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STUDY OF HEAT EXCHANGE BETWEEN A MODEL PARTICLE AND A FLUIDIZED BED

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UDC 66.096.5

An experimental study is made of heat exchange between an unsecured model particle and a fluidized bed. The test data is generalized with a dimensionless relation.

Work is now being done here and abroad on the development of new burners for low-temperature combustion of solid fuel in a fluidized bed (FB). The new type of burner is capable of operating efficiently on low-grade, coarse-crushed coals and makes it possible to reduce harmful atmospheric emissions [1]. The fuel is fed into an FB of coarse, incombustible particles and comprises 1-3% of the bed weight. Heat transfer from the burning particles to the FB to a large extent determines its temperature and, thus, the kinetics of the reaction occurring on its surface, as well as the condition of the mineral part of the fuel. The last-mentioned factor is particularly important, since fusion of the ash may lead to sintering of the particles and disruption of the operation of the burner.

The literature contains extensive information on the heat exchange of an FB with stationary surfaces submerged within it [2, 3]. However, in the system being discussed here the particle is not stationary and participates in the complex circulating movement of the material in the FB. Meanwhile, its dimensions and density are different from the dimensions and density of the inert particles. The data in [4], obtained by measurement of the temperature and time of combustion of a single coke particle in a fluidized bed in a 40-mm-diameter column, showed that the dimensions of the fuel particle affects the rate of heat transfer.

We had the goal of taking a detailed look at the heat exchange between a movable model particle and an FB under "cold" conditions and determining how it is affected by the physical parameters of the particle itself and the material of the bed, as well as the hydrodynamics and scale of the system.

The tests were conducted in apparatuses of circular (150 mm diameter) and rectangular (400 × 250 mm) cross section. The apparatuses had transparent windows permitting observation of the bed. In both apparatuses, we used gas-distributing grates in the form of perforated metal plates with a layer of dense cloth pressed between them. To ensure uniform gas distribution, the chambers under the grates were filled with spherical packing. The bed was fluidized with room-temperature air. The rate of flow of the air was measured with an accuracy no worse than ±3% from the pressure drop after a standard diaphragm. Table 1 shows characteristics of the dispersed materials used. The initial height of the bed in the tests was not changed and was 150 and 250 mm, respectively, for the small and large apparatuses.

The heat-transfer coefficient was determined by the method of regular thermal regime [5]. The combination aluminum transducer and model particle with a caulked-in thermocouple was heated in molten tin to about 250°C and thrown into the FB. Since the value of the Bi criterion did not exceed 0.01 in the tests and the nonuniform temperature over the cross section of the transducer could be ignored [5], the average heat-transfer coefficient over the surface was equal to $\alpha = mcG/F$.

The temperature of the bed was measured with a thermocouple with an open junction. The temperature difference between the transducer and the FB was recorded by a potentiometer to within 0.5°K, which ensured

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TABLE 1. Characteristics of the Dispersed Materials

Serial number	Material	d_i, mm	ρ_i, kg/m³	φ
1	Sand	0,62	2500	0,85
2	Spherical silica gel	0,65	900	1,0
3	Glass beads	1,2	2500	1,0
4	Millet grains	2,0	1200	0,96
5	Rolled fireclay powder	2,0	2350	0,8
6	Fireclay powder	3,2	2300	0,7
7	Peas	6,2	1340	1,0
8	Porcelain beads	6,3	2400	1,0

TABLE 2. Characteristics of the Transducer–Model-Particles

Serial number	ρ_p, kg/m³	D_e, mm	D, mm	L, mm
		Spherical transducers		
1	2700	5,2	5,2	—
2	2700	7,7	7,7	—
3	2700	9,6	9,6	—
4	2700	12,3	12,3	—
5	2700	15,0	15,0	—
		Cylindrical transducers		
6	2700	16,0	13,1	12,9
7	1790	16,2	13,1	13,55
8	2700	10,9	9,35	9,0
9	1840	11,3	9,35	8,05
10	1150	9,8	8,05	8,0

