MODELING OF UNSTEADY HEAT TRANSFER IN DISPERSED SYSTEMS

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Using the two-zone model we consider unsteady heat transfer in filtrated dispersed systems. It is shown that the boundary zone must be taken into account.

Unsteady heat transfer in dispersed systems is important in many applied problems, such as the calculation of thermal regimes in various technological processes (drying and baking of dispersed materials, heat treatment of elements in beds of a dispersed heat carrier, the operation of chemical reactors and other devices with dispersed beds in transient and unsteady regimes, and so on). On the purely scientific level the study of unsteady heat transfer is evidently the best method of determining the mechanism of heat transport in dispersed systems, as well as constructing adequate models.

A rather detailed review of the phenomenological models of heat transfer has been given recently in [1]. Analysis of these models shows that there are two basic approaches to the study of the mechanism of heat transfer, which use different features of heat transfer in dispersed systems: a) the heterogeneity of the system over the entire volume of the sample is taken into account by introducing a temperature difference between the phases and allowing heat transfer between them [1-5]; b) the granular bed is treated as a quasi-homogeneous medium in which the temperatures of the different phases are equal and the dispersed structure of the bed is taken into account by introducing a zone of higher porosity near the heat-transfer surface [6-12].

There has not appeared in the literature up to the present time a well-founded discussion of which of these assumptions is necessary to describe heat transfer in a wide interval of experimental conditions. The purpose of the present paper is to analyze the possibilities of mathematical modeling of heat transfer from the point of view of the one-temperature, two-zone models (those belonging to group b) to determine the region of applicability of this group of models, and to compare the results with those of the two-temperature one-zone models (group a)).

The two-zone model of external heat transfer in filtrated dispersed systems was developed in [10, 12, 13] and shown to be applicable to steady-state heat transfer in fixed and fluidized beds. The basic feature of this model is that the zone of higher porosity near the heat-transfer surface is modeled as a gas layer whose effective thickness is determined from experiment. This model is used here to study unsteady heat transfer.

<u>Fixed Aerated Bed</u>. We consider the heating (cooling) of a vertical cylinder in a granular bed. As shown in [12], for large values of the Peclet number ($c_f\rho_f uR^2/L\lambda_s^h$) the equations of the heat transfer model can be written in the following dimensionless form:

$$\frac{\partial\theta_f}{\partial\,\mathrm{Fo}}=\frac{\partial^2\theta_f}{\partial\xi^2}+\frac{1}{\xi}\frac{\partial\theta_f}{\partial\xi}-\mathrm{Pe}\frac{\lambda^0}{\varepsilon}\theta_f \tag{1}$$

$$(\xi^0\leqslant\xi<\xi^0+1);$$

$$\frac{\partial\theta_s}{\partial\,\mathrm{Fo}}=a^0\left(\frac{\partial^2\theta_s}{\partial\xi^2}+\frac{1}{\xi}\frac{\partial\theta_s}{\partial\xi}-\mathrm{Pe}\,\theta_s\right)$$

$$(\xi^0+1<\xi\leqslant B) \tag{2}$$

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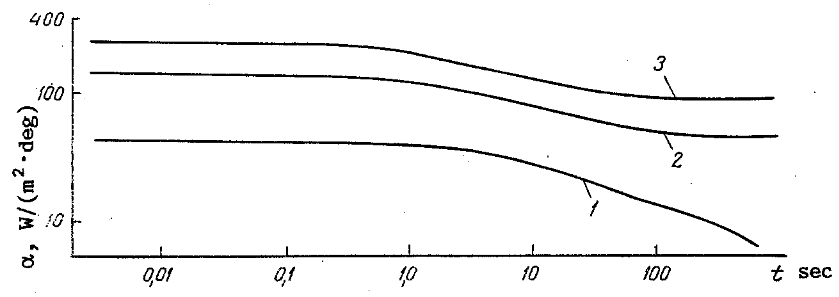

Fig. 1. Dependence of the heat-transfer coefficient of a fixed aerated bed on time: 1) u = 0; 2) 0.6; 3) 1.3 m/sec (peas).

