HEAT TRANSFER IN A RETARDED FLUIDIZED BED

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An oscillation model of a fluidized bed is used to examine the effect of the degree of retardation of the bed on external heat transfer.

To maintain constant temperatures in fluidized-bed units in which exothermic and endothermic reactions take place, heat-transfer surfaces are placed in the bed. Transferring heat by an agent circulating in a heat exchanger is the fundamental process in presently available methods of burning solid fuel in a fluidized bed [1]. Here, in accordance with the relation

$$Q = \alpha S \Delta t \tag{1}$$

an increase in heat removal Q can be achieved by growth of the heat-transfer surface S, since the coefficient of external heat transfer from the layer α is hard to increase under conditions of developed fluidization. The values of α that are reached are due to intensive mixing of the solid phase (grains) of the fluidized bed. Here, the determining scale of the bed L is the minimum size of the bed — either the diameter D_u in the high, narrow laboratory columns used to select the parameters of the process or the height of the bed H_{bd} in a full-size commercial unit. Development of the heat-transfer surface S in the unit by the placement of bundles of tubes or coils inside the bed interferes with free circulation of the solid phase and retards it. Thus, we can expect that an increase in the load S/V and restraint of the bed by the heat-transfer surfaces will be accompanied by a corresponding reduction in the heat-transfer coefficient α. In this case, the product αS, which determines total heat transfer, will increase nonmonotonically with S and, after attaining a certain optimum value $(\alpha S)_{max}$, will decrease sharply with a further increase in the degree of retardation.

Individual studies which have examined the effect of the geometry of the tube bundle on external heat transfer [2, 3] have not noted any appreciable reduction in α in the case of relatively low degrees of retardation. Qualitative considerations based on representations regarding the possibility of the tubes affecting the rupture of gas bubbles moving rapidly through the bed [4] do not permit prediction of the detailed character of the effect of S/V on α due to the formal, qualitative character of the so-called two-phase fluidization model [5] on which these representations are based.

The monographs [6, 7] developed the notion that gravitational oscillations which take place in a fluidized bed have a significant effect on all technical characteristics of the unit. This idea made it possible to explain the effect of the scale of the unit on these characteristics. According to rough estimates, the determining frequencies of these oscillations are related to the principal scale of the unit by the equation

$$\omega_0 = \sqrt{g/L}. \tag{2}$$

These frequencies and their periods $\tau_0 = 2\pi/\omega_0$ are also manifest in measured pulsations of local density, pressure, and velocity [7] and in the periods of circulating flows of "packets" of solid phase. These packets alternately press against the surfaces of the heat exchanger in succession with "bubbles." Based on similar representations, the "packet model" of Mikley and Fairbanks [8] made it possible to obtain a theoretical relation for the coefficient of external heat transfer of the fluidized bed

$$\alpha = 2\sqrt{\lambda_{ef} c_s \rho_s (1 - m_0)/\tau_0}, \tag{3}$$

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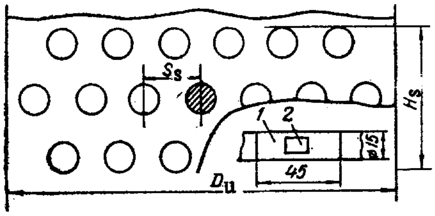

Fig. 1

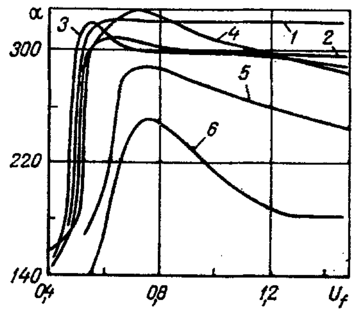

Fig. 2

Fig. 1. Geometry of the rod bundle: 1) heated section; 2) heat-flux transducer.

