, for the single stage column in different values of clear liquid height and superficial gas velocity at constant diameter, and correlated to the parameters, U_{sg} and H/D ratio in the ranges 0.005 < U_{sg} < 0.05 and 3.0 < H/D < 7.0, by the equation: $\varepsilon_g \propto U_{sg}^{0.83} (H/D)^{0.09}$. They also found that in the absence of gas redistributors, the values of k_{La} and D_{ax} decreased with the increase in liquid height [7]. Ekambara and Joshi [8] investigated the effect of height to diameter ratio (H/D) on mixing time in 0.2 and 0.4 m inner diameter of columns (i.d.column), for a gas–liquid system with variation in U_{sg} from 0.07 to 0.295 ms^{-1}, respectively. For 0.2 m i.d.column H/D ratio was varied from 3 to 10 and from 1 to 5 for 0.4 m i.d.column. However, the experimental observations of mixing time conform to $t_m \propto H^{0.4-0.66}$. They observed that the axial dispersion coefficient

increases with an increase in H/D ratio for both the columns, and the proportionality relationship was found to be $D_{ax} \propto H^{0.65-1.72}$ [8].

Urseanu et al. [9] measured total gas hold up and axial dispersion coefficients in the structured bubble columns and the values were compared with the experimental results obtained in the same work with empty bubble columns. The results reveal that the gas hold up in structured bubble columns is practically the same as in empty bubble columns when compared at the same superficial gas velocity based on open area available for gas–liquid dispersion. The presence of the structured elements in the bubble column reactor reduces the liquid phase back mixing by one order of magnitude. Field and Davidson have reported the relationship between dispersion coefficient, dispersion height and mixing time as [10]: $D_{ax} = \left(\dfrac{A.H_D^2}{t_m} \right)$

1.1- Liquid mixing time (t_m)

An important parameter used frequently to represent mixing in reactors is t_m. It expresses the overall mixing property of liquid and a shorter tm means less chance of a dead zone forming [11]. Mixing time is defined as the time required to achieve the desired degree of homogeneity (usually 90-95%) after the injection of an inert tracer impulse into the reactor. The so-called degree of homogeneity (I) is given by:

$$I = (C - C_\infty)/C_\infty \tag{1}$$

Where C is the maximum local concentration and C_∞ is the mean concentration of tracer at

complete mixing [12]. The higher the degree of mixing, the shorter t_m would be. Mixing time, t_m, is the time required to attain a given deviation from the fully mixed state from the instance of a tracer input [5].

1.2- Overall gas holdup

Overall gas holdup is one of the important parameters for bubble column design and scale-up. It is defined as the fraction of the reactor dynamic volume occupied by the gas. It characterizes the hydrodynamics of reactors, and depends mainly on the gas velocity, physical properties of the liquid and type of gas sparger. The overall gas holdup was determined by using the volume expansion technique. The aerated and un-aerated liquid heights in the reactor (H_d and H_l) were noted and the overall gas holdup was calculated as [13]:

$$\varepsilon_g = \frac{H_d - H_l}{H_d} \tag{2}$$

Bubble columns are generally operated with low liquid velocities, which have been reported to have little or no effect on overall gas holdup. Moreover, the spatial variation of gas holdup gives rise to pressure variation and eventually results in intense liquid phase motion [14]. The height of the sparger region depends upon the difference between d_{BP} and d_{BS}, the coalescing nature of the liquid phase and the liquid circulation in the heterogeneous regime. The relative proportion of the sparger region in the total column height decides the effect of H/D ratio on ε_g. If the sparger region is small, the effect of H/D ratio on ε_g is minimum and vice-versa [14].

1.3- Theory of tracer experiment

Differential bubble rise velocity and gas holdup distribution is responsible for the liquid phase dispersion and mixing, which in turn affect the gas-phase dispersion. Deckwer [15] has suggested that the radial dispersion coefficient is always less than one-tenth of the value of the axial dispersion coefficient. Therefore, the one dimensional (axial) model is sufficient to mimic the liquid phase mixing phenomena in a sectionalized bubble column and the longitudinal dispersion coefficient can then be used to express the liquid phase back-mixing characteristic of the bubble column [15].

