WAYS OF INTENSIFYING DRYING PROCESSES

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This paper gives some results of investigations aimed at finding ways of increasing the rates of heat and mass transfer in drying.

Simulation of naphthalene sublimation by means of a built-up naphthalene slab composed of narrow prisms was used to determine the local values of the heat-transfer coefficients by weighing or measurement of the reduction in thickness of the naphthalene slab [1-3]. The range of the investigation was varied by altering the nozzle slit width b = 5-40 mm and the flow velocity w_0 = 10-40 m/sec ($Re_0 = 6.04 \cdot 10^3$-$3.78 \cdot 10^4$). The distance between the plane of the nozzle mouth and the slab surface varied in the range s = 10-360 mm, and the dimensionless distance s/b varied in the range 0.25-40.

We obtained two types of characteristics (types A and B), shown in Fig. 1, for the local mass-transfer coefficients β_l. The behavior of β_l indicates that, except in the case of very small distances between the nozzle mouth and the surface (region s < b), the maximum values of β_0 occur in the plane of the flow axes. At such small distances we are no longer dealing with a flow perpendicular to the wall, but with a flow parallel to the wall – the "wall flow."

We found that the values of β_0 are not always inversely proportional to the distance between the nozzle mouth and the surface and that these relationships at small distances descend smoothly to a certain minimum. As the nozzle mouth approaches the surface the value of β_0 again increases. If we denote the maximum

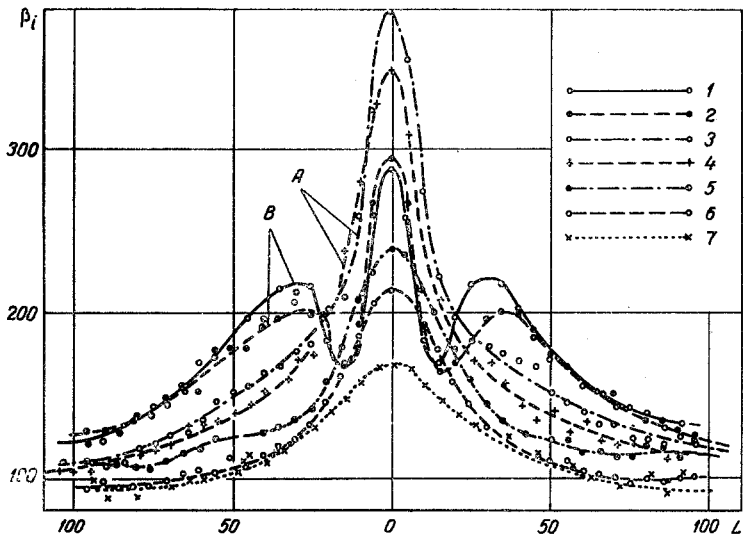

Fig. 1. Characteristics of local transfer coefficients β_l below slit nozzle (b = 5 mm) for different distances s (w_0 = 30 m/sec): 1) s = 1 cm; 2) 2; 3) 4; 4) 6; 5) 10; 6) 12; 7) 20.

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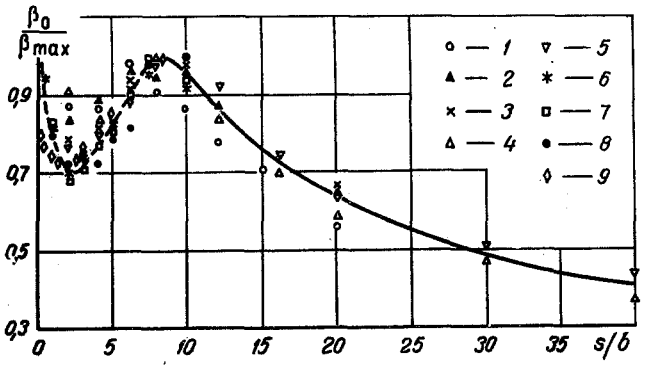

Fig. 2. Plot of β_0/β_{max} against ratio s/b: 1) w_0 = 10 m/sec; 2) 20; 3) 30 (b = 10 mm); 4) 20 m/sec; 5) 30 (b = 5 mm); 6) 10 m/sec; 7) 20; 8) 30 (b = 20 mm); 9) 18.85 m/sec (b = 40 mm).