an accuracy of ±5% for the measurement of the heat-transfer coefficient for a 5.2-mm-diameter transducer and an error no larger than ±4% for larger transducers. Heat outflow along the thermoelectrodes from the hot junction of the thermocouple was minimized by making a through dilametrical hole in the body of the transducer and caulking the thermocouple at the transducer outlet flush with the surface. Thus, a section of length D of the thermocouple traversed an isothermal surface. We used spherical and cylindrical transducers, characteristics of which are shown in Table 2. The effective diameter of the cylindrical transducer was determined as the diameter of an equivalent-surface sphere. Some of the cylindrical transducers were made hollow to determine the effect of the density of the model particle on the value of α. The ends of the hollow cylinders were glued to thin (0.4 mm) glass–Textolite coverings. End heat losses were evaluated with the approximation of a linear temperature distribution in the covering (heat transfer into the latter can be considered quasistatic, since the characteristic cooling time of the covering is 20-30 times less than the cooling time of the transducer). The size of the loss (15-23%) was considered in calculating the heat-transfer coefficient.

Despite the fact that the thermocouples caulked into the transducers had thin (0.2 mm) and fairly long electrodes, they could have had some effect on the character of motion of the model particle. To evaluate this effect, we conducted a special series of experiments. A model particle without a thermocouple was immersed in molten tin and heated for a certain time nearly to its temperature. It was then thrown into the FB. After a certain period τ, it was quickly extracted by means of a netted fiberglass cage submerged previously in the bed. The cells in the cage were smaller than the dimensions of the transducer but permitted passage of the bed material. In individual tests, the particle was suspended on a thin thread which offered almost no obstruction to its movement. The temperature of the particle extracted from the bed was measured by the tightly clamped junction of a low-inertia thermocouple. The temperature measurement error was evaluated as follows. As the transducer cooled in air, the caulked-in thermocouple pressed to its surface was used to record curves of the temperature change over time. This allowed us to correlate each temperature measured by the contact method with the actual temperature of the transducer. The resulting curves also allowed us to take into account the cooling of the transducer in air up to the moment of measurement of its temperature. The resulting set of values of transducer temperature, corresponding to different times τ with the same fluidization regime, made it possible to determine the cooling rate and heat-transfer coefficient.

It can be seen from the test data shown in Fig. 1 that the presence of the caulked-in thermocouple had almost no effect on the heat-transfer efficiency of the particle. This allowed us to use the thermocouple-equipped transducers in the main series of tests.

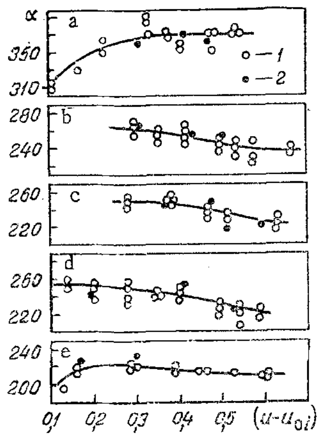
Fig. 1

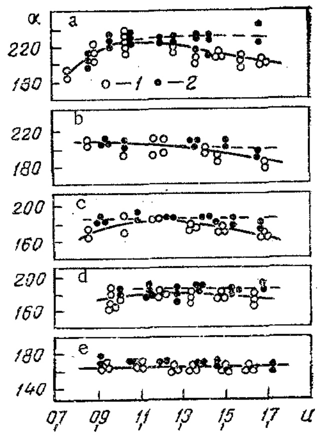
Fig. 2

Fig. 1. Experimental relations $\alpha = f(u - u_{0i})$: 1, 2) transducers with and without a caulked-in thermocouple; a) sand, transducer No. 1; b, c) silica gel, transducers Nos. 1 and 2; d, e) glass beads, transducers Nos. 1, 4. Diameter of apparatus 150 mm. α, $W/m^2 \cdot {}^\circ K$; $(u - u_{0i})$, m/sec.

Fig. 2. Experimental relations $\alpha = f(u)$: 1, 2) apparatuses of 150-mm diameter and 250 × 400 mm cross section; a-e) transducers Nos. 1, 2, 4, 7, and 10, respectively; bed material – fireclay, $d_i = 2$ mm, u, m/sec.