The time course of a tracer's concentration at some measurement location in a bubble column can be described by the following complete dispersion model:

$$\frac{\partial C}{\partial t} = D_{ax} \frac{\partial^2 C}{\partial z^2} - \frac{U_l}{\varepsilon_l} \frac{\partial C}{\partial z} + \frac{D_r}{r} \frac{\partial C}{\partial r} + D_r \frac{\partial^2 C}{\partial r^2} \qquad (3)$$

where C is the tracer concentration at time t, D_{ax} is the axial dispersion coefficient, z is the axial distance from the point of tracer input, D_r is the radial dispersion coefficient, r is the radial distance measured from the center of the column, U_l is the superficial velocity of the liquid in the column, and ε_l is the liquid holdup. Eq. (4) assumes an instantaneous input of the tracer pulse and it can be made dimensionless using the following definitions of the variables:

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

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

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

In Eqs. (5) to (7), y is the dimensionless axial distance and L is the height of gas–liquid dispersion in the reactor. The dimensionless form of Eq. (4) is

$$\frac{\partial C_T}{\partial \theta} = \frac{\partial^2 C_T}{\partial y^2} - \frac{U_l L}{\varepsilon_l D_{ax}} \frac{\partial C_T}{\partial y} + \frac{1}{x} \frac{\partial C_T}{\partial x} + \frac{\partial^2 C_T}{\partial x^2} \qquad (8)$$

Where C_T is the dimensionless tracer concentration defined as follows:

$$C_T = \frac{C}{C_\infty} \qquad (9)$$

In Eq. (9), C and C_∞ are the instantaneous and the final equilibrium concentrations of the tracer in the liquid batch. For a batch of liquid in a bubble column, there is no superimposed liquid flow and, hence, $U_l = 0$. Furthermore, considering the cylindrical symmetry of the tracer response and the fact that the wall, the upper surface and the bottom of the reactor are impermeable to the tracer, we have the following conditions:

(i) When $x = 0$, $\partial C_T / \partial x = 0$,

(ii) When $x = R/L\sqrt{D_{ax}/D_r} = \beta$ (i.e. at the wall when r=R) $\partial C_T / \partial x = 0$

(iii) When $\theta > 0$ and y = 0, $\partial C_T / \partial y = 0$ (i.e. at the surface of the dispersion),

(iv) When $\theta > 0$ and y = 1, $\partial C_T / \partial y = 0$ (i.e. at the bottom of the reactor)

Using the above noted initial and boundary conditions, Eq. (8) can be solved to obtain the following expression for the dimensionless tracer concentration:

$$C_T = \sum_{n=1}^{\infty} \frac{J_0(\nu_n x)}{J_0^2(\nu_n \beta)} e^{(-\nu_n^2 \theta)} \times \left(1 + 2 \sum_{m=1}^{\infty} \cos(m\pi y) e^{(-m^2 \pi^2 \theta)}\right) \quad (10)$$

Where J_0 is the zero-order Bessel function and ν_n is the nth root of the first-order Bessel function, J_1. A similar two dimensional dispersion model has been used to characterize mixing in liquid–solid fluidized beds. Note that when the C_T in Eq. (10) is radially invariant (i.e. $D_r = \infty$), ν_n, β and x become zero and $J_0^2(\nu_n \beta) = J_0(\nu_n x) = 1$

In this case Eq. (7) reduces to

$$C_T = 1 + 2 \sum_{m=1}^{\infty} \cos(m\pi y) e^{(-m^2 \pi^2 \theta)} \quad (11)$$

which is identical to the solution of the axial dispersion model that was reported by Ohki and Inoue without considering the radial component of the dispersion [16].

