DIMENSIONLESS CORRELATION OF LIQUID-PHASE DISPERSION COEFFICIENT IN BUBBLE COLUMNS

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Recently, the authors¹⁾ measured the liquid-phase dispersion coefficient in 10 cm and 19 cm I.D. bubble columns with single-nozzle gas spargers using air as the gas and using water and aqueous solutions of methanol and cane sugar as liquids. All the experimental data could be correlated well by the following empirical equation:

$$E_L = (0.15 + 0.69 u_G^{0.77}) D_T^{1.25} (1/\mu_L)^{0.12} \quad (1)$$

where E_L is the liquid-phase dispersion coefficient in m²/sec, u_G the superficial gas velocity in m/sec, D_T the column diameter in m, and μ_L the liquid viscosity in cP. This equation, however, is dimensionally inconsistent. In this communication, a new dimensionless correlation is proposed.

Dimensionless Correlation

Factors which affect the liquid-phase dispersion coefficient E_L in bubble columns with single-nozzle gas spargers may be considered to be the column diameter D_T , the superficial gas velocity u_G , the density ρ_L , the viscosity μ_L and the surface tension σ of the liquid, and the gravitational acceleration g . The effects of the nozzle diameter d_0 and the aerated liquid height Z_F on E_L were found in the previous work¹⁾ to be negligible. Thus, by dimensional analysis, we obtain:

$$\frac{E_L}{\sqrt{D_T^3 g}} = f\left(\frac{u_G}{\sqrt{D_T g}}, \frac{\mu_L^4 g}{\rho_L \sigma^3}, \frac{\rho_L D_T^2 g}{\sigma}\right) \quad (2)$$

Figure 1 presents the experimental data for the water-air system in both columns, 10 cm and 19 cm in diameter, as a log-log plot of $E_L/\sqrt{D_T^3 g}$ vs. $u_G/\sqrt{D_T g}$. The data points for two columns of different diameters are seen to fall on a single solid line represented by

$$\frac{E_L}{\sqrt{D_T^3 g}} = 0.077 + 0.39 \left(\frac{u_G}{\sqrt{D_T g}} \right)^{0.72} \quad (3)$$

This indicates that the effect of the dimensionless group $\rho_L D_T^2 g / \sigma$ on $E_L / \sqrt{D_T^3 g}$ is unimportant and can be neglected. In Fig. 2, the experimental data obtained with the methanol solution-air system and the cane sugar solution-air system in the 10 cm column

are shown as a log-log plot of $E_L/\sqrt{D_T^3 g}$ vs. $u_G/\sqrt{D_T g}$. As can be seen in the figure, the data points for both systems lie about 5 to 30% below the dotted line representing the data for the water-air system. The

Fig. 1 Liquid-phase dispersion coefficient for water-air system

Fig. 2 Liquid-phase dispersion coefficient for methanol solution-air and cane sugar solution-air systems

Fig. 3 Effect of physical properties of liquids on liquid-phase dispersion coefficient

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Table 1 Physical properties of liquids investigated

Liquid	Temp. [°C]	Density $\rho_L \times 10^{-3}$ [kg/m ³]	Viscosity $\mu_L \times 10^3$ [kg/m·sec]	Surface tension $\sigma \times 10^3$ [kg/sec ²]	$\mu_L^4 g / \rho_L \sigma^3$ [—]
Water	20.0	1.00	1.00	72.0	2.63×10^{-11}
8 wt % aq. methanol sol.	18.1	0.985	1.33	59.5	1.48×10^{-10}
15 wt % aq. methanol sol.	16.0	0.974	1.71	51.0	6.49×10^{-10}
53 wt % aq. methanol sol.	12.5	0.911	2.02	38.2	3.21×10^{-9}
35 wt % aq. cane sugar sol.	14.0	1.156	5.18	73.6	3.80×10^{-9}
50 wt % aq. cane sugar sol.	16.4	1.233	19.20	75.5	2.51×10^{-6}

vertical difference among the lines in Fig. 2 may be due to the difference in the value of the dimensionless group $\mu_L^4 g / \rho_L \sigma^3$. The values of $\mu_L^4 g / \rho_L \sigma^3$ for the liquids investigated are given in Table 1. In Fig. 3, the values of $E_L / \sqrt{D_T^3 g}$ read from Fig. 2 at a constant $u_G / \sqrt{D_T g}$ value of 0.10 are plotted against $\mu_L^4 g / \rho_L \sigma^3$ on logarithmic coordinates. Data points fall on a straight line with a slope of -0.03 , indicating that $E_L / \sqrt{D_T^3 g}$ should be proportional to $(\mu_L^4 g / \rho_L \sigma^3)^{-0.03}$.

From the results described above, the following equation correlating the experimental data was obtained:

$$\left(\frac{E_L}{\sqrt{D_T^3 g}} \right) \left(\frac{\mu_L^4 g}{\rho_L \sigma^3} \right) = 0.037 + 0.188 \left(\frac{u_G}{\sqrt{D_T g}} \right)^{0.72} \quad (4)$$

Figure 4 shows the comparison of the observed values of E_L presented in Figs. 1 and 2 with the values of E_L calculated from Eq. (4). The observed values are in good agreement with the calculated values, with an average deviation of 3.2%.

Nomenclature

D_T	= column diameter	[m]
d_0	= diameter of gas inlet nozzle	[cm]
E_L	= liquid-phase longitudinal dispersion coefficient	[m ² /sec]

Fig. 4 Comparison of observed and calculated values of liquid-phase dispersion coefficient

g	= gravitational acceleration	[m/sec ²]
u_G	= superficial gas velocity	[m/sec]
Z_F	= aerated liquid height	[m]
ρ_L	= density of liquid	[kg/m ³]
μ_L	= viscosity of liquid	[kg/m·sec] or [cP]
σ	= surface tension of liquid	[kg/sec ²]

Literature Cited

- 1) Hikita, H. and H. Kikukawa: *Chem. Eng. J.*, **8**, 191 (1974).

ON THE LOW REYNOLDS NUMBER MASS TRANSFER IN PACKED BEDS

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Considerable experimental information is available in the literature on the particle-fluid heat and mass transfer in packed beds^{1,10}. For the most part, these efforts are confined to the measurement of transfer rates to gases and liquids flowing at moderate

and high flow rates. Low rate data, especially for solid-liquid systems, are scanty and are mainly for spherical particles. In the present work new mass transfer rate data are obtained by measuring the rate of dissolution of compressed benzoic acid pellets into water and 60% aqueous propylene glycol solutions.

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