INTERPHASE MASS TRANSFER IN A FLUIDIZED BED WITH A PACKING

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Measurements are made of interphase mass-transfer coefficients in a fluidized bed with small- and medium-volume packings. Our experimental data and material from other studies are generalized on the basis of a proposed model.

The occurrence of adsorptive and catalytic processes in a fluidized bed (FB) with a packing is governed to a considerable extent by mass exchange between the gas and particles. The literature contains experimental data on the interphase mass-transfer coefficients in an FB with small-volume packings in the form of beds of coiled wires or netted cylinders [1-4]. Some information on interphase mass transfer in an FB with medium-volume packings occupying 10-50% of the volume of the layer can be found in [5] (spherical packing) and [6] (Raschig rings). This information is fragmentary, however, and there are no generalizing correlations for the mass-transfer coefficient in an FB with a packing. Also, the data in these studies were obtained indirectly by the model-chemical-reaction and gas-tracer methods. With these methods, the empirical output concentrations are analyzed by means of earlier-adopted approximate models.

The main laws governing interphase mass transfer in an FB with a packing are not yet established, which makes it difficult to construct theoretical models. Here we attempt to obtain information on interphase mass transfer (gas—solid) in an FB with small- and medium-volume packings, including certain promising modifications on which there is not yet any data [7, 8]. We will also attempt to determine the laws governing mass transfer in the gas phase of such a system.

The nonuniform FB consists of two phases: a continuous phase comprised of particles suspended in a gas and having a porosity close to the porosity of the FB at the beginning of fluidization; a discontinuous phase comprised of gas bubbles. At high gas filtration velocities, a "phase inversion" takes place in an FB with a packing. So-called suspended packets (aggregates) of particles float in the gas flow [9]. Momentum is transferred, energy is dissipated, and a component is transferred from the gas flow to the suspended particles at the surface of the aggregates. All this occurs in a boundary layer by a molecular mechanism. For the diffusion boundary layer, in which the concentration profile can be assumed linear in a first approximation, we have

$$\beta^* = D/\delta_{\mathrm{d}} \tag{1}$$

We write the following relation for changing over from β^* to the volumetric mass-transfer coefficient, referred to a unit volume of the fluidized system:

$$\beta = \beta^* \frac{S_{\mathrm{s}}}{V_{\mathrm{s}}} \frac{V_{\mathrm{s}}}{V_{\mathrm{p}}} \frac{V_{\mathrm{p}}}{V}. \tag{2}$$

It can be suggested, on the basis of geometric considerations, that $S_s/V_s \sim 1/\delta_k$. Also considering that the size of the particle aggregates is limited by the size of the channels created by the packing, we can assume that $S_s \sim S_p$ and

$$\frac{V_{\mathrm{s}}}{V_{\mathrm{p}}} \sim \frac{S_{\mathrm{s}}}{S_{\mathrm{p}}} \frac{\delta_{\mathrm{k}}}{l_{\mathrm{p}}} \sim \frac{\delta_{\mathrm{k}}}{l_{\mathrm{p}}},$$

where $V_p/V = \varepsilon_p$ by definition. Inserting these relations into (2) gives us

$$\beta \approx \beta^* \frac{\varepsilon_{\mathrm{p}}}{l_{\mathrm{p}}}. \tag{3}$$

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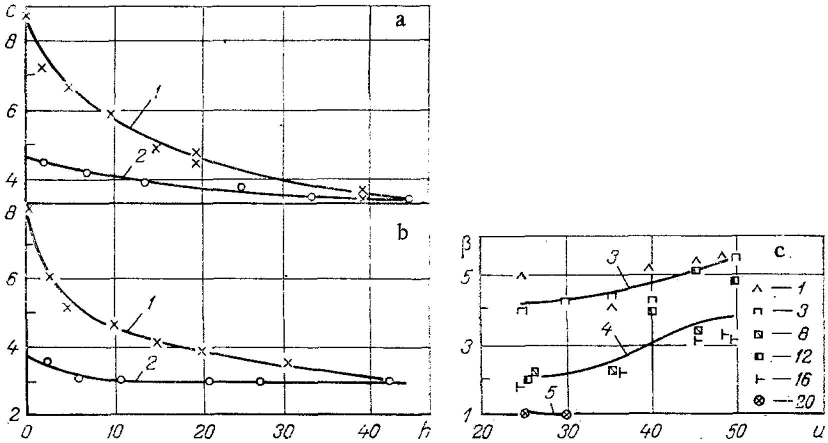

