HEAT TRANSFER IN A THREE-PHASE FLUIDIZED BED

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The authors present results of a theoretical and experimental investigation of heat transfer between a fluid of a fluidized bed and the solid surface under conditions where air is bubbling through the bed.

For a number of chemical technology processes, such as solution, extraction, etc. It is efficient to use a fluidized bed. In a case where one has exhausted the possibilities of heterogeneous heat-transfer processes in a two-phase system it is expedient to apply additional perturbations to the system to intensify the heat and mass transfer processes. These perturbations can be obtained by bubbling gas through the fluid of the bed.

Such three-phase fluidization is also used in a number of heat-transfer facilities with two-phase heat-transfer agents – vapor and liquid-gas, which allows not only intensification of heat transfer, but also the added technological effect of self cleaning of the surfaces due to the action of the fluidized particles.

The laws of three-phase fluidization have not been studied sufficiently as yet, and the available sparse experimental data are contradictory. Since the presence of the gas phase appreciably complicates the behavior of the system, only semiempirical methods are applicable for a quantitative description of the transfer processes under these conditions.

At a low volumetric gas content one can have a stable bubble regime of fluidization in which the bubbles are almost uniformly distributed over the bed volume and behave practically independently. For this situation one can make some theoretical estimates of the influence of the gas phase motion on the intensity of heat transfer between the bed and the surface due to the special features of the behavior of an individual bubble.

The continuous oscillations of the volume and shape of a rising bubble leads to the generation of wave motions of the surrounding medium [1]. The ensemble of rising gas bubbles leads to a random superposition of these wave motions, thereby causing added convective dispersion in the bed. This intensifies the transfer processes of the passive mixture, which is dictated by temperature in the case of heat transfer. The appropriate phenomenological coefficient is analogous to the turbulent diffusion coefficient (or the thermal diffusivity).

The self oscillations of the bubbles are described by a set of spherical harmonics with frequency spectrum ω_K, defined by the Rayleigh formula [2]:

$$\omega_k^2 = K(K+1)(K-1)(K+2)\,\frac{\sigma R^{-3}}{(K+1)\rho_g + K\rho_l}, \tag{1}$$

where the second harmonic (K = 2) makes an appreciable contribution to deformation of the bubble.

The velocity fluctuations of the liquid caused by the bubbles are on the order

$$V' \sim \omega R \tag{2}$$

or, taking account of Eq. (1) and with K = 2,

$$V' \sim \left(\frac{24\sigma}{3\rho_g + 2\rho_l}\right)^{1/2} R^{-1/2}. \tag{3}$$

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As is true of pure turbulence, the random velocity fluctuations V' make an added contribution to the transfer of the mixture, which, in analogy with turbulent diffusion [3], can be described by a diffusion coefficient

$$D^* = L(R)V', \tag{4}$$

where L(R) is the characteristic scale length of the perturbations caused by a bubble of radius R. To a first approximation one can assume that $L = k_1R$, and it follows, therefore, from Eqs. (3) and (4) that

$$D^* \sim k_1 \left(\frac{24\sigma}{3\rho_g + 2\rho_l}\right)^{1/2} R^{1/2}. \tag{5}$$

The first such analysis of the influence of bubble fluctuations on convective diffusion in the gas fluidized bed was made in [4]. The estimates made there showed that $D^* \sim R^{3/4}$. Such a substantial difference from Eq. (5) is explained by the action of quite different causes, generating oscillations of bubbles in the liquid and gas of fluidized beds. In the first case capillary forces play an appreciable role, and in the second case it is fluctuations of the dynamic head of the gas stream washing the bubble. Expressing the bubble radius in Eq. (5) as

$$R = \left(\frac{3}{4\pi n}\right)^{1/3} \left(\frac{\beta}{1-\beta}\right)^{1/3}, \tag{6}$$

where β is the volume gas content

$$\beta = \frac{V_g}{V_g + V_l} = \frac{4/3\pi R^3 n}{1 + 4/3\pi R^3 n}, \tag{7}$$

and introducing the coefficient of proportionality k_2, we obtain

$$D^* = k_1 k_2 \left(\frac{3}{4\pi n}\right)^{1/3} \left(\frac{24\sigma}{3\rho_g + 2\rho_l}\right)^{1/2} \left(\frac{\beta}{1-\beta}\right)^{1/6}. \tag{8}$$

