TEMPERATURE OF PARTICLES HEATED IN A FLUIDIZED BED OF INERT MATERIAL

A. P. Baskakov, N. F. Filippovskii,
V. A. Munts, and A. A. Ashikhmin

UDC 662.62:66.096.5

The authors give formulas for calculating the heat- and mass-transfer coefficients for particles with a fluidized bed of inert material. A method is given for determining the temperature of the hot particle.

With known concentration C of oxygen in a gas and known temperature t of a bed of inert particles, the temperature of a hot particle t_p is given by the relation

$$t_p = t + \frac{CQ}{\alpha(1/K + 1/\beta)}. \tag{1}$$

In a fluidized bed of fine particles the rate of filtration of the gas through the solid layer (in the gaps between the particles) is very small, and therefore the gas convection does not appreciably influence the heat and mass transfer processes. Taking into account that the heated particle is separated from the inert particles of the same size by a spherical layer of volume equal to that of the pores pertaining to one particle, we can write [1]

$$\mathrm{Nu}_1 = \frac{\alpha d}{\lambda_g} = \frac{2}{1-(1-\varepsilon)^{1/3}} \simeq 10. \tag{2}$$

This agrees with the experimental and theoretical data of [2] obtained with heat transfer to a sensor and a bed of nonmetallic particles without blown gas.

The minimum value of mass-transfer coefficient from a single particle in a motionless bed is given by the relation Sh = 2. An analogous expression must be valid also for mass transfer from a single particle to a bed of inert particles without beam gas. The diffusion conditions in the pores of the solid material are worse than for a pure gas filling the pores, and therefore the value of Sh (with the coefficient of gaseous diffusion substituted into it) must be less than two. In the first approximation [3]

$$\mathrm{Sh} \simeq 2\varepsilon \simeq 1. \tag{3}$$

In the bed of very large particles ($\mathrm{Ar} > 10^6$) convective heat and mass transfer predominate, and there is an analogy between the processes of convective heat and mass transfer. For the case when the diameter of the hot particle is equal to that of the inert particles, we use the formula

$$\mathrm{Nu}_1 = 0{,}4\,(\mathrm{Re}/\varepsilon)^{2/3}\,\mathrm{Pr}^{1/3}, \tag{4}$$

proposed by the authors of [1] to calculate the interphase heat transfer in a fluidized bed of large particles ($\mathrm{Re}/\varepsilon > 200$). Taking into account that the intensity of interphase heat and mass transfer in a bed of these particles depends slightly on the fluidization velocity [4-6], we can simplify Eq. (2), reducing it to the conditions corresponding to the onset of fluidization. This approach also has the physical basis that the particles are found mainly in the dense phase with blowing velocity close to that of the onset of fluidization. Taking ε = 0.48 and calculating Re from the simplified relation [7] for large Archimedes numbers

$$\mathrm{Re} = 0{,}25\,\mathrm{Ar}^{1/2}, \tag{5}$$

S. M. Kirov Ural Polytechnic Institute, Sverdlovsk. Translated from Inzhererno-Fizicheskii Zhurnal, Vol. 52, No. 5, pp. 788-793, May, 1987. Original article submitted March 24, 1986.

 0022-0841/87/5205-0574$12.50

we obtain for heat transfer

$$Nu_2 = 0{,}26\,(Ar\,Pr)^{1/3} \tag{6}$$

and for mass transfer

$$Sh_2 = 0{,}26\,(Ar\,Sc)^{1/3}. \tag{7}$$

Over a wide range of particle diameter $d_p = d$ we can propose interpolation relations that coincide with the limit for small ($Ar \simeq 0$) and large ($Ar \simeq 10^8$) values of the Archimedes number

$$Nu_3 = 10 + 0{,}23\,(Ar\,Pr)^{1/3}, \tag{8}$$

$$Sh_3 = 1 + 0{,}26\,(Ar\,Sc)^{1/3}. \tag{9}$$

We should stress that in a bed of fine particles ($Ar < 10^2$) the ratio $Nu_3/Sh_3 \simeq 10$, and in a bed of coarse particles ($Ar > 10^6$) $Nu_3/Sh_3 \simeq 1$.

