CHARACTERISTICS OF EXTERNAL HEAT TRANSFER IN A FLUIDIZATION BED OF COARSE PARTICLES

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Results are shown of an experimental study concerning the heat transfer in a fluidization bed (2-13 mm) of particles, with a cylindrical thermoprobe used for measurements. A predominance of the convective component in the heat-transfer coefficient has been confirmed in the case of coarse particles. A criterial formula is proposed for calculating the coefficient of external heat transfer in a fluidization bed with the Archimedes number $Ar = 1.4 \cdot 10^5 - 3.0 \cdot 10^8$.

There are only very few published data available on the heat transfer in a fluidization bed of particles larger than 2 mm [1, 2]. Meanwhile, a study of such a bed would be quite valuable not only theoretically but also in many practical applications [3].

The authors have made such a study, using as the apparatus a cylindrical channel of acrylic glass with an inside diameter of 123 mm and a height of 1000 mm. The gas distributor was a perforated mesh with 1.5 mm (diameter) holes and a 29.2% active cross section. The fluidization beds were made up of steel balls 2.0, 3.97, 5.96, 7.94, or 9.53 mm in diameter, lead shot 2.0, 4.0, or 5.9 mm in diameter, and alundum balls 2.7, 5.0, 6.7, 10.0, or 12.92 mm in diameter. The beds were poured to a 160 mm height in all tests.

The steady-state method was used for measuring the heat transfer from the cylindrical surface of a copper probe, 40 mm in outside diameter and 100 mm in active length, with Hetinax deflectors at the ends for reducing the heat losses. Prior to the experiment, the probe was placed coaxially inside the apparatus with the aid of a special device. The misalignment between thermoprobe axis and apparatus axis was not more than 1-2 mm. The temperature difference between the thermoprobe surface and the fluidization bed was measured with two Chromel–Alumel differential thermocouples made of thermocouple-grade microwire 1.5 mm in diameter. The thermoprobe was a cylindrical tube, inside diameter 25 mm, containing a Nichrome heater coil with a maximum power of 270 W. Two grooves 1.5 mm deep were milled at opposite locations on the probe surface, 50 and 75 mm long, respectively, from the top end down. The thermocouple conductors with the hot junctions were then buried in these grooves under a layer of solder each and trimmed flush with the surface, while the cold junctions were installed inside the core of the fluidization bed. The thermocouple emfs were measured with a model PP-63 class 0.05 potentiometer and the heater power was measured with a model D-522 class 1.0 wattmeter. The beginning of steady state was checked by the readings of a model ÉPP-09M class 0.5 recording potentiometer. The heater power was regulated so as to maintain a temperature difference of not more than 45-50°C between the cylinder surface and the bed core. The mean heat-transfer coefficient was calculated without accounting for the heat losses through the probe ends and with the parameters of the fluidizing agent for the bed were adjusted according to the bed temperature and the pressure above the distributor grid. The thermoprobe was assumed to cover a negligibly small portion of the total channel section.

The heat-transfer coefficient is shown in Fig. 1 as a function of the dimensionless velocity of the fluidizing agent. Along the abscissas axis has been plotted the dimensionless velocity in terms of the Reynolds number, inasmuch as a change in the gas flow rate causes also a change in the gas temperature and pressure above the distributor grid, due to an increased drag in the bed and in the channel behind it.

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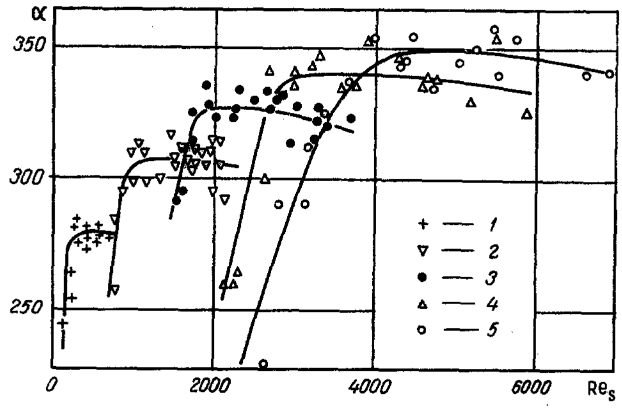

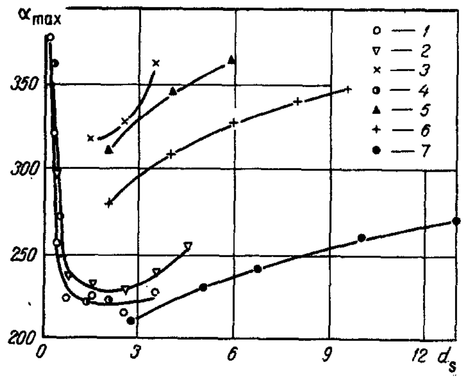

Fig. 1 Fig. 2

Fig. 1. Heat-transfer coefficient α (W/m$^2\cdot$deg) in a fluidization bed of steel particles, as a function of the Reynolds number: 1) $d_S = 2.0$ mm; 2) 3.97; 3) 5.96; 4) 7.94; 5) 9.53.

