

INVESTIGATION OF INTERPHASE HEAT TRANSFER IN A FLUIDIZED BED

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A method is described for calculating heat transfer in a fluidized bed on the basis of the calorimeter equation. Results are given of investigations performed by a technique developed by the authors. A relation is obtained for calculating the heat transfer between gas and particles.

A rigorous choice of dimensionless numbers based on the equations of mathematical physics and precisely formulated boundary conditions is not yet possible for heat transfer in a fluidized bed. The usual method of processing the data in the form of the parametric relation $Nu = f(Re)$ indicates acceptance of a physical model of the heat-transfer process in which the main thermal resistance is the boundary film of gas phase. Obviously, this model of the heat-transfer process does not correspond to the true situation. This disagreement is one cause of the large discrepancies observed when the test data of various authors are correlated.

There are two ways of viewing heat transfer between the gas and the particles. In the first case the object of investigation is the heat transfer between a single particle in the fluidized bed and the gas. In this approach the heat-transfer coefficient α (the true value is usually designated) naturally serves as a measure of heat transfer, while the validity of an approximate description of the process with the aid of the relation $Nu = f(Re)$ is indisputable. In the second case object of investigation is the heat transfer between the solid phase as a whole and the gas. Attempts to describe this process, as usual, on the basis of the heat-transfer equation

$$Q = \alpha F \Delta T \quad (1)$$

lead to an apparent contradiction. The so-called "effective" heat-transfer coefficients prove to be two or three orders below the true values [1]. Moreover, we find that the effective values of α depend on the height of the static bed H_0 , i.e., on the heat-transfer surface [2], and this points directly to the inapplicability of (1) to interphase heat transfer.

Some authors [3, 4] apply (1) not to the whole bed, but only to the active heat-transfer zone, since there is practically no heat flux outside its limits. Then the examination is restricted to regimes in which the entire available temperature head $\Delta T_0 = |T_g' - T_M|$ is expended in the bed. It should be noted that these regimes are the most favorable for heat transfer only, but are not optimum in general. For instance, increase of gas velocity in the equipment may lead to loss of part of the available head ΔT_0 [4, 5], but in return a decrease of equipment diameter may be achieved. Therefore the limitation of the region of investigation

so far accepted by the condition $T_g'' = T_M$ should be relaxed.

According to the data of Kazakova et al. [4], at sufficiently high gas velocities w , the gas temperature at the bed outlet T_g'' differs considerably from the temperature of the material T_M . At the same time the active heat-transfer zone is sharply defined. Thus, in the upper, "passive" region of the bed, under the above conditions, the temperature head $\Delta T' = |T_g'' - T_M|$ does not vary with height. This fact, at first sight paradoxical, is due to an inflow of particles from the active zone into the upper part of the bed, in which dynamic thermal equilibrium is established.

It is clear from what has been said that determination of the heat-transfer coefficients from the surface area of the particles in the active zone and the corresponding mean integral temperature head is essentially a means of reconciling the values of α found from experiment with the true values. In this sense it is not at all obligatory to calculate α exactly from the active area F_{act} , determined by some condition or other; it may be affirmed a priori that the test values of α will be the closer to the true values, the smaller the part of the surface area of the particles in the bed (reckoning from the gas distributor) used in the calculation. Yet, even in the case in which it has been possible to determine the true heat-transfer coefficients with sufficient certainty, it will be impossible to calculate the heat transfer, since it is possible to determine the mean temperature head and the surface area only by experiment. Thus, the kinetic calculation of interphase heat transfer in a fluidized bed cannot be based on Eq. (1).

As a basic design equation it is expedient to take the calorimeter equation

$$Q = W_g (T_g' - T_g''), \quad (2)$$

transformed to the form

$$Q = W_g \psi \Delta T_0, \quad (2a)$$

where the dimensionless temperature function

$$\psi = \delta T_g / \Delta T_0 = (T_g' - T_g'') / (T_g' - T_M).$$

In the limiting case when the entire available temperature head is utilized in the bed, $\delta T_g = \Delta T_0$, and $\psi = 1$.

The possibility of determining the dependence of the quantity ψ on the conditions governing the course of the process is determined from the balance equation

$$W_g \psi (T_g' - T_M) = W_M (T_M - T_M'), \quad (2b)$$

from which it follows that

$$T_M = T'_M / \left(1 + \frac{W_g}{W_M} \psi \right),$$

or, expressed in excess temperatures,

$$\vartheta_M = \vartheta'_M / \left(1 + \frac{W_g}{W_M} \psi \right). \quad (3)$$

In a given regime of operation of the equipment, the temperature of the material $\vartheta_M = |T'_g - T_M|$ takes on a quite definite value, but then, according to (3), ψ is also determined. It is evident that ψ does not depend on the temperature, and in this sense is an important characteristic of the temperature field in the fluidized bed.