2- Experimental

Experiments were carried out in a cylindrical bubble column with an inner diameter of 11 cm and a total volume of 17 lit. Fig. 1 is the schematic diagram of the experimental setup. The column studied was a batch liquid system, while the liquid was tap water and the gas was compressed air. The system worked at ambient temperature and atmospheric pressure. A perforated plate sparger of 8cm in diameter was used in the column to distribute the gas phase, located 4 cm above the bottom of the column, having 12 holes with a 1mm diameter. The air flow rate was measured by a calibrated rotameter before entering the reactor. Furthermore, the overall gas holdup was determined using the volume expansion technique and the tracer method was used to measure the liquid phase residence time distribution and mixing time, defined as the time required to attain a 5% deviation from complete homogeneity from the instance of tracer addition. Since the reactor had been filled with the appropriate amount of fluid, the tracer, which is 10 ml of NaOH 1 N, was injected into the batch liquid phase as a pulse at the center of the surface of dispersion instantaneously. The change in pH with time was measured using a pH electrode located in the middle of the column at 70 cm below the upper edge of the bubble column. The pH electrode used had a virtually instantaneous response (approximately 1 s for full response) over the range of the experiments. The location of the electrode in bubble column is shown in Fig.1. The pH data were recorded and analyzed manually using a PC computer. Results showed that pH did not go over 12.0. The dimensionless concentration, C_T, of the tracer (hydroxide ion) was calculated as follows:

$$C_T = \frac{\left[OH^-\right]_{instant}}{\left[OH^-\right]_{final}} \quad (2)$$

The H/D altered in the range of 1.36 to 8.84 and the superficial aeration velocity varied from 0.14 to about 1.46 ms^{-1}.
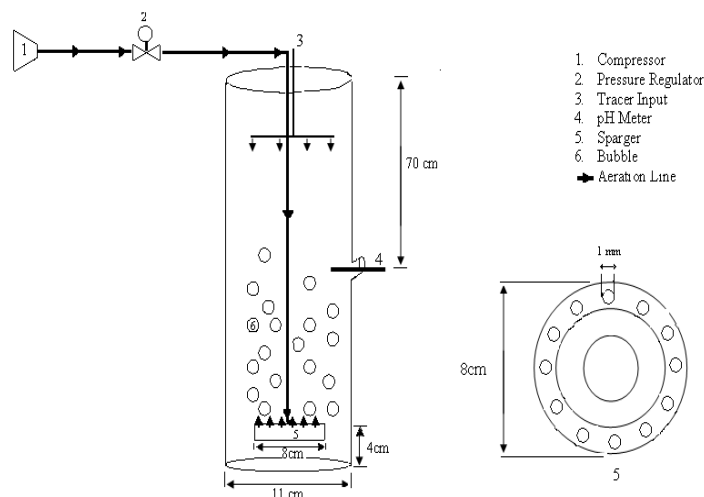
Figure 1. Schematic of experimental setup

3- Results and discussion
3.1- Overall gas holdup
3.1.1- Effects of superficial gas velocity

The gas holdup, which is one of the most important parameters characterizing the hydrodynamics of reactors, depends mainly on the gas velocity, physical properties of the liquid and type of gas sparger. The gas holdup increases with enhancement of superficial gas velocity. Many researches used gas holdup as a parameter from which the transition between regimes could be estimated. Gas holdup is an important parameter because it determines the amount of the gas phase retained in the system at any time. Fig. 1 is the plot of the gas holdup versus superficial velocity of the input gas. The rate of increase of the overall gas holdup (ε_{gr}) with superficial gas velocity depends on the existing two phase flow regime.

Therefore, the change of the slope in the dependence of the gas holdup versus superficial gas velocity in the column can be used for the estimation of the flow patterns from the homogeneous flow regime to the heterogeneous flow regime. The transition between the mentioned flow regimes can be estimated at $U_{sg} \approx 0.7$-0.9 ms^{-1} in our experiments (Fig.2).