In the heat-transfer case the quantity D* is analogous to the turbulent diffusivity. Therefore, following [5], we can write the effective thermal conductivity of the liquid of the fluidized bed in the form

$$\lambda_{ef} = cD_{ef} = c(D + D^*) = \lambda + cD^*. \tag{9}$$

The thermal conductivity $\lambda = cD$ of the liquid can be computed using empirical correlations [6] as a function of the Archimedes and Reynolds numbers.

To determine the heat-transfer coefficient in a three-phase bed one can use the results of [7]:

$$\alpha \simeq \sqrt{\frac{2}{9}\frac{\lambda_{ef}}{\tau}}. \tag{10}$$

This expression is analogous in structure to the well-known relations following from the packet heat-transfer model in the fluidized bed [8, 9]. However, Eq. (10) contains not the surface contact time of the packet of particles, but the characteristic damping time τ of correlations of the heat-transfer coefficient fluctuations, which, as has been shown theoretically and experimentally in [7], can be described by the correlation function

$$\varphi(t) = \langle \delta\alpha(0)\,\delta\alpha(t) \rangle = \exp(-|t|/2\tau)\cos\bar{\omega}t. \tag{11}$$

Here δα is the deviation of the heat-transfer coefficient from the mean value α defined by Eq. (10), and the ⟨⟩ denotes the operation of statistical averaging.

Using in Eq. (10) the effective thermal conductivity of the bed, Eq. (9), which accounts for convective dispersion of the heat initiated by the moving bubbles, we obtain

$$\alpha \simeq \sqrt{\frac{2}{9}\frac{\lambda c}{\tau}K} \left(\frac{\beta}{1-\beta}\right)^{1/12}, \tag{12}$$

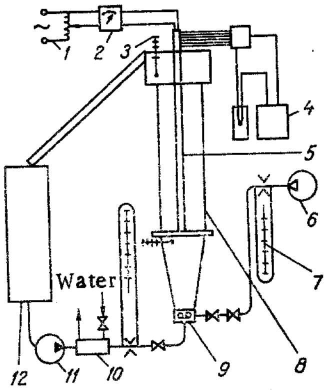

Fig. 1

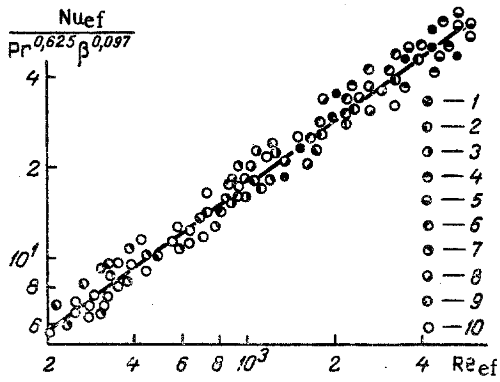

Fig. 2

Fig. 1. Schematic diagram of the experimental facility: 1) autotransformer; 2) type K-50 measuring unit; 3) thermometer; 4) type F-30 measuring unit; 5) electrical calorimeter; 6) compressor; 7) differential manometer; 8) working channel; 9) mixer; 10) cooler; 11) pump; 12) tank.

Fig. 2. Correlation of the test data on stabilized heat transfer: water–air–glass: 1) d = 3.61 mm; 2) 2.72; 3) 2.36; water–air–Alundum: 4) 2.97; 5) 2.33; 20% solution of glycerine–air–glass: 6) 3.61; 7) 2.72; 20% solution of glycerine–air–Alundum: 8) 2.97; 35% solution of glycerine–air–glass: 9) 4.10; 10) 3.61; the line is Eq. (14).

where

$$K = k_1 k_2 \left(\frac{3}{4\pi n}\right)^{1/3} \left(\frac{24\sigma}{3\rho_g + 2\rho_l}\right)^{1/2} D^{-1}.$$

Equation (12) indicates that for intense motion of bubbles the thermal conductivity of the bed itself is less than the addition due to convective dispersion, i.e., $D^*/D > 1$.

