AXIAL DISPERSION IN AN APPARATUS WITH MOBILE PACKING

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The paper deals with modelling of the flow of liquid on a plate with mobile packing. The results of measurements have been interpreted by means of a simple dispersion model whose two parameters were determined from the nonideal step-input of a tracer and its response. It has been found that in the gas flow rate region followed (1.0 – 3.5 m s^{-1}), the liquid flow rate followed (5.36 – 12.5 . 10^{-3} m s^{-1}), and the static bed height followed (21 – 47 . 10^{-3} m) the diffusion Péclet number is independent of these quantities. The mean residence time of the liquid on the plate decreases with increasing flow rate of both gas and liquid, but it increases with increasing height of packing.

Promising ways of realization of mutual contact between a gas and a liquid include the processes adopting a three-phase fluidized layer. The main element of an apparatus working on this basis is a plate with a layer of light, most often spherical particles which are brought into intensive random motion by the flow of both gas and liquid. The three-phase fluidized layer formed is characterized by favourable conditions for heat and mass transfer. The high flow rates of gas and liquid can then cause considerable mixing of the liquid, which affects infavourably the driving force in the heat and mass transfer. For a correct calculation of the heat and mass transfer coefficient it is necessary to know the quantities characterizing the intensity of mixing of the liquid on the plate.

The apparatuses with mobile packing have been paid relatively great attention since their introduction[1]. As compared with the total number of papers published in the field of hydrodynamics, heat and mass transfer, the number of papers dealing with the modelling of flow of liquid in this type of apparatus is considerably limited[2 - 10]. Although some of the papers are systematic and relatively extensive[3,5,7,9], so far no reliable correlation has been suggested for the calculation of intensity of mixing of liquid on a plate with mobile packing involving a wide range of working conditions.

The aim of the present work is to determine the basic quantitative data characterizing the intensity of mixing of liquid on a plate with mobile packing which are necessary for correct calculations of the mass transfer coefficient in the liquid.

THEORETICAL

Most frequently, the flow of a liquid on a plate with mobile packing is described by a simple dispersion model whose parameters can be determined by the transient response technique. This model is represented by the equation

$$\frac{\partial c}{\partial \tau'} = \frac{1}{Pe}\frac{\partial^2 c}{\partial Z^2} - \frac{\partial c}{\partial Z} \tag{1}$$

with one initial and two boundary conditions

$$c(Z, 0) = 0 \qquad \text{for} \quad 0 \le Z \le 1 \tag{2}$$

$$c\Big|_{Z \to 0^+} - \frac{1}{Pe}\frac{\partial c}{\partial Z}\Big|_{Z \to 0^+} = c\Big|_{Z \to 0^-} \tag{3}$$

$$\frac{\partial c}{\partial Z}\Big|_{Z = 1} = 0 \,. \tag{4}$$

In Eq. (*1*) $Z = z/h_d$ is a dimensionless length coordinate, $Z = 0$ corresponds to the inlet into the layer, $Z = 1$ corresponds to the outlet out of the layer (the plane of plate). $\tau' = \tau/\tau_m$ is the dimensionless time, and $Pe = \bar{u}_L h_d/D_L$ is the diffusion Péclet number in which $\bar{u}_L = h_d/\tau_m$ is the mean rate of flow of liquid in the layer. In Eq. (*3*) the expression $c|_{Z \to 0^-}$ denotes the time change of concentration of the tracer at the inlet into the layer. For certain forms of this function, then, there exists an analytical solution of Eq. (*1*). For the step-input response technique, which is often used for the apparatus with mobile packing, the solution is given in refs[11,12].

EXPERIMENTAL

The experimental set-up adopted is represented in Fig. 1. The main part was a laboratory column of $58.5 \cdot 10^{-3}$ m inner diameter with one plate made of a wire net with square meshes (the wire thickness $0.4 \cdot 10^{-3}$ m, the length of mesh side $1.7 \cdot 10^{-3}$ m). The free area of plate was 65.5%. The height of the contact space, i.e. the distance between the plate and mist eliminator was 0.45 m. In the upper part of this space there was a distributor of liquid which was constructed in such a way that the liquid flowed on the wall of the column down to the plate. The packing consisted of polystyrene spheres of $6 \cdot 10^{-3}$ m diameter and 1 000 kg m^{-3} density. The choice of column diameter was motivated by the aim to further utilize the results obtained earlier[13] and concerning the interfacial area in this type of apparatus.