Chisti and Moo-Young proposed a correlation equation of gas holdup in bubble columns. There was a transition point around U_{sg} of 0.05 ms^{-1}. The correlation equation of gas holdup with respect to the superficial velocity was shown as follows [13]:

$$\varepsilon_g = \begin{cases} 2.47 U_{sg}^{0.97} & U_{sg} < 0.05 \, (m/s) \\ 0.49 U_{sg}^{0.46} & U_{sg} > 0.05 \, (m/s) \end{cases} \qquad (3)$$

As the gas flow rate increased, the bubbles became closely packed and coalesced to form larger bubbles, which steadily increased in size with the gas flow rate, causing a slower increase in the gas holdup. Because of the high turbulence fields the large bubbles often have little definition to their shape which fluctuates quite randomly.
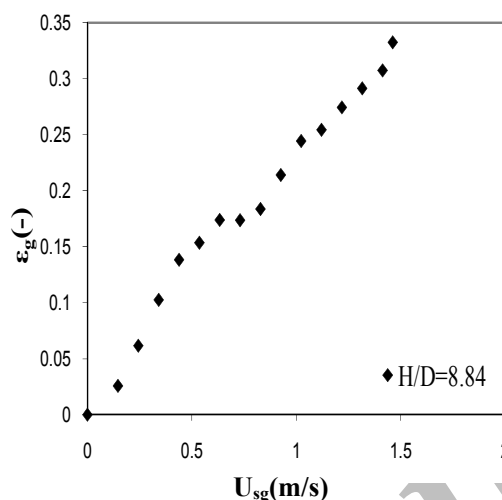
Figure 2. Overall gas holdup (ε_g) versus superficial gas velocity (U_{sg})

An example of ε_g changes with U_{sg} can be seen with U_{sg} in Fig. 2, when pure liquid height was at its highest (H/D=8.84).

3.1.2- Effects of H/D ratio
The gas phase moves in one of the two characteristic regimes depending upon the nature of dispersion. These are called homogeneous (or bubbly flow) and heterogeneous (or churn turbulent) regimes. The homogeneous regime is characterized by uniform sized bubbles. Further, the concentration of bubbles is also uniform, particularly in the transverse direction. The process of coalescence and dispersion are practically absent in the homogeneous regime and hence the sizes of bubbles are entirely dictated by the sparger design and the physical properties of the gas and liquid phases [14]. In contrast, in the heterogeneous regime, the role of sparger design diminishes depending upon the column height. In fact, the total column height can be divided into

two regions: the sparger region and the bulk region. The size of the bubble formed at the sparger (primary bubble size, d_{BP}) depends upon the sparger design, the local energy dissipation rate and the surface active contaminants. In the sparger region, the bubble size changes with respect to height depending upon the coalescence nature of the liquid phase, the extent of turbulence and the bulk motion. At the end of the sparger region, the bubbles attain an equilibrium size (called secondary bubble size, d_{BS}). The equilibrium is governed by the breaking forces due to bulk motion (turbulent and viscous stresses) and the retaining force due to surface tension. There was practically no effect of the H/D ratio on gas holdup as shown in Fig. 3.

The probable reason for this could be that the bubbles generated have a size very similar to the equilibrium bubble size. Because the relative proportion of the sparger region (region where the bubbles size change to

equilibrium one) to in the total column height decides the effect of H/D ratio on ε_g. Therefore, when the sparger region is small, the effect of H/D ratio on ε_g is minimized.

3.2- Axial dispersion coefficient
The residence time distribution curves were obtained and fitted to the model described above (Eq. 11) with only one adjustable parameter, the liquid axial dispersion

coefficient, D_{ax}. These curves were digitized manually and expressed as $C/C_\infty(-)$ versus time(s). Two typical fits of all RTD curves measured at three heights of liquid are shown in Fig. 4 for the bubble column. Similar excellent fits of the RTD curves were obtained for a whole range of gas velocities ($0.146 \leq U_{sg} \leq 1.46$ ms^{-1}) with three different H/D ratios (4.73, 6.36 and 8.84).
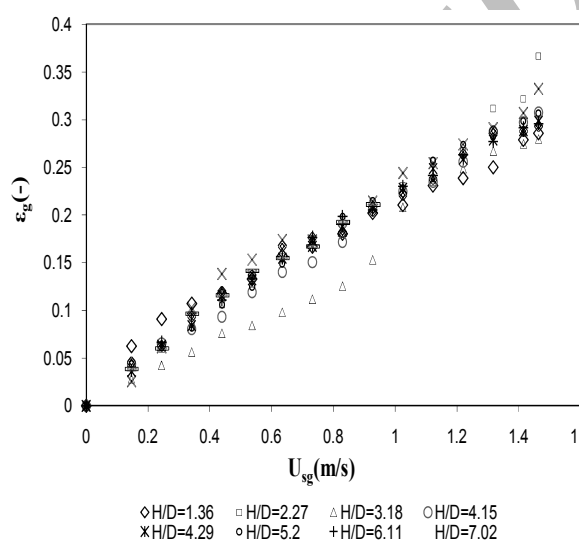

