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Chemical Engineering Science 60 (2005) 3573–3586

Chemical
Engineering Science

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Liquid phase axial mixing in bubble columns operated at high pressures

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Received 24 September 2004; received in revised form 18 January 2005; accepted 31 January 2005

Abstract

Liquid phase axial mixing was measured in a 100 mm i.d. bubble column operated in the pressure range of 0.1–0.5 MPa. Water, ethanol and 1-butanol were used as the liquid phase and nitrogen as the gas phase. The temperature and superficial gas velocity were varied in the range of 298–323 K and 0.01–0.21 m/s, respectively. The axial dispersion coefficient increased with an increase in the gas density due to pressure. The temperature had surprisingly a small effect. A CFD model was developed for the prediction of flow pattern in terms of mean velocity and eddy diffusivity profiles. The model was further extended for the prediction of residence time distribution and hence the axial dispersion coefficient (D_L). The predictions of axial dispersion coefficient agree favorably with all the experimental data collected in this work as well as published in the literature. The model was extended for different gas–liquid systems. The predicted values of axial dispersion coefficient were found to agree very well with all the experimental data.

Keywords: Dispersion; Mixing; Pressure; Gas density; CFD; Bubble columns

1. Introduction

Bubble column reactors find widespread use in industry. The majority of the bubble columns are operated at pressure above atmospheric (Shah et al., 1982). The obvious advantage of an increased pressure are an increase in the solubility of the gas phase reactants. High gas density has been observed to destabilize larger bubbles and thus provide higher values of gas hold-up and hence interfacial area for mass transfer.

Bubble columns have been extensively studied for the last several decades and fairly good information is available on the liquid phase axial mixing. However, most of the studies have been carried out at atmospheric pressure and especially for the air–water system whereas the investigations under high-pressure conditions are very scarce. Holcombe et al. (1983) determined the axial liquid dispersion coefficient in a 78 mm bubble column with 1.8 m height and a pressure range 0.3–0.72 MPa. The superficial gas velocity was varied up to 0.6 m/s. They used heat as a tracer to measure the thermal dispersion coefficient. They found that the effect of pressure on thermal dispersion coefficient was negligible in the pressure range of their study.

Tarmy et al. (1984) investigated liquid phase axial mixing in industrial coal liquefaction reactors using radioactive tracers. The operating pressure was varied up to 17 MPa. They found that the values of dispersion coefficient increased with an increase in pressure.

Houzelot et al. (1985) investigated the axial dispersion in a 50 mm i.d. column. The superficial gas velocity was varied up to 6 mm/s and pressure in the range of 0.1–0.3 MPa. They have observed an insignificant effect of pressure on the axial dispersion coefficient which is perhaps due to the limited range of variables covered in their work.

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0009-2509/$ - see front matter © 2005 Elsevier Ltd. All rights reserved.
doi:10.1016/j.ces.2005.01.032

Wilkinson et al. (1993) measured the axial dispersion coefficient in a batch-type 158 mm i.d. bubble column with nitrogen–water system at pressures between 0.1 and 1.5 MPa using the electrical conductivity method. It was found that the value of D_L increases with increasing pressure, especially under high gas velocity conditions ($V_G > 100\,\mathrm{mm/s}$).

Onozaki et al. (2000a,b) studied gas–liquid dispersion behavior in coal liquefaction reactors using a neutron absorption tracer technique. In a column of 1000 mm i.d., they found that the axial dispersion coefficients of the liquid phase under coal liquefaction conditions ($P = 16.8$–$18.8\,\mathrm{MPa}$) were much smaller than those estimated from literature correlations obtained at ambient conditions.

Yang and Fan (2003) have measured liquid axial dispersion coefficient at high pressure (up to 22 MPa) in 50.8 and 101.6 mm i.d. bubble columns. The superficial liquid and gas velocities were varied in the range of 3.4–10 mm/s and 20–400 mm/s, respectively. They have used heat as a tracer to measure the thermal dispersion coefficient. They have found the axial dispersion coefficient to decrease with an increase in pressure. Further, they have also observed that the distributor design does not have a significant effect on the liquid phase mixing.