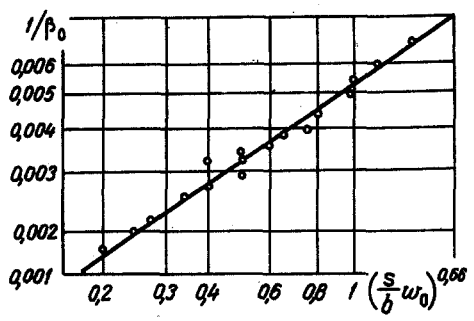

Fig. 3. Plot of $1/\beta_0$ against $(s/b)^{-0.66}$ for case b = 10 mm, and w_0 = 10-40 m/sec. Line obtained from formula $\beta_0 = 29.9 w_0^{0.66}(s/b)^{-0.66} b^{-0.33}$.

value of β_0 as β_{0max}, all the experimentally obtained values can be represented by a single characteristic – the relationship $\beta_0/\beta_{0max} = f(s/b)$ (Fig. 2). One extremum is close to s/b = 2.5 (minimum) and the other to 8.5 (maximum). The value s/b = 8.5 is also the boundary for differentiation of the type A and B curves, as Fig. 1 shows.

We paid particular attention to the region s/b > 8.5, for which we obtained an expression for β_0 as a function of w_0 and b (Fig. 3)

$$\beta_0 = 29.9 w_0^{0.66}\left(\frac{s}{b}\right)^{-0.66} b^{-0.33}.$$

The experimental results were later used to derive a dimensionless relationship for the mean mass-transfer coefficients β_m, which were obtained by planimetry of the β_l characteristic in strips of width L = 0.05, 0.08, 0.1, and 0.15 m (corresponding to L/b = 1.25-30).

The final result of this treatment is the diagram shown in Fig. 4 and the dimensionless equation

$$sh = k\,Re^{0.77}\,sc^{1/3},$$

k also depended on the width b:

$$k = 0.2 b^{-0.155}.$$

The results of this investigation were used to select drying regimes for some materials [4-7].

Spray dryers have been used successfully in various branches of industry, although a drawback of these dryers is the limitation on the initial concentration of material, which necessitates drying chambers of large and often unsuitable dimensions.

The drying rate in spray dryers can be increased in several ways:

a) by reduction of the size and attainment of maximum uniformity of the spray;

b) by an increase in spray density in the chamber, which, however, necessitates an increase in the flow velocity of the drying agent and, hence, a more efficient separation device;

c) by more effective interaction in the spraying device by improving the method of supply and flow of the drying agent in the drying chamber and reliable mixing of the sprayed emulsion with the drying agent;

d) by an increase in the initial temperature of the drying agent combined with a simultaneous increase in the difference between the entrance and exit temperatures of the drying agent;

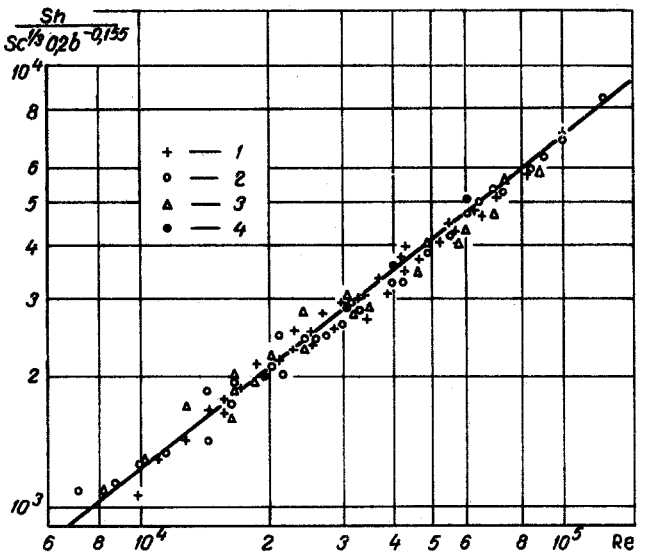

Fig. 4. Plot of $sh/sc^{1/3}$ against Re (line obtained from formula $sh = kRe^{0.77}sc^{1/3}$; $k = 0.2b^{-0.155}$): 1) b = 5 mm; 2) 10; 3) 20; 4) 40.

