HEAT AND MASS TRANSFER THROUGH A HORIZONTAL GRID GRATING IN A FLUIDIZED BED

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Apparatus with a limited fluidized bed are the most promising for application in many branches of engineering. In particular, sectioning a fluidized bed by horizontal baffles in the form of one or several grid gratings for particles permits a substantial reduction in the axial mixing of the disperse material, obtaining a temperature profile nonisothermal along the height, and in the long run, assuring realization of new technological processes in principle: fuel ignition with simultaneous deep utilization of heat, nonstationary catalysis, thermal and chemical treatment of disperse material in the counterflow phase. The heat and mass transfer through a horizontal grid grating is the governing scientific problem for modeling the mentioned processes.

This question is examined slightly in the literature. Thus, there is no theoretical research at all. Available experimental data obtained in apparatus of 0.06-0.40 m diameter, sectioned by gratings with up to 30% active section, concern mainly questions of the transmissivity of the sectioning baffles.

The grating transmissivity was investigated under conditions when there was no fluidized bed beneath them. The experiments were performed for cases of both single holes in the baffles and perforated gratings without [1, 2] and with [3, 4] an opposing gas flow.

In principle, the deduction of a presence of dynamic crests above the holes from which particles emerge is important in these works, and the emergence governs the mass flow rate of the disperse material through the holes. The most substantial result is that the mass flow rate of the particles getting through the holes diminishes as the linear velocity of the ascending gas flow increases. The flow rate in the holes, for which particle efflux ceases completely, considerably exceeded the twisting flow for the case of perforated gratings [3].

Up to now no single point of view has been asserted with respect to the dependence of particle circulation through the downcomer grating on the fluidization velocity. Thus, Martyushin and Golovin [5] considered the mass flow rate of particle circulation between adjacent sections of equal capacity of the downcomer grating in the absence of a fluidized bed beneath it. They obtained that the grating capacity (circulation intensity) decreased as the fluidization velocity increases. It is shown in [6] on the basis of measurements of the mixing intensity of a magnetic flare between adjacent sections of a fluidized sand bed separated by a downcomer grating that an increase in the fluidization velocity resulted, on the contrary, in an increase in the particle circulation intensity. And, finally, the existence of two domains of mixing intensity variation is detected [7] as the velocity of the cooling gas increases: The particle mixing intensity first grew to reach a maximum and then diminished. There is no other material that would supplement these contradictory results, which does not permit setting up a true mechanism for solid phase circulation through sectioned baffles in a fluidized bed.

The heat transfer in a fluidized bed sectioned by downcomer gratings has been studied to a still lesser extent. Results on the effective heat conductivity of a fluidized bed in the zone of wire downcomer gratings with more than 50% active section are obtained in a catalytic heat generator in [8]. The data of the experiment extended the dependence which is purely empirical in nature and cannot be used in the case of single-layer baffles.

An investigation of a multisection furnace-heat transfer apparatus is performed in [9] on the basis of mathematical modeling of the longitudinal heat transfer process. The longitudinal heat transfer intensity through reticulated gratings is measured experimentally here

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TABLE 1

Material	d_p, mm	ρ_p, kg/m^3	c_p, kJ/(kg·deg)	w_0, m/sec	w_b, m/sec
Fluvial sand	0,35	2500	0,80	0,10	2,97
» »	0,58	2560	0,80	0,22	4,68
» »	0.97	2550	0,80	0,45	6,69
Aluminum oxide	0,59	1250	1,05	0,12	3,08
» »	1,27	1250	1,05	0,37	5,38

to confirm the deductions that are qualitative in nature. However, the results on [9] do not clarify completely the general regularities of the heat transfer process and cannot be generalized in the form of computational dependences.

More detailed investigations of the features of fluidization and heat transfer in a sectioned bed were performed in [10, 11] on a "cold" test stand by using reticulated gratings with a 40 and 60% active section. On the basis of the experimental data on the influence of the thermophysical, regime, and geometric parameters of the sectioned system on the heat transfer intensity, a number of important deductions of physical nature are made, that are in principle of value for the construction of a heat transfer process model.

