INVESTIGATION OF HEAT TRANSFER BETWEEN A SURFACE AND A FLUIDIZED BED IN DRYING PROCESSES

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Results are presented of an experimental investigation of heat transfer between a surface and a moist fluidized bed. The test data have been reduced by means of criterial relations.

Since a material of low thermal stability cannot be dried in a fluidized bed with a high-temperature drying agent, the output of driers with a fluidized bed for such materials is low.

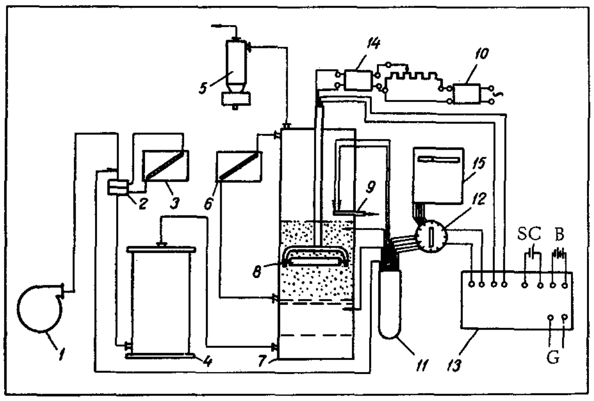

Fig. 1. Schematic of the apparatus: 1) air blower; 2) diaphragm; 3) differential manometer; 4) electric heater; 5) cyclone; 6) differential micromanometer; 7) test apparatus; 8) heat transfer tube; 9) psychrometer; 10) voltage stabilizer; 11) Dewar flask; 12) thermocouple switch; 13) R-330 potentiometer; (14) wattmeter; 15) potentiometric recorder; SC—standard cell; B—battery; G—galvanometer.

Increase of drier output while retaining a low initial temperature of the drying agent may be achieved by supplying additional heat through heat transfer surfaces located in the bed [1–3]. Additional heat supply during drying of MSN copolymer permitted the drier output to be increased by a factor of 2–2.5 [3].

During drying with additional heat supply, an increase in the intensity of heat transfer between the surface and the bed has been noted [2, 3]. It is therefore of interest to carry out a systematic investigation of heat transfer between a surface and a fluidized bed during the drying process.

The experimental investigations were conducted in an apparatus (Fig. 1) in which the heat transfer surface was a copper tube of diameter 26 mm and length 176 mm, located in the fluidized bed of a drier and with a spiral nichrome heater inside it. The hot junctions of a differential ten-channel thermocouple were embedded in five equally distributed slots on the surface of the tube. For electrical insulation the thermocouple junctions were sheathed in fiberglas, pressed into the slots in the tube, and covered by a copper cylinder of wall thickness 1 mm. To avoid air circulation in the slots, their ends were sealed with epoxy resin. The cold junctions of the thermocouples were fastened to a holder and distributed in the bed. The heating element was held in a frame with the aid of sharply pointed screws [5].

The investigations of heat transfer between the surface and the fluidized bed during the drying process was conducted in the unsteady regime. This allowed investigation of heat transfer under conditions of constant and decreasing rates of drying, and for a dry bed.

Drying curves and curves of variation of bed temperature are required to describe an unsteady drying process. When constructing drying curves we require to know the variation of moisture content of the material in the bed during the entire experiment.

The moisture content of the material was calculated from the resistance of the bed, using the initial weight and humidity of the bed, according to a method described in [4]. The bed resistance was measured by an inclined differential manometer, permitting calculation of moisture content to an accuracy of 0.3%. The air humidity at the bed outlet was measured directly above the bed by means of a psychrometer with wet and dry suction thermocouples.

In order to remove the hydrodynamic characteristics of the fluidized bed and to construct the thermal balance, measurements were made of the air flow rate, as well as of the temperatures of the bed and of the air at the metering diaphragm and below the distributor. Measurements of temperature were carried out with copper-constantan thermocouples and a double-range R-330 potentiometer of accuracy class 0.015%. The power supplied by the electric current passing through the spiral of the tube was measured by means of a multirange wattmeter of accuracy class 0.5%.

The heat transfer coefficient was calculated from the equation

$$\alpha = Q/F\,\Delta t. \tag{1}$$

The amount of heat transferred was determined from the current flowing through the heater of the heat transfer tube, this being all transmitted to the bed through the tube surface. It was assumed that the transmission of heat by conduction from the tube to the support elements may be neglected on account of the method of construction of the tube supports. Transfer of heat by radiation was not taken into account because of the low temperature (not above 343.2° K) of the tube surface.

With the aim of creating a constant thermal flux, the heater was fed through a voltage stabilizer, and controlled by an autotransformer.