Fig. 2. Dependence of the heat-transfer coefficient on filtration velocity (sand with d_4 = 0.72 mm): 1) free bed; 2) σ = 0.1; 3) 0.16; 4) 0.2; 5) 0.34; 6) 0.5.

determined by the product of the volumetric specific heat of the solid-phase packets $c_s\rho_s(1 - m_0)$, where m_0 is the so-called porosity of the granular bed (usually, this quantity is on the order of $m_0 = 0.4$), and the effective thermal conductivity $\lambda_{ef} = \lambda_g f(m_0)$. The volumetric specific heat of the solid phase for nonporous particles, determined by vibration of the atoms of the crystalline lattice, depends little of the material and is about $2 \cdot 10^6$ J/(m^3•K). The effective thermal conductivity of a solid-phase packet pressed against the surface is determined by the thermal resistance of the gas spaces between particles and with a typical value $m_0 = 0.4$ is roughly 14 times greater than the thermal conductivity of the gas. The period τ_0 in Eq. (3) shows the effect of the gradual decrease in the temperature drop between the surface and packet as the time of their contact increases.

The correctness of the above physical approach is confirmed by the agreement of the orders of magnitude of estimates made and actual operating data. Thus, given values $L \cong 0.1$ m and $g = 10$ m/sec^2 found by averaging the laboratory and commercial scales, Eq. (2) yields a mean frequency of bed pulsation $\nu_0 = 5$ Hz, while in practice frequencies in the range 1-10 Hz are seen. Moreover, all of the observations confirm the fact, following from (2), that the periods of the pulsations increase with an increase in the scale of the unit. A direct check of the quantitative prediction

$$\alpha \sim 1/\sqrt[4]{L} \tag{4}$$

was not made due to the difficulty of separating the fundamental frequencies from the broadband noise background created by small jets and small internal fountains pulsating from the holes in the gas-distributing grate. Naturally, given such broadband excitation, the fluidized bed will resonate at the natural frequencies (2). Moreover, the amplitudes of these pulsations will be maximal. A similar formulation of the above values in Eq. (3) leads to the correct order of magnitude for the maximum coefficients of external heat transfer 300-400 W/(m^2•K), which have been measured by many authors by means of heaters immersed in laboratory columns with a fluidized bed. Insertion of (2) into (3) leads to the conclusion that α decreases relatively little with the scale of the unit. In fact, somewhat lower values of α = 180-200 W/(m^2•K) are seen in commercial units. The oscillation model of a fluidized bed (FB) also yields satisfactory estimates for certain other parameters of the process. Thus, with $L = 0.1$ m and $\nu_0 = 5$ Hz, the velocity of the circulating solid-phase flows $v_c = L/\tau_0 = L\nu_0 \sim \sqrt{L}$ is on the order of 0.5 m/sec and decreases with a changeover to the narrower laboratory columns. The available empirical data confirms that under these conditions the circulation velocities are on the order of tens of centimeters a second.

By analogy with the theory of transport processes in molecular physics and turbulence theory, we evaluate the mixing (diffusion) coefficients of the solid phase

$$D \approx 0{,}1 v_c L. \tag{5}$$

The relation $D \sim L^{3/2}$ which follows from this has made it possible to generalize all available measurements of the diffusion coefficients of grains in an FB from laboratory columns with a diameter of 3 cm to metric commercial units [7, p. 113]. Since mixing can begin only

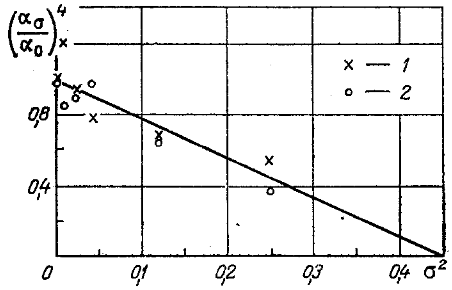

Fig. 3

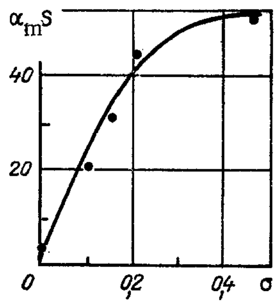

Fig. 4

Fig. 3. Change in the maximum heat-transfer coefficient with retardation (α_0 — free flow): 1) millet grain, d_4 = 2.0 mm; 2) sand, d_4 = 0.72 mm.