Fig. 1. Experimental characteristics of interphase transfer in an FB of particles of keramzit with a packing, $D_k = 15$ cm. a) Particles of moisture concentration over the height of an FB packing No. 2 from Table 1: d = 0.25 mm; $H_0 = 45$ cm; u = 45 cm/sec; 1) in gas phase; 2) in solid phase. b) The same, packing No. 4; d = 0.37 mm; $H_0 = 53$ cm; u = 30 cm/sec. c) Dependence of β on gas filtration velocity. d) 0.25 mm; 3) small-volume packings; 4) medium-volume packings; 5) spheres. See Table 3 for notation of points.

The groups (aggregates) of particles move continuously in the turbulent gas flow, and energy dissipation occurs in it on structures having the Kolmogorov scale δ_k. This scale is determined by the rate of energy dissipation ε_d and viscous properties of the flow ν [10]:

$$\delta_k^4 \sim \nu^3/\varepsilon_d. \tag{4}$$

Both processes (energy dissipation and diffusion of the component from the gas) occur at the surface of the aggregates in boundary layers. The first process occurs in the dynamic boundary layer, while the second occurs in the diffusion boundary layer. The relationship between the thickness of the diffusion (δ_d) and dynamic (δ_k) boundary layers on the surface of the particle aggregates is described by the expression [11]

$$\delta_d = \frac{\delta_k}{\mathrm{Sc}^{1/3}}. \tag{5}$$

In an FB with a packing, it is expedient to use the hydraulic radius of the channels formed by the packing as the measure of the dynamic boundary layer. The channel radius determines the size of the particle aggregates. Assuming that $\delta_k \sim l_p$, from (1)-(5) we arrive at the following relation after a transformation:

$$\beta \sim \frac{\varepsilon_p \varepsilon_d^{\frac{1}{3}} \delta_k^{\frac{1}{3}}}{l_p \mathrm{Sc}^{2/3}} \sim \frac{\varepsilon_p \varepsilon_d^{\frac{1}{3}}}{l_p^{2/3} \mathrm{Sc}^{2/3}}. \tag{6}$$

The rate of energy dissipation in the FB is equal to

$$\varepsilon_d = \frac{1}{\rho} \frac{dP}{dh} u = \frac{\rho_{\tau} g (1-\varepsilon) u}{\rho},$$

where dP/dh is the pressure drop per unit of height of the FB. Inserting this value ε_d into (6) and changing over to dimensionless coordinates, we arrive at the following relation to determine interphase mass transfer in an FB with a packing:

$$\frac{\mathrm{St}}{\varepsilon_p} \sim \mathrm{Fr}^{-\frac{1}{3}} \mathrm{Sc}^{-\frac{2}{3}}. \tag{7}$$

We studied interphase mass transfer in an FB with a packing experimentally. We used the adsorption method in [12] to measure β. Moisture was adsorbed from air in an FB of small-volume adsorbents (keramzit — a lightweight concrete aggregate — and fireclay), when the transfer of water vapor from the gas phase to the solid phase is governed by external mass transfer.

TABLE 1. Characteristics of the Packings

Number	Type	D_p, cm	H_p, cm	ε_p	l_p, cm
	Small-volume packings:				
1	wire cylinders	1,2	1,5	0,92	0,3*
2	same.	2,7	2,4	0,95	0,5 †
	Medium-volume packings:				
3	spheres, $D_s = 1.3$ cm	—	—	0,47	0,15
4	rings with slits [7]	2,2	2,3	0,68	0,29
5	Rasching rings	2,5	2,5	0,67	0,38
6	ceramic packing [8]	3,0	3,0	0,70	0,50

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*d_{wire} = 1.3 mm, number of turns 6.