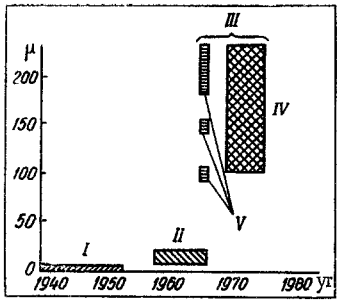

Fig. 5. Possible increase in rate (μ) of process in tube dryers (specific evaporation rate, $kg/m^3 \cdot h$): I) first types of tube dryers; II) experimental types with improved internal aerodynamics; III) spray-type tube dryers; IV) prospective increases; V) results of preliminary tests.

e) by a reduction of the initial moisture content of the sprayed material, which leads to the use of pastelike suspensions.

In Czechoslovakia, attention has been centered on spray drying with an increased flow velocity of the drying agent. The choice of a distributive coil with guide blades for the air feed resulted in a specific evaporation rate of 17 $kg/m^3 \cdot h$ at working temperatures of 165-180°C. The use of spray dryers of this type for the drying of dyes and intermediate dyestuff products improved the quality of the final product [8].

The use of spray-type tube dryers, which should lead to a further increase in the specific evaporation rate (Fig. 5), has recently been investigated. Figure 6 shows the design of an installation in which regulation of the relative amounts of primary air for atomization in the pneumatic nozzle, secondary air for guiding the spray cone, and tertiary air for prevention of contact of the moist particles with the wall, provides a very versatile method of drying. The flow in an apparatus with a spray nozzle of diameter D_t = 2.5 mm, pressure of spray air up to 7 kg/cm^2, and diameter of drying tube 180 mm, has been studied and the spraying and drying of some materials (ceramic slurries, fodder yeast) in such an apparatus have been investigated.

The operation of the spray nozzle and the flow in the drying tube were investigated. Figure 7 shows the velocity distribution of the spray air at pressure 1.2 kg/cm^2, and Fig. 8 shows the ratio of the axial velocity of the free beam $w_{\varkappa,0}$ at distance $\varkappa$ and the velocity in the center of the beam w_0 as a function of the ratio $\varkappa/D_t$.

During the investigation of the evaporation process the temperature of the secondary air was 282-288°C, and that of the tertiary air 120-127°C. This led to a specific evaporation rate of 214-238 $kg/m^3 \cdot h$ and in tests with a washed ceramic mass at 50% concentration the rate was 148.5-156 $kg/m^3 \cdot h$.

