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A Tower-type Fermenter with a Gas Entrainment Process

(III) Determination of Axial Dispersion Coefficient of Liquid Flowing in Column

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Abstract

The degree of the axial dispersion of the liquid flowing in the vertical column of the equipment was investigated in order to obtain data needed to complete the equations derived in a previous paper.

Residence time distribution data at different operating conditions were obtained by means of pulse tracer injection. Liquid flow was found to exhibit behavior quite like that of plug flow. The axial dispersion coefficients of liquid were determined from the data. The values were found to be roughly constant for any operating conditions.

Introduction

In the preceeding paper,[1] thoretical analyses of gas entrainment process were developed. It was suggested in the paper that the motion of the liquid at the region where the liquid film plunged had an influence on the gas entainment, and this effect was represented by an axial dispersion coefficient of the liquid.

In this study the axial dispersion coefficient of liquid flowing in the column of the equipment was determined by means of the measurement of residence time distribution at different operating conditions.

Experimental Apparatus and Procedure

Three different column of 3.0, 4.04, and 5.5 cm i.d., were used for the experiments. The length of each column was 110 cm. All column were made of acrylic resin to allow visual observation. At the top of the column, a liquid chamber of 20 cm i.d. was mounted. Four pressure manifolds were located on the columns. A schematic diagram of the assembly is shown in Fig. 1.

In order to approach the conditions assumed in theoretical analysis, considerable care for ensuring a stable liquid film flow was taken. The size of the liquid chamber at the top of the column was determind so as to diminish the velocity of the liquid in it, and two entrances for liquid were located just above the bottom in opposition to each other. Furthermore, wire mesh was also inserted to suppress

the disturbance of the liquid flow.

The liquid was pumped to the upper chamber through an orifice flow meter, by which the liquid flow rate was measured. The flow meter was calibrated previously over the experimntal range of the flow rate. The liquid in the chamber enters a vertical column in the form of film flow. When the liquid film plunges into the liquid in the column, air is entrained into the liquid to form a mixed-phase of air bubbles and liquid. The mixed-phase flow of air-liquid was then introduced to a gas-liquid separator (cylindrical, ca. 26×16 cm i.d.), where the air was separated from the liquid. The liquid held in the separator was recirculated in the same way by a centrifugal pump. The air from the separator was conducted to an orifice meter, by which the volumetric flow rate of the air entrained was measured. Varying the liquid flow rate and flowing distance of the liquid film, the measurements were carried out.

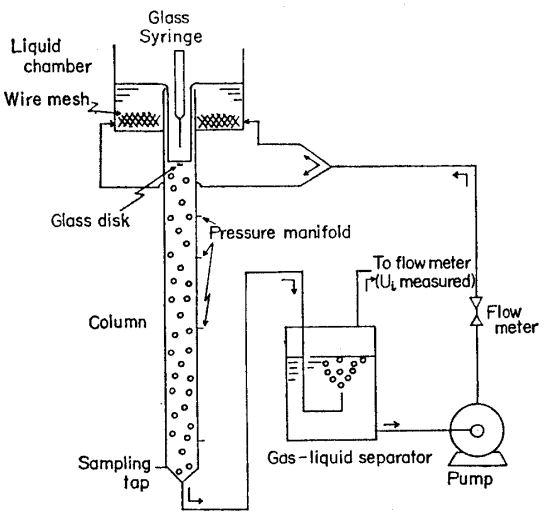

Fig. 1. Flow diagram of experimental apparatus

The liquids used in the experiments were water and 50% (by weight) glycerol solution. The working volume of the liquid ranged from 5.7 to 8.0 *l* according to the column size. All experiments were carried out at 30°C; the temperature was maintained by a flexible electric heater attached around the gas-liquid separator.

Gas hold-up was measured according to the method of a previous paper.[2)]

In the measurement of the response to an instantanous pulse tracer input, a solution of sodium fluorescein (dissolved in the same liquid as that circulating in the apparatus in order to eliminate differences in specific gravity) was used as the tracer; the concentration of the tracer was about 250 ppm. A small glass disk was placed just below the liquid level in the center of the column. The diameter of the disk was smaller than that of the column. In each run approximately 1.0 ml of the tracer was fed on the disk to prevent its forced dispersion due to the injection, using a glass syringe. The syringe was placed at the center of the column, about 1 cm above the level of the liquid.

