5. S. P. Malyshenko and V. I. Mika, Teplofiz. Vys. Temp., 12, No. 3, 735 (1974).
6. A. Z. Patashinskii and V. L. Pokrovskii, Fluctuation Theory of Phase Transitions [in Russian], Nauka, Moscow (1975).
7. J. Cahn and J. Hillard, J. Chem. Phys., 28, 258 (1958).
8. V. G. Barantsev, Author's Abstract of Candidate's Dissertation, Leningrad State University (1974).
9. K. G. Wilson, Physica, 73, 199 (1974).
10. I. A. Fakhretdinov and A. V. Chalyi, Izv. Vyssh. Uchebn. Zaved., No. 11, 35 (1976).
11. L. D. Landau and E. M. Lifshits, Mechanics of Continuous Media [in Russian], GTTI, Moscow (1944).
12. I. Z. Fisher and Yu. P. Krasnyi, Ukr. Fiz. Zh., 11, 104 (1966).

HEAT EXCHANGE IN THE FLOW OF A GAS SUSPENSION IN A LONG HORIZONTAL PIPE

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UDC 536.244

The results of an experimental investigation of the heat exchange of a gas-suspension stream in different lengths of horizontal pipes are presented.

It is known [1] that two-phase streams of the gas–solid-particle type are distinguished by a number of high qualities as the heat-transfer agent and working substance in power engineering. The inadequate degree of study and the complexity of the mechanism of heat-exchange processes in the flow of such streams result in the necessity of the systematic accumulation of test data.

The results of an investigation of heat exchange in the motion of a gas suspension in a horizontal pipe are presented in the report being offered. Air served as the carrier medium of the stream, while particles of white marble 110-220 μm in size were used as the solid phase.

The investigation* was carried out on an experimental installation for which a schematic diagram and the measurement procedure was described in [2]. But in contrast to [2], the start-up and heated sections were made of one common pipe without joints or connections. The inner diameter of the pipe was 14 mm. The heating was accomplished with radiative electric furnaces in sections on a pipe length of 4 m.

The separate determination of the transferred heat flux for the sections from the difference in the power supplied and lost through the thermal insulation of the furnaces made it possible to find in each test the characteristics of the heat exchange for different pipe lengths x = 1, 2, 3, and 4 m.

In order to study the influence of the dynamic conditions of entrance of the stream on the heat exchange we provided for the delivery of solid particles to the carrying air at distances of 0.07, 0.3, 0.75, and 1.37 m from the heated section. The start-up section had special fitting in its upper part for this. A calculated estimate of the velocity of the solid particles at the entrance to the heated section, made on the basis of a one-dimensional flow model, showed that with such start-up lengths under the conditions of the experiments the particle velocity was about 0.2-0.85 of the velocity of the transporting gas.

The investigation was carried out at flow-rate concentrations of the solid phase of from 0.17 to 8 kg per 1 kg of air. The Reynolds number of the gas was varied in the interval of $Re_{W;d}$ = 1950-18,000. In this case the transportation of the particles was stable with a stable drop of stream pressure over the pipe length in each test. The stream temperature at the entrance to the heated section was in the range of t_0 = 10-60°C. The wall temperature along the length of the pipe was kept nearly constant and was varied from 130 to 690°C from test to test. Measurements in the cross sections x/d = 35.7 and 207 showed that it also hardly varied over the perimeter of the pipe in each test.

*Engineer V. I. Rednikov took part in conducting the tests.

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TABLE 1. Values of the Multiplier c and the Exponent n in the Empirical Equations

Equation	$x=1$ m ($x/d=71.5$)		$x=2$ m ($x/d=143$)		$x=3$ m ($x/d=214.5$)		$x=4$ m ($x/d=286$)	
	c	n	c	n	c	n	c	n
(5)	0,0034	0,9	0,00156	0,95	0,0009	0,9755	0,00065	0,9878
(6)	0,0106	0,74	0,00566	0,79	0,00387	0,8155	0,00323	0,828
(7)	2,96		3,63		4,3		4,97	

The coefficient of heat exchange was determined, as in [2], from the initial temperature difference with a limiting relative error of 6%.

The tests with a gas suspension were preceded by tests with pure air.