†d_{wire} = 1.3 mm, number of turns 10.

TABLE 2. Characteristics of Fluidized Materials

Number	Material	d, mm	$\rho_t \cdot 10^{-3}$, kg/m^3	Source
1	Keramzit	0,13	1,5	
2	«	0,19	1,5	
3	«	0,25	1,5	
4	«	0,37	1,5	
5	Fireclay	0,19	2,1	
6	Catalyst	0,95	1,3	[2]
7	«	0,50	0,9	[2]
8	«	0,15	1,0	[2]
9	«	0,13	3,5	[2]
10	Silica gel	0,21	1,1	[3]

A portion of dry adsorbent was poured into a column with a packing. We provided a supply of moist air and took air samples of about 100 cm^3 through tubes over about 20 sec simultaneously from several points along the column axis. The sampling time was fairly short, so that the moisture concentration in the layer during the sampling period remained nearly constant.

We then cut off the supply of fluidizing air, and the gas-sampling tubes sucked samples of air from several points along the column axis (about 40 cm^3 from each point). The moisture concentration in the air removed from the intergranular channels corresponded to the mean equilibrium moisture content at the surface of the particles located on a radius of about 2 cm from the sampling points.

The air samples were analyzed with coulometric hygrometers [13]. The measurement data was used to construct profiles of moisture concentration in the gas phase (air) and equilibrium concentrations at the surface of the fluidized particles. In constructing the second of these profiles, we consider the expansion of the FB.

Figure 1a, b shows experimental profiles for two types of packing (small- and medium-volume). The top curve reflects the change in concentration in the gas phase, while the bottom curve shows the same in the solid phase. The area bounded by these curves was used to determine the effective interphase mass-transfer coefficient, referred to a unit volume of the FB with packing, with allowance for expansion of the bed

$$\beta = \frac{u(c_{\text{in}} - c_{\text{out}})}{\int\limits_0^H (c - c_p)\,dh}. \tag{8}$$

The accuracy of the measurement of β is determined by the error of the calculation of the integral in (8). The measurement error of the integral is 20% at a moisture-content measurement error of 3%. The coefficient of variation for the mass-transfer coefficient obtained under identical test conditions was no greater than 25%, which verifies the estimate made here.

We studied interphase mass transfer in an FB with packings in the form of wire cylinders of two sizes, ceramic spheres, Raschig rings, rings with slits [7], etc. Tables 1 and 2 show characteristics of the packings and the materials fluidized. The tests were performed in columns 15 and 30 cm in diameter. Air velocity was changed from the velocity corresponding to the beginning of fluidization to 60 cm/sec. The moisture content of the air at the bed

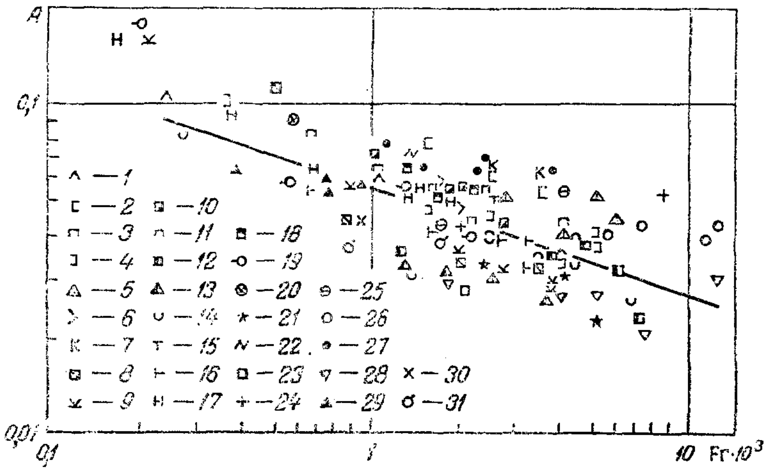

Fig. 2. Dependence of the complex $A = St \cdot Sc^{0.73}/\varepsilon_p$ on the Froude number. See Table 3 for the designation of the points.