Figures 1 and 2 show typical empirical dependences of α on the filtration rate. It can be seen from the figures that the heat-transfer coefficient changes negligibly (within 5-10%) with filtration rate. The functions $\alpha = f(u)$ generally are monotonically decreasing for smaller model particles, with the maximum near the beginning of fluidization. For larger transducers – especially in fluidized beds of relatively small particles – the maximum of the curve shifts in the direction of high gas velocities. This is possibly related to the low mobility of the transducer, or even to its lying on the grate, at low values of u. The mobility of the transducer increases as the velocity increases. This was confirmed by both visual observations and the more intensive tapping of the transducer against the wall of the column. At high filtration rates, the periodic ejection of the model particles above the surface of the bed was particularly evident. The small transducers were mobile nearly throughout the range of fluidization velocities.

It can be seen from Fig. 2 that the values of α_{max} obtained in columns of different cross section, other conditions being equal, agree to within 3%. This is evidence of the independence of the heat-transfer coefficient on the scale of the apparatus.

Figure 3 shows the dependence of α_{max} on the size of the bed particles. With a constant-size transducer, heat-transfer efficiency decreases with an increase in particle diameter but, already at $d_i = 1$-2 mm, stabilizes. It then begins to increase. This is evidently due to the fact that there are two heat-transfer mechanisms in the system: a conductive mechanism connected with adsorption and heat transfer by the particles; a convective mechanism (heat transfer by the moving gas), which predominates in fluidized beds with coarse particles [2]. It should be noted that the coarse-particle range, being the least studied, is of great practical interest from the point of view of the combustion and processing of solid fuel in fluidized beds. Taking this circumstance into account, we generalized the test data for bed particles with $d_i \geq 1.2$ mm.

It can be seen from Fig. 3 that, with a constant d_i, heat-transfer efficiency increases with a decrease in transducer size.

The test data was generalized in the form of the functional relation

$$\mathrm{Nu}_{max} = f(\mathrm{Ar},\ d_i/D_e,\ \rho_i/\rho_p,\ \varphi). \tag{1}$$

We chose Eq. (1) on the basis of the consideration that it should reflect the main properties of the system, the conditions of suspension of the particles by the gas – determined by the criterion Ar – and individual parameters of the model particle – relative values of its diameter and density. Since the criterion Ar (as Nu_{max}) is based on the diameter and density of the bed particles, the quantities d_i and ρ_i were taken as measures for the corre-

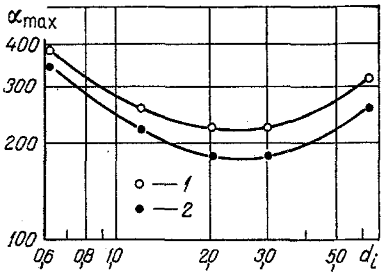

Fig. 3

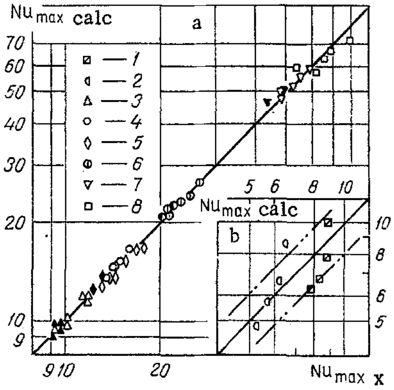

Fig. 4

Fig. 3. Experimental relations $\alpha_{max} = f(d_i)$: 1, 2) transducers Nos. 1 and 4; apparatus of diameter 150 mm, d_i, mm.

Fig. 4. Comparison of experimental data with data calculated with Eq. (1) – a, and the equation in [4], b: 1-8) bed materials Nos. 1-8) bed materials Nos. 1-8, respectively (the darkened points correspond to the cylindrical transducers).

sponding characteristics of the model particle. As shown above, the dimensions of the apparatus do not affect the value of α_{max} and were not included in Eq. (1).

In fluidized beds of coarse material, the shape of the particles may affect the heat-transfer coefficient. There is not sufficient time for thorough heating of the particles during their stay at the surface [2], and most of the resistance to conductive heat flow is exerted by the contact gas interlayer. The effective thickness of the interlayer is greater for particles of irregular form than for spherical particles of the same mean diameter. This caused us to insert the form factor φ into Eq. (1). This factor is equal to the ratio of the surface of a sphere of equivalent volume to the surface of the particle. The form factor of the particles was determined by analyzing particle micrographs and then averaging.

