

Dispersion coefficients in water-air bubble co-current flow through packed columns

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ABSTRACT

The dispersion coefficient in the liquid phase (D_L) has been studied in a tubular vertical column (packed and unpacked) with water flowing upward and air bubbling co-currently. This has also been done without any air inside the column. Different flow rates of water have been applied in each case and the tracer responses of a pulse injection were measured by a conductivity cell. Experimental results for a packed column agree well with the predicted dispersion model presented here. The deviation in the case of an empty column was due to short circuiting of fluid, recycling of fluid or the existence of stagnant regions of fluid. The size of the packing was found to have a clear effect on the dispersion coefficient.

The results are useful in predicting the relationship between mixing parameters of a co-current dispersion model.

NOTATION

- C_L = water-phase tracer concentration, mol/m³
 D_L = axial dispersion coefficient of water based on water-occupied area of a column cross section, m²/s
 D_p = nominal size of solid packing, m
 F_L = water volumetric flow rate, m³/s
Pe = Peclet number = $(D_p U_L)/(\varepsilon_L D_L)$
 Q_L = amount of liquid-phase tracer imposed, mol
Re = Reynolds number = $(D_p u_L \rho)/\mu$
 t = time, s
 u_L = superficial water velocity (F_L /column cross section), m/s
 x = axial distance variable, m
 X = distance between tracer input point and response measuring point, m

GREEK LETTERS

- ε_L = volume fraction of water
 ρ = water density, kg/m³
 μ = water viscosity, Pa s

INTRODUCTION

A. Background

Three phase beds are used in coal liquefaction and gasification processes, and they have applications in petrochemical synthesis such as catalytic hydrogenation of petrochemical fractions. In all these cases, as well as in other mass transfer processes (e.g. absorption and extraction), counter-current flow of fluids is preferred as no accurate co-current dispersion model has been developed. Wakao & Funazkri (1978) listed a number of authors who did not consider the effect of dispersion on particle-to-fluid mass transfer coefficients in packed beds for counter-current flow. Similarly Shulman & Molstad (1950) and Calderbank & Moo-Young (1961) neglected the influence of axial dispersion. On the other hand, Towell *et al.* (1965), as well as Akita & Yoshida (1974), assumed total dispersion in the liquid phase and plug flow in the gas phase. Wakao & Tanaka (1974), Chang (1969), Reuss (1970) and many others took into account the dispersion in the liquid phase. Michelson & Ostergaard (1970) reported the relationship of the dispersion and mass-transfer coefficients to liquid flow rates. The dispersion coefficients increased with liquid flow rate. Considering all these evaluation procedures, Mangartz & Pilhofer (1981) discussed a dispersion model of mass transfer in addition to Lewis' (1922) and Whitman's (1923) *Film Theory* and Higbie's (1935) *Penetration Theory*. Danckwerts (1953) carried out experiments with water in a packed column and formulated axial boundary conditions which were verified by Wehner & Wilhelm (1956). Wakao & Funazkri (1978) corrected the gas phase and liquid phase mass-transfer data considering the effect of axial dispersion coefficients.

Similar studies were performed by others (Abou-El-Hassan *et al.* 1982; Burck 1975). However, their experiments were performed using counter-current flow of liquid and gas phases, unlike the present work which deals with co-current flow.

B. Method of approach

A number of dispersion models were given in a paper by Bischoff & Levenspiel (1962). Among such models, the axial dispersed plug-flow model is the most widely used. In this paper we present the measurement of the dispersion coefficient in the liquid phase as it is well established that the dispersion in the gas-phase has a negligible influence on the volumetric mass transfer coefficient.

Concentration profile measured by response techniques is widely used for calculating dispersion coefficients. The responses are created by injecting tracer substances. Different injection techniques were discussed by Abu-El-Hassan *et al.* (1982) and others (Perry & Chilton 1973; Reid *et al.* 1977). There the authors proposed a technique called the *Middle Injection Technique* to evaluate mixing by considering dispersions with appreciable back mixing. Since we are using a packed column and considering a dispersed plug-flow model, we cannot use this technique. Rather, a comparatively simple technique called the *One-shot Injection and Downstream Response Technique* has been utilized. The technique is also called the *Impulse Response Technique*. We used KCl solution (10% by wt) as a tracer liquid, the injected amount of which was so chosen that its concentration in the column was <0.001 molar, thus fulfilling the condition of proportionality of its concentration to conductivity. The dispersion coefficients have been determined using back calcu-

lations in the predicted model with the help of a computer program. The curves of Peclet number (dispersed) v . Reynolds number have been drawn for different packings both with and without air to show the clear picture of dispersion coefficients.

