The data presented above was obtained under steady-state conditions. The process of heating bodies with an initial temperature below t_s can be divided into three stages. First the body is heated to t_s at almost the same rate as in pure vapor. The vapor flow moves to the surface and presses dry particles against it. These particles are moistened by condensate, forming a coat. The coat subsequently dries and is destroyed, and the temperature of the body remains nearly constant and equal to the saturation temperature. After complete destruction of the coat, the temperature of the body, as in a conventional bed, is increased further by the noncondensing gas. Thus, a 30-mm-diameter copper cylinder is heated in a bed of 0.12-mm corundum particles from 40 to 100°C in 10 sec. Then the cylinder temperature rises by a total of 1°C over the next 70 sec, after which it begins to exponentially approach the bed temperature, equal to 140°C. The heat-transfer coefficient in the third stage of heating is determined by Eq. (1).

NOTATION

d, diameter, mm; F, area, m^2; Q, heat flow, W; q, heat flux, W/m^2; r, heat of vaporization, J/kg; t, temperature, °C; v'', specific volume of the saturated vapor, m^3/kg; α, heat transfer coefficient, $W/(m^2 \cdot K)$; Ar, Archimedes number; Nu, Nusselt number; Pr, Prandtl number; Indices: co, coil; d, condensate; s, saturation; f, surface; ss, super-heated steam; fb, fluidized bed; c, calculated; t, coat.

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INTENSIFICATION OF HEAT EXCHANGE BETWEEN A FLUIDIZED BED AND CONTAINING WALLS WITH A NONUNIFORM GAS DISTRIBUTION

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The effect of a nonuniform initial gas distribution on heat exchange between a bed and containing walls is studied.

The problem of intensifying heat transfer between a fluidized bed and solids immersed in the bed is very familiar, as are several engineering methods of solving it. However, in certain processing operations it is important to increase heat transfer not to bodies submerged in the bed, but to the walls containing the bed. The rate of heat transfer with external surfaces is usually significantly lower than the rate of heat transfer with the surfaces of bodies located in the core of the bed [1, 2]. This has to do with important qualitative differences in physical patterns of flow of the bed over the respective surfaces [3]. Thus, in practice, increasing the coefficient of heat transfer from the walls of a processing unit to the fluidized bed within often requires the use of special techniques.

It was proposed in [4] that this be done by delivering the gas only to the region of the wall. This led to gushing near the wall and a rapid decrease in heat-transfer coefficient with an increase in filtration velocity, with the usual processing advantages of fluidization being lost. Similar (and more successful) attempts have been to intensify heat exchange with the walls for beds fluidized with droplets of liquid (e.g., see [5]). Heat transfer can also

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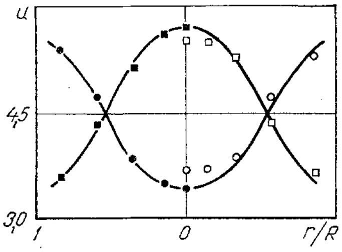

Fig. 1. Characteristic concave and convex air-velocity profiles at a height of 36 mm above the gas-distributing grate: dark and clear points) H = 0.5D and H = D, respectively; N = 4.0. u, m/sec.

be effectively increased by using special mixers [6] or by installing "baffle"-type guides at the wall [7], as well as by changing the shape of the process vessel itself [8]. However, such methods complicate the design of the apparatus and the manufacturing process and sometimes increase operating power costs.

The article proposes that heat exchange between a bed and vertical wall be improved by using a nonuniform gas distribution. Here, the bed as a whole is fluidized, but an excess gas flow is introduced at the wall. It follows from the theory in [9] and was confirmed by the experiments in [10, 11] that this increases both the velocity of the large-scale circulating motion and the mean relative velocity of the phases at the wall, the latter determining the intensity of the chaotic pseudoturbulent pulsations. What is evidently the most complete analysis of the effect of nonuniformity of the gas distribution on heat transfer with bodies of various shapes submerged in the core of a fluidized bed is presented in [10].

Our studies were conducted on a cylindrical unit with an inside diameter of 172 mm equipped with a sectional gas distributer. The design of the unit was similar to that used in [10, 11]. The main gas distributer consisted of a mass of lead shot separated into four annular sections and compressed on two sides by rigid grids. Particles of different sizes were used in the different sections, making it possible to change the fluid resistance within fairly broad ranges and so vary the velocity profile of the gas at the distributor outlet. Special attention was given to equalizing the gas velocity before the distributor. Attention was also paid to avoiding the adverse effects of "refraction" of the flow in the distributor, as well as the opposite effect of the bed, with a finite fluid resistance, on the effective characteristics of the flow at the distributer outlet. Gas velocity and pressure gradients at different sections of the bed were measured by methods developed earlier [10, 11].