Figure 3. Overall gas holdup (ε_g) versus superficial gas velocity (U_{sg}) with H/D ratio as a parameter

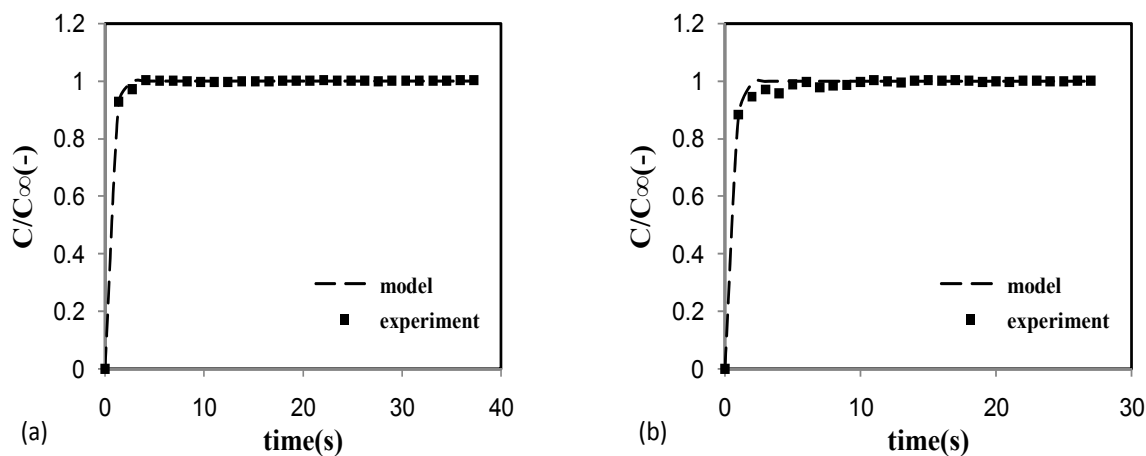**Figure 4.** Typical plots for experimental and calculated C/C_∞ versus time, U_{sg}=0.146 ms^{-1},
(a) H/D=6.36, D_{ax}=0.06 ms^{-2} (b) H/D=8.84, D_{ax}=0.07 ms^{-2}

Fig.5 shows the variation of D_{ax} with superficial air velocity for air-water systems. The graph clearly demonstrates the essential effect of gas dispersion mode (flow regime) on the extent of gas phase mixing in the bubble column (Fig.5). Results of our experiments proved the essential effect of flow regime on the extent of gas axial mixing in bubble column reactors. Within the homogeneous regime, the liquid axial mixing was significantly lower than under conditions of the heterogeneous regime and its extent shows fluctuations with increasing gas flow rate up to an appropriate value D_{ax}, marking the transition to the heterogeneous regime. The point where the axial mixing coefficients start to decrease is usually defined as the onset of transition from bubbly flow to churn-turbulent flow regimes, which can also be seen at 0.7-0.9 ms^{-1}. Under the conditions of the heterogeneous regime, the extent of axial gas mixing raises linearly with the gas flow rate.

It can be seen from Fig.5 that the value of the dispersion coefficient increases with the enhancement of the H/D ratio. A possible reason for an increase in the dispersion coefficient is due to an increase in the liquid height and dispersion height, which subsequently increases the mixing time (due to longer loop need to homogenization) at constant superficial gas velocity.