Fig. 4. Effect of retardation on the integral heat-transfer coefficient.

after the bed has made the transition from the stationary to the fluidized state, i.e., when the velocity of the flow u reaches the critical velocity corresponding to the beginning of fluidization u_c (the fluidization number $W = u/u_c = 1$), a more accurate design formula was proposed

$$D = \frac{1}{60}\sqrt{gL^3}(W-1)^n. \tag{6}$$

Measurements performed at the ITMO of the Academy of Sciences of the Belorussian SSR confirmed this relation with an exponent n = 0.85 [9].

Employing the correct sequence to evaluate the main parameters of an FB makes it possible to attempt to use the premises of the model to explain and evaluate the effect of retardation on the external heat transfer of the FB. Thus, it follows from (3) that α is proportional to the square root of the angular frequency of the pulsations ω. The retardation of the bed by the heat-transfer surfaces should lower this frequency due to friction of the oscillating masses against these surfaces. The elementary theory of vibrations for a system with one degree of freedom shows that the frequency is described by the relation

$$\omega = \sqrt{\omega_0^2 - \beta^2} \tag{7}$$

and changes from zero when the vibrations have been completely damped to the natural frequency ω_0. The parameter β is proportional to the total friction coefficient, i.e., in our case, the specific surface of the tubes placed in the bed or the degree of retardation of the bed

$$\beta = \gamma\frac{S}{V} = \omega_0 z. \tag{8}$$

It follows from this that α depends on S/V

$$\alpha \sim \sqrt{\omega} \sim \sqrt[4]{1-z^2}. \tag{9}$$

It is evident from this relation that the heat-transfer coefficient decreases by only 2.5% with an increase in the parameter z to a value equal to one-tenth of the maximum value $z_m = 1$, at which mixing of the solid phase has been completed suppressed. Even at z = 1/2, the reduction is no greater than 7%. Only immediately near the limiting value $z_m = 1$ does α decrease sharply. It is natural to suggest that z = 1 should be reached when the ratio V/S becomes comparable to the diameter of the grains of the bed and the latter begin to become wedged between the tubes in the bed, forming stationary "bridges" between them.

To qualitatively check the validity of prediction (9), we conducted a series of tests on a laboratory unit 200 × 200 mm in cross section. We fluidized particles of sand with a mean diameter of 0.7 mm and millet grain with d_4 = 2.0 mm. The sand and grain were fluidized with air at room temperature. A retarding insert consisting of horizontal rods with ϕ 15 mm, arranged in staggered fashion, was placed in the unit. The relative spacing of the tubes ranged from 1.16 to 4 in both the horizontal and vertical directions. A rod with an electric heater was placed in the center of the bundle (Fig. 1). The temperatures of the rod and bed were measured with Chromel–Alumel thermocouples. Heat flux was measured at the same time by a transducer — an LITMO calorimeter [10] — placed on the surface of the heating rod. The heat-transfer coefficient was calculated from Eq. (1). The degree of retardation was determined as the percentage of the volume of the fluidized bed occupied by retarding elements in the zone in which they were located:

$$\sigma = \frac{nF}{D_u H_s} = \frac{dS}{4\pi V} = z\sigma_{\mathrm{lim}}. \tag{10}$$

The heat-transfer coefficient α was measured for each packet with an increasing degree of retardation and different velocities of the fluidizing air flow u from 0.4 to 1.5 m/sec. The velocity of the air flow was calculated on the free section of the unit with retardation of the bed up to $\sigma = 0.5$. The character of the curves showing the dependence of α on u for retarded beds was the same as for the free fluidized beds: the heat-transfer coefficient increased sharply when the critical velocity corresponding to the beginning of fluidization u_c was reached. Here, α eventually attained a maximum value. Then, with further expansion of the bed and a decrease in the volumetric specific heat of the circulating two-phase medium, α began to decrease. The curves depicting the dependence of α on u for all of the tested values of σ are shown in Fig. 2.

