INVESTIGATION OF LIQUID-PHASE-LIMITED MASS TRANSFER ON GRID TRAYS

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Studies of mass transfer in the liquid phase have been carried out in columns with inside diameters of 192 and 784 mm, in the desorption of carbon dioxide from water by air. The procedure used in conducting these experiments has been described in [1]. The geometric characteristics of the test trays are listed in Table 1. In all, about 800 experiments were performed.

It was shown experimentally in [2] that, on shower-type trays, the liquid is almost completely mixed with respect to height of the bubbled layer. However, data reported in [3] indicate that, under certain tray operating conditions, liquid mixing may be substantially less than complete. One of the explanations of this fact [3] is the significant residence time of the liquid in the descending jets, in comparison with the average residence time of the liquid on the tray.

Data are listed in Table 2 for the quantity of liquid found in the jets, and on the total retention of liquid on the tray, under certain operating regimes of the column (Tray No. 1). Both quantities were determined by a "cutoff" method in a column with the trays and in a complete column. As can be seen from these data, a considerable part (up to 50%) of the total stock of liquid on the trays may be held in the descending jets.

There is no basis for supposing that the character of mixing in the descending jets will differ substantially from plug flow. Hence, in describing the mixing process on grid trays, we have adopted a model of a total-mixing cell with a plug-flow stream. The fraction of the plug flow p was determined on the basis of Eq. (1), which was obtained by integrating the area occupied by the jet of liquid with gravitational flow of the jet from the slots at an initial velocity v_0, through the height of the separation space H:

$$p = \frac{L}{gh}\left[\sqrt{v_0^2 + 2gH} - v_0\right], \quad (1)$$

where g is the acceleration of gravity, m/sec^2. The initial velocity of fall was assumed to be equal to the flow velocity of the liquid through the tray slot:

$$v_0 = \frac{L}{F\tau}, \quad (2)$$

where τ is the fraction of the cross-sectional area of the grid-tray slots occupied by the liquid downflow.

An attempt to describe our experimental data by equations proposed in [4-12] for downcomerless trays proved to be unsuccessful. The best agreement with experiment was given by calculations using the equations of [11]. The standard error in calculating the tray efficiency using this equation was 24.6% for all experiments, and the error in calculating the number of transfer units was 90.8%.

The equations of [4-12] describe the process of mass transfer in a layer of froth on the assumption that the elementary act of the process is the transfer of mass from the bulk liquid, which is the continuous phase, to a gas bubble. For systems in which the gas content of the layer is relatively low and hence the gas phase is primarily the disperse phase, such an approach is justified. In the tray operating regimes that we have studied, the froth consisted of a suspension of films, sprays, and drops of liquid; and the continuous-phase

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TABLE 1. Geometric Characteristics of Test Trays

Tray No.	Column diam D, mm	Slot width a, mm	Open area F, %	Tray thickness δ, mm	Coefficient of resistance of dry tray ξ	Tray spacing H_t, mm	Tray material
1	192	4,0	23,3	3,0	1,5	500	Clear plastic
2	192	6,0	23,3	3,0	1,63	500	" "
3	192	8,0	22,0	3,0	1,79	500	" "
4	192	10,2	23,9	3,0	1,84	500	" "
5	192	6,1	5,0	2,0	1,90	500	Steel
6	192	6,1	15,0	2,0	1,76	500	" "
7	192	6,1	24,7	2,0	1,62	500	" "
8	192	6,1	35,0	2,0	1,18	500	" "
9	784	5,4	26,0	3,0	1,68	700	Clear plastic
10	784	4,0	33,0	3,0	1,25	700	" "

TABLE 2. Quantity of Liquid Retained on Trays and in Inter-Tray Space

Liquid load, $m^3/(m^2 \cdot h)$	Gas velocity w, m/sec	Height of static liquid layer on trays h, mm	Fraction of plug flow p	Calculation according to Eq. (1)	Height of static layer corresponding to quantity of liquid present in separation space h_e, mm	Average residence time of liquid on tray θ, sec
5,0	1,0	9,3	0,485	0,065	4,9	6,69
5,0	1,55	13,2	0,334	0,044	4,9	9,64
5,0	2,1	13,0	0,332	0,045	4,9	9,36
5,0	2,9	13,4	0,285	0,04	4,9	9,65
5,0	3,25	17,1	0,206	0,03	4,9	12,28
5,0	3,53	23,0	0,132	0,03	4,9	16,52
20,0	1,10	24,4	0,281	0,092	8,45	4,37
20,0	1,56	33,5	0,182	0,061	8,45	6,03
20,0	2,0	35,8	0,165	0,056	8,45	6,44
20,0	2,7	37,6	0,103	0,045	8,45	6,77
20,0	3,2	48,4	0,056	0,03	8,45	6,69

approximation for the liquid is purely arbitrary in this instance. Hence, an attempt has been made to describe the process of mass transfer in such systems on the assumption that the liquid is the disperse phase under these conditions.

It is known [13] that mass transfer in liquid drops is best described by the Kronig–Brink model [14], which takes into account internal circulation of liquid in the drops. According to this model, the mass concentration in a drop is a function of the time of drop existence:

$$C(t) = \frac{c(t)}{c(0)} = \frac{3}{8} \sum_{n=1}^{\infty} A_n^2 \exp\left(-\mu_n \frac{16Dt}{R^2}\right), \quad (3)$$

where C(t) is the dimensionless mass concentration at the moment of time t; c(t) is the concentration of component being transferred in the drop at the moment of time t, moles/liter; c(0) is the concentration of component being transferred in the drop at the moment of drop formation, moles/liter; D is the molecular diffusivity, m^2/sec; R is the drop radius, m; A_n and μ_n are calculated, and for the first two terms of the series are $A_1 = 1.32$, $A_2 = 0.73$, $\mu_1 = 1.678$, and $\mu_2 = 9.83$.

