DETERMINATION OF HEAT TRANSFER COEFFICIENTS IN A FLUIDIZED BED AT VARIABLE FLOW VELOCITY

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On the basis of an empirical expression for calculating the local heat transfer coefficients the authors obtain an equation for the average heat transfer coefficients over the height of a fluidized bed.

In various cases when intense mass transfer processes take place in fluidized-bed heat exchangers, the gas velocity is not constant but varies over the height of the apparatus. In most practical cases it can be assumed that the variation of mass velocity is linear [1]:

$$G = bz. \tag{1}$$

In designing such heat exchangers it is desirable to know the average heat transfer coefficient over the bed height, for which it is necessary to obtain the generalized dependence of the local heat transfer coefficient (at a given level in the bed) on mass velocity over the entire range of variation of the former. The local heat transfer coefficients may be determined by the correlation method

$$\mathrm{Nu} = A\{\exp[-\mathrm{Re}(1-\varepsilon)] - \exp(-cW')\}. \tag{2}$$

Calculations were made for different fluidized systems (see the table) and the following generalized equation was obtained:

$$\mathrm{Nu} = 0.021\left(\frac{c_{\text{т}}\rho d^{1.5}g^{0.5}}{\lambda_g}\right)^{0.75}\times$$
$$\times\{\exp[-\mathrm{Re}(1-\varepsilon)] - \exp(-cW')\}, \tag{3}$$

where

$$c = 0.012\mathrm{Ar}^{0.6},$$
$$\mathrm{Re} = (G - G_{mf})\,d/\mu, \tag{3'}$$
$$W' = (G - G_{mf})/G_{mf}.$$

Equation (3) satisfactorily, with a mean accuracy of ±20%, describes the experimental data on heat transfer between the wall and the bed (see the table) in the laminar region of fluidization: $\mathrm{Re} < 10$; $20 < \mathrm{Ar} < 3000$ (Fig. 1).

Thus, correlation (3) describes the entire curve of variation of the local heat transfer coefficient with variation of mass velocity from a value corresponding to the limit of stability to the value at which solids are carried out of the heat exchanger.

The equation for determining the average coefficient of heat transfer over the height of the heat exchanger between the heating surface and the fluidized bed can be found by integrating expression (3) over the height of the bed:

$$\overline{\mathrm{Nu}} = 0.021\left(\frac{c_{\text{т}}\rho d^{1.5}g^{0.5}}{\lambda_{\text{г}}}\right)^{0.75}\times$$
$$\times\frac{1}{h}\int_0^h\{\exp[-\mathrm{Re}(1-\varepsilon)] - \exp(-cW')\}\,dz. \tag{4}$$

Considering (1) and (3), replacing the true value of the particle concentration $1-\varepsilon$ by its mean value $(1-\varepsilon)_m$ and introducing the notation

$$\mu/d(1-\varepsilon)_m = M,\quad G_{mf}/c = N,$$
$$(bh - G_{mf})\,d/\mu = \overline{\mathrm{Re}},\quad (bh - G_{mf})/G_{mf} = \overline{W}',$$

we finally get

$$\frac{\overline{\mathrm{Nu}}}{A} = \frac{1}{bh}[M\{\exp\mathrm{Re}_{mf} - \exp[-\mathrm{Re}(1-\varepsilon)_m]\} +$$
$$+ N\{\exp c - \exp(-cW')\}]. \tag{5}$$

Equation (5) was checked in experiments performed by the authors with a series of substances, heat treatment of which in a fluidized bed is accompanied by chemical decomposition with copious evolution of gas (Fig. 2).

To confirm relation (1) we performed a special series of experiments to determine flow velocity as a function of bed height. For this purpose various amounts of decomposable material were poured onto the screen of the same apparatus at constant heat load. The gas velocity was measured at the outlet from the bed with a pneumometric tube. The experiments showed that the gas velocity varies linearly over the height of the heat exchanger.

Fluidized Systems Used to Verify Eq. (3)

Reference	System	Density, kg/m³	Mean particle diameter, μ
[5]	Air-quartz sand	2600	128
[2]	Air-quartz sand	2590	140
[3]	Air-quartz sand	2660	150
[4]	Decomposition products-ammonium bicarbonate	1540	120
[4]	Decomposition products-ammonium carbonate	1625	72
[3]	Air-quartz sand	2660	258