To check prediction (9), we used these curves to determine the maximum values of the heat-transfer coefficient α_{max}. Also, to more clearly show the relationship, it was best that both parts of the equality be raised to the fourth power. Then the quantity α^4_m should have decreased linearly with an increase in σ^2. Figure 3 shows the experimental points in these coordinates, and it proved possible to use these points to draw (by the least squares method) a straight line. There is appreciable scatter on the initial section, which is evidently connected with the fact that at z = 0 the circulating flows are still experiencing friction against the wall of the column. This friction disturbs the proposed linear relation (5). Extrapolation of the averaged straight line to the value $\alpha_m = 0$ gives the formal limitint value $\sigma_{limn} = 0.665$, at which the bed motion has been completely damped. Unfortunately, this limiting value σ_{limn} cannot be applied to any system because the wedging of the grains should depend on their size. In our experiment, extrapolation of the straight line gives a limiting value of three for the ratio of the space between the rods to the diameter of the grains. For comparison, we note that when bulk materials are being emptied through a hole in a silo, the ratio of the hole diameter to the grain diameter at which wedging begins is 3-5 [11].

Since the dimension z that we introduced above is equal to the ratio of σ to σ_{limn}, we can analyze in general form the expected dependence of the rate of heat removal in the fluidized bed on the degree of retardation of the bed

$$S\alpha \sim f(z) = z\sqrt[4]{1 - z^2} = z(1 - z^2)^{1/4}. \tag{11}$$

Differentiating this function with respect to z, we have

$$f'(z) = (1 - z^2)^{1/4} + \frac{z}{4}(1 - z^2)^{-3/4}(-2z) = (1 - z^2)^{-3/4} \times \left(1 - z^2 - \frac{1}{2}z^2\right)$$

and from the condition $f'(z_m) = 0$ we find the position of its maximum $z_m = \sqrt{2/3} \approx 0.816$. It is interesting to note that a geometric calculation yields a cross section which is completely filled with a staggered array of tubes $\sigma_{full} = 0.848$, without any gaps.

Figure 4 shows the dependence of $S\alpha_m$ on σ for sand obtained from our test data. It can be seen from the figure that the integral heat removal αS still does not reach its maximum value even with a spacing between the rods (1.16-1)•15 mm = 2.4 mm, i.e., with a spacing which is 2.4/0.7 ≈ 3.4 times greater than the grain size. No wedging is seen and the mixing of the phase has not been completely suppressed in the unit. The last finding correlates fully with the data in a study we published earlier regarding mixing in an FB. The intensity of mixing was measured [12] from the movement of magnetite grains added to the fluidized sand in a column 400 mm in diameter. The appearance of the magnetite at a certain level was recorded by magnetometers. When the same column was first completely filled with ping-pong balls packed tightly so that the gaps between them at the narrowest sections amounted to about 21% of the total cross section, the solid-phase mixing coefficient measured from the movement of the magnetite grains decreased by less than one order of magnitude.

Thus, the completed theoretical analysis and verifying experiment show that fluidized-bed furnaces being designed allow for a substantial increase in the internal heat-removal surface without an appreciable reduction in integral heat transfer.

NOTATION

Q, integral heat flux, W; Δt, surface—bed temperature gradient, K; D_u, cross-sectional dimension of the unit, m; V, volume of the bed, m^3; ω_0, angular frequency of gravitational oscillations, sec^{-1}; g, acceleration due to gravity, m/sec^2; τ_0, ν_0, period and frequency of gravitational oscillations; λ_g, thermal conductivity of the gas phase, W/(m•K); λ_{ef}, effective thermal conductivity of a packet of particles, W/(m•K); c_s, ρ_s, specific heat, J/(kg•K), and density, kg/m^3, of the solid phase; m_0, porosity of the bulk granular bed; v_c, velocity of circulating flows of the solid phase, m/sec; D, mixing coefficient of the solid phase, m^2/sec; γ, friction coefficient, m/sec; n, number of tubes in a bundle; F, cross-sectional area of tube, m^2; H_s, height of bundle, m; d, diameter of tubes, m; u, filtration velocity of the fluidizing agent, m/sec.

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