HEAT TRANSFER IN FLUIDIZED BEDS

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We propose a simplified model of external heat transfer in a fluidized bed. We compare calculated and experimental data, and recommend a computational procedure in polydisperse beds.

Recent research on the combustion and gasification of low-grade solid fuels in a fluidized bed indicates considerable practical interest in the development of models of external heat transfer in high-temperature beds with a high variation of particle sizes. In developing such a model it is quite natural to consider a fluidized bed as a layer of large particles, taking account of their special features, in particular the important role of the convective heat transfer component [1].

With this approach, for all practical purposes it is valid to assume that the temperature drop between the heat-transfer surface and the bed core occurs mainly in the layer of particles nearest the surface. This in turn makes it possible in the first approximation to consider mechanistically heat transfer between the surface and a layer of fluidized particles as a problem of heating a packet consisting of a gaseous layer of thickness $l_0$ and a quasihomogeneous medium (the fluidized bed). Unlike models proposed earlier (e.g., [2]), it is assumed that the whole packet is penetrated by gas filtering through it.

This problem can be formulated mathematically as follows:

$$\lambda_f \frac{d^2 T_f}{dy^2} - \frac{c_\rho \mu (T_f - T_0)}{H_e} = 0, \quad 0 < y < l_0,$$

(1)

$$\lambda_s \frac{d^2 T_s}{dy^2} - \frac{c_\rho \mu (T_s - T_0)}{H} = 0, \quad l_0 < y < l$$

(2)

with the boundary conditions
As in [3], the thermal conductivity of the gaseous film is taken as the sum of conductive and convective (filtration) components: \( \lambda_f + \lambda_k + n \frac{cpud}{v} \). We express the thickness \( \ell_0 \) of the gaseous layer as a function of the mean free path of the gas [2], i.e., in the form \( \ell_0 = m(1 - \epsilon)^{-2/3} \).

The external heat-transfer coefficient is

\[
\alpha = \frac{\lambda_T}{T_s(l) - T_w} \frac{dT_f(0)}{dy}. \tag{4}
\]

By substituting into Eq. (4) the values of \( T_s(l) \) and \( dT_f(0)/dy \) found by solving (1)-(3), we obtain

\[
\alpha = \frac{\lambda_f}{\ell_0} \frac{1 + \varphi}{1 - \psi}, \tag{6}
\]

where

\[
\varphi = -\frac{k^2\delta(1 - e^\delta)}{\lambda_f(1 - e^\delta)}; \quad \psi = -\frac{(1 - e^\delta)^2}{\lambda_f(1 - e^\delta)}.
\]

Calculations showed that \( \varphi \) and \( \psi \) vary over very narrow ranges: \( 0.001 < \varphi < 0.01; 0.01 < \psi < 0.05 \) for practically all gases and solids used in fluidization technology, and that with an error of no more than 6% Eq. (6) can be replaced by

\[
\alpha = \frac{\lambda_f}{\ell_0}. \tag{7}
\]

Since \( \alpha \) can generally be measured to within 5-6%, writing Eq. (7) instead of (5) and (6) is quite acceptable.

Using the expression for \( \lambda_f \), Eq. (7) takes the form

\[
\alpha = \frac{\lambda_f}{\ell_0} \frac{1 - e^\delta}{e^\delta \ell_0}, \tag{8}
\]

which is the form commonly assumed for heat-transfer coefficients as the sum of conductive and convective components. Substituting \( \ell_0 \) into (8), we obtain

\[
\alpha = \frac{\lambda_f}{\ell_0} \frac{1 - e^\delta}{e^\delta \ell_0}, \tag{9}
\]