Thus, in the case where the process is not restrained by the kinetics, i.e., in the purely diffusion region, for an increase of particle diameter, the temperature drop between the hot particle and the bed will increase by an order in the interval $Ar = 10^2$–10^6, and remain almost unchanged outside those limits. In fact the temperature drop will be less because of the influence of radiation.

For an increase of diameter d_p of the body (for example, a hot particle) above the diameter d of the inert particles composing the main mass of the bed, the intensities of heat and mass transfer first drop, and then stabilize. For bodies of diameter d_p = 10-60 mm we can use the empirical relations [8] describing heat and mass transfer by convection of a gas and particles:

$$Nu_4 = 0{,}85\,Ar^{0{,}19} + 0{,}006\,Ar^{0{,}5}\,Pr^{1/3}, \tag{10}$$

$$Sh_4 = 0{,}009\,Ar^{0{,}5}\,Sc^{1/3}. \tag{11}$$

The gradually attenuating influence of the ratio of diameters of the body and the particles corresponds qualitatively to an exponential form of dependence of the heat transfer coefficient on this factor. Therefore, over a wide range of the ratio d_p/d we can suggest the calculation formula in the form

$$Nu = Nu_4 + (Nu_3 - Nu_4)\exp\left(-\frac{d_p}{4d}\right). \tag{12}$$

An indirect justification for Eq. (12) is the analysis in [9] of the influence of the wall height on the intensity of heat transfer to a bed of large particles, although undoubtedly future direct experiments may require it to be modified.

At high temperatures we must add the radiant coefficient [10]

$$\alpha_r = 7{,}3\sigma_0\varepsilon_p\varepsilon_b t_p^3 \tag{13}$$

to the conductive and convective heat transfer coefficients.

For mass transfer in a bed of fine particles the influence of the ratio d_p/d can be estimated from the condition that $Sh' \simeq 1$ and for the case $d_p > d$, if we take d_p as the characteristic dimension in the Sherwood number Sh'. But if we take d as the characteristic dimension, then with $d_p/d > 1$ we obtain $Sh = d/d_p$ in a bed of large particles the influence of the body diameter on the mass transfer is reduced [6]. Over a wide range of particle diameters and of the ratio $d_p/d > 1$ to calculate the mass transfer we can suggest the interpolation relation

$$Sh = \frac{d}{d_p} + Sh_4 + (Sh_2 - Sh_4)\exp\left(-\frac{d_p}{4d}\right). \tag{14}$$

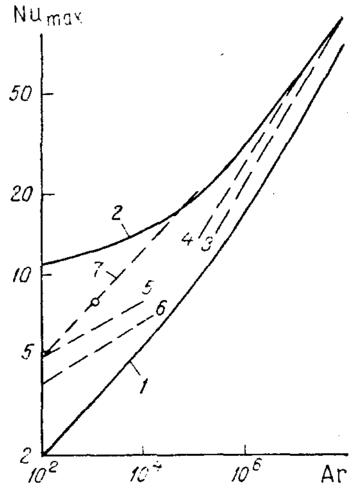

Fig. 1

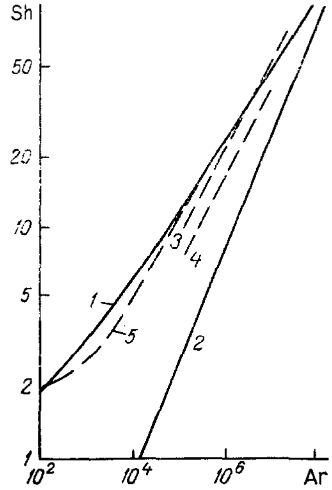

Fig. 2

Fig. 1. Dependence of the dimensionless coefficient of heat transfer $Nu_{max} = \alpha_{max} d/\lambda_g$ between a fluidized bed and a body immersed in it, as a function of the number Ar: 1, 2, 3) calculations using Eqs. (10), (8) and (6); 4, 5, 6, 7) data of [5], [12], [12] and [11], respectively. The body dimensions are: 1) d_p = 10-60 mm; 2, 3, 4) d_p = d; 5) d_p = 3d; 6) d_p = 10d; 7) wire of d_p = 0.2 mm (the point corresponds to the condition d_p = d).