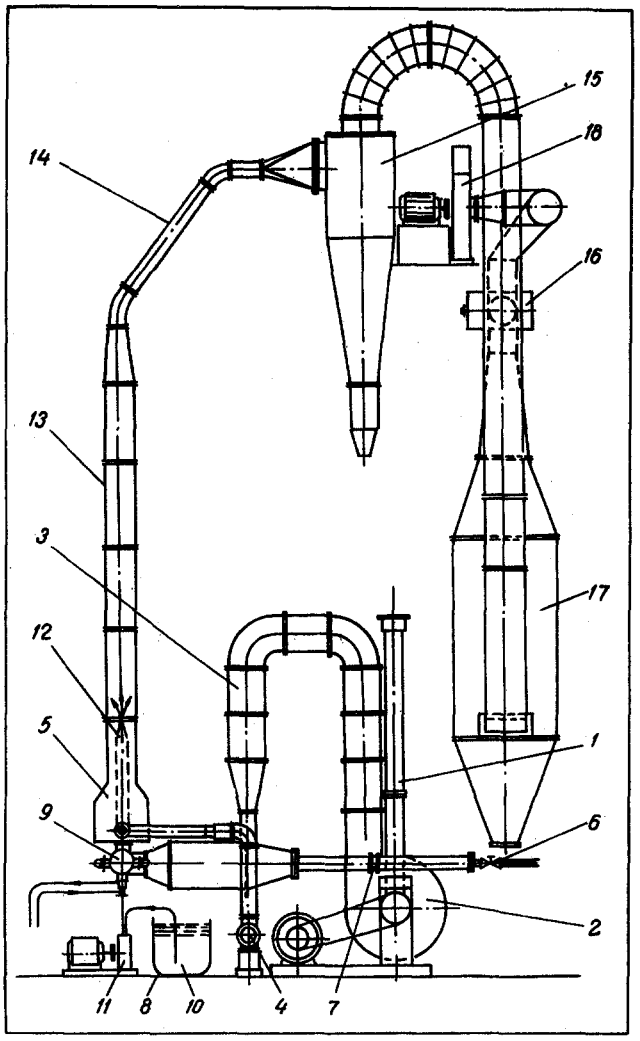

Fig. 6. Experimental installation for investigation of spray drying in tube dryer: 1) secondary-air metering orifice; 2) secondary-air blower with electric motor; 3) electric storage heater; 4) secondary-air supply pipe; 5) secondary-air distributing head; 6) throttle valve for pressurized primary air; 7) primary-air metering orifice; 8) electric storage heater; 9) primary-air distributing head; 10) reservoir for suspension; 11) suspension metering pump; 12) spray jet; 13) demountable drying chamber (tube); 14) connecting pipe; 15) cyclone separator; 16) slide valve; 17) cloth filter; 18) main blower with electric motor.

We experimentally investigated the motion of a heterogeneous gas–solid mixture in the drying region of a tube dryer and its effect on the rate and uniformity of drying.

The drying regime in a tube dryer differs from that in pneumatic conveying not only in the effect of the change of particle mass due to evaporation of moisture, but also in the nonisothermal nature of the process, the lower weight concentration μ (in tube dryers $\mu < 1$ kg/kg, in pneumatic conveying μ = 4-12 kg/kg), and in the very pronounced effect of the acceleration region at relatively small heights of the vertical drying tube.

The experimental investigation was carried out on a tube dryer with a glass tube of diameter D = 100 mm. To measure the absolute particle velocities we used photoelectric sensors located at selected points

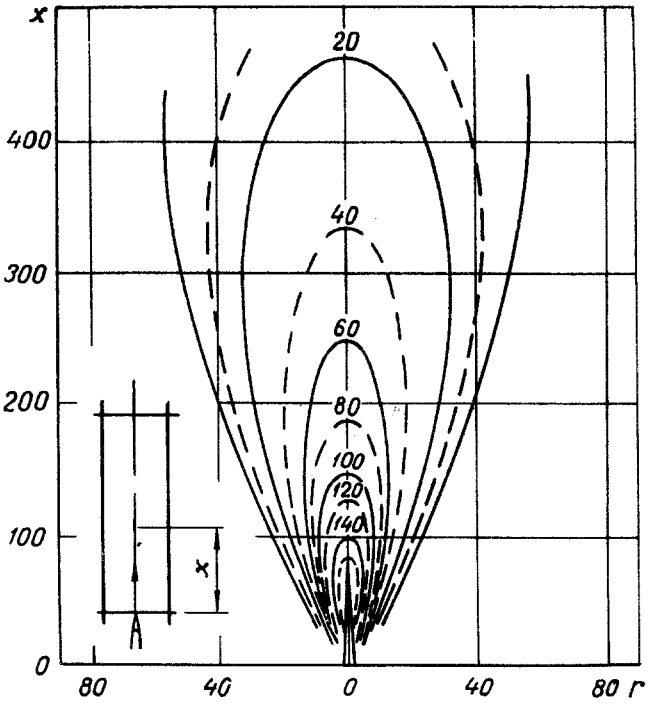

Fig. 7. Spray-air velocity distribution in drying tubes; x is the distance from the nozzle, mm; r is the distance from the axis, mm.