THEORETICAL MODEL

Consider a column in which water and air bubbles are co-currently flowing upward (Fig. 1). A liquid tracer is injected, as a delta function, uniformly over a cross section of the column at a distance $x = x_1$ in the column and the response signal is measured downstream away from the column exit at $x = x_1 + X$. If the tracer is non-volatile, a mass balance gives

$$\epsilon_L \frac{\partial C_L}{\partial t} + U_L \frac{\partial C_L}{\partial x} - \epsilon_L D_L \frac{\partial^2 C_L}{\partial x^2} = 0 \tag{1}$$

at $x = x_1, C_L = C_L(t)$
 where

$$\int_0^\infty C_L(t) dt = \frac{Q_L}{F_L}$$

$$C_L(t) = 0 \text{ for } t > 0$$

where D_L is the fluid dispersion coefficient, U_L is the superficial water velocity, ϵ_L is

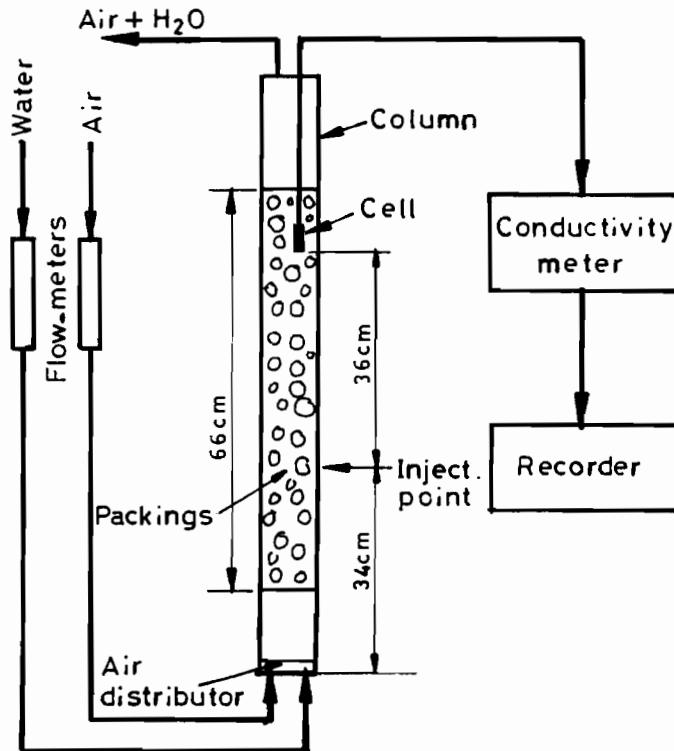


Fig. 1. Experimental set-up.

the volumetric fraction of water, F_L is the volumetric flow rate and Q_L is the amount of tracer injected.

Solving Eqn (1), the concentration C_L in the central portion of the column is expressed as

$$C_L = \frac{Q_L}{F_L} \frac{1}{\left[\frac{4\pi D_L t^3}{X^2} \right]^{1/2}} \exp \left[- \frac{\left(1 - \frac{tU_L}{\varepsilon_L X} \right)^2}{\frac{4D_L t}{X^2}} \right] \quad (2)$$

Once an impulse response, $C_L - t$ curve, is measured at any downstream point where no end effects exist in the column, the liquid phase dispersion coefficient may be determined by curve fitting in the time domain of the predicted models.

EXPERIMENTAL SET-UP

As shown in Fig. 1, the bubble column has a diameter of 7 cm and a height of 90 cm. The column was packed with glass packings consisting of hollow cylinders of outside diameter equal to the height. Two different sizes of packings were tested (4×4 mm and 5×5 mm). The injection port was placed at 0.34 m above the bottom of the column and the platinum type conductivity cell at 0.36 m above the injection port. Both air and water were passed from the bottom of the column (co-currently). Samples of 0.5 ml of 10% KCl solution at 25°C were injected as a tracer, the concentration of which is proportional to its conductivity. A normal syringe was used as an injector, the needle of which was pushed well into the perforated metallic distributor for a good planar distribution of tracer over the cross section of the column. The conductivity signal delivered by the cell was shown in the digital meter while being recorded in the connected recorder. C_L (concentration of tracer) in mol/m³ was calculated from conductivity readings. Then curves (C_L versus t) were compared with those theoretically predicted (with assumed values of dispersion coefficients) for calculating D_L . Theoretical curves were calculated by Eqn (2) of the original theory. Volume fraction of water was calculated both from flow meter readings and by direct measurement of outlet volumes.