The results obtained were compared with a calculation by the well-known equation

$$\mathrm{Nu}_{g,d}=0.018\mathrm{Re}_{g,d}^{0.8}(T_g/T_w)^{0.5}, \tag{1}$$

where the coefficient of heat exchange was determined from the logarithmic-mean temperature difference. Good agreement between the data obtained and the dependence (1) was noted for x = 1 and 2 m. For x = 3 and 4 m the test points had a larger scatter, which was explained by the increased error in the determination of the temperature difference with the small temperature heads at the end of the pipe.

With allowance for the relation

$$\Delta t_l=\Delta t_0\frac{4\mathrm{St}\dfrac{x}{d}}{\ln\dfrac{1}{1-4\mathrm{St}\dfrac{x}{d}}} \tag{2}$$

between the initial temperature difference and has logarithmic-mean difference with t_w = const [3], and for the temperature dependences

$$\lambda_g/\lambda_w=(T_g/T_w)^{0.82},\ \mu_g/\mu_w=(T_g/T_w)^{0.683} \tag{3}$$

of the coefficients of thermal conductivity and viscosity of air [4], one can transform (1) to the form

$$\mathrm{Nu}_{w,d}^{0}=\left\{1-\exp\left[-0.072\frac{x/d}{\mathrm{Re}_{w,d}^{0.2}\mathrm{Pr}_w}\left(\frac{T_g}{T_w}\right)^{0.7736}\right]\right\}\mathrm{Re}_{d,w}\mathrm{Pr}_w\frac{d}{4x}. \tag{4}$$

The coefficient of heat exchange determined from (4) is referred to the initial temperature difference.

The test data obtained satisfy (4) with an accuracy of ±4% for all the pipe lengths. Their approximation led to the simple dependence

$$\mathrm{Nu}_{w,d}^{0}=c\,\mathrm{Re}_{w,d}^{n} \tag{5}$$

where the multiplier c and the exponent n of the Reynolds number are variable and depend on x/d. The values of c and n are presented in Table 1.

The results of the tests on heat exchange with a gas suspension for different lengths of heated section are presented in Figs. 1 and 2.

It is seen from Fig. 1 that a change in the length of the preconnected isothermal start-up section does not affect the intensity of heat exchange. Separation of the test points into layers with respect to the start-up length is also absent in the coordinates of Fig. 2 (not shown in the figure). A decrease in the start-up length to 0.07 m with other parameters in the tests unchanged led to a decrease in the wall temperature at the start of the heated section. But a decrease in t_w was recorded only by the first and, to a lesser extent, by the second thermocouples, mounted on the wall at the cross sections x/d = 7.15 and 21.4. It comprised a few degrees or tenths of a degree and was not reflected in the indices of average heat exchange for pipe lengths x ≥ 1 m.

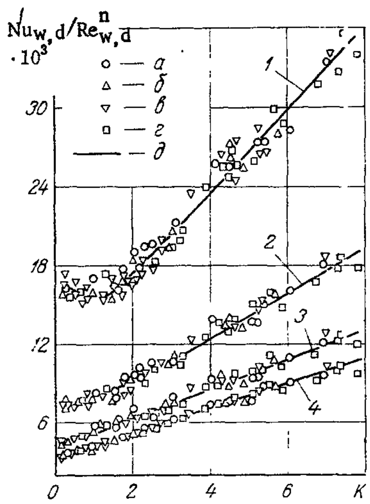

Fig. 1

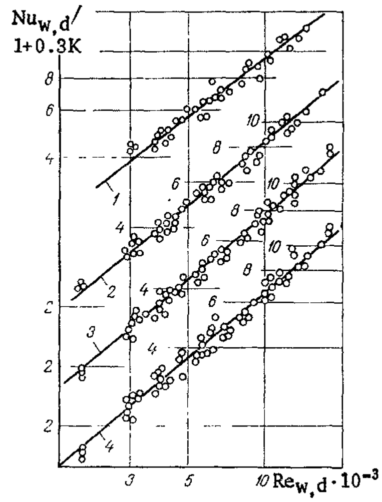

Fig. 2

Fig. 1. Dependence of $Nu_{w,d}/Re^n_{w,d}$ on flow-rate particle concentration K, kg/kg: 1) x = 1 m; 2) 2; 3) 3; 4) 4 m; a) x_{st} = 0.07 m; b) 0.3; c) 0.7; d) 1.37 m; e) from Eq. (6).