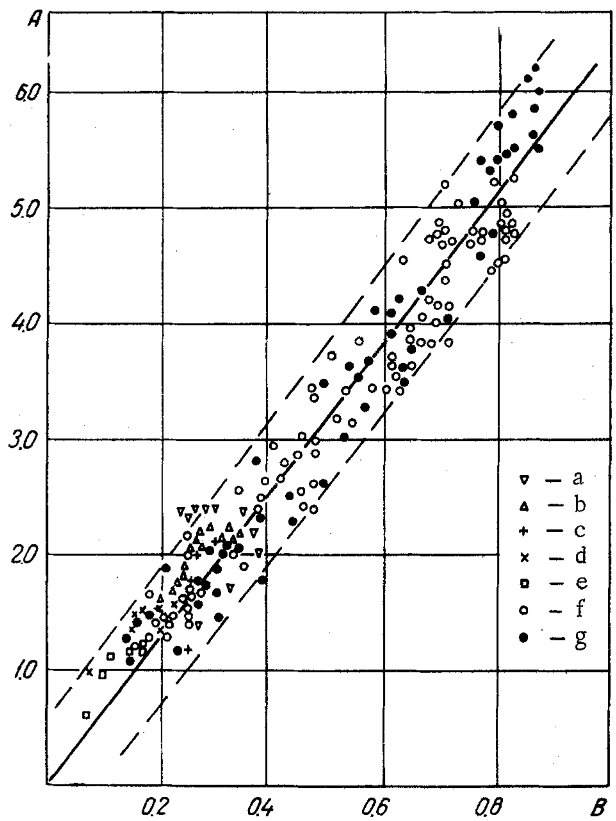

Fig. 1. Generalized correlation of local heat transfer coefficients ($A = Nu[c_T \rho\, d^{1.5}\, g^{0.5}/\lambda_g)^{0.75}]^{-1} \cdot 10^3$; $B = \exp[-Re(1-\varepsilon)] - \exp(-cW'))$: a), b), c), d), e) for quartz sand with d = 128, 140, 150, 238, and 351 μ, respectively, from data of [5, 2, 3]; f) for ammonium bicarbonate with d = 120 μ according to [4], g) for ammonium carbonate with d = 72 μ according to [4].

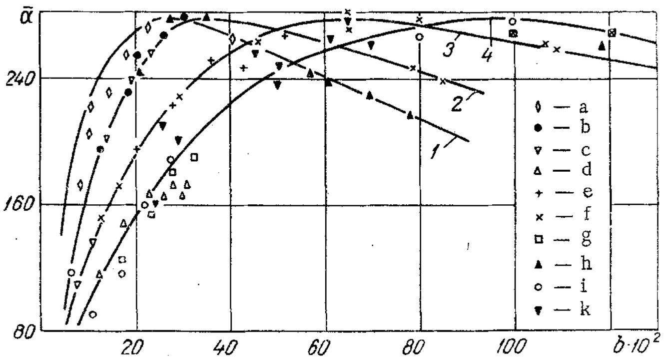

Fig. 2. Average heat transfer coefficient $\overline{\alpha}$, W/m^2·deg, as a function of the parameter b, kg/m^3 · sec, and height of heat exchanger, m: 1), 2), 3), 4) from Eq. (5) with h = 0.385, 0.25, 0.145, and 0.1 m, respectively; the points represent experimental values of $\overline{\alpha}$ [a) at d_1 = = 2 mm, p = 26 777 N/m^2; b), c), d), e) d_1 = 19, 12, 15, and 26 mm, p = 4000 N/m^2; f), g), h), i), k) d_1 = 26, 19, 15, 12, and 2 mm, p = 6670 N/m^2.

For the heat exchanger to operate under optimal conditions it is necessary to maximize the average heat transfer coefficient over the bed height. The maximum value can be determined from Eq. (5). For sufficiently fine particles $G_{mf} \ll M$ and $cW' > 5$, then $\exp Re_{mf} \approx 1$, and $\exp cW' \approx 0$.

Expanding Eq. (5) in series and retaining only three terms, after simple transformations we get

$$\overline{Nu}_m/A = 1 - G_{mf}(\exp c + c)/c\,(bh)_m - \overline{Re}\,(1-\varepsilon)_m/2, \quad (6)$$

where $\overline{Nu}_m$ is the Nusselt number calculated from the maximum average values of the heat transfer coefficient, $(bh)_m$ defines the optimal thermal conditions for the heat exchanger at which α_{max} are attained.

To find $(bh)_m$ it is necessary to differentiate (5) with respect to bh and equate the derivative to zero. Setting $bh - G_{mf} = x$, after differentiation and a series of transformations we get

$$x^2 - 2G_{mf}x - (\exp c + cG_{mf})\,2M/c = 0. \quad (7)$$

Solving quadratic equation (7) and neglecting G^2_{mf} as a small quantity, we finally get

$$(bh)_m = [2M(\exp c + c)\,G_{mf}/c]^{0.5}. \quad (8)$$

Equation (8) relates the parameter b with the height of the heat exchanger. Thus, having determined $(bh)_m$ from (8), we can always select a heat load and bed height such that the heat exchanger operates under optimal conditions. Calculations of $(bh)_m$ from Eq. (8) are satisfactorily confirmed by the experimental data (Fig. 2). Substituting (8) into (6), we get a theoretical relation for the average maximum heat transfer coefficient over the height of the bed:

$$\overline{Nu}_m/A = 1 - 1.41\,(1 - \exp c/c)^{0.5}\,Re^{0.5}_{mf}, \quad (9)$$

where $Re_{mf} = G_{mf}d(1-\varepsilon)_m/\mu$ is the modified Reynolds number corresponding to the velocity at onset of fluidization.

From Eq. (9) it follows that the average maximum heat transfer coefficient does not depend on the rate of filtration of the gas, but is determined by the thermophysical properties of the solids and the fluidizing agent.

The values obtained for the maximum average heat transfer coefficients are satisfactorily confirmed by the experimental data (Fig. 2).

NOTATION:

G—mass velocity, kg/m^2 · hr; b—quantity of gases formed in unit volume of bed, kg/m^3 · hr; z—coordinate along height of heat exchanger, m; h—height of heat exchanger, m; $\overline{\alpha}$—heat transfer coefficient averaged over a height of bed, W/m^2 · deg; α—local heat transfer coefficient, W/m^2 · deg; c_T—specific heat, J/kg · deg; ρ—density, kg/m^3; d—particle diameter, m; λ_g—heat conductivity of gas, W/m · deg; ν—kinematic viscosity of gas, M^2/sec; G_{mf}—mass velocity for minimum fluidization, kg/m^2 · hr; $(1-\varepsilon)$—particle concentration in bed; Re—Reynolds number, Ar—Archimedes number; $W' = (G - G_{mf})/G_{mf}$—fluidization number, d_1—diameter of tube in heat exchanger, mm; p—pressure in apparatus, N/m^2.

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