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Drying of suspension and pastes in fluidized bed of inert particles

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A fluid bed dryer with inert particles was used for the drying of suspensions and pastes. The effects of the operating conditions on the dryer throughput and on the product quality were investigated. Experiments were performed in a cylindrical column 215 mm in diameter and 1200 mm in height with 0.925 mm diameter glass spheres as the fluidizing media. Cineb fungicide, copper hydroxide and pure water were used as the feed material. With respect to the main efficiency criteria, *i.e.*, specific water evaporation rate, specific heat consumption and specific air consumption, a fluid bed dryer with inert particles represents a very attractive alternative to other drying technologies. A high drying efficiency results from the large contact area and from the large temperature difference between the inlet and outlet air. A rapid mixing of the particles, due to aggregative fluidization and mechanical agitation, leads to nearly isothermal conditions throughout the bed. In our experiments, suspensions and very dense pastes were successfully treated. Suspension and product hold-up in the bed varies between 6 and 8 % by mass and a product with the same particle size as the raw material is obtained.

Keywords: drying, suspensions, pastes, fluidized, inert particles.

INTRODUCTION

Many processes in the chemical, pharmaceutical and food processing industries involve the drying of solutions, suspensions and pastes. This unit operation is the most energy consuming industrial operation. Although there are hundreds of variants actually used in drying, research efforts all over the world are associated with the development of more sophisticated systems. In general, the trends in drying technology are associated with higher energy efficiency, enhanced drying rates and the development of more compact dryers, better control for enhanced quality and optimal capacity, developments of multi-processing units (for example filter-dryer), *etc*. Mujumdar[1] pointed out that numerous new or improved drying technologies are currently at various stages of development. Particularly, with respect to the drying of solutions, suspensions and pastes, the use of fluidized, spouted, vibrofluidized and agitated beds of inert particles will became very important alternatives to the classic drying technologies.

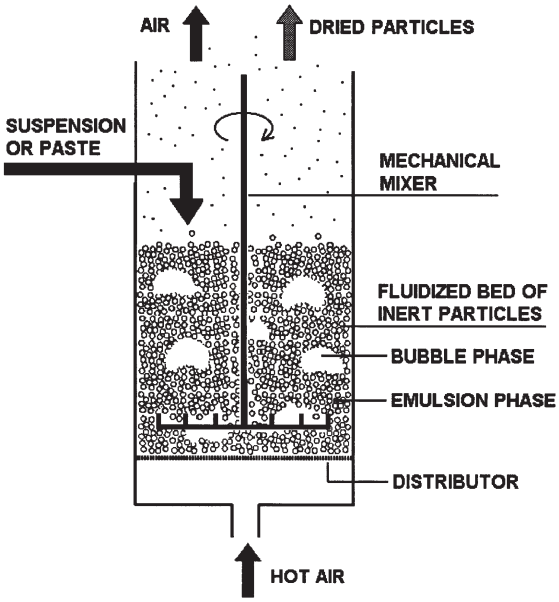

Fig. 1. The drying of suspensions in a fluidized bed of inert particles.

An efficient drying system should meet several conditions: high values of heat and mass transfer coefficients, high contact area, high input of heat carrier gas, uniform temperature distribution over the drying chamber, the use of concentrated suspensions (as high as possible) in order to minimize the amount of water to be evaporated and the use of a high inlet air temperature, as high as possible. However, many of these conditions are conflict with each other. The only drying concept that meets the majority of the mentioned conditions is drying in an agitated bed of inert particles. Several systems (fluidized, spouted, spout-fluidized and different modifications) are at various stages of development.[1–7]

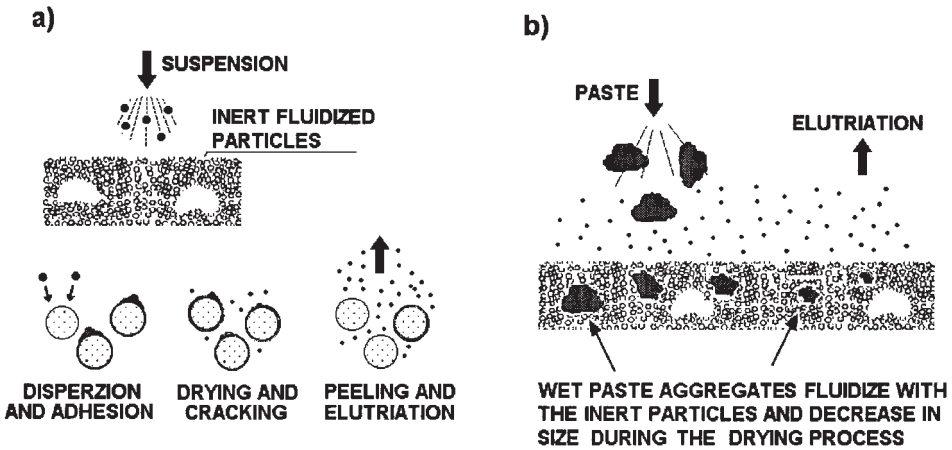

Fig. 2. The drying mechanism: a) feed - suspension, b) feed - paste.