The lateral wall of the apparatus was a cylindrical copper calorifier shell section with a Nichrome heater well insulated externally, keeping losses to no more than 10-20% of the total quanity of heat given off. The temperature of the surface of the heat exchanger was measured with eight Copper–Constantin thermocouples embedded in the apparatus at different heights over the distributor. The same type of thermocouples were used to measure temperature at different points inside the bed.

We used glass spheres 1.5-2.0 mm in size and 2.5-3.0 and 3.5-4.0 mm spherical particles of aluminosilicate catalyst fluidized with air. The minimum fluidizing velocity with a uniform gas distribution was 0.90, 0.79, and 0.92 m/sec, respectively. The static height of the bed H was changed within the range from 0.25D to 2D in increments of 0.25D. The tests were conducted with three types of initial gas-velocity profiles: plane, convex, and concave, corresponding to introduction of the excess volumes of gas into the core of the bed and in the region near the wall. Such profiles were described in [11], and a theory permitting evaluation of their change with increasing distance from the gas distributor was presented in [9]. The convex and concave profiles used in most of the tests are illustrated by the curves in Fig. 1. It is apparent that an increase in the static height of the bed (and hence in its fluid resistance) has almost no effect on the shape of these profiles.

As might be expected, with a convex initial gas-velocity distribution, the coefficient of heat transfer from the wall turns out to be quite a bit less than that corresponding to a uniform initial distribution and the same total gas consumption throughout the investigated range of number of fluidizations N. The best results in the sense of intensifying heat trans-

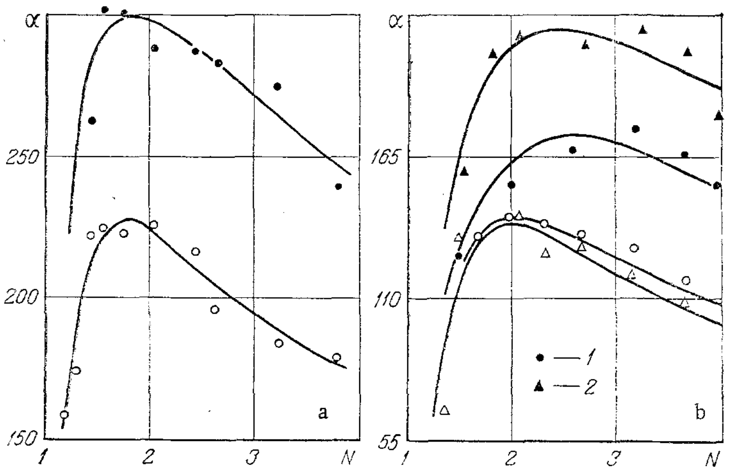

Fig. 2. Dependence of coefficient of heat transfer on number of fluidizations: a) glass spheres, d = 1.5-2.0 mm, H = 0.75D; b) particles of aluminosilicate catalyst, d = 3.5-4.0 mm, H = 1.5D (1) and H = 1.75D (2); dark and clear points — concave and plane initial air-velocity profiles, respectively, α, W/m^2•sec.

fer from the wall were obtained with the concave initial distribution, when the gas velocity at the wall was ~1.5 times as great as the velocity in the core of the bed (Fig. 1); these results are discussed below.

The relative increase in the coefficient of heat transfer from the wall compared to the bed with a uniform gas distribution, regarded as a function of the height of the bed, has a distinct maximum. For particles of diameter 1.5-2.0, 2.5-3.0, and 3.5-4.0 mm, this maximum is reached at H/D equal to 0.75, 1-1.25, and 1.5-1.75, respectively. Partial "breakthrough" of the gas and the formation of a fountain is seen with lower heights, and the resulting drop in α is similar to that noted in [4]. The effect of intensified heat transfer degenerates for relatively high beds, since the effect of the initial distribution decreases and ultimately disappears with increasing distance from the grate, and the velocity profile approaches a convex limiting profile independent of the initial profile [9, 11].

Figures 2 and 3 show characteristic dependences of α on the number of fluidizations N. It is apparent that, apart from a very substantial increase in α with the transition to a nonuniform gas distribution of the type being examined, there is a certain expansion of the range of N in which α is roughly equal to its maximum value. Also, Fig. 3 compares the effect of intensified heat transfer from the wall with the similar effect for bodies immersed in the bed. Along with the data shown in Figs. 2 and 3 for these cases, we have included the results of tests by M. B. Kliot [10] involving heat transfer from a bundle of tubes located in the central part of the bed (Fig. 3a) and from a coil located symmetrically relative to the axis of the apparatus (Fig. 3b). These tests were conducted on similar units; maximum heat transfer was seen in them when the maximum of the initial gas velocity was located under the immersed body, i.e., with the convex initial profile shown in Fig. 1 for the tube bundle and having a maximum at r/R ≈ 0.5 (i.e., under the coiled tube) for the curves in Fig. 3b.