The temperature difference between the surface and the bed was determined by the readings of the differential thermocouple, with correction for the embedded depth of the hot junctions of the thermocouple. It was assumed that the temperature of the fluidized bed was constant throughout the whole volume above the region of temperature stabilization of the incoming gas. The error in calculating heat transfer coefficient did not exceed 2.2%.

The tests were carried out with particles of silicagel of various sizes (see table). Critical fluidization velocities for dry and moist particles were found in preliminary tests, and the dependences were obtained of the mean porosity of a bed of dry and moist particles on the velocity of the air passing through.

The investigations showed that the gas velocity, the particle diameter, the relative humidity of the gas, and the position of the tube in the bed—in both transverse and longitudinal flow—all have an influence on the heat transfer between the surface and the fluidized bed during the drying process. No influence of moisture removal in the bed nor of the specific heat flux in the tube was observed, within the range of their variation in the experiments. The specific heat flux at the heat transfer surface was varied in the range 970–6800 W/m^2. Moisture removal in the bed was in the range 9.9–76 kg/m^2·hr.

The coefficient of heat transfer between the surface and the fluidized bed does not stay constant during the drying process. As may be seen from Fig. 2, it has its greatest value at the start of the drying for the maximum relative humidity φ_2 of the air, and retains its value for some time, after which it gradually decreases, as φ_2 drops, toward the value for the dry bed.

Characteristics of Silica Gel Fractions

Equivalent diameter, mm	Specific weight, kg/cm^3	Saturated weight, kg/cm^3	Porosity of bed at rest	Fluidization velocity, m/sec
0.670	870	410	0.529	0.140
0.875	870	415	0.523	0.176
1.237	870	425	0.511	0.310
2.45	870	450	0.483	0.551
3.50	870	470	0.460	0.770

The largest excess of the heat transfer coefficient α for the moist bed over α for the dry bed was in the range 1.15–1.31. It should be noted that the bed temperature began to increase while the drying rate was held constant. From the time of increase of bed temperature, a decrease in the coefficient of heat transfer between the surface and the bed was observed.

As may be seen from Fig. 3, the nature of the dependence of the experimental values of heat transfer coefficient on the gas velocity for silica gel particles of different diameters is similar to that for a dry bed [2]. The increase of the maximum heat transfer coefficient for a moist bed in comparison with a dry bed is not the same for particles of different diameter. The difference increases as particle size decreases. Thus, for example, for particles of diameter 0.67 mm the increase is 28.4%, and for particles of diameter 2.45 mm—16%.

Other conditions being equal, the heat transfer coefficients for a horizontal arrangement of the heated tube were higher than for a vertical one. The heat

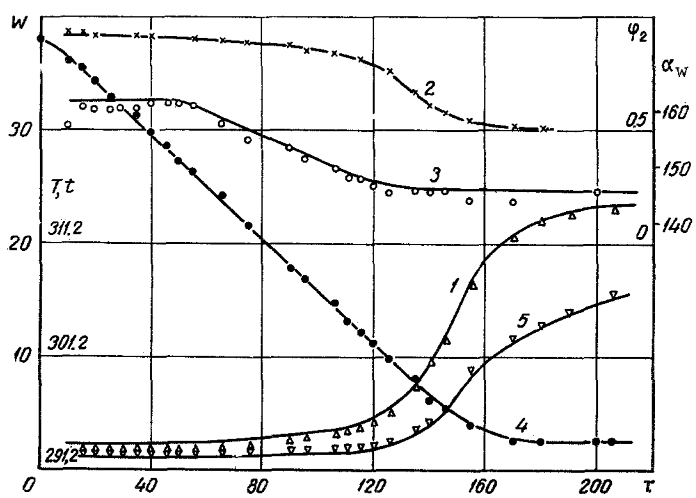

Fig. 2. Dependence of the dry bulb temperature, T, °K (1), relative humidity of the air at the bed outlet, φ_2, % (2), heat transfer coefficient, α_W, W/m^2·degree (3), humidity of the bed, W, % (4), and wet bulb temperature, t (5), on the drying time, τ, min.