It may be expected that under unsteady heating (or cooling) of the bed, the quantity ψ will not depend on time. In fact, the solution of the heat-balance equation found by us for unsteady heat transfer in the bed

$$T_M = T'_M + \frac{W_g}{G_M c_M} \int_0^t (T'_g - T'_g) dt \quad (4)$$

has the form

$$\begin{aligned} \psi &= \text{const}, \\ \varphi &= \psi \exp(-\psi t^*), \end{aligned} \quad (5)$$

where

$$\varphi = (T'_g - T''_g)/(T'_g - T'_M) = \vartheta''_g/\vartheta'_M,$$

and the dimensionless time is

$$t^* = \frac{W_g}{G_M c_M} t.$$

As may be seen from (5), the rate of change of the temperature field in dimensionless time is numerically equal to the quantity ψ . This permits ψ to be determined experimentally by simple means. Taking logs of both sides of (5), we obtain

$$\ln \varphi = \ln \psi - \psi t^*,$$

or

$$\ln \vartheta''_g = \ln \vartheta'_M + \ln \psi - \psi t^*. \quad (6)$$

In the coordinate system $\ln \vartheta''_g - t^*$ the last equation is a straight line with slope equal to $(-\psi)$.

To verify the relations obtained, we undertook tests on unsteady cooling of particles in a fluidized bed. Cooling studies were made of a fluidized monodisperse bed of steel balls of five sizes (1.3, 1.59, 2.0, 2.44, and 3.18 mm), of glass beads 0.45 mm in diameter, and of spherical silica gel particles 1.6 mm in diameter. Air was used as the fluidizing medium.

The equipment consisted of a thin-walled brass cylinder of inside diameter 88 mm. The calculated heat losses of the equipment were two orders of magnitude lower than the rate of heat transfer in the bed, and were therefore not taken into account. Measurements of gas temperature at various points over the cross

section of the equipment showed that there was no temperature gradient and confirmed that it was possible to neglect heat losses. The air entering the equipment had been previously dried with silica-gel in a nozzle. The flow rate of air was determined from the pressure drop in a standard orifice, and at small flow-rates, by an RS-7 rotameter. Measurement of air temperature differences at the boundaries of the bed was accomplished with a copper-constantan differential thermocouple with its hot junction protected by a mesh from contact with the particles. The current in the differential thermocouple circuit was measured continuously and was recorded on photosensitive paper with the aid of an N-700 galvanometer oscillograph. The paper feed rate was 3 mm/sec.

The particles, preheated in an electric oven to 600–700° K, were poured into the equipment, whereupon the supply of air was at once switched on. The time for the bed to reach an ordered state of cooling was not less than 10–15 sec in all tests, after which the measured temperature difference ϑ''_g began to fall exponentially. From the measured values of the rate of change in ϑ''_g , corresponding values of ψ were calculated for each test. It is interesting to note that whereas the arrangement of the thermocouple hot junction is very important in reducing the heat-transfer data on the basis of Eq. (1), this point is of no importance in the method examined. Temperature measurement of the material and of the gas were previously required to determine the temperature head used in the calculations. In the first case thermocouples with exposed hot junctions were used, and in the second case thermocouples with hot junctions protected from contact with particles and from gas suction. In using the technique based on determination of ψ , it is important to know not only the temperature of the gas and of the material but also their rates of change. However, the rate of change of these two temperatures is identical, since $\vartheta''_g = \psi \vartheta_M$. Therefore the very same result is obtained, independently of the construction of the thermocouple. This was confirmed by tests in which the protecting mesh was removed from the thermocouple hot junction.

The tests showed that ψ decreased with increase of gas velocity and with reduction of particle diameter. From the physical point of view this means that with more intense mixing the relative magnitude of the reverse heat flux increases, and so the fraction of the available temperature head not used in the bed must increase. In addition, the influence of other factors is also affected, for example, the nonuniformity of fluidization.

From reduction of the experimental material obtained, the following relation is proposed for calculating the degree of thermodynamic efficiency of the equipment:

$$\psi = 0.145 Fr^{-0.21} (1 - 1.12^{-Fr}) \left(\frac{\rho_M - \rho_g}{\rho_g} \right)^{0.1} \quad (7)$$

This relation requires further refinement on the basis of wider experimental material. It may be seen from

these expressions that the height of the static bed affects heat transfer only for thin beds ($H_0/d < 40$). For beds with $H_0/d > 40$, heat transfer does not depend on the height of the static bed. However, even for thick beds the equality $T_g'' = T_M$ is fulfilled only for specific values of the gas velocity. With further increase of the velocity the value of ψ falls off. If calculation according to (7) gives $\psi > 1$, it is necessary to take $\psi = 1$, i.e., to use the heat balance equation.

NOTATION

Q —quantity of heat; $W_g = \rho_g c_{pg} V_g$ —water equivalent of gas; $W_M = \rho_M c_M V_M$ —water equivalent of material; ρ_g and ρ_M —density of gas and material, respectively; c_{pg} and c_M —specific heat of gas and material; V_g and V_M —volume flow rate of gas and material; t —time; G_M —mass of material in bed; w —gas velocity in empty equipment; H_0 —height of static bed; d —particle diameter; T_g' and T_g'' —gas temperature at bed inlet and outlet; T_M' and T_M'' —material temperature at bed inlet and outlet; T_M —temperature of material in bed; ΔT_0 —available temperature head; $\vartheta_g'' = |T_g' - T_g''|$, $\vartheta_M = |T_g' - T_M|$, $\vartheta_M' = |T_g' - T_M'|$ —excess temperatures; $\psi = (T_g' - T_g'')/(T_g' - T_M)$, $\varphi = (T_g' - T_g'')/(T_g' - T_M')$ —dimensionless excess temperatures;

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