RESULTS AND DISCUSSION

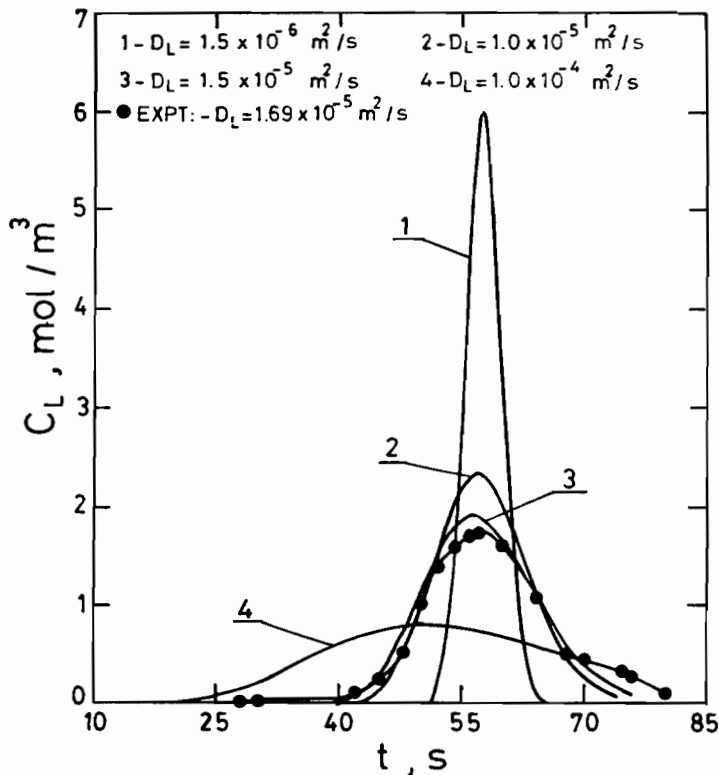
Results of dispersion coefficients in the liquid phase at different flow rates of water are shown in Table 1 for two different sizes of packing. These values of D_L were obtained by comparing the response curve measured with that predicted by Eqn (2). For simplicity, comparisons of only three experimental curves are shown in Figs 2a, 2b and 2c. Calculated results of Re and Pe are shown in Table 2 for these packings for different water-flow rates as before. Graphs of Pe versus Re are presented in Fig. 3, which reveals that Pe, based on D_p , may be considered to lie between 1.5 and 2.5. The dispersive Pe without air (which lies between 1 and 2) resembles those of single phase flow with axial mixing (Froment & Bischoff 1979). This indicates that the results for two-phase flow are reliable. Comparing with the experimental results of Carberry & Bretton (1958) and Ebach & White (1958) for single phase flow, the

Table 1. Calculation of D_L

$F_L \times 10^5$, m^3/s	$D_L \times 10^6, m^2/s$ Packings			
	4 × 4 mm		5 × 5 mm	
	with air	without air	with air	without air
1.79	7.4	11.0	14.2	15.5
2.04	8.8	11.5	15.1	14.9
2.29	11.1	11.5	16.9	15.2
2.54	12.8	10.9	17.7	16.7
2.78	14.6	15.0	19.2	20.5
3.04	12.4	15.5	19.5	20.4
3.25	16.8	14.2	20.4	22.8

limiting value of Pe (≈ 2) is well maintained in this experiment. The obtained values of Pe for two-phase flow may find useful application in column efficiency studies.

From the tabular and graphical data, it is observed that with the increase of the size of the packings in the column, D_L is generally increased. It is clear as well from Fig. 3 that the Peclet number is more stable with the air than without the air. This indicates that the dispersion coefficient with the air increases proportionately with

Fig. 2a. Calculation of D_L at $F_L = 2.29 \times 10^{-5}, m^3/s$.