Fig. 2. Maximum coefficient of heat transfer α_{max} (W/m$^2\cdot$deg) as a function of the particle size: 1) according to data in [5] for BAV catalyst particles $D_c = 49$ mm; 2) 73; 3) aluminum–nickel catalyst particles $D_c = 49$ mm; 4) according to data obtained by N. F. Filippovskii for corundum particles in an apparatus with a vertical calorimeter [6]; 5) according to data by these authors for lead particles; 6) steel particles; 7) alundum particles. Particle diameter d_S (mm).

As in a fluidization bed of small particles, the heat-transfer coefficient increases until the gas velocity reaches a critical value. At gas velocities higher than that, the heat-transfer coefficient changes little and remains almost constant up to a fluidization number W = 1.5-2. It is to be noted that in a bed of coarse particles operation becomes more difficult at higher fluidization numbers, as has been mentioned in [1] already, because the fluidization range for coarse and heavy particles is much narrower on account of the much closer proximity of the critical velocity to the ascent velocity [4]. An analysis of test data on critical velocities has shown a close agreement with calculations based on the well-known Todes formula [4] with $\varepsilon = 0.43$.

It is shown in Fig. 2 that the value of the maximum heat-transfer coefficient decreases, as the diameter of particles increases to 1-2 mm, and then again increases. Such a trend of the heat-transfer coefficient is explained by its increasing convective component. While in a bed of fine particles ($d_S < 1$ mm) most of the heat removed from a surface is lost on heating those particles, in a bed of coarse particles ($d_S > 1$-2 mm) most significant is the direct transfer of heat to the gas which filters through near the heat-transfer surface [7, 8]. As the size and the density of particles are increased, the gas filtration velocity, which is almost equal to the critical fluidization velocity, also increases and according to [8], therefore, the coefficient α_{conv} increases proportionally to $d_S^{0.38}$ and $\rho_S^{0.46}$. The dimensions and, evidently, the shape of the calorimeter have a definite effect on the heat-transfer rate in a bed of coarse particles. In our test we increased the calorimeter diameter from 10 to 40 mm and noted an 8-15% decrease in α_{max}. According to Fig. 2, the values of α_{max} obtained in a bed of aluminum–nickel catalyst [5] are higher than our values obtained in a fluidization bed of heavier particles.

The large role of convection in the total heat transfer in a bed of coarse particles allows one to hypothesize that both the horizontal and the vertical components of the velocities of particles have little effect on the heat-transfer rate in the boundary layer and that, consequently, the heat-transfer coefficient here must be the same as in a dense bed of particles through which the gas flows at the critical velocity. At the same time, since the temperature gradient in the core of our fluidization bed is zero, unlike in a dense bed, hence all the thermal resistance to heat transfer is lumped in the boundary layer. The magnitude of this resistance will depend entirely on the properties and the flow mode of the gas, being determined by its molecular thermal conductivity and the turbulence.

Our test values for the heat-transfer coefficients in a fluidization bed are compared in Fig. 3 with those in a packing [9-12], with the Reynolds number referred here to critical fluidization velocities. At high values of the Reynolds number the formulas by various authors [9-12] yield approximately the same values for the heat-transfer coefficient at the packing surface and these are almost the same as our values of α for a fluidization bed. At small values of the Reynolds number the various formulas yield appreciably

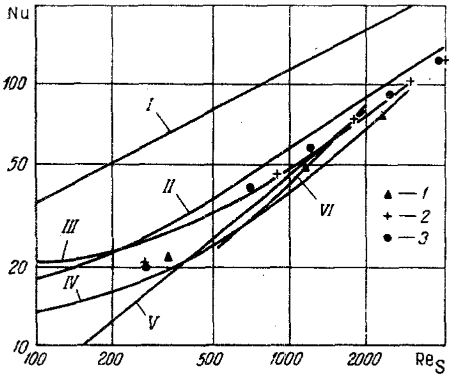

Fig. 3

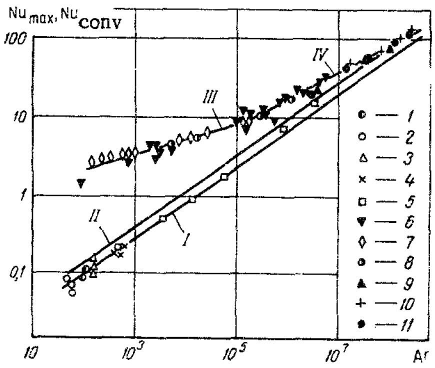

Fig. 4

Fig. 3. Comparison between heat-transfer coefficients in a fluidization bed and boundary-layer heat-transfer coefficients in a packing: I) according to the equation in [13]; II) the equation in [10]; III, IV) the equation in [12]; V) the equation in [11]; VI) the equation in [9]. 1) Test data obtained by these authors for a fluidization bed of lead particles; 2) steel particles; 3) alundum particles.