Fig. 2. Dependence of $Nu_{w,d}/(1 + 0.3K)$ on Reynolds number $Re_{w,d}$: 1) x = 1 m; 2) 2; 3) 3; 4) 4 m; solid curve) from Eq. (7).

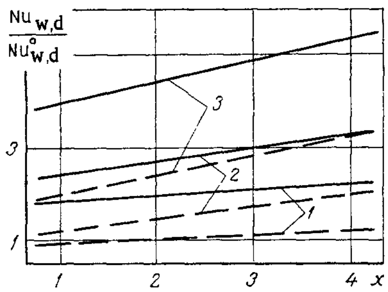

Fig. 3. Dependence of $Nu_{w,d}/Nu^0_{w,d}$ on pipe length x, m: 1) K = 1 kg/kg; 2) 3; 3) 7; solid curve) $Re_{w,d}$ = 3000; dashed curves) $Re_{w,d}$ = 15,000.

The absence of an influence of the initial particle velocity on the heat exchange indirectly confirms the concept [2] that two structural zones of flow are present in the thermal initial section of the pipe: a boundary layer of gas with an increased viscosity and the disperse core of the stream. With the concentration of the solid particles in the core of the stream a peculiar "cold" transport zone forms where there the further speed-up of the particles in the axial direction occurs. The results obtained in the tests also allow one to conclude that the formation of the indicated zones of the stream is the same at the entrance to the heated section in the investigated range of particle velocities.

It is seen from Figs. 1 and 2 that the character of the dependence of the heat exchange on the controlling factors undergoes some changes over the length of the pipe. Whereas for x = 1 m there is a minimum of the heat-exchange intensity in the vicinity of K = 1.2 kg/kg (Fig. 1), for x ≥ 2 m such a minimum is absent. But the presence of a section of a reduction in heat-exchange intensity with an increase in particle concentration for x = 1 m shows up out to x = 3 m for K < 1.8 kg/kg. In this region of concentrations for x = 2 and 3 m the curve of the dependence of $Nu_{w,d}/Re_{w,d}$ on K is flatter than that for K > 1.8 kg/kg.

The experimental data obtained are generalized by the empirical dependence

$$Nu_{w,d} = cRe^n_{w,d}(1 + 0.3K), \tag{6}$$

which for x = 1, 2, and 3 m is valid in the region of concentrations K = 1.5–8 kg/kg, while for x = 4 m it is valid for K = 0.17–8 kg/kg. The values of c and n for the investigated x are presented in Table 1.

Dividing (6) by (5), we obtain the dependence

$$\frac{Nu_{w,d}}{Nu^0_{w,d}} = c\,Re_{w,d}^{-0.16}(1+0.3K), \tag{7}$$

for the relative heat exchange, where the values of the multiplier c varies linearly with x (Table 1).

The dependence of the intensity of relative heat exchange on the pipe length is shown graphically in Fig. 3.

The results obtained were compared with the data of [2, 5], where a pipe 13 mm in diameter and the same particles of the solid phase of the stream were used. While the qualitative agreement was good, there were some quantitative differences, which came down to the following: The minimum of the heat-exchange intensity for x = 1 m is more weakly expressed and is shifted toward lower particle concentrations for a pipe of larger diameter (in [2, 5] it corresponded to K = 1.3 kg/kg). The levels of heat-exchange intensity obtained in the present work exceed the analogous data of [5] for x = 1 m by 20–24%, for x = 2 m by 6–7%, and for x = 3 m by 3–4%.

These disagreements can be partly explained by a difference in the conditions of creation of the thermal boundary layer at the entrance to the heated section. In contrast to [2, 5], where the start-up section was made of glass pipe, in the present work an earlier and more gradual creation of the boundary layer of viscous gas in the start-up section took place owing to the spread of heat along the pipe wall. This could affect the interaction of particles with the gas layer, lead to its thickening, and increase the heat-exchange intensity as a consequence.

But the influence of the pipe diameter, which shows up mainly in the initial section, evidently must be considered as the main reason for the disagreements obtained. With an increase in the pipe diameter the cross section of the "cold" core increases, and its porosity grows at the same values of K. The particles obtain a relatively large capacity for radial motions. This and the more considerable level of gas turbulence in the core lead to its earlier "smearing out," as a result of which the intensity of heat transfer increases. As the particles spread out over the entire pipe cross section beyond the thermal initial section [2] the influence of the pipe diameter on the heat exchange decreases.