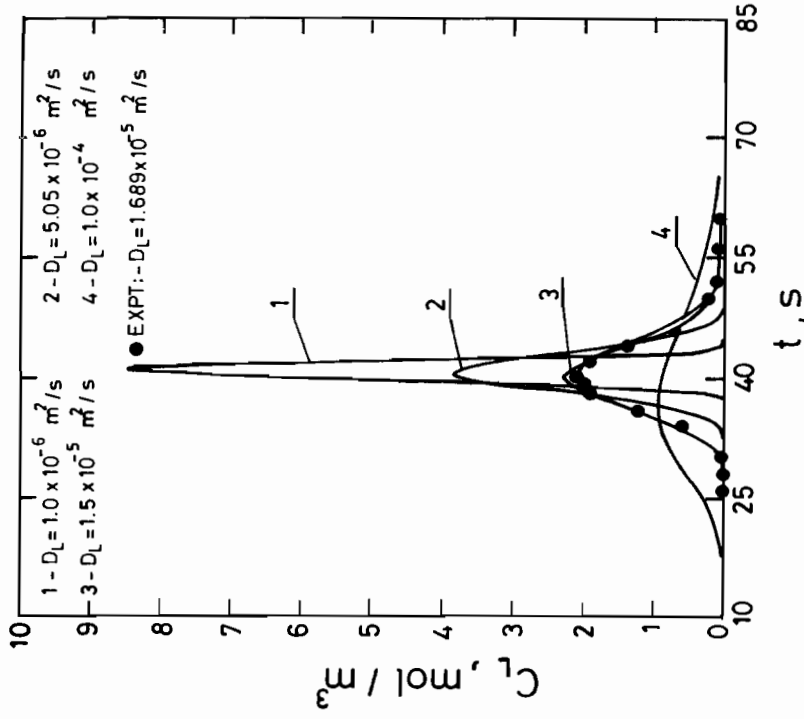


Fig. 2c. Calculation of D_L at $F_L = 3.25 \times 10^{-5} \text{ m}^3/\text{s}$.

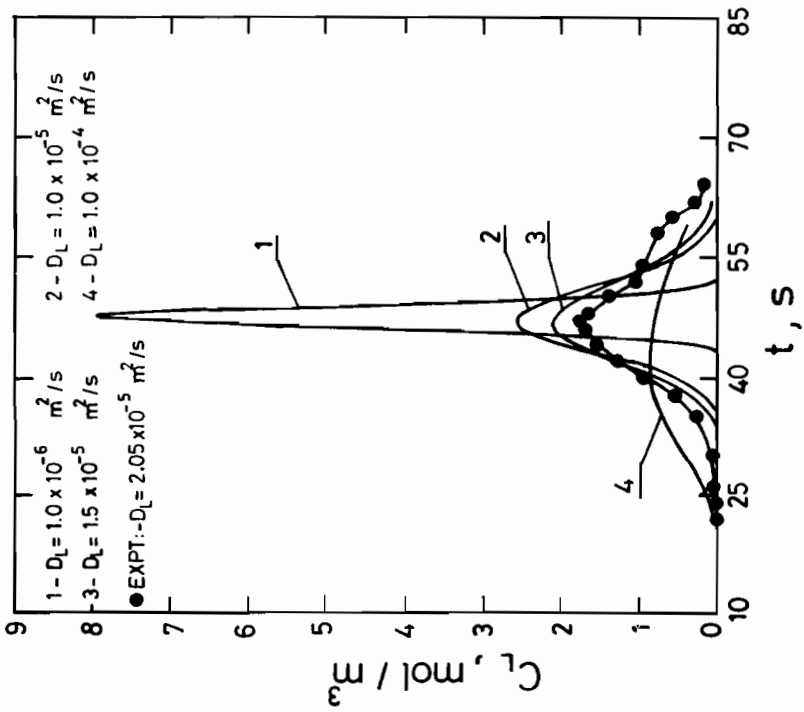


Fig. 2b. Calculation of D_L at $F_L = 2.78 \times 10^{-5} \text{ m}^3/\text{s}$.