Fig. 4. Dimensionless heat-transfer coefficient in a fluidization bed, as a function of the Archimedes number: 1–4) data for Nu_{conv} in [14]; 5) in [2]; 6) for Nu_{max} in [16]; 7) in [15]; 8) for Nu_{max} obtained by N. F. Filippovskii with a vertical calorimeter [6]; 9) for Nu_{max} obtained by these authors with lead particles; 10) steel particles; 11) alundum particles. I) According to the equation in [14] with Pr = 0.7; II) according to the equation in [2]; III) according to the equation in [15]; IV) according to Eq. (2).

different values for the heat-transfer coefficient. These discrepancies are, apparently, due to difficulties in estimating the conductive component of the total heat-transfer coefficient α.

While analyzing all the few known published studies on the heat transfer in dense and in fluidized beds, one of these authors [14] tried to estimate the convective Nusselt number for fluidization beds of coarse particles. In a later special study [2] the convective component was determined on the basis of mass transfer from a vertical naphthalene cylinder and basic agreement was found with the relation derived in [14]. For near or above optimum velocities calculated by the Todes formula, this equation becomes

$$Nu_{conv} = 0.0175\,Ar^{0.46}\,Pr^{n}, \tag{1}$$

with n = 1 according to [14] or n = 0.33 according to [2].

It is evident, according to Fig. 4, that in a bed of coarse particles the value of the total heat-transfer coefficient approaches that of its convective component calculated according to Eq. (1), with this equation remaining valid at least up to $Ar = 3.0 \cdot 10^8$. As the size of particles is decreased, the increasing role of the conductive component in the total heat transfer causes an increasing deviation of the test data from values based on formula (1). It is to be noted that heat transfer via the particles is somehow accounted for in most formulas for the boundary-layer heat-transfer coefficient in dense beds. In [9], for example, Nu_{conv} varies from 1.2 to 25 depending on the test conditions as well as on the material and the size of particles, while in [12] $Nu_{conv} = 18$.

Considering the data in [2, 5, 14, 15] and our data, the maximum coefficient of heat transfer from a surface to a fluidization bed of coarse particles is best calculated according to the following formula:

$$Nu_{max} = 0.21\,Ar^{0.32}, \tag{2}$$

valid for $Ar = 1.4 \cdot 10^5 – 3.0 \cdot 10^8$.

NOTATION

d_s is the diameter of particles in a bed;
D_c is the diameter of the test channel;

ρ_S	is the density of particle material;
α, α_{conv}, α_{cond}, α_{max}	are the total, convective, conductive, and maximum heat-transfer coefficient;
ε	is the porosity;
W	is the fluidization number;
Nu, Nu_{conv}, Nu_{cond}, Nu_{max}	are the Nusselt number based on α, α_{conv}, α_{cond}, and α_{max}, respectively;
Pr	is the Prandtl number;
Re_S	is the Reynolds number, referred to particles;
Ar	is the Archimedes number, referred to particles.

LITERATURE CITED

1. I. P. Mukhlenov, D. G. Traber, and V. B. Sarkits, in: Study of a Suspension Bed [in Russian], Vol. 54, Trudy Lensovet LTI, Gostekhizdat (1959).
2. A. P. Baskakova and V. M. Suprun, Khim. i Neft. Mashinostr., No. 3 (1971).
3. E. I. Bulakhov, in: Application of the Fluidization Bed in the Metallurgical Industry [in Russian], No. 21 (1970).
4. V. D. Goroshko, R. B. Rozenbaum, and O. M. Todes, Izv. Vuzov, Neft i Gaz, No. 1 (1958).
5. D. G. Traber, V. B. Sarkits, and I. P. Mukhlenov, Zh. Prikl. Khim., 33, No. 10 (1960).
6. N. F. Filippovskii and A. P. Baskakov, Inzh.-Fiz. Zh., 12, No. 2 (1972).
7. S. S. Zabrodskii, Hydrodynamics of and Heat Transfer in a Fluidization Bed [in Russian], Gosenergoizdat (1963).
8. A. P. Baskakov, Rapid Nonoxidizing Heatup and Heat Treatment in a Fluidization Bed [in Russian], Metallurgiya (1968).
9. S. Yagi and D. Kunii, AIChEJ, 6, No. 1, 97-104 (1960).
10. M. É. Aerov and D. A. Narinskii, Papers of the Eighth All-Union Interacademic Conference, Odessa (1968).
11. S. Yagi and N. Wakao, AIChEJ, 5, No. 1, 79-85 (1959).
12. A. D. Caldwell, Chem. Eng. Sci., 23, No. 4, 393-395 (1968).
13. S. Yagi and D. Kunii, Internat. Developm. in Heat Transfer, Part 4, ASME, New York (1961), pp. 750-759.
14. A. P. Baskakov, Izv. Akad. Nauk SSSR, Énergetika i Transport, No. 3 (1966).
15. N. N. Varygin and I. G. Martyushin, Khim. Mashinostroenie, No. 5 (1959).
16. I. P. Mukhlenov, D. G. Traber, and V. B. Sarkits, Zh. Prikl. Spektr., 33, No. 10 (1960).