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EXTERNAL HEAT TRANSFER IN INFILTRATED GRANULAR BEDS. EXPERIMENTAL STUDY

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Experimental data was obtained on external heat transfer in infiltrated granular beds of coarse particles. It was determined that calcualtions with the proposed model agree satisfactorily with the experimental results.

A two-phase model was formulated in [1] for heat transfer in stationary granular beds injected with gas. The model was ued to obtain simple expressions for the coefficient of heat transfer in the case of two different surfaces:

a) plane surface

$$\alpha = \frac{\lambda_f}{R} \frac{\lambda s + f^2 \xi_0}{1 + \lambda s \xi_0}, \tag{1}$$

b) cylindrical surface

$$\alpha = \frac{\lambda_f}{R} \frac{\lambda s + f^2 \xi_0 / K^*}{1 + \lambda s \xi_0 K^*} K^*. \tag{2}$$

Equations (1) and (2) are suitable for calculations of α at Pe > 500 and 100, respectively. To check the validity of the above relations, we used experimental data obtained from the literature and our own tests.

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TABLE 1. Characteristics of Dispersed Materials Used in the Tests

Dispersed material	d, mm	u_0, m/sec	ε_0	ρ_s, kg/m³	Notation in text
Glass spheres	1,75	0,85	0,4	2650	s. 1
Fireclay	3,0	0,95	0,48	2300	s. 2
Peas	5,7	1,35	0,42	1400	s. 3

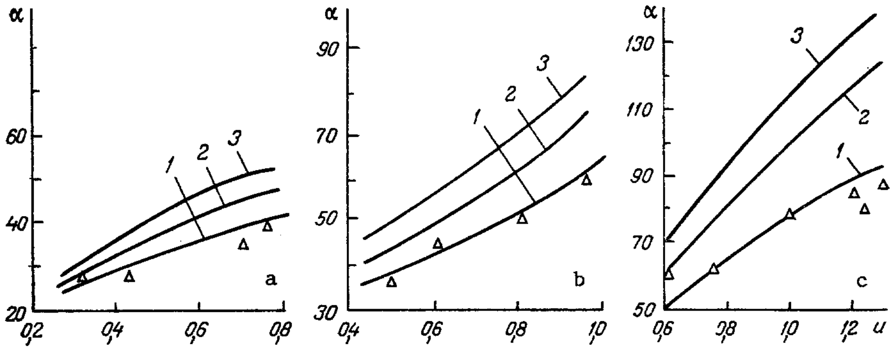

Fig. 1. Dependence of the heat-transfer coefficient of the infiltrated granular bed on air filtration velocity: a) Pe = 310-500, s. 1; b) Pe = 290-360, s. 2; c) Pe = 210-240, s. 3; 1) calculation from (2); 2) from (5); 3) from (4); points show experimental data from the present study. α, W/m²·K; u, m/sec.

The tests were conducted in a cylindrical column 300 mm in diameter. The column was equipped with a perforated gas distributor. A copper sensor with an outside diameter of 30 mm and a length of 150 mm was located vertically on the axis of the column 200 mm above the gas-distributing grate. The height of the bed of dispersed material ranged from 510 to 720 mm. The characteristics of the dispersed materials used in the tests are shown in Table 1. The tests were conducted at air filtration velocities $u < u_0$. Temperature was measured with chromel–copel thermocouples with wires 0.2 mm in diameter. The coefficient of heat transfer between the dense in filtrated bed of solid particles and the surface of the sensor was determined by a standard method [2] with the aid of the formula

$$\alpha = \frac{W}{S\,[T_w - T_s(R)]}\,. \tag{3}$$

The error of the determination of α from (3) was no more than 5%. The resulting test data is shown in Fig. 1. Also shown are values of α calculated from Eq. (2). As noted above, this equation can be used at Pe > 100. Equations (5) in [1] were used to determine the thermal conductivity of the gas film [3] and infiltrated bed and the thickness of the film in (2).

It can be seen that Eq. (2) satisfactorily describes the results we obtained. Figure 1 also shows values of α calculated from Gabor's formula [4]:

$$\alpha = \sqrt{\frac{4c_f\rho_f u\lambda_s}{\pi H}} + \frac{\lambda_s}{2a}\,, \tag{4}$$

which is valid for the case $Pe_a = c_f\rho_f ua^2/\lambda_s H \geq 1$.

Equation (4) gives results which are 20-30% higher than the actual values. This can evidently be attributed to the fact that it was obtained on the basis of a model of heat transfer which does not take into account the presence of a gas film at the heat-transfer surface and which presumes that porosity is constant over the entire volume of the bed [4].

Figure 2 compares values of α calculated from Eq. (2) and experimental data [5] obtained with different pressures of infiltrating gas. An appreciable difference is seen between these

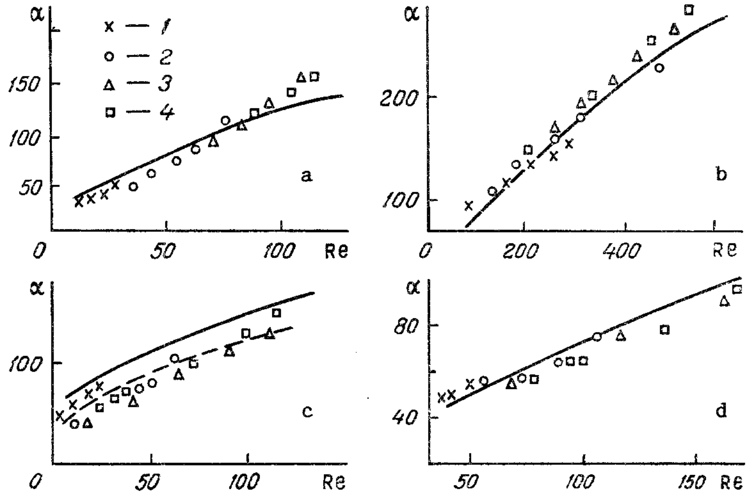

Fig. 2. Comparison of test data [5] and data calculated from the two-zone model: a) sand (d = 1.02 mm) – air, Pe = 52-100; b) sand (d = 2.37 mm) – air, Pe = 65-90; c) copper (d = 0.62 mm) – air, Pe = 60-100; d) sand (d = 1.02 mm) – CO_2, Pe = 64-130; dashed curve in Fig. 2c – calculation from (2), where H is the height of the fixed bed (380 mm); (1 – p = 0.1 MN/m^2; 2 – 0.38; 3 – 0.65; 4 – 0.93).