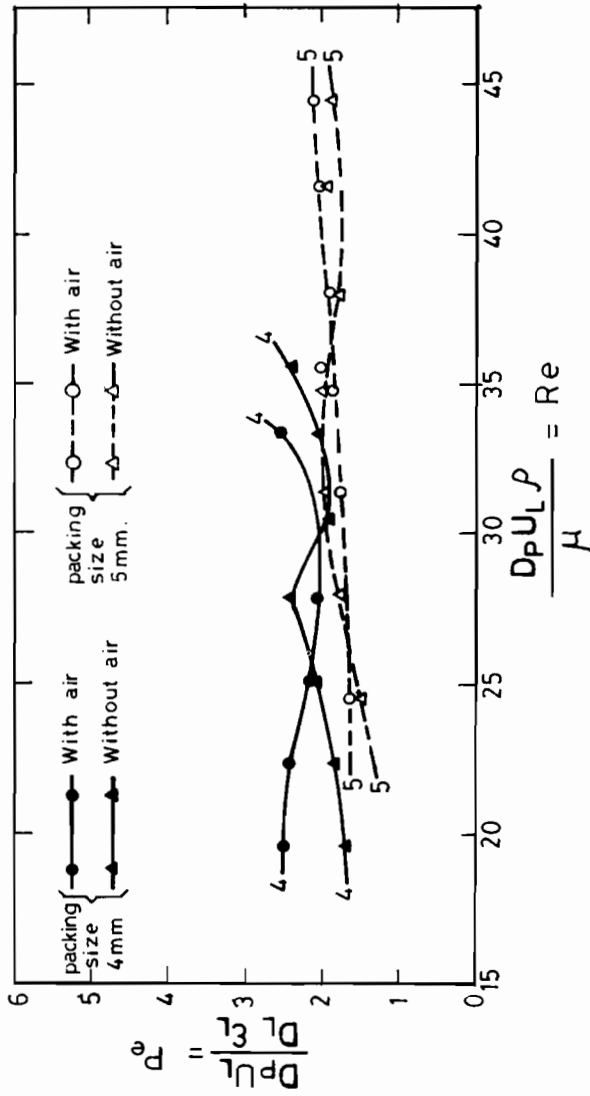


Fig. 3. Peclet No. (Pe) versus Reynolds No. (Re).

Table 2. Calculation of $Re = \frac{D_p U_L \rho}{\mu}$ and $Pe = \frac{D_p U_L}{D_L \varepsilon_L}$

$F_L \times 10^5$, m ³ /s	$U_L \times 10^3$, m/s	$D_p = 4 \times 4$ mm			$D_p = 5 \times 5$ mm		
		Re	Pe with air	Pe without air	Re	Pe with air	Pe without air
1.79	4.65	19.58	2.50	1.69	24.47	1.63	1.50
2.04	5.30	22.34	2.41	1.84	27.93	1.76	1.78
2.29	5.95	25.05	2.14	2.06	31.32	1.76	1.96
2.54	6.60	27.81	2.06	2.42	34.76	1.87	1.98
2.78	7.23	30.44	1.98	1.92	38.05	1.88	1.76
3.04	7.90	33.26	2.54	2.03	41.58	2.03	1.94
3.25	8.44	35.56	2.00	2.38	44.45	2.06	1.85

the superficial velocity and hence is more reliable. This is also a proof of uniform mixing of the components in the presence of air bubbles.

Curve fitting in the time domain shows that the theoretical model provides an accurate representation of the experimental data.

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قياس معامل التشتت في مزيج من الماء وفقااعات الهواء متحرك باتجاه واحد داخل أعمدة محشوة

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خلاصة

تم دراسة معامل التشتت في مزيج من الماء وفقااعات الهواء متحرك باتجاه واحد داخل أعمدة محشوة وأخرى خالية . كذلك درس هذا المعامل في وسط مائي خالٍ من فقاعات الهواء . ولقد تم قياس هذا المعامل عن طريق حقن المزيج المتحرك بمحلول كلوريد البوتاسيوم وتسجيل اثر ذلك بواسطة خلية التوصيل الكهربائي . وكانت نتائج التجارب متفقة مع نتائج النموذج الرياضي المقترح . والاختلاف في النتائج في حالة الاعمدة الخالية يمكن أن يكون بسبب سريان السائل بطريق مباشر دون مزج ، أو دوران السائل داخل العمود ، أو وجود مناطق ركود . ولقد تبين من الدراسة أن معامل التشتت يعتمد بطريقة مباشرة على حجم مادة الحشو . والنتائج التي توصلت اليها الدراسة مفيدة لحساب العلاقة بين عوامل الخلط المستخدمة في نموذج التشتت الرياضي للسوائل المتحركة باتجاه واحد .