A complex investigation of the heat and mass transfer processes through a single section horizontal downcomer grating in a fluidized bed is performed in this paper and a physical mechanism of the processes is set up on whose basis relationships are obtained for computation of the circulation intensity of disperse material and the heat transfer between sections.

1. DESCRIPTION OF THE EXPERIMENT METHODOLOGY

The experimental investigations were performed on a "cold" test stand in an apparatus with a 0.2 × 0.4 m section whose construction is described in [10]. Grids with 2.1 and 3.1 mm cell dimensions with active section fractions $\varphi = 0.4$ and 0.6 were used as downcomer gratings. Air was the liquefying agent. The characteristics of the disperse material are represented in Table 1.

The effective thermal resistance of the bed in the zone of the sectioned grating R taken as characteristic of the heat transfer intensity was investigated in the stationary thermal regime by the method in [11]. A heat source (electrical heater) was placed in the upper section of the bed and from which the heat was transmitted to the lower section and expended in heating the air arriving for fluidization. There was a temperature difference of ΔT between the sections, from which the effective thermal resistance was indeed determined.

The intensity of particle circulation through the grating was measured by the following methodology: A definite quantity of tagged particles, different from the majority just by color, was introduced into the upper section of the bed. Because of the high mixing intensity in the free bed, the tagged particle concentration in the bulk of the section was equilibrated practically instantaneously and then diminished in time relatively slowly because of the circulation of the disperse material. On the other hand, the tagged particle concentration in the lower section grew, remaining uniform over the whole section bulk. Under these conditions the circulation flux density of the disperse material through the grating j was calculated according to the change in tagged particle concentration in the sections on the time.

Indeed, considering that the disperse material in the sections is mixed ideally and instantaneously, we can write

$$M_1 d\mu_1 = F(\mu_2 - \mu_1) j d\tau, \quad (1)$$

where M_1 is the mass of the disperse material in the lower section, μ_1, μ_2 are the tagged particle mass concentrations in the lower and upper sections, F is the transverse section area of the apparatus and τ is the time. Expressing μ_2 in terms of μ_1 and integrating (1), we obtain

$$j = \frac{M_1 M_2}{F(M_1 + M_2)\tau} \ln \frac{\mu_{20} M_2 + \mu_{10} M_1 - \mu_1 (M_1 + M_2)}{(\mu_{20} - \mu_{10}) M_2} \quad (2)$$

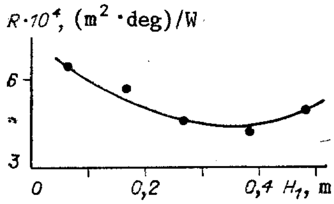

Fig. 1

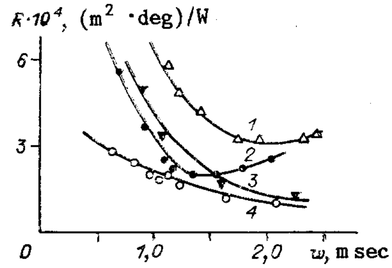

Fig. 2

(M_2 is the mass of the disperse material in the upper section, and μ_{10}, μ_{20} are the initial tagged particle mass concentrations in the lower and upper sections).

The quantity μ_1 was determined directly in disperse material samples selected from the middle of the lower section while M_1, M_2 were determined as the product of the volume of the fluidized bed by its density measured according to the pressure drops in each section. Moreover, the transfer flux density was calculated by means of the measured effective thermal resistance on the basis of the relationship

$$j_R = 1/(Rc_p), \tag{3}$$

which is valid if the heat transfer between opposite particle circulation fluxes is missing in the zone of the sectioning grating (c_p is the particle specific heat).

The particle efflux density through the grating in the absence of a fluidized bed in the lower section j_c was found by direct measurement of the quantity of disperse material ΔM passing through the grating in the time $\Delta\tau$ during which the height of the bed above the grating varied insignificantly

$$j_c = \Delta M/(\Delta\tau F). \tag{4}$$

2. RESULTS OF THE EXPERIMENTAL INVESTIGATIONS

As experiments showed, the introduction of horizontal downcomer gratings in a bed results in a substantial change in the fluidization pattern. A sectioned bed remains continuous only in the domain of liquefying gas velocities close to the minimal fluidization velocity. As the velocity increases a rarefied zone occurs under the baffle (a gas interlayer) whose presence has been noted repeatedly in the literature [12]. The free boundary of the rarefied zone fluctuates constantly during fluidization so that its height does not remain identical. An ejection of particle mass is observed at the site of gas bubble emergence into the rarefied zone, whereupon the fluidized bed is joined to the grating from below at this site. Simultaneously, at those sites over the section where the rarefied zone is conserved at this time, particle efflux occurs from the upper section in the form of individual jets, according to visual observations. As is shown below, the rarefied zone plays a substantial part in the process of particle mixing between adjacent sections.