The test data was generalized in accordance with Eq. (1) in the form of an exponential function. The generalization was done on an ES-1022 computer by the method of multiple linear regression analysis [6]. As a result, we obtained the following correlation:

$$\mathrm{Nu}_{max} = 0.41\mathrm{Ar}^{0.3}\left(\frac{d_i}{D_e}\right)^{0.2}\left(\frac{\rho_i}{\rho_p}\right)^{-0.07}\varphi^{0.66}. \tag{2}$$

Figure 4 compares theoretical relation (2) with the test data. The mean relative error of the approximation was 3.9%, which is close to the mean accuracy of the measurement of α. Table 3 shows the ranges of the parameters in Eq. (2) and the standard errors of the corresponding exponents (regression coefficients).

It should be noted that there was a clear discrepancy (of about 30%) between the results of for the irregular particles and the spherical particles when we attempted to generalize the test data by either introducing the form factor into the criteria in (1) in the form of the produce φd_i or by not allowing for the factor. Equation (2) also generalizes the test data for the spherical and cylindrical transducers (as a result of introduction of the effective diameter of the latter). This permits us to recommend the above correlation for evaluating the rate of heat transfer of model particles of irregular form.

Since α changes only slightly with an increase in filtration rate (see Figs. 1 and 2), Eq. (2) can be used for approximate calculations of heat-transfer rate at filtration rates differing from u_{opt}, which corresponds to the maximum on the curve $\alpha = f(u)$.

It follows from Eq. (2) that the heat-transfer coefficient increases with an increase in the size of the bed particles ($\alpha \sim d_i^{0.1}$) and decreases with an increase in the diameter of the model particle ($\alpha \sim D_e^{-0.2}$). Since the size of the model particle affects α more strongly, then an increase in d_i with a fixed value of the ratio d_i/D_e (i.e., with a simultaneous proportional increase in D_e) will lead to a decrease in α. In the special case when the model particle is no different from the bed particles (when $d_i/D_e = \rho_i/\rho_p = 1.0$ and the particles are spherical), correlation (2) takes the form:

TABLE 3. Ranges of the Parameters in Eq. (2) and the Standard Errors of the Corresponding Exponents

Parameter	Range	Regression coefficient (exponent)	Standard error of regression coefficient
Ar	$1{,}55\cdot10^5$—$2{,}2\cdot10^7$	0,302	±0,007
d_i/D_e	0,074—1,21	0,197	±0,02
ρ_i/ρ_p	0,44 —2,07	0,071	±0,018
φ	0,7 —1,0	0,66	±0,044

$$\mathrm{Nu}_{max} = 0.41\mathrm{Ar}^{0.3}. \tag{2a}$$

Equation (2a) is close in form to the correlation presented in [7] for heat transfer between a fluidized bed of coarse particles and a stationary vertical tube of large diameter, although the values of Nu_{max} obtained from Eq. (2a) are 40-55% higher than the values obtained from the correlation in [7]. This evidently has to do with the small size and mobility of the model particle. An intensification of heat transfer with a decrease in the diameter of secured cylindrical transducers to $D \le 2$-3 mm was seen earlier in fluidized beds of small particles [8, 11]. In [11], this phenomenon was analyzed on the basis of features of conductive heat transfer near a cylindrical surface of large curvature. The analysis led to a semiempirical formula (see [3]), which considers the effect of D on the heat-transfer coefficient. Comparison of the results calculated from the correlation in [7] with allowance for this formula and from Eq. (2a) showed that an increase in the conductive component can explain only about 60% of the increase in α_{max} at $d_i = D_e = 1$ ($\mathrm{Ar} \simeq 10^5$). The corresponding percentage are 17 and 6%, respectively, when we increase $d_i = D_e$ to 3 and 5 mm ($\mathrm{Ar} \simeq 10^6$ and 10^7). This shows that the effect of the size of the movable particle in its heat exchange with a fluidized bed of coarse particles is due mainly to the predominant contribution of convective heat transfer.