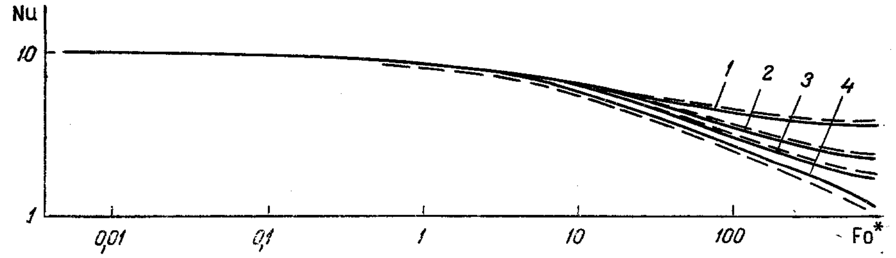

Fig. 2. Intensity of unsteady heat transfer in a fixed aerated bed: 1) $\lambda^0 = 6.32$; $a^0 = 0.95 \cdot 10^{-2}$; $Pe = 0.13 \cdot 10^{-2}$ (peas, u = 1.3 m/sec); 2) $\lambda^0 = 6.27$; $a^0 = 0.79 \cdot 10^{-2}$; $Pe = 0.13 \cdot 10^{-2}$ (chamotte, u = 0.9 m/sec); 3) $\lambda^0 = 5.20$; $a^0 = 0.54 \cdot 10^{-2}$; $Pe = 0.68 \cdot 10^{-3}$ (glass balls, u = 0.7 m/sec); 4) $\lambda^0 = 3.30$; $a^0 = 0.51 \cdot 10^{-2}$, Pe = 0 (peas, u = 0); solid curves: numerical solution of (1)-(4); dashed curves: calculation using (6).

with the conditions

$$\theta_f(0, \xi) = \theta_s(0, \xi) = 0;\ \theta_f(0, \xi^0) = 1;\ \theta_f(\mathrm{Fo}, \xi^0 + 1) = \theta_s(\mathrm{Fo}, \xi^0 + 1);$$
$$\frac{\partial \theta_f(\mathrm{Fo}, \xi^0 + 1)}{\partial \xi} = \lambda^0 \frac{\partial \theta_s(\mathrm{Fo}, \xi^0 + 1)}{\partial \xi};\ \frac{\partial \theta_s(\mathrm{Fo}, B)}{\partial \xi} = 0;\ \frac{\partial \theta_f(\mathrm{Fo}, \xi^0)}{\partial \mathrm{Fo}} = \Lambda \frac{\partial \theta_f(\mathrm{Fo}, \xi^0)}{\partial \xi}. \tag{3}$$

The last boundary condition in (3) corresponds to zero-gradient heating (cooling) of the cylinder.

The system of equations (1)-(3) was solved numerically, using the conservative, absolutely stable difference scheme of [14] with implementation by trial and error. Because the thermal properties of the gas film and dispersed bed are different, it was necessary to impose the following condition in order to obtain the required accuracy on the film-bed boundary:

$$\frac{a_f^h \Delta t}{(\Delta r_1)^2} = \frac{a_s^h \Delta t}{(\Delta r_2)^2},$$

where Δt, Δr_1, Δr_2 are the (dimensional) stepsizes in time and coordinate r in the film and bed, respectively. From this condition we obtain a relation between the dimensionless stepsizes: $h_1 = \sqrt{a^0} h / (B - \xi^0 - 1)$, h is the stepsize of a uniform network in the film, $\bar{\omega}_h = \{\xi_i = \xi^0 + ih,\ \xi_i \in [\xi^0, \xi^0 + 1], i = 0, 1, 2, \ldots, N;\ N = 1/h\}$, is the stepsize between the first and second nodes of a nonuniform network in the dispersed bed, $\hat{\bar{\omega}}_h = \{\xi_j \in [\xi^0 + 1, B], j = 0, 1, 2, \ldots K;\ h_j = 1.5 h_{j-1}\}$.