It can be noted on the basis of the experiments performed that the transverse dimension of the apparatus, exactly as the height of the fluidized bed above the downcomer grating, exerts no effect on the heat transfer between sections. At the same time the height of the location of a sectioning grating above the gas distributing grating H_1 influences the effective thermal resistance, where the dependence of R on H_1 is complex enough in nature (Fig. 1). The presence of a minimum in the dependence is associated with the change in the fluidization regime as H_1 increases from nucleate to piston (the data in Fig. 1 correspond to experiments with sand particles: $d_p = 0.58$ mm, velocity of fluidization $w = 0.8$ m/sec, $\varphi = 0.4$). The features of piston formation in a sectioned system and the particle mixing mechanism in this regime are described in greater detail in [11]. All the experimental results in this paper are obtained in the nucleate regime.

Figure 2 demonstrates the influence of the fluidization velocity on the effective thermal resistance for different disperse materials and kinds of downcomer gratings and the dependence of R on the particle parameters (the equivalent diameter d_p, the density ρ_p, and specific heat c_p) and the grating characteristics (the cell size and fraction of the active section φ): line 1 is for aluminum oxide particles with $d_p = 1.27$ mm, $\varphi = 0.4$; 2 is for the

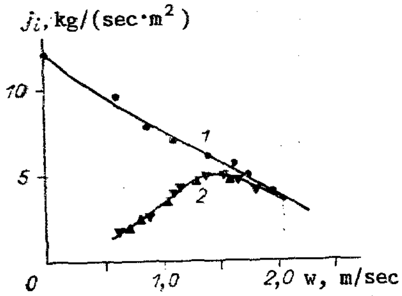

Fig. 3

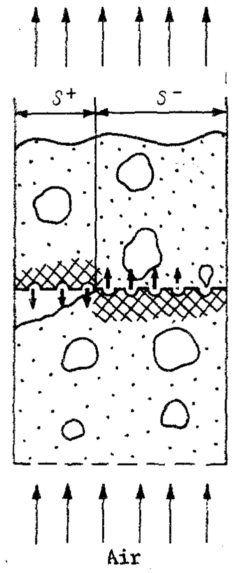

Fig. 4

same with d_p = 0.59 mm, φ = 0.4; 3 is for the same with d_p = 1.27 mm, φ = 0.6; and line 4 is for sand particles with d_p = 0.58 mm, Φ = 0.6. As the particle diameter increases and other conditions remain the same R increases (lines 1 and 2) while as the fraction of the active section and the grating cell dimensions increase the thermal resistance, on the other hand, diminishes (1, 3). It is seen that as the fluidization velocity increases in a number of cases (lines 1 and 2), the thermal resistance in the baffle zone first diminishes, reaches minimal values, and later grows. As experiments show, the minimal R are observed here in the domain of such fluidization velocities for which the mean gas flux velocity over the section in the grating holes corresponds approximately to the particle absorption velocity. In those cases when the critical gas flux velocities are not achieved (lines 3 and 4), the dependence R = f(w) is monotonically decreasing.

Experimental dependence of the capacity and intensity of disperse material circulation through the downcomer grating on the fluidization velocity are represented in Fig. 3. All the data are obtained for aluminum oxide particles with d_p = 0.59 mm and a grating with Φ = 0.4, H_1 = 0.265 m.

The particle efflux density through the grating f_c (capacity) was determined by means of relationship (4) (line 1). As is seen from the graph, the values of f_c diminish in the absence of disperse material in the lower section as the fluidization velocity increases in the whole range of variation of w. The result is in agreement with available literature ([5], say).