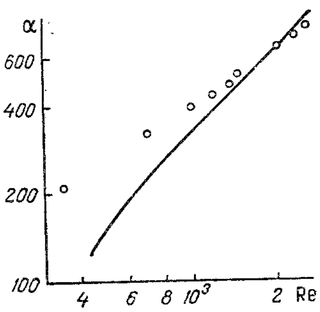

Fig. 3. Comparison of experimental data [6] with data calculated from Eq. (2). Glass spheres d = 3.1 mm; Pe = 75-90. Compacted fixed bed.

results only for the beds of copper spheres (Fig. 2c). The agreement between the empirical and theoretical data is improved considerably if we replace the height of the sensor in (2) by the height of the entire granular bed.

Figure 3 compares calculated and experimental [6] data on the rate of heat transfer in a packed bed of glass spheres. It can be seen from the figure that the values of α calculated from (2) agree fairly well (error no larger than 13%) with measurements made at Re $\geq$ 1200. The deviations increase with a decrease in Re and reach 45%. This can be attributed to certain features of the experiment. In particular, the thermocouple which measured bed temperature was located a short distance from the sensor. This naturally led to exaggerated values of α. The degree of exaggeration was particularly great at low filtration velocities (low Re).

It is interesting to analyze the effect of the gas interlayer on the value of the coefficient α. At $\ell_0 = 0$, we write Eq. (2) in the form

$$\alpha = \frac{\lambda_g s}{R} K^*. \qquad (5)$$

Figure 1 shows values of α calculated from (5). They agree best with the estimates obtained from Gabor's formula (4), which also does not take into account the additional thermal resistance created by the high-porosity zone (gas film) at the heat-transfer surface. As the data in [4] and the data obtained here show, the thermal resistance of the gas interlayer can be ignored for fine ($d < 0.8$ mm) particles and Eqs. (4) and (5) can be used. This obviously has to do with the fact that a reduction in particle size is accompanied by a reduction in the effective thickness of the gas film and, thus, the contribution of the latter to the total thermal resistance.

We compared the available experimental data on external heat transfer in infiltrated fixed beds with results calculated from theoretical relation (2). It was found that this formula describes the experimental data with an accuracy sufficient for practical purposes. Here, the model parameters in the equation are determined by Eqs. (5) [1].

It should be emphasized that the introduction of gas layer adjacent to the heat-transfer surface into the model and allowance for the effect of the thermal conductivity of the granular bed (the coefficient λ_s) on heat transfer makes the model sufficiently universal and also allows description of the heat-transfer coefficients in fluidized beds. Heat transfer in a fluidized bed was described in [3] by means of a theoretical scheme which is essentially a unidimensional variant of the model proposed in the present study. It was shown that at large values of the ratio $\lambda = \lambda_s/\lambda_f$ ($\lambda \sim 10^4$-10^5) — typical of developed fluidized beds due to intensive mixing of solid particles (high values of λ_s) — the value of α generally ceases to depend on λ_s and is determined only by the quantities ℓ_0 and λ_f:

$$\alpha = \lambda_f/l_0, \tag{6}$$

where $\ell_0 = 0.14d(1 - \varepsilon)^{-2/3}$, while λ_f is calculated from (5) [1].

Comparison in [6] of experimental values of α and values of α calculated from Eq. (6) showed that exceedingly simple relation (6) well describes the heat-transfer coefficient in fluidized beds within a broad range of experimental conditions.

Thus, the two-zone model developed in the present study to describe external heat transfer in infiltrated disperse systems correctly reflects the basic features of the process and makes it possible to describe the laws governing heat transfer within a broad range of experimental conditions in both fixed and fluidized beds. Simple formulas (1), (2), and (6), obtained on the basis of this model, make its use convenient for practical calculations.

NOTATION

a, radius of the heat-transfer surface; c, specific heat; d, particle diameter; $\bar{f} = \sqrt{\mathrm{Pe}\lambda/\varepsilon_0}$; H, length of the sensor; K_0, K_1, modified 0-th and 1-st order Bessel functions of the second kind (MacDonald functions); $K^* = K_1(s(\xi_a + \xi_0))/K_0(s(\xi_a + \xi_0))$; R, width (or radius) of the bed of solid particles; S, surface of heat-transfer sensor; $s = \sqrt{\mathrm{Pe}}$; T, temperature; u, filtration velocity; u_0, initial fluidization velocity; ℓ_0, thickness of the gas film; W, power of the sensor heater; α, heat-transfer coefficient; ε, ε_0, porosity at u and u_0; $\xi_0 = \ell_0/R$; $\xi_a = a/R$; ρ, density; $\lambda = \lambda_s/\lambda_f$; ν_f, viscosity of the gas; $\mathrm{Pe} = c_f\rho_f uR^2/\lambda_s H$, Peclet number; $\mathrm{Re} = ud/\nu_f$, Reynolds number. Indices: f, gas; s particles.

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