Analysis of the relations in Fig. 3 permits the following two conclusions. First, the effect of an increase in heat transfer from the wall in the transition to a nonuniform gas distribution is considerably greater than the corresponding effect for submerged bodies. Second, the difference in the mean coefficients of heat transfer from the external (wall) and internal (body) surfaces in the bed disappears with appropriate gas distributions. This is because of the fact, which follows directly from the physical analysis in [3], that, with the delivery of excess volumes of gas into the wall region, the gas flow over the wall should be of exactly the same nature as flow over the surface of a vertical plate immersed in the bed. In particular, the mass of slowly moving particles adjacent to the wall, which normally "shields" the wall from the core of the bed in beds with a uniform gas distribution, is destroyed.

Figure 4 shows the dependence of the ratio of the coefficient of heat transfer from the wall with a concave initial gas-velocity profile (Fig. 2) to the same coefficient with a

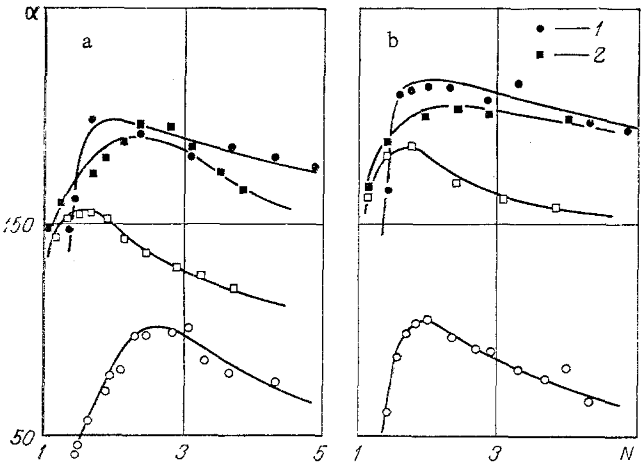

Fig. 3. Dependence of α on N for wall of apparatus (1) and submerged bodies (2) in bed of particles of aluminosilicate catalyst, d = 2.5-3.0 mm, with plane and nonuniform (concave for the wall, convex for the submerged bodies) initial velocity profiles (clear and dark points, respectively); a) H = 0.75D; 1) data in [10] for tube bundle; b) H = D; 2) data from [10] for coil.

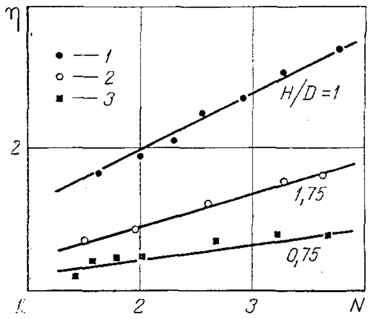

Fig. 4. Ratio η of coefficients of heat transfer from wall of apparatus with concave and plane initial velocity profiles as a function of the number of fluidizations N; 1, 2) particles of aluminosilicate catalyst, d = 2.5-3.0 and 3.5-4.0 mm, respectively; 3) glass spheres, d = 1.5-2.0 mm.

plane initial profile on the number of fluidizations. It can be seen that η is an increasing function N which is nearly linear in the investigated range of N. Being a function of particle size, η has a distinct maximum at the values of d at which α is minimal. To a certain degree, this is due to the fact that the acceleration of the phases near the wall with a nonuniform gas distribution is very slightly dependent on particle size, and the increase in heat-transfer coefficient is determined mainly by the distribution of the mean velocities of the phases.

NOTATION

D, diameter of apparatus; d, particle diameter; H, static height of bed; N, number of fluidizations; r/R, relative radial coordinate; u, vertical component of gas velocity; α, heat-transfer coefficient; η, ratio of coefficients of heat transfer corresponding to concave and plane initial velocity profiles.

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HETEROGENEOUS EQUILIBRIUM BETWEEN THE CONDENSED PHASE AND VAPOR OF APROTONIC SOLVENTS AND ELECTROLYTES BASED ON THEM

II. SOLVENT MIXTURES

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Experimental results for the saturated vapor pressure of mixtures of organic solvents with electrolytes based on them are given, together with thermodynamic characteristics of the vaporization process calculated on this basis.

Earlier, in [1], the factors stimulating the study of the equilibrium between condensed phase and vapor in mixtures of organic solvents and electrolytes based on them were formulated. Using the methods of sample preparation and of experiment described in [1], the saturated vapor pressure was measured for two-component mixtures of tetrahydrofuran (THF) with γ-butyrolactone (γ-BL) and propylene carbonate (PC) of two compositions (by volume) — 1:1 and 3:7 — and also of electrolytes based on them. As before, lithium perchlorate ($LiClO_4$) was used as the ionogenic component of the electrolyte.

As an example, in Fig. 1, values of the saturated vapor pressure of the given solvent mixture and, for comparison, data for the individual solvents of which they are composed are given.

The primary thing to notice is that the experimental data for each mixture, despite the significant difference in boiling point of the components (338.85, 474.45, and 509.35°K [1]

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