The growth of heat-exchange intensity along the length of the pipe obtained in the tests (Fig. 3) allows one to judge the significant role of radial motions of particles in the mechanism of heat transfer from the wall, and it points to the peculiar flows of nonisothermal gas-suspension streams.

NOTATION

d	is the pipe diameter;
x	is the length of heated section;
x_{st}	is the length of start-up section;
$K = G_s/G$	is the flow-rate concentration of solid particles;
G_s and G	are the mass-flow rates of particles and gas;
t, T	are the temperature;
Δt_0, Δt_l	are the initial and logarithmic-mean temperature heads;
λ	is the coefficient of thermal conductivity;
μ	is the coefficient of dynamic viscosity;
Nu	is the Nusselt number;
Re	is the Reynolds number;
Pr	is the Prandtl number;
St	is the Stanton number.

Indices

w	is the parameter at wall temperature;
g	is the parameter at gas temperature;
0	is the parameter under conditions at the entrance.

LITERATURE CITED

1. Z. R. Gorbis and V. A. Kalender'yan, Heat Exchangers with Dispersed Flowthrough Heat-Transfer Agents [in Russian], Énergiya, Moscow (1975).
2. V. G. Kashirskii and Yu. Ya. Pechenegov, Inzh.-Fiz. Zh., 25, No. 1 (1973).
3. V. M. Legkii and A. S. Makarov, Inzh.-Fiz. Zh., 20, No. 2 (1971).
4. B. S. Petukhov, Heat Exchange and Resistance in Laminar Fluid Flow in Pipes [in Russian], Énergiya, Moscow (1967).
5. Yu. Ya. Pechenegov and V. G. Kashirskii, Izv. Vyssh. Uchebn. Zaved., Énerg., No. 9 (1974).

STABILITY OF OPERATION OF APPARATUS CONTAINING HEATED FLUIDIZED BEDS

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The influence of temperature on the conditions of disturbance of stability and the establishment of a self-oscillatory mode of fluidization is investigated.

It was shown earlier that the fluidization process becomes unstable under certain conditions in apparatus containing fluidized beds which are characterized by a "free" subgrid chamber of relatively large volume and a gas-distribution grid of low hydraulic resistance [1]. A disturbance of stability leads to "driving" of the system with the ultimate establishment of a self-oscillatory mode of fluidization. Such a mode was studied theoretically and experimentally in detail in [2, 3] under ordinary conditions, i.e., with fluidization by a cold gas.

Sometimes, especially in apparatus of large size containing large amounts of granular material, self-oscillations are an undesirable phenomenon, since they lead to the appearance of cyclic impact loads on the gas-distribution grid and other elements of the structure of the apparatus with their possible destruction. At the same time, self-oscillations promote better mixing of the granular material and intensification of various bulk processes carried out in a granular bed, and in a number of cases they can be considered as a natural and easily attainable means of improving the working characteristics of the apparatus. The investigation of the stability of the fluidization process and the parameters of self-oscillation cycles under conditions reflecting the operation of real apparatus therefore turns out to be very important in a practical respect.

First of all, the temperature dependence of the region of stability in the space of the parameters of the process is of undoubted interest. In fact, real apparatus usually operate at elevated temperatures, and some of the important parameters vary quite strongly with a change in temperature. The results of the experimental research in [2, 3], carried out on "cold" laboratory installations, obviously cannot be employed directly in an analysis of "hot" industrial apparatus; but qualitative conclusions about the influence of temperature on the stability of the fluidization process can be drawn on the basis of the general theory in [1], which remains valid as before.

For this it is sufficient to use the equation determining the boundary of the region of stability [1],

$$v_{1,2}=\frac{1}{2n}\left[\frac{1}{Nn}-1-n\right]\left\{1\pm\left[1-\frac{4}{N}\,\frac{1+n}{[(Nn)^{-1}-1-n]^2}\right]^{1/2}\right\}. \tag{1}$$

Here the following dimensionless parameters are introduced:

$$n=\frac{k_2}{k_1},\quad N=\frac{2\rho Sk_1}{\nu m},\ v=\nu ck_1V,\ \nu=\frac{H_0}{H^2}\,(Q_b-Q_0). \tag{2}$$

Only four quantities from (1), (2) vary significantly upon the heating of a bed: ρ, ν, k_1, and k_2; by properly allowing for their temperature dependence in Eqs. (1) and (2) it is easy to evaluate the influence of the latter

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