The particle circulation flux density through a downcomer grating j in the presence of a fluidized bed in both sections of the apparatus (line 2) was computed by means of relationships (2) and (3). The agreement between the curves j(w) and $j_R(w)$ is quite important for clarification of the heat transfer mechanism since it proves that heat transfer between opposing particle fluxes circulating through the grating is missing. As experiments show, the dependence of the particle circulation intensity on the fluidization velocity has a maximum. Qualitatively this result agrees with the data from [7]. An increase in the particle circulation intensity occurs in the range of velocities w not exceeding the velocity corresponding to the particle absorption velocity in the grating holes. It is interesting to note that exchange by particles between sections does not cease during the passage through this border but only starts to attenuate.

3. HEAT AND MASS TRANSFER MODEL AND GENERALIZATION OF THE RESULTS

The main statements of the physical model of heat and mass transfer processes through a horizontal downcomer grating in a fluidized bed can be formulated as follows according to [11] and the present paper.

1. At each time particle transport from the lower to the upper section and back is realized through different grating holes: through holes above the rarefied zone in the lower section and through the compressed interlayer of disperse material overlapped from below in the upper section. This is verified by visual observation as well as by the agreement between the quantities j and j_R (see Fig. 3).

2. Since the descending disperse material particle flux is spilled into the rarefied zone in the lower section in the presence of a fluidized bed, under identical fluidization velocities the particle flux densities in each individual hole will be identical with or without the presence of the bed in the lower section.

3. The magnitude of the transfer flux density by the particles, referred to the whole section of the apparatus, depends on the relationship between the number of holes operating at a given time during passage of the particles downward and upward. As follows from a comparison of lines 1 and 2 in Fig. 3, for low fluidization velocities the fraction of holes through which the particles move downward is here small; their fraction grows and approaches one as the velocity w increases (the descending branch of lines 2 and 1 in Fig. 3 are practically coincident).

4. Because the thermal interaction between opposing particle circulation fluxes is absent (this follows from the agreement between j and j_R), the heat transfer through a single downcomer grating in a fluidized bed is determined completely by the particle circulation between the sections and is described by the relationship $q = jc_p\Delta T$.

The particle circulation pattern through a downcomer grating can be represented schematically as is shown in Fig. 4. Letting j^+ and j^- denote the mass flow rates of the particles arriving per unit hole area in the descending and rising fluxes, and S^+ and S^- the fractions of holes through which the disperse material passes from the upper to the lower section and from the lower to the upper sections, we write $j = j^+S^+\varphi = j^-S^-\varphi = j^-(1 - S^+)\varphi$, from which

$$j = \frac{j^+j^-}{j^+ + j^-}\varphi. \tag{5}$$

In conformity with the data of Fig. 3, we extract three ranges of variation of the particle circulation intensity for which on the basis of (5)

$$\begin{gathered} w < w_{\max},\ j^+ \gg j^- \to j \approx j^-\varphi; \\ w \approx w_{\max},\ j^+ \approx j^- \to j \approx j^+\varphi/2 \approx j^-\varphi/2; \\ w > w_{\max},\ j^+ \ll j^- \to j \approx j^+\varphi \end{gathered} \tag{6}$$

(w_{max} is the velocity of the liquefying gas corresponding to the maximal circulation intensity). As already noted, we can take $w_{max} \approx \varphi w_b$.

We select the first range for examination since it corresponds to the most widespread gas velocities in fluidization practice. It follows from (6) that the problem of describing particle circulation through a downcomer grating for fluidization velocities $w < w_{max}$ reduces to the solution of a problem on particle transport through a hole by an ascending gas flux from a compact sublayer of fixed particles adjacent to the grating holes from below.

Let us assume that just as has been established for the case of disperse material efflux from a hole subjected to gravitational forces [1-4], the compact particle sublayer under the grating is separated from the holes by dynamic crests during particle motion through a hole from the bottom up under the action of friction forces, as is also displayed schematically in Fig. 4. The stage of particle emergence from the crest will here also be the limiting stage of the transfer process by analogy.