Preliminary calculations on different networks showed that the values h = 0.1 and $\Delta\mathrm{Fo} = 10^m \cdot 0.001(\mathrm{Fo} \in (10^{m-1}, 10^m)$, m = 1, 2, ..., for $\mathrm{Fo} \in (0.1) \Delta\mathrm{Fo} = 0.001)$ are optimal.

The heat-transfer coefficients were found using the equations

$$\alpha_1 = -\frac{\lambda_f^h}{l_0} \frac{\partial \theta_f(\mathrm{Fo}, \xi^0)}{\partial \xi [\theta_f(\mathrm{Fo}, \xi^0) - \theta_s(\mathrm{Fo}, B)]}, \tag{4}$$

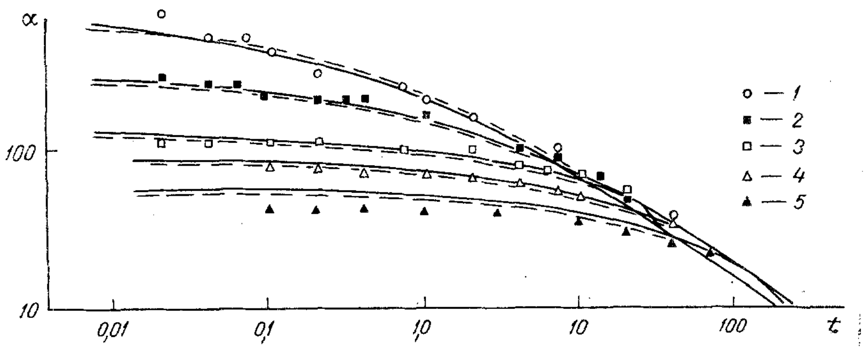

Fig. 3. Comparison of the experimental data of [17] with the values calculated in the two-zone model (glass balls—air): 1) d = 0.39 mm; 2) 0.93; 3) 2.07; 4) 3.05; 5) 5.05; solid curves: numerical solution of (1)-(3); dashed curves: calculation using (9).

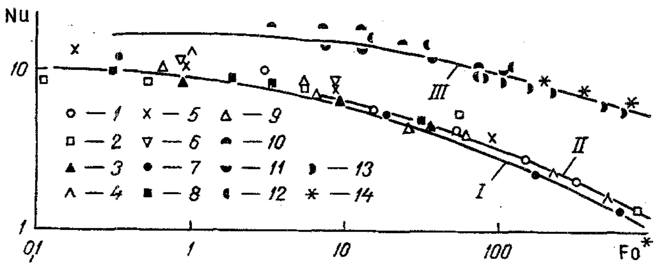

Fig. 4. Comparison of the experimental and calculated data for unsteady heat transfer in fixed beds: 1), 2), 3) d = 0.39 mm; 2.07; 5.05 3.05 (glass balls—CO_2 [17]); 7), 8), 9), d = 0.39 mm; 0.93; 2.07 (glass balls—freon-12 [17]); 10), 11), 12), 13), 14) d = 0.74 mm; 0.65; 0.44; 0.21; 0.17 (copper balls—air [3]); I: numerical solution of (1)-(3) for freon-12; II; for air and CO_2; III: calculation using (9) for Nu $= \frac{1}{\mathrm{Fo}^*}\int_0^{\mathrm{Fo}^*} \mathrm{Nu}\, d\mathrm{Fo}^*$

$$\alpha_2 = -\frac{\lambda_f^h}{l_0}\,\frac{\partial\theta_f(\mathrm{Fo},\ \xi^0)}{\partial\xi\theta_f(\mathrm{Fo},\ \xi^0)}\,.$$