Correlation (2) allows us to calculate the heat-transfer coefficient of a fuel particle in a fluidized bed when the radiant component is negligible. The criterion Ar entering into this correlation accounts for the effect of the thermophysical properties of the fluidizing gas and, thus, the temperature and pressure, on the rate of heat transfer. As was shown in [9], a formula of similar structure obtained on a "cold" model for calculating the heat transfer between a fluidized bed of coarse particles and a tube bundle ($\mathrm{Nu}_{max} \sim \mathrm{Ar}^{1/3}$) is equally valid for full-scale conditions in a burner with a fluidized bed.

The CO combustion reaction may affect the rate of heat transfer between a fluidized bed and a burning particle [4]. In the opinion of the authors of [10], this reaction, at least in the combustion of coarse particles, occurs either on or very near the surface. Thus, the heat-transfer coefficient is not affected. Figure 4b compares the data obtained in the present work for a fluidized bed of relatively fine particles (d_i = 0.62; 0.65 mm) with data calculated from the formula proposed in [4] for heat transfer between a fluidized bed heated to 1023°K and a coke particle burning in the bed. The formula in [4] is valid at $\mathrm{Ar} \le 4 \cdot 10^4$ and $2.3 \le D_e/d_i \le 14$. The deviation of the experimental points from the theoretical curve does not exceed the limits of accuracy of the formula. This shows that Eq. (2) can be used to mathematically model and calculate combustion in fluidized-bed burners in the ranges of parameters shown in Table 3.

NOTATION

c, specific heat of the transducer material; D_e, effective diameter of the transducer–model-particle; D, L, diameter and length of the transducer; d_i, mean diameter of bed particles; F, heat-exchange surface of the transducer; G, weight of the transducer; g, acceleration due to gravity; m, transducer cooling rate; u, filtration rate; u_{0i}, rate at beginning of fluidization of the bed of inert particles; α, average heat-transfer coefficient over the surface of the transducer; α_{max}, value of α corresponding to the maximum on the curve $\alpha = f(u)$; λ, λ_g, thermal conductivities of the transducer material and gas; ν, kinematic viscosity of the gas; ρ_p, ρ_g, ρ_i, effective density of the model particle, density of the gas, and density of the bed particles; φ, form factor of the bed particles; $\mathrm{Ar} = gd_i(\rho_i - \rho_g)\nu^{-2}\rho_g^{-1}$; $\mathrm{Bi} = \alpha D/2\lambda$; $\mathrm{Nu}_{max} = \alpha_{max}d_i/\lambda_g$. Indices: calc, calculated; x, experimental.

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SOME RELATED PROBLEMS OF FILTRATION AND HEAT CONDUCTION IN POROUS BODIES

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An examination is made of a dynamically similar boundary-value problem describing the uniform motion of a liquid or gas initiated by intensive heat flow to a porous body.

1. We will examine the motion of a uniform liquid in a porous body under nonisothermal conditions. We will assume that the time required to establish local thermal equilibrium is short and that we can use a one-temperature model. We then have the system of equations (see [1], for example):

$$-\frac{k\rho}{\mu}\nabla p = \Phi(J)\,\mathbf{J}/J, \tag{1}$$

$$\frac{\partial(m\rho)}{\partial t} + \operatorname{div}\mathbf{J} = 0, \tag{2}$$

$$\frac{\partial(i\rho m + CT)}{\partial t} + \operatorname{div}(i\mathbf{J}) + \operatorname{div}\mathbf{q} = 0, \tag{3}$$

$$\mathbf{q} = -\lambda\nabla T,\quad i = i(p,\ T),\quad \rho = \rho(p,\ T),\quad \lambda = \lambda(p,\ T),\quad \mu = \mu(p,\ T). \tag{4}$$

The function $\Phi(J)$ describes the filtration law. With the chosen form, system (1) covers a variety of cases of nonisothermal motion of liquids and gases in a porous medium with both a linear and a nonlinear filtration law. It was used in [1] to study temperature changes connected with the Joule–Thompson effect in the nonsteady flow of gas to wells. Below we examine what is in a sense the opposite problem: the motion of a liquid or gas initiated by intensive heat flow to a porous body.

2. Let a half-space $x \geq 0$ at the initial moment of time be filled with a moving gas with a constant pressure p_0 and temperature T_0, and let certain new values of pressure and temperature p_1 and T_1 be established at the boundary beginning with the moment $t = 0$. The resulting unidimensional motion satisfies the conditions

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