Fig. 2. Dependence of the dimensionless coefficient of mass transfer $Sh = \beta d/D$ between a fluidized bed and a body immersed in it as a function of the number Ar: 1) calculation for a small body with d_p = d using Eq. (9); 2) calculation using the empirical formula (11) for a large body with $d_p > 15$-20 mm; 3, 4) experimental relations of [6] for d_p = d and d_p = 10d; 5) experimental correlations of [13], [14] for d_p = d (in the calculations we assumed Sc = 0.7).

Figures 1 and 2 show a comparison of the calculations using the above formulas and the empirical data. It should be noted that, according to [12] and the analysis of Figs. 1 and 2, the intensity of heat and mass transfer from anchored and freely flying bodies is practically the same.

As they burn up the hot particles become fewer than the inert, but they continue to stay in the bed roughly until their dimension has decreased so much that the feed rate becomes less than the gas velocity in the equipment. In a bed of fine particles of inert material the rate of filtration of the gas is small, and therefore as before we can use the relation $Sh_1 = Sh_1' \simeq 1$. The higher value $Nu_1 = Nu_1' \simeq 10$ of Eq. (2) is associated with contact of the hot particle and the inert particles. With decrease of the ratio d_p/d there is a decrease of the number of points of contact of the hot particle with the inert particles. Thus, for spheres with d_p/d = 1; 0.732; 0.414; 0.225; 0.155; 0 the maximum possible number of points of contact is N = 12, 8, 6, 4, 3, 2. Taking into account Nu_1' is roughly proportional to the maximum possible number of points of contact, for the case of very fine particles of inert material and even finer hot particles ($d_p < d$) we obtain

$$Nu_1' = 2 + 8d_p/d. \tag{15}$$

Located between large particles of inert material, the finer hot particles are washed by a gas with velocity corresponding to the onset of fluidization of the inert particles, as given by Eq. (5). To calculate the heat and mass transfer in this case (besides for $d_p < d$) we use Eq. (4), taking as the characteristic dimension there the hot particle diameter:

$$Nu_2' = 0{,}26\,(d_p/d)^{2/3}\,(\mathrm{Ar\,Pr})^{1/3}, \tag{16}$$

$$Sh_2' = 0{,}26\,(d_p/d)^{2/3}\,(\mathrm{Ar\,Sc})^{1/3}. \tag{17}$$

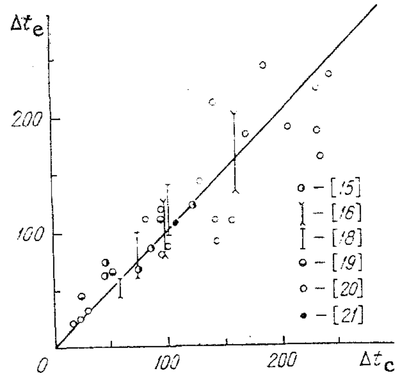

Fig. 3

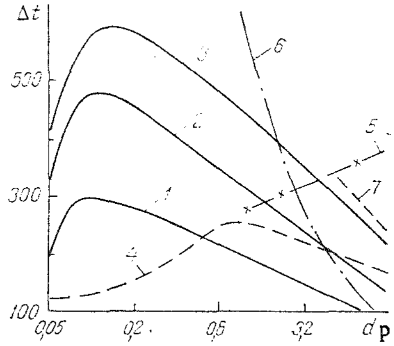

Fig. 4

Fig. 3. Comparison of the calculated Δt_c and experimental Δt_e values of overheat temperature of hot coal particles, °C.

Fig. 4. The overheat temperature of a coal particle Δt, °C, as a function of its diameter d_p, mm, in a bed of sand fluidized with air, t = 900°C: 1, 2, 3) calculation using the given method for brown coal with inert material particle diameter of d = 0.4, 0.8, and 1.5 mm, respectively; 4) the same for Ekibasuzskyi coal with d = 1.5 mm; 5, 6) calculations according to [16, 17] with d = 0.8 mm; 7) the data of [20] with d = 1.5 mm.

