ON THE THEORY OF EXTERNAL HEAT TRANSFER IN A FLUIDIZATION BED

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The fundamental equations of the "packet" theory are transformed and critically analyzed, whereupon the concepts pertaining to the mechanism of external heat transfer in a fluidization bed are further refined on the basis of new test data.

It is well known that the external heat transfer in a fluidization bed with a surface immersed in it represents a complex process and, for this reason, no explanation has been found yet to many of its aspects. This applies especially to the basic models of the mechanism and to the structural–hydrodynamic characteristics of the boundary layer which directly determine the quantitative and the qualitative indicators of that process.

In view of this, the authors have made experimental studies [1-4] aimed at refining the model of external heat transfer and at determining the structural–hydrodynamic characteristics of a fluidization bed with immersed bodies of various shapes.

These studies involved a plate, a sphere, a cylinder, a wedge, and bodies of arbitrary shapes and of various sizes, with the fluidization number W varying from 1 to 10 in a bed of particles of the 0.32-2.20 mm size range and 1050-2430 kg/m^3 density. The bed was fluidized with air at a temperature of 15-20°C.

In order to ensure an adequate reliability of test data, the bed porosity was determined by three independent methods: by illumination with a narrow beam of x-rays, by x-radiogram photometry, and by high-speed photography. The air velocity and the fluctuation frequency in the stream were measured with a slot-type pneumometric tube.

Over 2500 tests have established that the mean porosity and the air velocity in the boundary layer of a fluidization bed are, respectively, 1.1-1.4 and 1.2-2.0 times higher than those mean over the entire bed section. Furthermore, the porosity of the boundary layer in a fluidization system was found to be $\varepsilon_b \geq 0.70$-0.85 and the structural–hydrodynamic characteristics to be nonuniform across the immersed surface.

For a vertical plate

$$\frac{1-\varepsilon_b}{1-\varepsilon_0} = 2.4W^{-0.347}\mathrm{Ar}^{-0.077}(y/l)^{0.830} \tag{1}$$

and

$$\frac{w_b}{w_f} = 4.1W^{-0.23}\mathrm{Ar}^{-0.10}(y/l)^{-0.47}. \tag{2}$$

These relations are valid within the range $1.2 \leq W \leq 8.0$ and $2780 \leq \mathrm{Ar} \leq 406{,}000$.

These studies have shown that, along with packets and air bubbles, thin gaseous interlayers of nonuniform thickness form along a surface immersed in a fluidization bed. Their thickness fluctuates from 0 to 5d, the statistical mean thickness being by one order of magnitude larger than the effective gap width between the first row of particles and the heat-transfer surface. For particles of the 0.32-2.20 mm size

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this thickness was found to be 0.06-0.18 mm and could be accurately enough approximated by the relation $\delta_m = (0.1 + \sqrt{d}) \cdot 10^{-4}$ m (d measured in mm). It has also been found that the boundary layer exists not in a two-phase but in a three-phase state: as a continuous phase, a discrete phase, and as a dilute phase; also that the dilute phase, which is formed during the periodic breakup of air jets at the surface as well as in the hydrodynamic trails behind air bubbles, strongly affects the heat-transfer rate.

According to the temperature and velocity fluctuation simultaneously plotted on the oscillogram and according to the phase alternations at the surface of the thermoanemometer probe (as shown on the high-speed photograph frames), this effect is manifested by a sudden drop in the foil temperature every time the probe finds itself in the trail of an air bubble or in a large-scale vortex or in a local disperse stream, the heat-transfer coefficient being maximum ($\alpha = \alpha_{max} = \alpha_{conv}$) exactly then and not during a contact with packets of solid particles [5, 6].

During a weak contact between the probe and particle packets moving along the immersed surface (behind the dilute phase), the change in the foil temperature is already insignificant. This corresponds to an insignificant change in velocity during the time that packets remain exposed at the probe surface. An appreciable heating of particles under such conditions is hardly possible and, as during contact between the immersed surface and the dilute phase, the heat transfer during this period is determined by convection.

The relation between temperature fluctuations and velocity variations in a flowing boundary layer confirms our earlier conclusion [3] that thermoanemometer ~~probes in a f~~luidization bed record not only phase alternations but also velocity fluctuations. All three phenomena are here, undoubtedly, interrelated and represent the effects of the same cause: the nonhomogeneity of a fluidization system.

The results obtained here are evidence that the actual mechanism of external heat transfer in a fluidization system is rather complex and nonuniform across the heat-exchanger surface.

Apparently, at the lower part of the surface, where air streams break up and eddies form most often while the local porosity approaches unity, heat is transmitted only convectively by air and by loose particles of fine-grain material. From the upper part of the surface, on the other hand, heat is transmitted mainly by conduction through the layer ascending along it, inasmuch as this occurs here in a gravity-flowing aerated layer without fluidization. The heat transfer is most complex at the middle of a vertical surface. Here the porosity is always $\varepsilon_b > 0.7$ and the dominant role in the process is undoubtedly played by the convection of air and particles, but conduction is effective to some extent. Such a high rate of heat transfer within this zone of a developed fluidization bed is explainable, in the light of the author's model [7], by an intrusion of fluidized material into the laminar air layer, by a breakup of air jets along the surface, and by distortions of the temperature field.