From the foregoing discussion, it is clear that contrasting observations have been made and these have not been rationalized in a systematic manner on the basis of well understood fluid mechanics in bubble columns. The prevailing unsatisfactory state of the art is principally because of the following reasons: (i) The most common approach to account for the non-ideality has been based on a one-dimensional (1-D) axial dispersion model (ADM). Because of its simplicity and ease of use, the ADM has remained popular in the reactor design, and numerous correlations for the axial liquid dispersion coefficient in bubble columns have been developed over the years. However, the ADM makes an attempt to cover all the types of non-idealities without understanding the detailed fluid mechanics behind each non-ideality. (ii) Most of the experiments have selected air–water as the gas–liquid system. The results with organic liquids at higher temperatures and pressures have been contradicting. For instance, Panvelkar et al. (1982) and Wilkinson et al. (1993) have observed an increase in axial dispersion coefficient (D_L) with an increase in pressure. However, Yang and Fan (2003) have reported exactly the reverse.

In view of the above-mentioned status, it was thought desirable to undertake a systematic experimental investigation of the effect of pressure on axial mixing in bubble columns using organic liquids as the liquid phase. Further, in order to understand the apparent discrepancy of the effect of pressure on D_L, a CFD code has been developed to understand the variation in the flow pattern with respect to pressure. An attempt has been made to investigate the parametric sensitivity over a fairly wide range of temperature, pressure, superficial gas velocity and different gas–liquid systems on the flow pattern and hence on the axial dispersion coefficient.

2. Experiments

Liquid phase dispersion in the bubble column (Fig. 1) was studied in water, ethanol (96%) and 1-butanol, respectively, varying the pressure (0.1–0.5 MPa), the temperature (298 and 323 K) and the superficial gas velocity (0.01–0.21 m/s). The column diameter was 0.1 m; the dispersion height was kept constant at $H = 2.1\,\mathrm{m}$. The gas velocity and density are referred to half the dispersion height taking into account the vapor pressure of the liquid. The liquid properties are listed in Table 1.

The gas hold-up was determined from the pressure difference between ports located at 0.65 and 1.2 m from the gas sparger. The dispersion coefficients were measured for different gas–liquid systems with wide range of temperature and pressure. For determination of the dispersion coefficient, pulses of the same liquid, about 20 K higher or lower in temperature, were injected at the top of the dispersion (2.1 m from the sparger) and the time courses of the local temperatures at several lower positions were measured with Pt-100 sensors (time constant $< 3\,\mathrm{s}$). The records (Fig. 2) were analyzed with the 1-D dispersion model (Ohki and Inoue, 1970):

$$T = \left(1 + 2\sum_{n=1}^{\infty}\left[\left(\cos\frac{n\pi}{H}z\right)\exp\left\{-\left(\frac{n\pi}{H}\right)^2 D_L t\right\}\right]\right) \times (T_0 - T_\infty)\mathrm{e}^{at} + T_\infty. \tag{1}$$

Heat transfer at the wall was accounted for by using the hypothetical initial temperature T_0 and the constant a as additional fitting parameters. The analysis was based on the signal of the lower sensor, only, since these results showed less scatter. The reported data points are mean values for 2–6 (usually 3) measurements.

For the given experimental conditions, the drift flux constants have been calculated for each system. The experimental values of $\bar{\varepsilon}_G$, C_0 and C_1 are given in Table 2. It may be pointed out that the drift flux plot was not found to be a straight line and the values of C_0 were found to decrease and C_1 to increase with an increase in V_G.

3. Mathematical model and method of solution

For two-phase gas–liquid flow in bubble columns, Ekambara and Joshi (2003) have given the details pertaining to the CFD procedure for the prediction of 3-D flows. Mass and momentum balances for the phases get satisfied during the numerical integration of the equations of continuity and motion. Fig. 3 shows the excellent agreement between the CFD predictions and the experimental measurements of flow pattern reported by Hills (1974), Menzel et al. (1990), Grienberger and Hofmann (1992), Yu and Kim (1991) and Yao et al. (1991) over a wide range of column diameter (D), column height (H_D), superficial gas velocity (V_G), sparger