Thus, for analysis of the process of heat transfer between the solid surface and the three-phase fluidized bed using Eq. (12) we need information both on the transfer coefficients (λ, D), and on the nature of the damping of correlations of fluctuations $\alpha(t)$, obtained under the same conditions. At present there are no such data. Nevertheless, the model ideas developed have a known practical interest, since Eq. (12) determines explicitly the influence of the volume gas content β on the heat transfer. The nature of this dependence can be checked experimentally, if the data on heat transfer in the bed is represented in the traditional form

$$\mathrm{Nu} = A\,\mathrm{Re}^m \mathrm{Pr}^n \left(\frac{\beta}{1-\beta}\right)^{1/12} = A\,\mathrm{Re}^m \mathrm{Pr}^n \beta^{0.083}. \tag{13}$$

The experiments were conducted in a facility (Fig. 1) which includes the following basic elements: the circuit of the fluidizing medium (liquid phase), the gas circuit, the supply circuit for the electrical calorimeter, and the low-voltage thermometer circuit. The working section of the facility is a vertical annular channel, the external cylinder of which is made of lucite to allow visual observation of the behavior of the system, and the internal coaxial electrical calorimeter is mounted on a distribution grid.

To determine the coefficients of heat transfer to the three-phase bed we used the method of a steady heat flux generated by the heater of the electrical calorimeter. The wall temperatures as a function of height were measured with the aid of copper–constantan thermocouples, and that of the gas–liquid mixture was measured by mercury thermometers. At the working section inlet the mixture temperature was held constant by varying the flow rate of cooling water in the cooler. In the tests we used narrow fractions of Alundum and glass particles (d = 0.91-4.10 mm), water and aqueous solutions of glycerine (20% and 35% by weight), and also air.

The liquid and gas velocities, reckoned to the total cross section of the equipment, were varied in the range 0.07-0.25 m/sec, with which we could identify the range of existence of the bubble regime of fluidization of the three-phase bed for the above-mentioned test conditions.

The experiments established that the heat-transfer coefficients stabilize with height of the heat-transfer surface, which can probably be explained by dynamic equilibrium between the perturbing motion of the solid particles and the gas bubbles in the wall zone, promoting removal of heat from the wall, and heat generation of the surface of the electrical calorimeter. By correlating the test data on stabilized heat transfer we obtain the relation

$$Nu_{ef} = 0.08\, Re_{ef}^{0.788} Pr^{0.625} \beta^{0.097}, \qquad (14)$$

which is valid in the following range of variation of the governing similarity numbers: Re_{ef} = 200-6200; Pr = 5.4-19.9; β = 0.05-0.35. The test data are approximated by Eq. (14) with an rms error of 9.6% (Fig. 2). As governing parameters we used the equivalent diameter of the channel, and the actual liquid velocity and temperature.

Comparison of Eqs. (13) and (14) shows that the theoretical and experimental results agree, if we allow for the approximate nature of the estimates of α and the accuracy of the empirical formulas (14). From this we can hope that the developed model ideas on the influence on the heat transfer of convective dispersion initiated by the gas bubble fluctuations are basically adequate to describe the processes occurring in the three-phase fluidized bed.

NOTATION

w, frequency; K, harmonic number; σ, surface tension; R, bubble radius; ρ, density; V', velocity of propagation of perturbations in the liquid; D, coefficient of turbulent diffusion; L, scale length of perturbations; k, K, A, coefficients of proportionality; n, concentration of gas bubbles per unit volume; β, volume gas content; V, volume flow rate; c, volume heat capacity; λ, thermal conductivity; α, heat-transfer coefficient; τ, time; φ(t), correlation function; d, particle diameter; Nu, Nusselt number; Re, Reynolds number; Pr, Prandtl number. Subscripts: g, gas; ℓ, liquid; ef, effective; *, fluctuating quantity.

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