3.3- Liquid-phase mixing time

Mixing time is an indicator of degree of mixing and it is one of the most important process parameters during a batch mixing process. Mixing time is typically considered as the time required for the tracer concentration to reach the desired degree of homogeneity [8]. It can be seen from Fig.6 that the mixing time decreases with an increase in the superficial gas velocity. The reason is that the more the U_{sg}, the greater the Dax would be, and so the time needed to reach the desired degree of homogeneity is reduced, which means a lower mixing time. As mixing time is a global index of mixing and it is affected by axial and radial mixing and the effects of bulk flow, in higher air flow rates the driving force decreases because of the existence of liquid axial and radial mixing, this results in a descending rate of decrease in t_m.
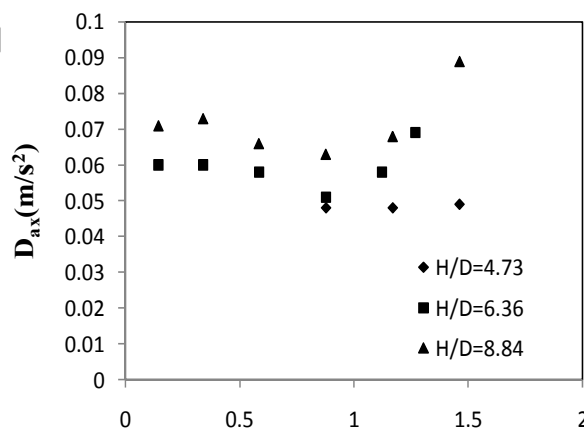

Figure 5. Axial dispersion coefficient (D_{ax}) versus superficial gas velocity (U_{sg}) with H/D ratio as a parameter
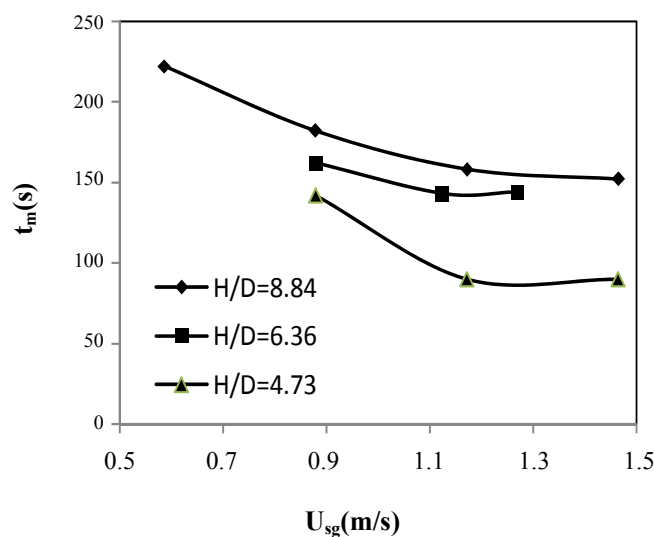
Figure 6. Mixing time (t_m) versus superficial gas velocity (U_{sg}) with H/D ratio as a parameter

Figure 6 also shows that mixing time increases with an increase in the H/D ratio. Actually, an enhancement in the H/D ratio means an elevation in the clear liquid height, so that the time required for the tracer to homogenize over a longer distance and over a higher liquid volume also increases, which cause an increase in the mixing time.

4- Conclusions

The combined effect of superficial gas velocity and height to the diameter ratio of a bubble column was investigated. With an increase in U_{sg}, the holdup was found to increase. On one hand, the change of slope in ε_g versus superficial gas velocity (U_{sg}) indicated the transition point from homogeneous to heterogeneous regime at $U_{sg}= 0.7$-0.9 ms^{-1}. In addition, the bubbles being generated just on the exit of the sparger had a similar size to the equilibrium bubble size, so that the H/D ratio had a negligible effect on holdup values. On the other hand,

the axial dispersion model was used with a semi-closed boundary condition and it also has a significant contribution in the prediction of RTD, and hence in calculating the axial dispersion coefficients (D_{ax}). Fitting results of the model and experimental RTD curves were achieved from the tracer injection method and as a result of the model parameter, D_{ax} in three H/D ratios of 4.73, 6.36 and 8.84 were attained. Consequently, results showed that an elevation in H/D ratio causes an increase in D_{ax}. Also, within the bubbly flow regime D_{ax} was low and had a slow increase, starting to decrease at the beginning of the transition regime to churn turbulent, since in the heterogeneous regime D_{ax} linearly increased with U_{sg}. Moreover, the mixing time data declined with an ascending trend of Usg, but enhancement of H/D ratio resulted in a rise in t_m. The result is that a longer distance was required for the liquid to homogenize.