The Archimedes numbers in Eqs. (16) and (17) are calculated from the characteristics of the inert material, and Nu' and Sh' are calculated from the diameter of the hot particles. Over a wide range of diameters of the inert particles with $d_p/d \leqslant 1$ we can write relations analogous to Eqs. (8) and (9):

$$\mathrm{Nu}_3' = 2 + 8\frac{d_p}{d} + 0{,}23\,(d_p/d)^{2/3}\,(\mathrm{Ar\,Pr})^{1/3}, \tag{18}$$

$$\mathrm{Sh}_3' = 1 + 0{,}26\,(d_p/d)^{2/3}\,(\mathrm{Ar\,Sc})^{1/3}. \tag{19}$$

Figure 3 compares the calculations using the proposed formulas with the known experimental data of [15-21]. Unfortunately, most of the authors do not give the values of the kinetic constants of the coals used. In these cases the rate constants of burning of brown and were calculated from the formula K = 8610 exp (−11908/T) (as for Irsh-Borodinskyi [22]), and for cokes and Ekibastuzkyi coal we took K = 11060 exp (−13555–T) [23]. The thermophysical parameters entering into the equations were taken for air at an average temperature between that of the bed and the hot particles. The scatter of the experimental values of the hot temperature stemmed to a considerable extent from the difficulties of experimentally determining the temperature of a hot coal particle. The measurement of a thermocouple temperature [16, 20] can lead to substantial errors because of the added errors due to the fact that the coal particle has quite high electrical conductivity. In temperature measurement by photographic methods [15, 19] we actually measure the temperature of that part of the particle which is not in the bed. The melted wire method, used in [18], allows one to determine only the range of hot temperatures. In addition, difficulties also arise with determining the oxygen concentration, if there are enough hot particles in the bed [20].

To estimate the conditions for possible sintering of ash in a bed, it is important to known the overheat temperature as the coal particles burn up, i.e., as their diameter decreases for a constant diameter of inert material. Calculations show that in the diffusion regime a decrease of the diameter of the coal particles leads to an increase of the overheat temperature (Fig. 4). This is in good agreement with direct measurements of particles in the process of burnup [15], and is also confirmed by experimental data obtained on coal particles of different diameters (Fig. 4, curve 7 [20]), and by theoretical calculations (Fig. 4, curve 6 [17]). In some experiments contrary data were obtained. For instance, in the empirical formula of the authors of [16], $\Delta t \sim d_p^{0,15}$ (Fig. 4, curve 5). The cause of the contrary data obtained by different authors may be the above-mentioned difficulties in accurate measurement of Δt, and with differences in properties of hot materials. In particular, in the kinetic region a decrease of coal particle diameter leads to a decrease of Δt, since the heat transfer coefficient continues to grow, but the intensity of heat release per unit area does not change. According to the data presented in Fig. 4 (curves 1-3) at a fluidized bed temperature of 900°C this region of combustion of brown caoal is observed for $d_p \lesssim$ 1 mm, and for Eikbastuzskyi coal, for $d_p \lesssim$ 0.2 mm. The maximum overheat temperature corresponds in this case to a zone of transition from the diffusion combustion regime to the kinetic regime, i.e., for $K \sim \beta$.

With the proposed method one can calculate the temperature of hot particles, with accuracy sufficient for engineering purposes, as required for calculating the rate of burnup of coal in fluidized bed furnaces and for estimates of possible conditions of slag formation.

NOTATION

C, mass concentration of oxygen in the gas; d, d_p, diameter of a particle of inert material and of coal, respectively; D, coefficient of molecular diffusion of CO_2 in air; K, rate constant of the reaction $C + O_2$; N, number of points of contact of the hot particle with inert particles; Q = 12.3 MJ/kg, heat of the reaction $C + O_2$ in calculating 1 kg of oxygen; t, t_p, temperature of the fluidized bed and the hot particle, respectively; α, coefficient of heat transfer; β, coefficient of mass transfer; ε, porosity of the fluidized bed; ε_b, ε_p, emissivity of the fluidized bed and of the coal particle, respectively; λ_g, thermal conductivity of air; σ_0, Stefan–Boltzmann constant; ν, kinematic viscosity of air; Ar, Nu, Pr, Re, Sc, Sh, the Archimedes, Nusselt, Prandtl, Reynolds, Schmidt and Sherwood numbers, respectively.