A sampling tap was provided at the bottom of the column. Samples were taken from the central part of the column. At the instant of the injection of the tracer, a stop watch was started and samples were withdrawn into glass tubes at constant intervals. The samples were analyzed for fluorescence by a Hitachi Fluorescence Spectro-photometer Type 204.

Residence Time Distribution Function

The residence time distribution function for the case of an instantaneous pulse tracer input has been given by the following equations.[3,4]

$$E(\phi)=\frac{Vc}{v_0 c_0} \quad \cdots\cdots(1)$$

where

$$\phi=\frac{\theta}{\theta_m} \quad \cdots\cdots(2)$$

$$\theta_m=\frac{V}{f} \quad \cdots\cdots(3)$$

In the present study on two-phase flow, the following definition was employed for V in Eqs. (1) and (3), taking into account the gas hold-up.

$$V=(1-\varepsilon)V_t \quad \cdots\cdots(4)$$

Results and Discussion

Over thirty experimental runs were carried out at different flow rate of liquid and levels of liquid in the column, using air-water and ais-glycerol solution (50% by weight) systems in three different columns; measurements of the tracer concentration in the samples taken were performed for each run.

The residence time distribution function, $E(\phi)$, was calculated from Eq. (1) using the results of the measurement. An example of an $E(\phi)$ curve is shown in Fig. 2. Curves of quite similar shape were also obtained from other runs. Fig. 2 indicates that, as expected from visual observation of the flow pattern of the tracer added, the liquid in the column exhibits a very close approach to plug flow, and the curves of $E(\phi)$ resemble closely the curves in the diffusion model.[4] In order to obtain the axial dispersion coefficient from the data, a convenient method was employed in the present study; this was to equate the measured variance with that in the diffusion model. The variance in the model is given by[4,5]

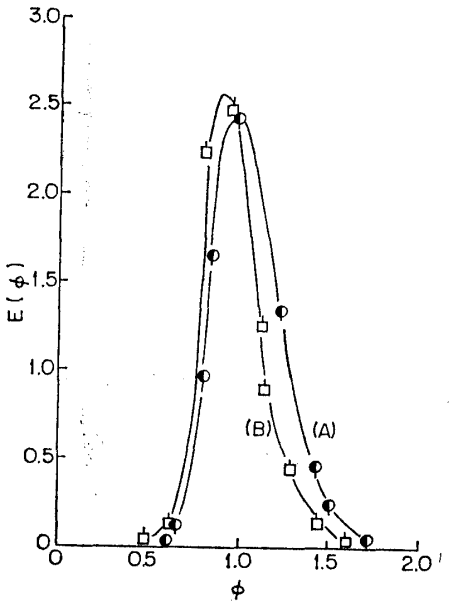

Fig. 2. Example of residence time distribution curves. (Cf. Fig. 4 for mark plotted)

$$\sigma_d^2=\frac{2}{(PeB)^2}(PeB-1+e^{-PeB}) \quad \cdots\cdots(5)$$

When the value of PeB is large, Eq. (5) can be reduced to

$$\sigma_\alpha^2 = \frac{2}{PeB} \quad \cdots\cdots(6)$$

where

$$PeB = \frac{FL}{D_L} \quad \cdots\cdots(7)$$

On the other hand, a plot of the cumulative fraction of $E(\phi)$ against ϕ on logarithmic probability paper yield a straight line as shown in Fig. 3. The cumulative fraction of $E(\phi)$ was obtained by the graphical integration of the $E(\phi) \sim \phi$ curve; Fig. 3 was thus derived from Fig. 2. The variance is given by

$$\sigma^2 = \frac{1}{2} \ln \frac{\phi_2}{\phi_1} \quad \cdots\cdots(8)$$

where ϕ_1 and ϕ_2 are the values of ϕ corresponding to the cumulative fraction of $E(\phi)$ being 15.9% and 84.1%, respectively. Equating Eq. (8) with Eq. (6), PeB can be calculated, and then the axial dispersion coefficient, D_L, is obtained using Eq. (7).