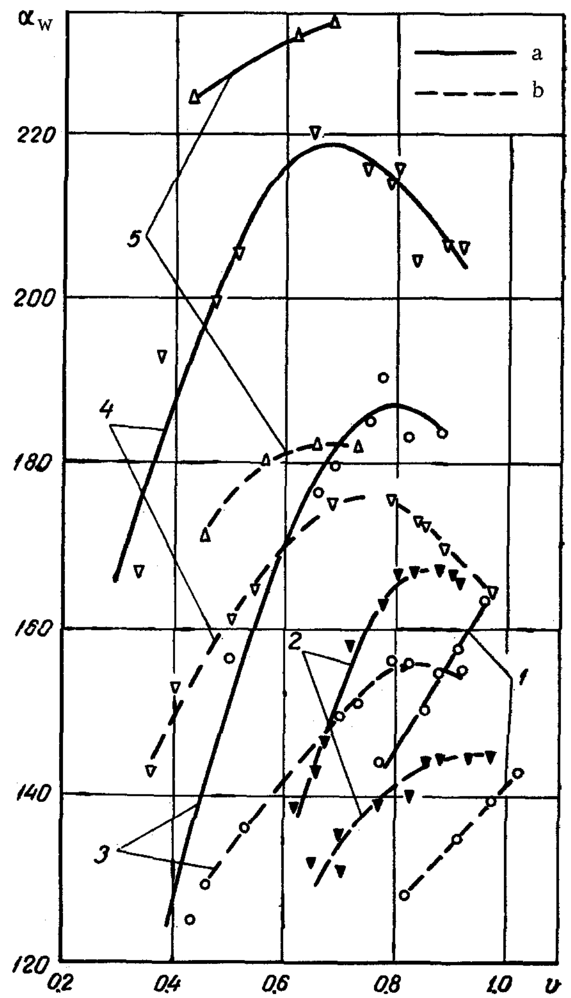

Fig. 3. Dependence of the coefficient, α_w, ($W/m^2 \cdot °C$), of heat transfer between a vertical tube and the bed on the gas velocity, v, (m/sec), for particles of silica gel of diameter: 1—3.50 mm; 2—2.45; 3—1.237; 4—0.875; 5—0.670; a)—in the first drying period; b) for the dry bed.

transfer coefficient for the heater in a horizontal position increased with distance from the distributor, as has also been noted by other investigators [5]. The heat transfer coefficient for a vertical tube was somewhat greater when the heater was positioned on the bed axis than when it was at the periphery.

The results of the tests to determine α have been reduced to criterial form in the shape of the relation

$$\mathrm{Nu} = f\left[\mathrm{Ar};\ \frac{m^3}{(1-m)^2};\ \mathrm{Gu};\ \mathrm{Pr};\ l\right]. \qquad (2)$$

Since the Re number does not uniquely describe the hydrodynamics of a fluidized bed, we have taken as a characteristic criterion the group $\mathrm{Ar_e}(1 - m)$ [6, 7], which may be transformed, by replacing d_e by the diameter of the equivalent channels, to the form

$$\frac{g\left(\frac{2}{3}\varphi_\kappa d_e \frac{m}{1-m}\right)^3 \gamma_M}{\nu^2\gamma_g}(1-m) \sim \mathrm{Ar}\,\frac{m^3}{(1-m)^2}. \qquad (3)$$

The influence of mass transfer during the drying process on the heat transfer is taken into account by the thermal Gukhman number

$$\mathrm{Gu} = (T_d - T_w)/T_d. \qquad (4)$$

The simplex l in (2) takes into account the influence of the geometrical parameters on heat transfer.

Since all the tests were carried out with air, the Prandtl number, which in this case has a constant value, may be omitted. The final equation used to reduce the experimental heat transfer data has the form

$$\mathrm{Nu} = f\left[\mathrm{Ar};\ \frac{m^3}{(1-m)^2};\ \mathrm{Gu};\ l\right]. \qquad (5)$$

In the period of constant drying rate and heat transfer coefficient, the air leaving the bed was practically saturated. In this case there is self-similarity of the heat transfer process, and the Gu number drops out of Eq. (5).

Equations for the horizontal and vertical arrangements of the tube in the bed were obtained by a method of successive introduction of parameters. For $\mathrm{Gu} < 4\cdot10^{-3}$ these equations have the form:

for a horizontal tube arrangement

$$\mathrm{Nu} = 0.220\,\mathrm{Ar}^{0.32}\left[\frac{m^3}{(1-m)^2}\right]^{0.445}\left(\frac{h}{D}\right)^{0.15}, \qquad (6)$$

for a vertical tube arrangement

$$\mathrm{Nu} = 0.268\,\mathrm{Ar}^{0.28}\left[\frac{m^3}{(1-m)^2}\right]^{0.54}\left(1-\frac{r}{R}\right)^{0.02}. \qquad (7)$$

Equation (7) is represented by curve 1 of Fig. 4, from which it may be seen that the experimental points are grouped in a satisfactory way about the straight line. The maximum deviation from the results of calculation according to (7) does not exceed 10%. Equations (6) and (7) have been verified in the ranges $2.07\cdot10^3 \le \mathrm{Ar} \le 410\cdot10^3$; $0.53 \le m \le 0.83$; $0.135 \le h/D \le 0.625$; $0.286 \le r/R \le 0.85$.