The coefficient λ_s^h was calculated using the equations for a fixed aerated bed [15], λ_f^h was calculated from the formula $\lambda_j^c + 0.0061 c_f \rho_f \frac{u}{\varepsilon} d$, given in [10]. Calculations were carried through for the following materials: glass balls ($d = 1.75$ mm; $u_0 = 0.85$ m/sec; $\varepsilon_0 = 0.40$); chamotte ($d = 3.0$ mm; $u_0 = 0.95$ m/sec; $\varepsilon_0 = 0.48$); peas ($d = 5.7$ mm; $u_0 = 1.35$ m/sec; $\varepsilon_0 = 0.42$). Because of the smallness of the quantity $\theta_s(\mathrm{Fo}, \mathrm{B})$, the quantities α_1 and α_2 were equal in all cases. The thickness ℓ_0 of the gas film was taken to be 0.1d [3]. Figure 1 shows some typical results of calculations based on (1)-(3). It was established that at a certain time the heat-transfer coefficient reaches its minimum possible steady-state value, which coincides with the value calculated from the equation

$$\alpha_{ss} = \frac{\lambda_f^h}{R}\,\frac{\lambda^0\sqrt{\mathrm{Pe}^*} + (\lambda^0\,\mathrm{Pe}^*/\varepsilon)\,\xi_0/K^*}{1 + \lambda^0\xi_0\sqrt{\mathrm{Pe}^*}\,K^*}\,K^*, \tag{5}$$

obtained in [12] for steady-state heat transfer.

A more graphic picture of the features of heat transfer in fixed aerated beds is given in Fig. 2, where the calculated values of α are constructed in dimensionless form. It is easy to see that as a^0 increases the steady-state value of Nu is reached more rapidly and the ratio Nu/Nu_{lim} increases and approaches unity (Nu_{lim} is the value of Nu in the limit $Fo \to 0$). The following simple interpolation formula provides a representation of the results which is convenient for analysis:

$$Nu = Nu_{ss}(1 - e^{-2a^0\sqrt{\pi Fo^*}}) + \frac{\lambda^0/\sqrt{a^0}}{k\lambda^0/\sqrt{a^0} + \sqrt{\pi Fo^*}}, \quad (6)$$

where k is the coefficient in the expression $\ell_0 = kd$ determining the thickness of the gas film; $Nu_{ss} = \alpha_{ss}d/\lambda_f^h$, and α_{ss} is calculated using (5). In the absence of aeration (u = 0) we have $Nu_{ss} = 0$.

To test the correctness of the numerical calculation, and using the fact that when the heat transfer is highly unsteady the intensity does not depend on the shape of the heat-transfer surface [15],* we obtained the analytical solution for the heating (cooling) of a plate in a fixed bed without aeration (u = 0):

$$\frac{\partial \theta_f}{\partial Fo} = \frac{\partial^2 \theta_f}{\partial \xi^2}, \quad 0 \leq \xi < 1; \quad (7)$$

$$\frac{\partial \theta_s}{\partial Fo} = a^0 \frac{\partial^2 \theta_s}{\partial \xi^2}, \quad 1 < \xi \leq B, \quad (8)$$

subject to conditions analogous to (3). The expression for the coefficient α has the form:

$$\alpha = -\frac{\lambda_f^h}{l_0} \frac{\sum_{n=1}^{\infty} \frac{\gamma_n}{\Delta_n} e^{-\mu_n^2 Fo}}{\sum_{n=1}^{\infty} \frac{\sigma_n}{\Delta_n} e^{-\mu_n^2 Fo}},$$

$$\gamma_n = \frac{\lambda^0 \mu_n}{\sqrt{a^0}} \sin\frac{\mu_n(B-1)}{\sqrt{a^0}} \cos\mu_n + \mu_n \cos\frac{\mu_n(B-1)}{\sqrt{a^0}} \sin\mu_n,$$