LITERATURE CITED

1. N. I. Gel'perin, V. G. Ainshtein, and V. B. Kvasha, Fundamentals of Fluidization Technology [in Russian], Moscow (1967).
2. A. P. Baskakov, Inzh.-Fiz. Zh., 6, No. 11, 20-25 (1963).
3. M. M. Avedesian and J. F. Davidson, Trans. Inst. Chem. Eng., 51, 121-131 (1973).
4. C. G. Wen and T. H. Chang, "Particle to particle heat transfer in air-fluidized beds," Proc. Int. Symp. on Fluidization, Eindhoven (1967), pp. 491-506.
5. G. I. Pal'chenok and A. I. Tamarin, Inzh.-Fiz. Zh., 45, No. 3, 427-433 (1983).
6. G. I. Pal'chenok and A. I. Tamarin, Inzh.-Fiz. Zh., 47, No. 2, 235-242 (1984).
7. O. M. Todes and O. B. Tsitovich, Equipment with a Fluidized Granular Bed [in Russian], Moscow (1966).
8. A. P. Baskakov, B. V. Berg, A. F. Ryzhkov, and N. F. Fillippovskii, Heat and Mass Transfer Processes in a Fluidized Bed [in Russian], Moscow (1978).
9. N. F. Filippovskii, A. P. Kaskakov, and A. A. Zharkov, Heat and Mass Transfer in Technical Processes and Equipment [in Russian], Minsk (1985), pp. 106-110.
10. O. M. Panov, A. P. Baskakov, Yu. M. Goldobin, et al., Inzh.-Fiz. Zh., 36, 409-415 (1979).
11. N. N. Varygin, Trudy Mosk. Inst. Khim. Mashinostr., No. 26, 33-38 (1964).
12. W. Prins, W. Draijer, and W. P. M. Van Swaiij, "Heat transfer to immersed spheres fixed or freely moving in a gas-fluidized bed," Rep. No. N41 at 16th Int. Symp. on Heat and Mass Transfer, Dubrovnik, Yugoslavia, 3-7 Sept. (1984).
13. T. M. Hsiung and G. Thodos, Chem. Eng. Sci., 32, No. 6, 581-592 (1977).
14. T. M. Hsiung and G. Thodos, Int. J. Heat Mass Transfer, 22, No. 7, 1003-1008 (1979).
15. I. B. Ross, M. C. Potel, and J. F. Davidson, Trans. Inst. Chem. Eng., 59, No. 2, 83-88 (1981).
16. A. I. Tamarin and D. M. Galershtein, Problems of Heat and Mass Transfer in Combustion Processes Used in the Energy Industry [in Russian], Minsk (1980), pp. 104-121.
17. I. R. Howard, Fluidized Beds: Combustion and Applications, London (1983).
18. J. G. Yates and P. R. Walker, New Developments in the Theory and Practice of Fluidization [Russian translation], Moscow (1980), pp. 7-16.
19. R. K. Chakraborty and J. R. Howard, J. Inst. Fuel, 51, No. 12, 220-224 (1978).
20. Yu. G. Pavlov, "Investigation of the nature and development of the method of calculating the mechanical underfire of fuel on combustion of Kanskoachinskii coal in a fluidized bed," Author's Abstract of Doctoral Dissertation, Technical Sciences, Moscow (1985).
21. R. B. Reel and F. J. Canstos, Inst. of Energy Symp. Ser. No. 4, Fluidized Combustion: Systems and Applications, London (1980), Sections II-B-2-1 to II-B-2-9.
22. S. L. Shagalova and I. N. Shnitser, Combustion of Solid Fuel in Vapor Regenerator Furnaces [in Russian], Leningrad (1976).
23. T. V. Vilenskii and D. M. Khmalyan, Dynamics of Combustion of Dusty Fuels [in Russian], Moscow (1977).