All the measurements were carried out in the system air-tap water. The tracer was an almost saturated KCl solution which was injected into the inlet pipeline for liquid before the distributor of liquid. The KCl concentration in both streams of liquid was measured conductometrically. One conductometric cell was placed at the end of the distributor of liquid ($Z = 0$) and the other one immediately under the plate ($Z = 1$).

The height of the distributor above the plate was always adjusted at the value corresponding to the height of the expanded layer of packing which was estimated visually. For the conductometry we used an OK-102/1 apparatus (Radelkis). The signal from the conductometer in each branch was led to a 12 bite A/D converter which was interfaced with a PC/AT computer. The levels of digital signals were recalculated to the concentration data. The parameters of the dispersion model were determined separately from the data collecting using the procedure described in the next section. The time change of the tracer concentration at the inlet ($Z \rightarrow 0^-$) was realized as a nonideal step change. The flow rate of KCl solution was chosen in such a way as not to affect the total flow rate of liquid significantly (less than 1%) and to ensure a sufficient change of conductivity of the liquid. The sampling frequence was constant in all the cases, viz. 5 Hz. Within one measurement we obtained 120 – 190 triads of data (time, KCl concentration at $Z \rightarrow 0^-$, and that at $Z = 1$), and each measurement was repeated 5 – 10 times.

CALCULATIONS

The parameters of the model given by Eq. (*1*), i.e. the quantities Pe and τ_m, were determined by the procedure based on numerical integration of the differential equation with additional optimization procedure. The whole procedure can be summarized in the following steps:

1. The initial assessment of the mean residence time of liquid and that of the diffusion Péclet number (in all the experiments $\tau_m^{(0)} = 7$ s, $Pe^{(0)} = 5$).
2. The numerical integration of the partial differential equation (*1*). Obtaining of the values $c_{i\,\mathrm{calc}}$ for $Z = 1$.
3. The calculation of the objective function defined by the equation

$$F(Pe, \tau_m) = \sum_{i=1}^{m} \left(c_{i\,\mathrm{exp}} \Big|_{Z=1} - c_{i\,\mathrm{calc}} \Big|_{Z=1} \right)^2 . \qquad (5)$$

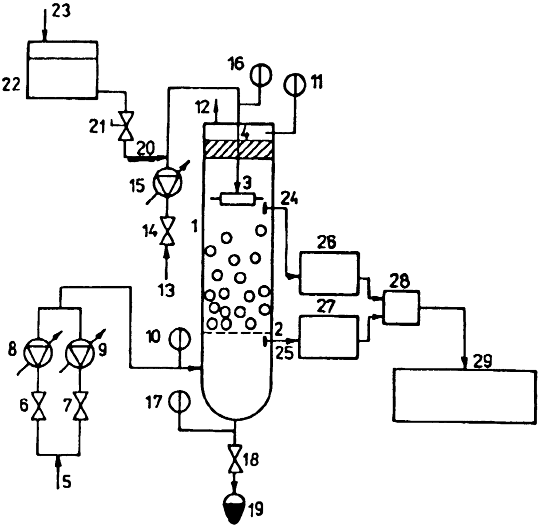

FIG. 1

Scheme of the experimental set-up: 1 column; 2 plate with mobile packing; 3 distributor of liquid; 4 mist eliminator; 5 air inlet; 6, 7, 14, 18 valves; 8, 9, 15 rotameters; 10, 11, 16, 17 thermometers; 12 air outlet; 13 water inlet; 19 waste; 20 capillary; 21 cock; 22 reservoir for KCl solution; 23 pressure air; 24, 25 conductometric cells; 26, 27 conductometers; 28 A/D converter; 29 PC/AT computer

4. The carrying out of one step of minimization of the objective function (*5*) by the modified simplex method – the algorithm by Nelder and Mead[14]. In this step, corrected values of *Pe* and τ_m are obtained.

5. The procedure beginning from step 2 was repeated until the minimum of the objective function (*5*) was reached.

Integration of Partial Differential Equation (1)