TABLE 3. Conditions of Mass-Transfer Tests

Designation in Fig. 2	Number of packing in Table 1	Number of material in Table 2	H_0, cm	Designation in Fig. 2	Number of packing in Table 1	Number of material in Table 2	H_0, cm	Remarks
1	1	3	47	16	6	3	60	
2	2	2	40	17	6	3	65	D_C = 30 cm*
3	2	3	50	18	6	5	60	
4	2	3	80	19	3	3	80	
5	2	4	40	20	3	3	50	
6	2	5	40	21	2	4	46	
7	5	1	30	22	2	3	48	D_C = 30 cm†
8	5	3	47	23	2	4	80	
9	5	3	80	24	—	6†	81	[2]‡
10	5	5	37	25	—	6†	81	[2]‡
11	4	2	61	26	—	6†	89	[2]‡
12	4	3	55	27	—	6†	96	[2]‡
13	4	3	80	28	—	7†	78	[2]‡
14	4	4	53	29	—	9†	52	[2]‡
15	6	2	60	30	—	8†	75	[2]‡
				31	—	10†	100	[3]‡

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*In the remaining cases, D_C = 15 cm.

†The packing consisted of wire spirals.

‡The mass-transfer coefficients were recalculated per unit volume of the fluidized system using a formula for calculating expansion from [14].

inlet was 6-10 g/m³. We conducted about 200 tests. The test conditions are described in greater detail in Table 3, which also shows data from other studies.

Figure 1c shows the dependence of β on filtration velocity according to the test data for small- and medium-volume packings. Interphase transfer is greater in the case of small-volume packings, which is evidently due to the higher concentration of particles in the system.

We used Eq. (7) to generalize our test data and the data from other studies [2, 3] and introduced additional complexes reflecting the effect of the scale of the unit (D_C/H_0, H_0/l_p) and the size of the particles (Ar):

$$\mathrm{St} = a_0 \,\mathrm{Fr}^{a_1} \,\mathrm{Sc}^{a_2} \,\varepsilon_p^{a_3} \left(\frac{D_C}{H_0}\right)^{a_4} \mathrm{Ar}^{a_5} \left(\frac{H_0}{l_p}\right)^{a_6}. \tag{9}$$

A multidimensional regression-analysis program [15] was used on a computer to calculate values of the coefficient a_0 and exponents a_i (i = 4-6) and their standard deviations S_i. We also determined confidence intervals for these values with a confidence level α = 0.05. The results, shown in Table 4, indicate that for exponents a_i (i = 4-6) the inequality $(a_i - 0) <$

TABLE 4. Values of the Coefficient a_0 and Exponents a_i (i = 1-6) in Eq. (9) and Their Confidence Intervals $t_{\alpha/2,n-7}S_i$

i	a_i	$t_{\alpha/2,n-7}S_i$	i	a_i	$t_{\alpha/2,n-7}S_i$
0	0,083	—	4	0,119	0,311
1	—0,337	0,068	5	0,060	0,061
2	—0,753	0,234	6	—0,277	0,384
3	0,960	0,436			

$t_{\alpha/2,n-7}S_i$ ($t_{\alpha/2,n-7}$ = 1.984 with a number of test points n = 110), i.e., these exponents are reliably trivial. Thus, the corresponding factors in (9) should be omitted. The exponent a_3 is reliably equal to unity, since $(a_3 - 1) < t_{\alpha/2,n-7}S_3$. Finally, the relation determining interphase mass transfer in an FB with small- and medium-volume packings has the following form:

$$\frac{St}{\varepsilon_p} = 0{,}0058\,\mathrm{Fr}^{-0.34}\,\mathrm{Sc}^{-0.73}. \tag{10}$$

This relation is shown in Fig. 2 together with all of the experimental points. The mean relative deviation of the test points from the approximation is 25%. Equation (10) is valid for the following ranges of the variables: $0{,}166\cdot10^{-3} \leq \mathrm{Fr} \leq 14.4\cdot10^{-3}$, $25 \leq \mathrm{Ar} \leq 3.3\cdot10^4$, $0.65 \leq \mathrm{Sc} \leq 1.44$, $0.47 \leq \varepsilon_p \leq 0.96$, $1.5 \leq H_0/D_c \leq 6$, $40 \leq H_0/l_p \leq 520$.