$$\sigma_n = -\frac{\lambda^0}{\sqrt{a^0}} \sin\frac{\mu_n(B-1)}{\sqrt{a^0}} \sin\mu_n + \cos\frac{\mu_n(B-1)}{\sqrt{a^0}} \cos\mu_n - 1,$$

$$\begin{aligned}\Delta_n = &-\mu_n\left(\frac{\lambda^0(B-1)}{a^0} + 1\right)\sin\mu_n \cos\frac{\mu_n(B-1)}{\sqrt{a^0}} + \\ &+\left(\frac{\Lambda\lambda^0(B-1)}{a^0} + 1 + \Lambda\right)\cos\mu_n \cos\frac{\mu_n(B-1)}{\sqrt{a^0}} - \\ &-\left(\frac{(1+\Lambda)\lambda^0}{\sqrt{a^0}} + \frac{\Lambda(B-1)}{\sqrt{a^0}}\right)\sin\mu_n \sin\frac{\mu_n(B-1)}{\sqrt{a^0}} - \\ &-\mu_n\left(\frac{\lambda^0}{\sqrt{a^0}} + \frac{B-1}{\sqrt{a^0}}\right)\cos\mu_n \sin\frac{\mu_n(B-1)}{\sqrt{a^0}};\end{aligned} \quad (9)$$

μ_n are the roots of the characteristic equation

$$\frac{\lambda^0}{\sqrt{a^0}} \sin\frac{\mu_n(B-1)}{\sqrt{a^0}} (\Lambda\cos\mu_n - \mu_n \sin\mu_n) + \cos\frac{\mu_n(B-1)}{\sqrt{a^0}} (\mu_n \cos\mu_n + \Lambda\sin\mu_n) = 0.$$

*In the framework of the two-zone model this experimental fact is easily explained. Indeed, for small times of contact of the particles with the surface, the particles are not able to change the surface temperature significantly. Hence practically the entire thermal resistance is concentrated in the gas film and it does not depend on the shape of the heat-transfer surface because of the small thickness of the film.

An example of the comparison between the experimental data [17], the numerically calculated values using (1)-(3), and the values obtained using the exact solution (9) is given in Fig. 3. We see that in the highly unsteady region the numerical calculation agrees closely with the results calculated using (9) and also with the experimental data.

The data of [17] and also the experimental results of [3] on heat transfer of a moving bed of copper balls with a plane surface are shown in Fig. 4 in dimensionless coordinates. We see that the experimental points of [17] at small Fo^* are grouped about the straight line $Nu \simeq 10$, while the data of [3] are grouped about the line $Nu \simeq 16$. Hence the effective thickness of the gas film can be obtained for the two experiments. At small Fo^* the heat-transfer coefficient is completely determined by the contact thermal resistance (the resistance of the gas film $R_f = l_0/\lambda_f^h$): $Nu_{lim} \simeq d/R_f\lambda_f^h = d/\ell_0 = d/kd = 1/k$. Hence $k \simeq 1/Nu_{\ell im}$ and therefore $\ell_0 = 0.01/d$ for the first group of data and $\ell = 0.06d$ for the second. For these values of ℓ_0 there is good agreement between the experimental and calculated data.†

The experimental data cited above are graphed in Fig. 5 in the $\tilde{Nu}$–$\tilde{Fo}$ plane, which is often used to analyze unsteady heat transfer data. We see that there is a rather significant scattering of the points at small $\tilde{Fo}$. Hence the conclusion that the quantity $\tilde{Nu}$ is independent of the thermal properties of the gas and the particles in the small Fo region is of restricted validity, as is the equation Nu = 2 in this region (see [16], for example). This is probably due to the fact that the experimental results were analyzed without the use of a system with a highly heat-conducting material (such as copper), for which the effect of the gas layer [3]) must be taken into account.