The extraordinary complexity of the mechanism of external heat transfer in a fluidization bed casts some doubt on the conclusions drawn by the authors of [5] and [8] concerning the adequacy of simplified models in the "packet" theory, as weighed against test data. Skepticism is obviously justified here, if one examines the analytical solution in [8]:

$$\alpha_c = 2\,\frac{1-f_0}{\sqrt{\tau_c}}\sqrt{\frac{\lambda_S\, c_M \rho_S}{\pi}}, \tag{3}$$

according to which the average over the period $(\tau_c + \tau_0)$ heat-transfer coefficient α_c should strongly depend on the time of contact between packets and the surface (τ_c) and on what fraction $(1-f_0)$ of its fluctuation period this contact time constitutes [5, 9].

The values of parameters τ_c and f_0 in Eq. (3) are usually found experimentally on the basis of thermoanemometer readings. In [6, 9] these values were represented in terms of correlations

$$\tau_c = 0.44\mathrm{Fr}^{-0.14}(W-A)^{-0.28}(d/D)^{0.225} \tag{4}$$

and

$$f_0 = 0.33\mathrm{Fr}^{0.14}(W-A)^{0.28}, \tag{5}$$

applicable to particles of the 0.12-0.65 mm size at a fluidization number W from 1 to 5. It follows from (4) and (5) that τ_c and f_0 are related as follows:

$$\tau_c = 0.145\,\frac{1}{f_0}\,(d/D)^{0.225}, \tag{6}$$

so that the critical group becomes

$$\frac{1-f_0}{\sqrt{\tau_c}}=2.63\,(1-f_0)\sqrt{f_0}\,(d/D)^{-0.112}. \tag{7}$$

The value of f_0 in this group usually varies, for a fluidization bed, within the range from 0.2 to 0.5, hence $(1-f_0)\sqrt{f_0} = 0.37 \pm 0.01$. Then (3) with (7) yields, finally,

$$\alpha_c=1.95\sqrt{\frac{\lambda_S\, c_M \rho_S}{\pi}}\left(\frac{d}{D}\right)^{-0.112}. \tag{8}$$

It is quite evident that parameters τ_c and f_0, on which α_c should basically depend, have been completely eliminated from the fundamental equation of this theory. In Eq. (8) is also missing the filtration velocity and the particle size appears to a small power only (−0.112).

This inconsistency between the fundamental equation of the "packet" theory and the model on the basis of which it has been derived becomes apparent also when Eq. (9) is examined in terms of numerical values of the said parameters given in [5, 10] (there $(1-f_0)/\sqrt{\tau_c} \to 1$).

This inconsistency is not removed even by the modification of the equation according to [5]:

$$\alpha_{\text{cond}}=\frac{1-f_0}{R_K+0.5\sqrt{\dfrac{\pi\tau_c}{\lambda_S\, c_M\rho_S}}}, \tag{9}$$

since it is based on (3) and the empirical correction factor R_K does not compensate for the missing parameters τ_c.

However, since the heat-transfer coefficient in (3) and (9) for a developed fluidization bed has turned out to be independent of τ_c and f_0, an experimental verification of this equation with respect to these parameters [5, 9, 10] is ridiculous. This can be illustrated on the following example. In [10] a thermo-anemometer probe in a bed of 0.12-mm corundum particles with $w_f = 0.14$ m/sec read: $\tau_0 = 0.115$ sec, $\tau_c = 0.13$ sec, and $R_K = 0.0006\ \text{m}^2\cdot\text{deg/W}$. According to these data, Eq. (9) would yield $\alpha_{\text{cond}} = 409\ \text{W/m}^2\cdot\text{deg}$. Let us assume that τ_c had been determined with a 400% error ($\tau_c = 0.65$ sec). Then $f_0 = \tau_c/(\tau_c+\tau_0) = 0.15$ and $\alpha_{\text{cond}} = 433\ \text{W/m}^2\cdot\text{deg}$; thus, a 400% error in τ_c results in only a 6% difference in the value of α_{cond}, which is well within the accuracy of measurements.

The preceding analysis as well as the test results indicate that any simplification of the complex mechanism of external heat transfer, if carried to exaggeration, will not be successful; further progress in the theory of external heat transfer in fluidization systems will depend on more rigorous models describing the mechanism of this process, and must be based on reliable test data pertaining to local structural–hydrodynamic characteristics of a fluidization bed around immersed bodies of specific shapes.

NOTATION

d	is the diameter of particles;
c_M	is the specific heat of particles;
ρ_S	is the density of the continuous phase;
λ_S	is the thermal conductivity of the continuous phase;
ε_0	is the porosity of the bed bulk;
ε_b	is the porosity of the boundary layer;
w_b	is the velocity at the immersed plate surface;
w_f	is the filtration velocity;
δ_m	is the effective thickness of a gaseous interlayer;
l	is the plate length;
y	is the distance from the lower plate edge;
D	is the diameter of the calorimeter;
τ_0	is the time of contact between air bubbles and the plate surface;
α, α_c, α_{conv}, α_{cond}	are the heat-transfer coefficients;
W	is the fluidization number;
Ar	is the Archimedes number;

Fr is the Froude number;

τ_c is the time of contact with packets;

R_K is the thermal resistance of contacts;

A is a correction factor;

f_0 is the relative time of contact between air bubbles and the plate surface.

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