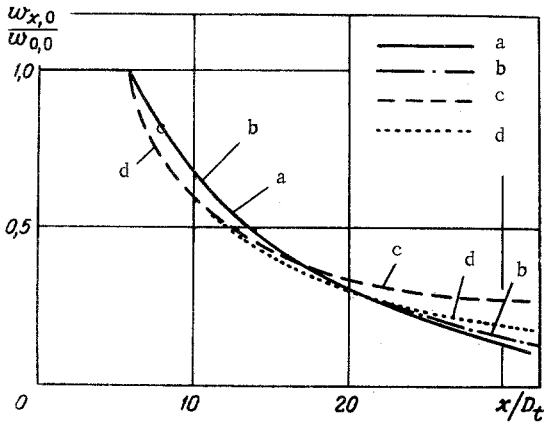

Fig. 8. Plot of spray-air velocities in axis of drying tube: a) measurement, primary air only; b) measurement, primary and secondary air; c) from [9]; d) from [10].

along the drying tube. The sensors were germanium photodiodes, on which a light beam from a source mounted in the opposite side of the tube was incident. The intensity of the flux on the photodiode depended on the passage of material through the drying tube and was recorded by an oscilloscope.

The results of the experimental investigation enabled us to compare the particle velocity characteristics with the characteristics calculated from the equation

$$\frac{d\bar{w}}{d\bar{s}} = \frac{1}{\bar{w}_k}\left[\left(\frac{\bar{v}-\bar{w}}{\bar{w}_k}\right)^{2-\varkappa} - 1 - a_1\bar{w}^2\right]\frac{\bar{w}}{a_2-\bar{u}}\,\frac{d\bar{u}}{d\bar{s}},$$

which was derived for the flow of material in a drying tube with due regard to the change in particle mass due to drying [11].

An example of the experimentally determined and calculated velocity characteristics for a polystyrene suspension in the region 0-5 m from the entrance to the drying tube is shown in Fig. 9. A comparison shows that, in the case of particles of diameter d = 0.9 mm, there were definite differences at the start of the acceleration region; these differences gradually decreased and on attainment of the steady state the calculated velocities were practically the same as the measured velocities. In experiments with particles of diameter d = 2.6 mm the difference between the measured and calculated velocity characteristics was much greater (approximately 30% greater at a distance of 5 m). We subsequently found that the acceleration region was greater, the greater the flow velocity v of the medium and the greater the descent velocity w_k of the material; for particles with descent velocity w_K = 4 m/sec and flow velocity v = 25 m/sec a steady state was attained in a distance of 4.5 m.

Subsequent experimental work was directed towards a test of the effect of polydispersity of the material on the uniformity of drying. The experiments were conducted with polystyrenes, PVC, and common

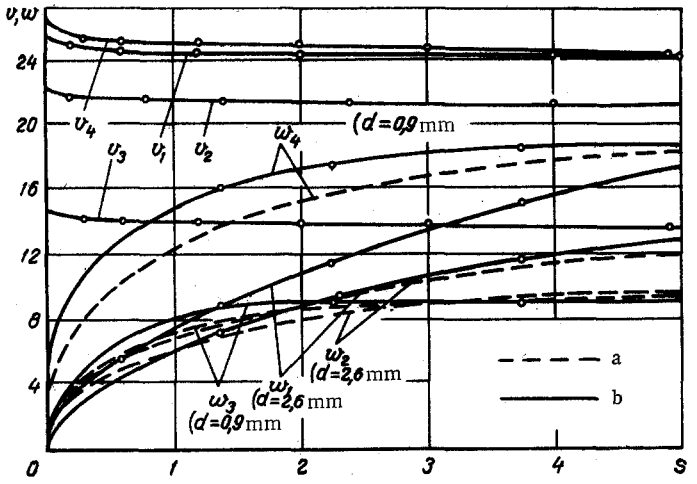

Fig. 9. Comparison of calculated (a) and experimentally determined (b) particle velocities (s, m; v, w m/sec^{-1}).

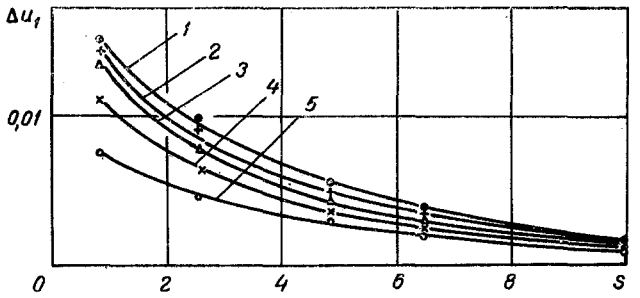

Fig. 10. Experimentally determined relationship for difference in specific moisture content of smallest and largest polystyrene fractions (Δu_1, kg·kg^{-1}; s, m): 1) μ_A = 0.22 kg·kg^{-1}; 2) 0.175; 3) 0.135; 4) 0.1; 5) 0.085.