Developed Fluidized Bed. The system of equations (1)-(3) was also used to model unsteady heat transfer in a fluidized bed. The horizontal thermal diffusivity of the bed and its porosity were found from the equations established in [18] and [19], respectively. The quantity ℓ_0 was determined from the formula $\ell_0 = 0.14d(1-\varepsilon)^{-2/3}$ [10], and λ_f^h was determined from the expression given above, as in the case of a fixed bed. The first equation of (4) was used to determine the quantity α. Calculations of α using (1)-(3) were carried through for the materials listed above for different rates of air filtration. It was established that in all caes the coefficient α reaches its steady-state value (λ_f^h/ℓ_0) almost instantaneously, and remains constant thereafter. This is natural if we take into account the conclusion of [10] that the thermal resistance of the gas film dominates the heat transfer of a fluidized bed with a surface. This is also consistent with the calculations for a fixed bed and with (6), from which it follows (as noted above) that as a^0 increases the steady-state value of Nu is reached more rapidly and $Nu/Nu_{\ell im} \to 1$. Unfortunately, there is no reliable experimental data in the literature on unsteady heat transfer in fluidized beds, and hence the conclusions obtained here cannot be checked.

Discussion of the Results. Our analysis of the two-zones model shows that it quite accurately describes not only steady heat transfer [10, 13], but also unsteady heat transfer in dispersed systems. This universality, in our view, shows that the introduction of a zone of higher porosity is the most significant factor in the formulation of a model of heat transfer. It makes it possible to correctly take into account the following extremely important fact: the porosity and thermal conductivity in the boundary region (which, to a significant degree, determine the quantity α) can differ significantly from the corresponding quantities far from the heat-transfer surface [3, 9] (this is particularly true for the thermal conductivity in developed fluidized beds). It is not difficult to see that the actual continuous variation of these quantities near the heat-transfer surface is taken into account in the model with the only change being replacement by a step-like variation. It turns out that it is not essential to take into account the heterogeneity over the entire volume of the bed, as was done in [1], for example. It is true that in the framework of the model considered here such a treatment has some significance. As noted above, in such a treatment one of the basic parameters of the model (the coefficient k) is independent of the thermal characteristics of the dispersed material in the case of a fixed bed.

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*We emphasize once again that the quantity k characterizes only the effective thickness of the gas film. This is possibly the explanation for the different values of k for glass and copper balls. Use of the two-particle model to take into account the heterogeneity of the granular bed leads to the value $k \approx 0.1$ for both glass and copper balls (see [3], Fig. 4-23).

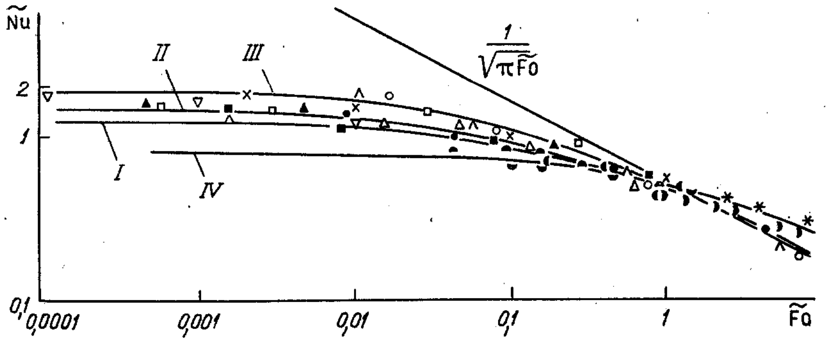

Fig. 5. Dependence of $\tilde{N}_u$ on $\tilde{F}_0$ for different media: I), II), III) numerical solution of (1)-(3) for freon-12, CO_2, and air respecticely; IV) calculation using (a); the meaning of the different points is the same as in Fig. 4.