For the degree of extraction E(t), we can obtain from (3) the following equality [15]:

$$E(t) = \left[1 - \exp\left(-2.25 \frac{\pi^2 Dt}{R^2}\right)\right]^{1/2}. \quad (4)$$

Assuming that the time of existence of the drops t is proportional to the average residence time of the liquid on the tray θ, let us now expand the exponential in Eq. (4) in a Maclaurin series; and, for residence times that are not too large ($\theta \leq 50$ sec), let us limit ourselves to the first two terms of the series. We obtain:

$$E(t) = \frac{1.5\pi}{R}\sqrt{D\theta}\,. \qquad (5)$$

Taking into account the fact that, with total mixing of the liquid in the froth layer, the mass transfer coefficient referred to unit area of the tray, $\beta_i a$, is related to the degree of extraction by the equation

$$\beta_i a = L\frac{E}{1-E}, \qquad (6)$$

we find that

$$\beta_i a = L\frac{1.5\pi\sqrt{D\theta}}{1.5\pi\sqrt{D\theta}}.^{*} \qquad (7)$$

Equation (7) was used in working up the experimental data satisfying the condition $\theta \leq 50$ sec; here the drop radius was found by minimizing the sum of the squares of the deviations of the calculated values from the experimental values for the mass transfer coefficients. The calculations showed, however, that the best agreement with experiment is obtained if we assume that the time of drop existence is proportional to the 0.23 power of the average residence time of the liquid on the trays. The final equation for the mass transfer coefficient has the form

$$\beta_i a = \frac{1.43L(D\theta)^{0.115}}{R^{0.23} - 1.43(D\theta)^{0.115}}. \qquad (8)$$

Generalizing Eq. (8) to extend its validity to times greater than 50 sec, we obtain for the degree of extraction an equation analogous to (4):

$$E = \left[1 - \exp\left(-2.25\frac{\pi^2 D\theta}{R^2}\right)\right]^{0.115}. \qquad (9)$$

Here the drop radius in the system under investigation proved to be 0.00417 m. The standard error of calculation with Eq. (9) for all experiments was 12.47%, and the coefficient of correlation between the calculated and experimental values of tray efficiency was 0.91.

The relatively low exponent for the complex $(D\theta/R^2)$ in Eq. (8) reflects, in our opinion, inadequate accounting for the effect of incomplete mixing of liquid on the trays. As can be seen from Table 2, Eq. (1) gives values for the fraction of plug flow p that are unduly low in comparison with the measured values; for each liquid load, as the average residence time of liquid on the tray is decreased, the error of calculation increases. The deviations between the calculated and experimental values of p are apparently related to the phenomenon of wetting of the column walls and rundown of liquid as a film on the walls; this is not taken into account in Eq. (1). Nonetheless, as shown by the results from calculation of the experimental data obtained in the 784-mm diameter column, in which the fraction of liquid running down the walls must be substantially lower, the exponent for the complex $(D\theta/R^2)$ is no greater than 0.2.

The average residence time of liquid on the trays in our experiments varied from 2 to 150 sec. The experimental data reported in [10] for the desorption of carbon dioxide from dimethylformamide solutions on 180-mm diameter grid trays with an open area of 13% were obtained with an average residence time of the liquid on the trays amounting to 0.3-0.5 sec. Calculations have shown that these data follow Eq. (9) with an accuracy of ± 3.8, Eq. (9).

Treatment of the experimental data of [16] on the joint desorption of carbon dioxide and helium from water by air on shower-type trays has shown that the relationship between the degree of extraction and the molecular diffusivity follows Eq. (9).

A relation between the liquid-phase mass transfer coefficient and the average residence time of the liquid on the trays has been recognized for sieve trays and bubble-cap trays [17, 18].

As can be seen from Eq. (8), the liquid-phase mass transfer coefficient is very highly dependent on the liquid load. This fact has been noted repeatedly in experimental studies [19]. Meanwhile, if we assume that the liquid phase on the trays is continuous, then the rate of relative movement of the bubble, which enters into

*As in Russian original; denominator apparently should be $R - 1.5\,\pi\sqrt{D\theta}$ – Translator.

the equation for the diffusional flux, must be defined not in terms of the relatively low linear velocity of the liquid, but in terms of strong circulating flows taking place on the trays [20].

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HYDRAULIC RESISTANCE IN PNEUMOTRANSPORT BY RETARDED DENSE BED

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The characteristics of pneumotransport by a retarded dense bed (RDB) and the determination of the rate of movement of the transported material have been examined previously [1]. Here we are presenting results from a study of hydraulic resistance in pneumotransport by a RDB.

As was shown previously [1], the void fraction of a retarded dense bed is constant from bottom to top of the transport pipe and is somewhat (6-8%) higher than the void fraction of a nonmoving bed of the material being transported. It was also established that the particles of the material execute shifts of the pulsation type relative to each other. From these facts we can conclude that the previously proposed hypothesis of Berg and Vel'shof [2] on plug flow of the dense bed is not confirmed by experiment. Hence, for the description of the resistance of a retarded dense bed, we are proposing here a different model, taking into account the above-noted structural characteristics of the bed in RDB pneumotransport.

In this model, movement of an element of the bed is regarded as a shift of a gas-permeable porous "piston" that has the same void fraction as that of the moving bed; the solid particles filling the volume of the "piston" are not rigidly interconnected, rather executing small relative shifts within the volume of the element.

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