The rectangular region for which the solution of the partial differential equation was to be found with one initial and two boundary conditions was divided by the integration steps h and k in the directions Z and τ', respectively, into N and M nodal points, respectively (in all the experiments it was $h = 0.05$ and $k = 0.2/\tau_m$). In order to ensure stability of solution, we adopted the Crank Nicolson method which necessitated, for computation of the j-th concentration profile ($j = 2, 3, \ldots, M$), a simultaneous solution of a set of N linear algebraic equations. This set involves the equation

$$(1 + \frac{3}{2h\,Pe})\, c_{1,j+1} - \frac{2}{h\,Pe}\, c_{2,j+1} + \frac{1}{2h\,Pe}\, c_{3,j+1} = c_{1,j+1}\Big|_{Z \to 0^-} ,$$

$$i = 1 \qquad (6)$$

which results from discretization of the boundary condition (*3*) in which the derivation in the left limit point is replaced by a three-point formula. The discretization of the differential equation (*1*) gives (N – 2) linear equations

$$- \frac{k}{2h}\left(\frac{1}{h\,Pe} + \frac{1}{2}\right) c_{i-1,j+1} + (1 + \frac{k}{h^2\,Pe})\, c_{i,j+1} -$$

$$- \frac{k}{2h}\left(\frac{1}{h\,Pe} - \frac{1}{2}\right) c_{i+1,j+1} = \frac{k}{2h}\left(\frac{1}{h\,Pe} + \frac{1}{2}\right) c_{i-1,j} +$$

$$+ (1 - \frac{k}{h^2\,Pe})\, c_{ij} + \frac{k}{2h}\left(\frac{1}{h\,Pe} - \frac{1}{2}\right) c_{i+1,j}$$

$$i = 2, 3, \ldots N - 1 . \qquad (7)$$

The last, N-th, equation is obtained by discretization of the partial differential equation (*1*) for $Z = 1$ with taking into account the boundary condition (*4*)

$$-\frac{k}{h^2 Pe} c_{N-1,j+1} + \left(1 + \frac{k}{h^2 Pe}\right) c_{N,j+1} = \frac{k}{h^2 Pe} c_{N-1,j} + \left(1 - \frac{k}{h^2 Pe}\right) c_{Nj}$$

$$i = N . \qquad (8)$$

The matrix of coefficients of the set of Eqs (*6*) – (*8*) is characterized by a large number of zero elements. Therefore, the method of factorization was adopted to solve the equation set mentioned.

A typical course of the input and output (both experimental and optimized) signals is given in Fig. 2 for one measurement. The graphical comparison of calculated and experimental concentrations of the tracer at the outlet shows that the procedure used leads to a satisfactory agreement between the model and experiment.

RESULTS AND DISCUSSION

The experimental work was directed to estimation of the effects of the gas and liquid flow rates and of the static bed height on the degree of mixing of liquid on the plate with mobile packing. The gas flow rate was varied between the limits from 1.0 to 3.5 m s^{-1}, the liquid flow rate was 5.36, 8.93, and 12.5 . 10^{-3} m s^{-1}. The effect of static bed heights was monitored at three levels: 21, 32, and 47 . 10^{-3} m.

Figure 3 presents the dependence of the diffusion Péclet number on the gas flow rate for the static bed height h_0 = 47 . 10^{-3} m. From the picture it is obvious that the obtained

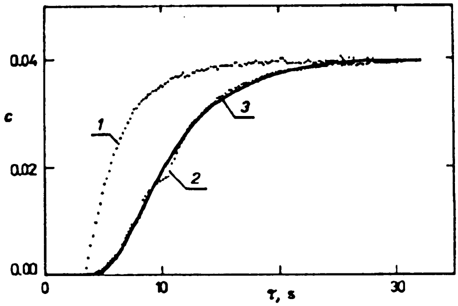

FIG. 2

Time dependence of concentration of tracer *c* (kmol m^{-3}): *1* experimental values (inlet), *2* experimental values (outlet), *3* calculated values (outlet)

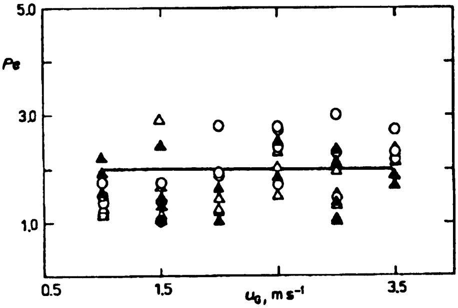

FIG. 3

Dependence of diffusion Péclet number on gas flow rate for h_0 = 47 . 10^{-3} m and u_L: ○ 5.36 . 10^{-3} m s^{-1}, ▲ 8.93 . 10^{-3} m s^{-1}, Δ 12.5 . 10^{-3} m s^{-1}

values of the diffusion Péclet number show a large scattering, the effect of the liquid flow rate on the Péclet number being statistically insignificant. The values of the *Pe* number vary within the limits from 1.1 to 3.3, the mean value being 2.0 ± 0.5. The effect of the gas and liquid flow rates on the diffusion Péclet number can be evaluated in similar way for other values of the static height of packing, too.