Equation (10) generalizes test data from different studies conducted by different methods. It is similar to Eq. (7), which was obtained on the assumption that mass transfer and the rate of energy dissipation in a fluidized bed are related to each other. It is interesting to note that interphase mass transfer in an FB with a packing is independent of the size of the fluidized particles. This indicates that mass transfer in an FB occurs not on the surface of individual particles but on the surface of particle aggregates, the size of which is determined by the hydraulic radius of the packing.

Commercial units with a fluidized bed operate at high filtration velocities when an "aggregate" regime exists [9]. Semiempirical correlation (10) can be recommended for calculating interphase mass transfer in such units.

NOTATION

c, c_{in}, c_{out}, c_p, concentration of moisture in the air, respectively, in the fluidized bed, inlet to the bed, and outlet, and the equilibrium concentration at the surface of the solid particles; d, mean diameter of fluidized particles; d_{wire}, diameter of wire; D, coefficient of diffusion of component in the gas; D_s, D_p, D_c, diameters of sphere, element of packing, and column; g, acceleration due to gravity; h, H_0, H, height of fluidized bed corresponding to running, initial, and working values, respectively; H_p, height of element of packing; $l_p = V_p/S_p$, hydraulic radius of packing; P, pressure inside FB; S_p, S_s, surface area of packing and viscous boundary layer at the surface of the aggregates; $t_{\alpha/2,n-7}$, tabulated value of Student's criterion with a specified confidence level $1 - \alpha$ and a specified number of experimental points n; u, gas filtration velocity; V, V_p, V_s, volumes of the fluidized system, packing, and diffusion boundary layer at the surface of the aggregates, respectively; β^*, β, coefficients of mass transfer between the gas and particles, referred to a unit volume of the boundary layer and the entire system, respectively; δ_d, δ_k, thicknesses of the diffusion and dynamic boundary layers on the surface of the particle aggregates, respectively; ε, ε_p, porosity of the fluidized bed and packing; ε_d, unit rate of energy dissipation; ν, viscosity of the fluidizing gas; ρ, density of the fluidizing gas; ρ_t, apparent density of the fluidized particles; $\mathrm{Ar} = d^3g(\rho_t - \rho)/\nu^2\rho$, Archimedes number; $\mathrm{Fr} = u^2\rho/gl_p\rho_t$, modified Froude number; $\mathrm{Sc} = \nu/D$, Schmidt number; $\mathrm{St} = \beta l_p/u$, Stanton number.

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NONLINEAR FILTRATION IN CRACKED POROUS MATERIALS

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Stationary filtration to a well and to a gallery in a cracked porous medium is investigated with the strong dependence of structural–mechanical properties of the medium on the pressure of the filtering fluid being taken into account.

Motion in cracked-porous materials is usually described on the basis of a continual model within whose framework the material is considered as two coexistent fictitious porous media; filtration therein corresponds to independent progress over the porous modules and over the system of cracks in the presence of mutual fluid transfer [1]. The equations for the unknown mean pressures in the cracks and in the modules are obtained in [2]; they are analogous in structure to the heat-conduction equations in a heterogeneous medium (see [3-5], for example).

However, in contrast to the majority of heat-conduction processes, a strong nonlinear dependence of the cracked porosity and permeability (analogous to the specific heat and the heat conduction in the corresponding thermal problem) on the pressure within the cracks is characteristic for filtration processes in cracked and cracked porous media. This pressure dependence results in the appearance of a number of qualitatively new effects (finiteness of the

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