Let us represent the balance of forces acting on a particle moving uniformly in a crest in the form

$$P_f - P_e - P_g = 0 \tag{7}$$

(P_f, P_e, P_g are the friction force between the gas and the particle, between the particle and the surrounding disperse material, and the gravity force). According to the Darcy law

$$P_f = k_0\nu_g\rho_g w_g d_p, \tag{8}$$

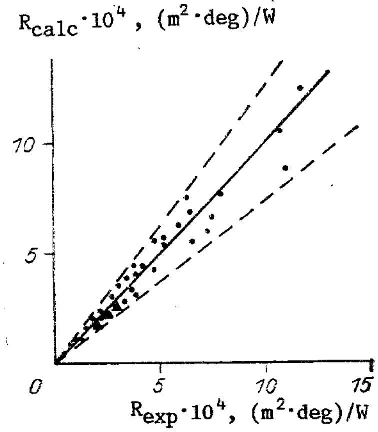

Fig. 5

where k_0 is a dimensionless coefficient; ν_g is the coefficient of kinematic viscosity of the gas, ρ_g is the density of the surrounding gas, and w_g is the gas velocity at the crest surface. We analogously express the gravity force in terms of the minimal fluidization velocity w_0

$$P_g = k_0 \nu_g \rho_g w_0 d_p. \tag{9}$$

Analysis of the experimental data shows that

$$P_e = k^* \rho_p u_p \tag{10}$$

(k^* is the dimensional drag coefficient m^3/sec).

Considering the dynamic crest a hemisphere with diameter equal to the hole diameter, we write for the gas velocity on the crest surface

$$w_g = w/(2\varphi). \tag{11}$$

Taking account of (7)-(11)

$$u_p = \frac{k_0 \nu_g \rho_g d_p (w - 2\varphi w_0)}{k^* \rho_p \varphi}. \tag{12}$$

Upon emergence from the crest the particle traverses a distance on the order of d_p in a time d_p at a velocity u_p, i.e., $\tau \sim d_p/u_p$. The frequency of particle emergence from the crest area on the order of d_p^2 equals $1/\tau \sim u_p/d_p$, and from unit crest surface $1/(\tau d_p^2) \sim u_p/d_p^3$.

The quantity of particles proceeding into the hole per unit time from the whole crest is $\omega_p \sim u_p d_{oh}^2/d_p^3$ (d_{oh} is the equivalent diameter of the grating hole). Then the particle mass flow rate into the hole (we consider the particle spherical) per unit area is $j^- \sim d_p^3 \rho_p \omega_p / d_{oh}^3 \sim \rho_p u_p$ from which we obtain, by taking account of (6) and (12), a relationship for the particle mass flow rate (circulation flux density) referred to unit area of the apparatus section, in the form

$$j^* = k \nu_g \rho_g d_p (w - 2\varphi w_0),\ k \sim 1/k^*. \tag{13}$$

Besides the conditions governing particle emergence from the fixed sublayer through the dynamic crests, a number of factors still affect the intensity of disperse material efflux from the lower to the upper section in the general case. This is a blocking effect characterized by the ratio d_{oh}/d_p as well as the volume and velocity of the disperse material being ejected by the bubbles under the grating, whose influence can be reflected by the introduction of the Reynolds criterion constructed according to the diameter d_b and the rate of rise w_b of the bubble and the Froude criterion constructed according to the particle diameter and characterizing relationship of the inertial and gravity forces into (13) ($Re_b = d_b w_b/\nu_g$; $Fr = (w - w_0)^2/(g d_p)$, and g is the acceleration of gravity). Taking the above into account

$$j = j^* (d_{oh}/d_p)^a \mathrm{Re}_b^b \mathrm{Fr}^c. \tag{14}$$

The values of d_b and w_b can be found from dependences available in the literature. For instance, according to [13] the gas bubble diameter emerging into the rarefied zone under the downcomer baffle is $d_b = 0.54(w - w_0)^{0.4}(H_1 + 4\sqrt{A_0})^{0.58}g^{-0.2}$, where A_0 is the "entrapment" area per bubble while the rate of bubble rise is $w_b = 0.711\sqrt{gd_b}$. Substituting these dependences into (14), we calculate the values of the constants $a = 2.0$, $b = 0.3$, $c = -0.09$, $k = 18\cdot10^5$ sec/m^3 on the basis of the experimental data obtained.