The same conclusions were arrived at also by Uysal and Ozilgen[10] who estimated the values of the diffusion Péclet number in a column of 0.195 m diameter. The only parameter which affected the *Pe* values was the diameter of the packing particles. The authors explain the large scattering of the diffusion Péclet number referred to the diameter of packing particles (approximately from $1.3 \cdot 10^{-2}$ to $1.7 \cdot 10^{-1}$) as a result of intensive and random motion of spherical particles in this type of apparatus.

Wasowski and Mlodzinski[8] found that in the region of developed fluidization the mean values of the Péclet number varies within the limits from 1.5 to 2.5. These values approach the *Pe* values obtained in the present work although the diameter and density of the spherical particles used are different, viz. $10 - 20 \cdot 10^{-3}$ m and 260 – 280 kg m^{-3}, respectively.

The values of the diffusion Péclet number published in other papers differ from those given in this report. Kuprianov, Kan and Jatskov[2] found extremely low values of the diffusion Péclet number (0.11 – 0.43) wherefrom they deduce the conclusion of ideal mixing of liquid on the plate with mobile packing. The *Pe* values found by Koval, Bespalov and Kuleshov[4] lie approximately in the same range as those published by Chen and Douglas[3], i.e. from 2.5 to 19.5. For particles with very low density (53 and 112 kg m^{-3}) and for $h_0/D_c = 2.4$, the authors[7] found the *Pe* values from 7 to 14 for spherical particles and from 3 to 18 for irregular particles. Then for the ratio h_0/D_c =

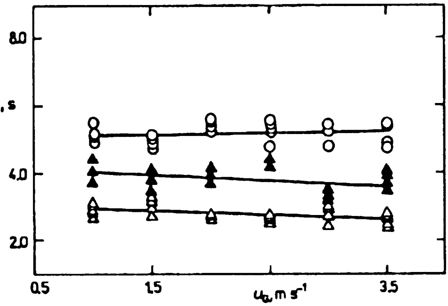

FIG. 4

Dependence of mean residence time of liquid on gas flow rate for $h_0 = 21 \cdot 10^{-3}$ m and u_L: ○ $5.36 \cdot 10^{-3}$ m s^{-1}, ▲ $8.93 \cdot 10^{-3}$ m s^{-1}, Δ $12.5 \cdot 10^{-3}$ m s^{-1}

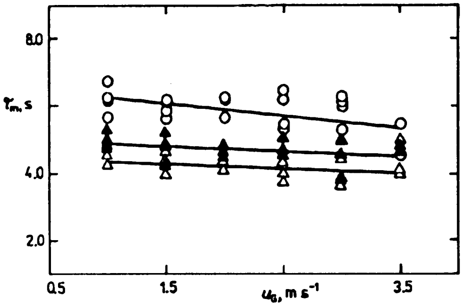

FIG. 5

Dependence of mean residence time of liquid on gas flow rate for $h_0 = 32 \cdot 10^{-3}$ m and u_L: ○ $5.36 \cdot 10^{-3}$ m s^{-1}, ▲ $8.93 \cdot 10^{-3}$ m s^{-1}, Δ $12.5 \cdot 10^{-3}$ m s^{-1}

0.78 the following *Pe* ranges were obtained: 1 – 6 (spherical particles) and 2 – 6 (irregular particles).

Figures 4 – 6 show the dependences of the mean residence time of liquid on the gas flow rate; the parameter of individual lines is the liquid flow rate. In contrast to the diffusion Péclet number, the mean residence time of liquid is affected by the gas and liquid flow rates and by the static bed height as well. From the dependences given it follows that the mean residence time of liquid decreases with increasing liquid flow rate, and also the dependence $\tau_m = f(u_G)$ (except for the dependence for $u_L = 5.36 \cdot 10^{-3}$ m s^{-1} and $h_0 = 21 \cdot 10^{-3}$ m) shows a moderately decreasing trend, this dependence being approximatable by the straight lines

$$\tau_m = a_0 + a_1 u_G, \qquad u_L = \text{const}, \qquad h_0 = \text{const}. \qquad (9)$$

For the individual cases the constants a_0 and a_1 were estimated by linear regression and are given in Table I. From the comparison of the dependences $\tau_m = f(u_G)$ for the individual static bed heights and constant liquid flow rate it follows that an increase in the static bed height causes an increase in the liquid residence time on the plate of the apparatus.