The one-zone, two-temperature models, where the zone of higher porosity near the heat-transfer surface (which implies a change in ε and λ_s near the surface) is not taken into account, cannot provide a quantitative description of heat transfer for a wide interval of experimental conditions. The calculations of [3, p. 209] in the two-particle model showed that the effect of a gas layer with $\ell_0 = 0.1$ d had to be taken into account in order to obtain agreement with the experimental data for copper balls. For beds of particles with small thermal conductivity (such as glass balls) inclusion of a glass layer leads only to a slight increase in α (Figs. 4-23a and 4-23b in [3]). Ignoring the boundary zone leads to serious difficulties in describing heat transfer in mixed dispersed beds, for which high values of the effective thermal conductivities are typical (a fluidized bed, for example). Use of the one-zone, two-temperature model to analyze steady heat transfer in such systems involves the artificial introduction of the exposure time of a "packet" of particles near the heat-transfer surface [1]. The thermal conductivities of the gas and particles inside the "packet" are obtained as in the case of a fixed bed.

Hence some method of taking into account the significant change of the porosity and thermal conductivity of a dispersed bed near the heat-transfer surface is necessary for any model claiming to provide a universal description of the phenomena. Also, as shown by our analysis, the inclusion of a temperature difference between the phases and heat transfer between them over the entire volume of the granular bed is not necessary for a dispersed layer, which can be treated as a quasi-homogeneous medium.* The introduction of a temperature difference between the phases together with a thin boundary zone significantly complicates the model and makes it difficult to use for analysis [3], while only marginally improving the description of the process.

NOTATION

$a^0 = a_s^h/a_f^h$; a, thermal diffusivity; $B = R/\ell_0$; c, specific heat; d, particle diameter; $\mathrm{Fo} = a_f^h t/l_0^2$; $\mathrm{Fo}^* = a_f^h t/d^2$; $\tilde{\mathrm{Fo}} = a_s^h t/d^2$; K_0, K_1, modified Bessel functions of the second kind of orders zero and one; $K^* = K_1(\sqrt{\mathrm{Pe}^*}\,\xi^*)/K_0(\sqrt{\mathrm{Pe}^*}\,\xi^*)$; $k = \ell_0/d$; ℓ_0, thickness of the gas film; L, length of the heat-transfer surface; M, thickness of the plate; $\mathrm{Nu} = \alpha d/\lambda_f^h$; $\tilde{\mathrm{Nu}} = \alpha d/\lambda_s^h$; $\mathrm{Pe} = c_f\rho_f u l_0^2/c_s\rho_s(1-\varepsilon) L a_s^h$; $\mathrm{Pe}^* = c_f\rho_f u R^2/c_s\rho_s(1-\varepsilon) L a_s^h$; R, radius of the dispersed fill; r^0, radius of the cylinder; t, time; T, temperature; T_0, inlet temperature of the gas; u, rate of filtration; u_0, rate at the start of fluidization; r, x, coordinates; α, heat-transfer coefficient; ε, porosity; $\xi = r/\ell_0$ (cylinder); $\xi = x/\ell_0$ (plate); $\xi^0 = r^0/\ell_0$; $\xi_0 = \ell_0/R$; $\xi^* = (r^0 + \ell_0)/R$; $\theta = (T - T_0)/T_w - T_0)$; λ, thermal conductivity; $\lambda^0 = \lambda_s^h/\lambda_f^h$; ρ, density; $\Lambda = 2c_f\rho_f\ell_0/\rho_b c_b M$ (plate); $\Lambda = 2c_f\rho_f\ell_0/\rho_b c_b r^0$ (cylinder). Subscripts: b, heat-exchanger; f, gas; s, solid particles; w, heat-transfer surface; superscripts: h, along the horizontal.

*In general, the introduction of a temperature difference between the phases is necessary in the formulation of models of heat transfer inside a granular bed (for example, in the case where the direction of the heat flux is opposite to the drift velocity of the heat carriers (20]).

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