Finally, the equation to compute the circulation flux density of the disperse material between adjacent sections is written in the form

$$j = 18\cdot10^5 \nu_g \rho_g d_p (w - 2\varphi w_0) \times (d_{oh}/d_p)^{2,0}\, \mathrm{Re}_b^{0,3} \mathrm{Fr}^{-0,09}.$$

We represent the effective thermal resistance of the fluidized bed in the zone of the downcomer baffle on the basis of (3) as

$$R = 1/\left(18\cdot10^5 \nu_g \rho_g d_p c_p (w - 2\varphi w_0)\, (d_{oh}/d_p)^{2,0}\, \mathrm{Re}_b^{0,3} \mathrm{Fr}^{-0,09}\right). \quad (15)$$

The results of experiments obtained by us (points) and in [8] during fuel combustion in a sectioned apparatus (triangles) with the results of a thermal resistance computation by using (15) under corresponding conditions are compared in Fig. 5. The written relation assures a computation accuracy not below 25% (the dashed lines) in a broad range of variation of the thermophysical properties of the liquefying gas and disperse material, the fluidization velocities and also the geometric parameters of the sectioned system. Moreover, the data of [8] corresponding to measurements on a "hot" test stand are processed well by using (15).

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FREE CONVECTION HEAT TRANSFER IN AN OPEN SYSTEM OF VERTICAL RODS

V. I. Eliseev and Yu. P. Sovit

UDC 536.25

Investigations of free convection heat transfer on vertical surfaces are a well-developed section of the theory of natural convection flows. Extensive handbook and bibliographic material on this topic can be found in [1]. However, the problems of developing effective heat transfer apparatus, the necessity to compute the temperature regimes of complex rod systems possessing heat liberation, the selection of effective methods of protecting packets of electrical cables from overheating determine the urgency of formulating and solving problems on the hydrodynamics and heat transfer of different sets of rods. An effective model that permits reflection of the hydrodynamic and thermal interaction of rods between themselves and the bundle as a whole with the environment is a filtration flow model. It is used extensively at this time for heat transfer computations under forced convection in anisotropic rod structures [2-5]. Critical relationships obtained on the basis of processing experimental data are used here to determine the thermal and hydrodynamic forces of solid and liquid phase interaction per unit volume of a porous body. Considerably less attention is paid to questions of mathematical modeling of free convection heat transfer in such media. Existing researches are mainly experimental in nature [6-9]. Consequently, application of the filtration flow model in a porous medium to the description of free convection processes in rod bundles and execution of numerical computations of the heat transfer of rod collections with an external cooling medium are of great interest. Meanwhile the lack of critical dependences for bulk friction and heat liberation in such a flow specifies the urgency of the problem of a theoretical determination of the desired quantities. Solution of these problems is indeed the purpose of this paper.

1. FUNDAMENTAL EQUATIONS AND BOUNDARY CONDITIONS

Let us consider the axisymmetric free convective incompressible fluid flow in a vertical bundle of rods. We assume that the flow mode is laminar and the viscosity, heat conductor, and specific heat are independent of the temperature. We direct the x axis along the longitudinal axis of the bundle, then the r axis will lie in a plane perpendicular to the x axis while the angle φ is measured from a certain initial position of the xor plane. Let us extract a small space element $\Delta V = r\Delta\varphi\Delta x\Delta r$, containing a sufficiently large quantity of rods in addition to fluid (Fig. 1). In the presence of the rods the space configuration is characterized by the quantities

$$\varepsilon = \Delta V_f/\Delta V,\quad \varepsilon_x = \Delta S_{fx}/\Delta S_x,\quad \varepsilon_r = \Delta S_{fr}/\Delta S_r,\quad \varepsilon_\varphi = \Delta S_{f\varphi}/\Delta S_\varphi,$$

where ΔV_f is the volume of space occupied by the fluid, ΔS_j is the area of a side of the element with normal along the appropriate axis, and ΔS_{fj} is the area of the flow-through part of the appropriate side of the element. The flow field is determined by the velocity vector $\mathbf{V} = \mathbf{i}u + \mathbf{j}v$ as well as by mass and volume force vectors acting on the extracted element. Taking into account the axisymmetry of the motion and using the standard procedure for deriving the conservation equations for a continuous medium [10], we have

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