References

1. Lefebvre, S., and Guy, Ch., "Characterization of bubble column hydrodynamics with local measurements", Chem. Eng. Scien., 54, 4895 (1999).

2. Alvaré, J., and Al-Dahhan, M.H., "Liquid phase mixing in trayed bubble column reactors", Chem. Eng. Scien., 61, 1819 (2006).

3. Doshi, Y.K., and Pandit, A.B., "Effect of internals and sparger design on mixing behavior in sectionalized bubble column", Chem.Eng. Journal., 112,117 (2005).

4. EI-Temtamy, S.A., Khalil, S.A., Nour El-Din, A.A., and Gaber A., "Liquid-phase axial mixing in a bubble column bioreactor containing yeast suspensions", Appl. Microbiol. Biotechnol., 21, 65 (1985).

5. Miron, A.S., Ceron Garcia, M.C., Camacho, F.G., Grima, E.M. and Chisti, Y., "Mixing in bubble column and airlift reactors", Trans. of Instit. of Chem. Eng., 82 (10),1367 (2004).

6. Al-Dahhan, M.H., Mills, P.L., Gupta, P., Han, L., Dudukovic, M.P., Leib, T.M. and Lerou, J.J., "Liquid-phase tracer responses in a cold-flow counter-current trayed bubble column from conductivity probe measurements", Chem. Eng. and Process., 45, 945 (2006).

7. Thaker, K. and Rao, D.P., "Effects of gas redispersion and liquid height on gas liquid hydrodynamics in multistage bubble column", Trans. of Instit. of Chem. Eng., 85(10),1362 (2007).

8. Ekambara, K. and Joshi, J.B., "CFD simulation of mixing and dispersion in bubble columns", Trans. of Instit. of Chem. Eng., 81(A), 987 (2003).

9. Urseanu, M.I., Ellenberger, J., and Krishna, R., "A structured catalytic bubble column reactor: hydrodynamics and mixing studies", Catalys. Tod., 69,105 (2001).

10. Field, R.W., and Davidson, J.F., "Axial dispersion in bubble column", Trans. of Instit. of Chem. Eng., 58(4),228 (1980).

11. Bando, Y., Hayakawa, H., and Nishimura, M., "Effects of equipment dimensions on liquid mixing time of bubble column with draft tube", Chem. Eng. Japan., 31(5), 765 (1998).

12. Michael, J.C., and Gluz, M., The encyclopedia of bioprocess technology: bioreactors, airlift reactors, John Wiley & Sons Inc., USA, p. 320 (1996).

13. Chisti, M.Y., and Moo-Young, M., "Gas holdup in pneumatic reactors", Chem. Eng., 38, 149 (1988).

14. Thorat, B.N., Shevade, A.V., Bhilegaonkar, K.N., Aglawer, R.H., Parasu Veera, U., Thakre, S.S., Pandit, A.B., Sawant, S.B. and Joshi, J.B., "Effect of sparger design and height to diameter ratio on fractional gas holdup in bubble columns", Trans. of Instit. of Chem. Eng., 76(A),823 (1998).

15. Deckwer, W. D., Bubble column reactors, Wiley, Chichester, p. 113 (1992).

16. Camacho, R. F., Sanchez, M. A., Ceron Garcia, M.C., Garcia Camacho, F., Molina G. E., and Chisti, Y., "Mixing in bubble columns: a new approach for characterizing dispersion coefficients", Chem.Eng. Scien., 59, 4369 (2004).