Axial dispersion coefficient thus calculated for all runs are shown in Fig. 4. It was noted that the values of D_L were approximately constant in error by no more than ±10% within the range of experiments. It may be concluded from the result that the degree of the axial dispersion of the liquid flowing in the column is not influenced by the liquid flow rate, liquid viscosity, liquid level in the

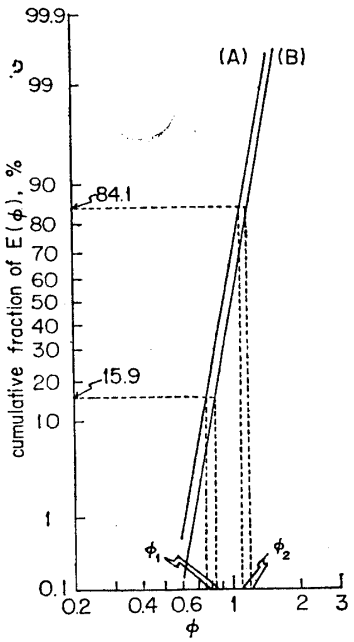

Fig. 3. Cumulative fraction of $E(\phi)$ described from data of Fig. 2.

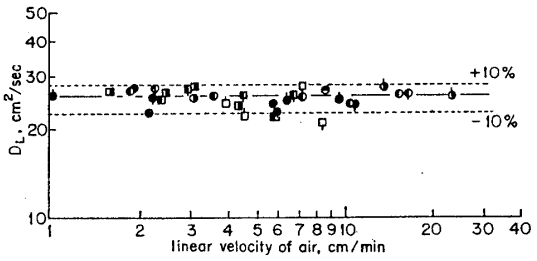

Fig. 4. Variation of D_L with linear velocity of air actually entrained.

Key

System	Column dia. (cm)	Mark	Liquid flow rate (kg/min)
air-water	4.04	◑	13.6
		◐	14.2
		⊖	14.7
		◑	15.3
		⚲	16.1
	5.5	●	24.5
		◕	25.4
		●	26.4
air-50% gly. soln.	3.0	□	9.0
		⚲	9.8
	4.04	◨	14.3
		◧	15.1
		⚲	16.0

column, or the column diameter within the range of the operation of this equipment.

There have been many studies on the mixing or axial dispersion of fluid in tubular reactors. Hikita[6] has reviewed some of the data on the axial dispersion coefficient in a bubble column. According to the review, the dispersion coefficient is independent of liquid flow rate, liquid viscosity, and liquid depth, but it increases with the increase of column diameter and gas flow rate. The same results have been reported by Kato.[7] Although there seem to be some difference between the characteristic of bubble columns and those of the column in the present investigation, the fact that the axial dispersion coefficient is independent of the liquid flow rate, liquid viscosity, and liquid depth agree with the result obtained in this study. No effect from the gas flow rate was found in the present study probably because of the low gas flow rate in this column compared to that in the bubble column.

Acknoledgement

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Notation

B = ratio of length to diameter in column ($=L/D$), dimensionless
c = concentration of tracer in sample, ppm
c_0 = concentration of tracer added, ppm
D = column diameter, cm
D_L = axial dispersion coefficient, cm^2/sec
$E(\phi)$ = residence time distribution function
f = volumetric flow rate of liquid, cm^3/sec
L = actual column length occupied by liquid ($=V/S$), cm
Pe = Peclet number based on column diameter, (FD/D_L), dimensionless
S = cross-sectional area of column, cm^2
v_0 = volume of tracer added, ml
V = actual volume of liquid in column ($=(1-\varepsilon)V_t$), ml
V_t = volume occupied by air-liquid phase, ml

Greek letters

ε = gas hold-up, cm^3/cm^3
θ = time elapsed, sec
θ_m = mean residence time, sec
ϕ = dimensionless time ($=\theta/\theta_m$)
$\sigma_\alpha^2, \sigma^2$ = variance of $E(\phi)$ in the diffusion model and in the experimnt, respectively.

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