TABLE I
Constants a_0 (s) and a_1 (s^2 m^{-1}) of Eq. (9)

$u_L \cdot 10^3$ m s^{-1}	$h_0 = 21 \cdot 10^{-3}$, m		$h_0 = 32 \cdot 10^{-3}$, m		$h_0 = 47 \cdot 10^{-3}$, m	
	a_0	a_1	a_0	a_1	a_0	a_1
5.36	5.038	0.057	6.591	-0.355	9.502	-0.378
8.93	4.196	-0.163	5.027	-0.156	7.934	-0.588
12.50	3.036	-0.130	4.458	-0.133	7.092	-0.539

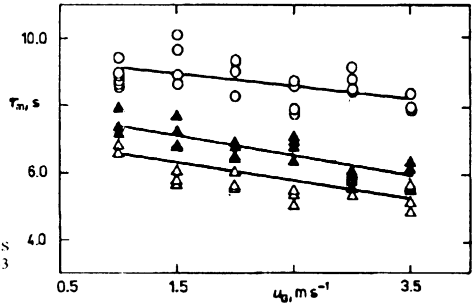

FIG. 6
Dependence of mean residence time of liquid on gas flow rate for $h_0 = 47 \cdot 10^{-3}$ m and u_L: ○ 5.36 . 10^{-3} m s^{-1}, ▲ 8.93 . 10^{-3} m s^{-1}, △ 12.5 . 10^{-3} m s^{-1}

The following equation was suggested for a quantitative evaluation of the effect of the gas and liquid flow rates and of the static bed height on the mean residence time of liquid:

$$\tau_m = D_1\, u_G^{p_1}\, u_L^{p_2}\, h_0^{p_3} \,, \tag{10}$$

where the constant D_1 and the exponents p_i (i = 1, 2, 3) were determined by nonlinear regression. The final form of the correlation (*10*) reads as follows:

$$\tau_m = 7.64\, u_G^{-0.094}\, u_L^{-0.469}\, h_0^{0.749} \,, \tag{11}$$

where $[\tau_m]$ = s, $[u_G]$ = m s^{-1}, $[u_L]$ = m s^{-1} and $[h_0]$ = m.

The suitability of this equation was judged on the basis of the mean relative quadratic error of correlation

$$\bar{\delta} = \sqrt{\frac{1}{n}\sum_{i=1}^{n}\left(\frac{\tau_{\mathrm{mi\,exp}} - \tau_{\mathrm{mi\,calc}}}{t_{\mathrm{mi\,exp}}}\right)^2} \,.\, 100 \,. \tag{12}$$

This error is 10.4% for a set of n = 392 measurements.

CONCLUSIONS

The experiments aimed at estimation of mixing intensity of a liquid on a plate with mobile packing. The results were treated with application of a simple dispersion model. The parameters of the model – the diffusion Péclet number and the mean residence time – were estimated from the nonideal step change of tracer concentration (KCl solution) at the inlet and from its response by the method of numeric integration of the principal partial differential equation and subsequent optimization of parameters. It was found that in the investigated ranges of gas and liquid flow rates and static bed heights the diffusion Péclet number shows a constant value (Pe = 2.0). The mean residence time of liquid on the plate decreases with incresing gas and liquid flow rates, whereas it increases with increasing static bed height. The results of investigation are presented graphically and, for the mean residence time of liquid, also in the form of an empirical correlation.

SYMBOLS

a_0	constant in Eq. (9), s
a_1	constant in Eq. (9), s^2 m^{-1}
c	molar concentration, kmol m^{-3}
D_c	column diameter, m
D_L	dispersion coefficient in liquid, m^2 s^{-1}
D_1	constant, s$^{(1 - p_1 - p_2)}$ m$^{-(p_1 + p_2 + p_3)}$

F	objective function, kmol^2 m^{-6}
h	integration step in the direction Z
h_0	static bed height, m
h_d	height of expanded packing layer, m
k	integration step in the direction τ'
M	number of nodal points in the direction τ'
m	number of concentration data in a time series
N	number of nodal points in the direction Z
n	number of measurements
Pe	diffusion Péclet number
p_i	$i = 1, 2, 3$; exponents in Eq. (*10*)
u	velocity, m s^{-1}
u_L	mean flow rate of liquid in layer, m s^{-1}
Z	dimensionless length coordinate
z	length coordinate, m
$\bar{\delta}$	mean relative quadratic error, %
τ	time, s
τ'	dimensionless time
τ_m	mean residence time of liquid, s

Subscripts

calc	calculated value
exp	experimental value
G	referring to gas
L	referring to liquid

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