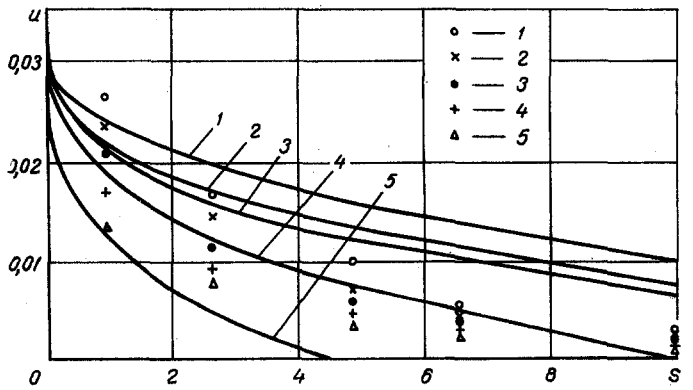

Fig. 11. Drying characteristic for individual polystyrene fractions (t_A = 50°C; v_A = 23 m/sec; μ_A = 0.22 kg/kg): 1) $d_1 = 3.25 \cdot 10^{-3}$ m; 2) $d_2 = 2.75 \cdot 10^{-3}$ m; 3) $d_3 = 2.00 \cdot 10^{-3}$ m; 4) $d_4 = 1.125 \cdot 10^{-3}$ m; 5) $d_5 = 5.00 \cdot 10^{-3}$ m.

salt. The uniformity of drying of individual fractions was determined from samples of material taken from the tube dryer at different distances and graded by sieving. We calculated the drying process for the conditions in the experiments by the method which takes into account the polydispersity of the material [12, 13], and compared the calculated characteristics with the results obtained by measurement.

The experimental results indicate that the difference between the specific moisture content of the largest and smallest fractions was greatest in the region of input of the material (Fig. 10). With increase in distance the difference decreased and the drying of the individual fractions became more uniform.

We also investigated the effect of the weight concentration μ on the uniformity of drying of a polydisperse material. The results indicated that an increase in μ increased the difference between the moisture-dependent maximum and minimum fraction in the region of input of the material. With increase in the distance the effect of μ decreased and 7-8 m from the input point the effect of the mixing ratio on the uniformity of drying of individual fractions was negligible.

A comparison of the experimentally determined and calculated characteristics of drying of individual fractions (Fig. 11) shows that the experimental values give a smaller range of specific moisture content of the individual fractions than the calculated values.

As regards drying in a fluidized bed we think that the more extensive use of this drying method in industry is hampered by some low parameters in comparison with the parameters in dryers based on other pneumatic principles. This is because the attainment of the fluidized state requires a greater amount of gaseous medium than is required in the drying process proper, so that part of the power consumption is completely wasted from the viewpoint of intensification of heat and mass transfer.

A very effective improvement in drying in a fluidized bed can be effected by additional vibrational loosening of the bed. Vibration results in attainment of the fluidized state at velocities below the critical fluidization velocity and, in addition, may have a favorable effect on the drying process.

NOTATION

$\bar{w} = w/v_A$	is the dimensionless velocity of material corresponding to entrance velocity v_A of drying medium;
$\bar{v} = v/v_A$	is the dimensionless velocity of drying medium;
$\bar{w}_k = w_k/v_A$	is the dimensionless descent velocity of particles;
$\bar{s} = sg/v_A^2$	is the dimensionless path;
$\bar{u} = (u_A - u)/u_A$	is the dimensionless specific moisture content of material corresponding to entrance moisture content u_A;
$\varkappa$	is the index;
$a_1 = 4\xi v_A^2/Dg$	is a constant;
$a_2 = (1 + u_A)u_A$	is a constant;
ξ	is the contact friction coefficient of particles.

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