The heat transfer coefficient during the drying process decreases following the period of the greatest value, reaching the value for the dry bed at $\mathrm{Cu} \sim 10.9\cdot10^{-3}$. The Gukhman number varied in the drying process in the range $4\cdot10^{-3}$ to $40.9\cdot10^{-3}$.

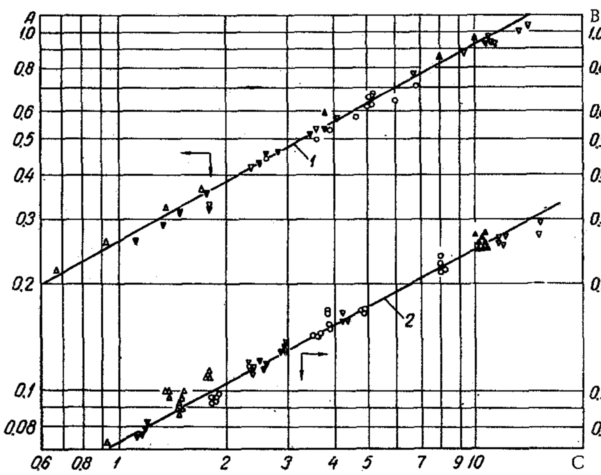

Fig. 4. Heat transfer between the vertical tube and the moist bed ($\mathrm{Nu}/\mathrm{Ar}^{0.28}(1-r/R)^{0.02} \equiv A$; $\mathrm{NuGu}^{0.24}/\mathrm{Ar}^{0.28}(1 - r/R)^{0.02} \equiv B$; $m^3/(1 - m)^2 \equiv C$); 1) with $\mathrm{Gu} < 4\cdot10^{-3}$; 2) $4\cdot10^{-3} < \mathrm{Gu} < 10.9\cdot10^{-3}$.

By the same method of successive introduction of parameters we obtained equations describing the heat transfer in this period of the drying process. They have the form:

for a horizontal arrangement of the heated element

$$\mathrm{Nu} = 0.0604\,\mathrm{Ar}^{0.32}[m^3/(1-m)^2]^{0.445}\,\mathrm{Gu}^{-0.24}(h/D)^{0.15}, \qquad (8)$$

for a vertical arrangement of the tube

$$\mathrm{Nu} = 0.074\,\mathrm{Ar}^{0.28}[m^3/(1-m)^2]^{0.54}\,\mathrm{Gu}^{-0.24}(1-r/R)^{0.02}. \qquad (9)$$

Equation (9) is represented by curve 2 in Fig. 4. The deviation of the test points from the results of calculation do not exceed 10.2% for the horizontal arrangement, and 15.2% for the vertical arrangement.

Equations (6)–(9), which describe heat transfer between the surface and the moist fluidized bed, were obtained using a bed porosity and a specific weight of particles for both the moist and the dry bed. It turned out that, to describe the heat transfer between the surface and the fluidized bed during drying, we can use the porosity of the dry bed and the specific weight of the dry particles.

The variation of the porosity of the bed and the variation of the specific weight of the particles as a function of bed humidity proved to have opposite effects on the heat transfer in the bed, and to compensate one another.

Equations of heat transfer between the surface and a dry bed of particles of silica gel were obtained by the same method as was used for the heat transfer equations during the drying process.

For a horizontal heater arrangement an equation was obtained in the form

$$Nu = 0.197\ Ar^{0.32}\,[m^3/(1-m)^2]^{0.445}\,(h/D)^{0.15}. \quad (10)$$

The greatest deviation of the experimental points from those calculated according to (10) did not exceed 10%.

For a vertical tube the heat transfer equation has the form

$$Nu = 0.226\ Ar^{0.28}\left[\frac{m^3}{(1-m)^2}\right]^{0.54}\left(1-\frac{r}{R}\right)^{0.02}. \quad (11)$$

The maximum deviation of the experimental points from those calculated according to (11) does not exceed 15.6%.

NOTATION

α—heat transfer coefficient; Q—amount of heat transferred; Δt—temperature difference; λ—thermal conductivity of medium; d_e—particle equivalent diameter; φ_k—particle shape factor; m—porosity of bed; F—area of tube heat transfer surface; γM—specific weight of material; γ_g—specific weight of gas; ν—kinematic viscosity of gas; h—height of horizontal tube above distributor; r—radius of position of axis of vertical tube; R, D—radius and diameter of bed; T_c—temperature of air leaving bed; °K; T_w—wet bulb temperature; $Nu = \alpha\delta/\lambda$—Nusselt number; $\delta = (2/3)\varphi d_e[m/(1-m)]$—diameter of equivalent channels; $Ar = g(\varphi_k d_e)^3\,\gamma_M/\nu^2\,\gamma_g$—Archimedes number; φ_2—relative humidity of gas leaving bed.

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