Drying of suspensions or pastes in a fluidized bed of inert particles

A schematic diagram of the proposed drying system is given in Fig. 1. The feed material is directly supplied into the column where the inert particles are fluidized by hot air. The product is separated from the exhaust air by a cyclone and/or bag filter. The drying mechanism depends of the feed slurry density and consistency, as illustrated shematically in Fig. 2. If the feed is a relatively dilute slurry (suspension) the drying mechanism consist of three steps, which occur simultaneously in the different regions of the bed. The charged suspension forms films which adhere to the surface of the inert particles. Because of the very large surface area of the inert particles and the intensive fluidization the moisture is removed in a very short time. The solids remaining on the surface of the inert particles are peeled off by friction and collision. Finally, the product powder is elutrated from the inert bed with the exhaust air. If the feed is a dense slurry (paste) then wet paste aggregates fluidize together with the inert particles. During the drying process the size of the aggregates decreaes due to the elutration of dried particles from the surface. Note that a typicall dry particle is about two orders of magnitude smaller than the inert particles in the bed. In our system, a mechanical mixer of the fluidized inert particles additionally prevents the formation of large aggregates. Due to the rapid mixing of the inert particles during fluidization and mechanical mixing, the bed temperature is approximately uniform.

EXPERIMENTAL

The drying system is schematically shown in Fig. 3. The drying chamber is a cylindrical column D_c = 215 mm i.d. in the lower part and 320 mm i.d. in the upper part. The overall column height is 1200 mm, where the effective column height (above the distributor) is 900 mm. The inert particles are glass spheres with a mean diameter d_p = 0.925 mm and a density 2620 kg/m^3. The inert bed mass is 4.52 kg and the static bed height is H_0 = 80 mm. The total inert particle area is 11.12 m^2. The minimum fluidization velocity was determined at ambient air temperature using standard procedure (U_{mF} = 0.61 m/s).

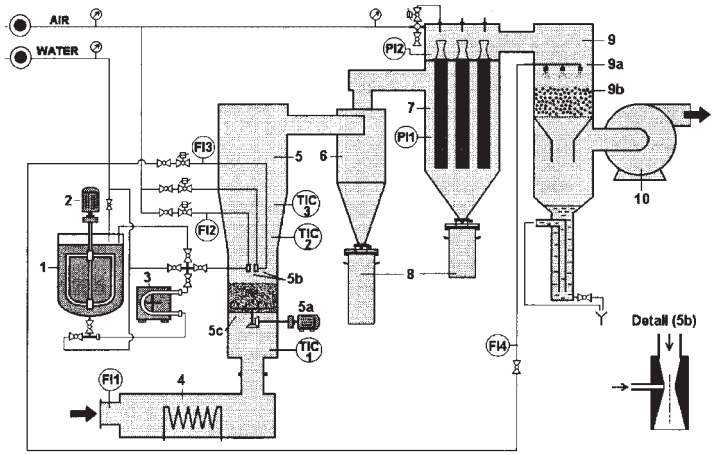

Fig. 3. Schematic diagram of the drying system: (1 – tank, 2 – agitator, 3 – pump, 4 – air heater, 5 – fluidization column, 5a – mechanical mixer, 5b – venturi tube, 5c – distributor, 6 – cyclone, 7 – bag filter, 8 – product containers, 9 – scrubber, 9a – nozzle, 9b – packing, 10 – blower, FI – flowrate indicator, PI – pressure indicator, TIC – temperature indication and control).

In order to prevent particle agglomeration, the bed is equipped with a mechanical mixer (≈100 rpm). The product is separated from the air in a subsequent step by a cyclone and bag filter. Before leaving the system, the air is passed through a packed bed scrubber. The system is equipped with a tank for the slurry with an agitator and with a peristaltic pump as the feeding device. A temperature controller TIC1 maintains the inlet air temperature at the desired level. A thermocontroller TIC2, which is situated 0.5 m above the bed and is connected with a feeding pump, serves to keep the drying temperature constant. Thermocontroller TIC3, which is also placed 0.5 m above the bed, is set at a temperature which is 20 ºC above the drying temperature (TIC2 +20 ºC). Its role is to prevent overheating of the bed, in the case of feed pump failure, by introducing pure water into the system. The feeding line is connected with a Venturi tube placed about 100 mm above the dynamic bed height. The role of the Venturi tube is to disperse the feed slurry into small droplets or aggregates.

During the experiments, the inlet air temperature and drying temperature were continuously recorded using a PC and data acquisition system.

Two groups of experiments were conducted. In the first group the feed was water, while in the second group the feed suspensions were CINEB fungicide [$(CH_2–NH–CS_2)_2–Zn$] and copper hydroxide [$(Cu(OH)_2$].

RESULTS AND DISCUSSION

The drying tests were performed continuously. For all runs, the desired air flowrate and air inlet temperature (TIC1) were selected. When the temperature above the bed (drying temperature) reached the set value (TIC2), the feeding process was begun. In the further process the drying temperature was constant since the thermocontroller TIC2 controls the feeding pump. The full stationary state was reached after several minutes since inlet air temperature had reached the set value TIC1. The system was very stable since during the operation the temperature variations were less than 2 ºC, as illustrated in Fig. 4. In all experiments the water contents in the suspension varied between $x = 0.65$ and $x = 0.75$ kg_{H_2O}/kg_{sus}. Several runs with copper hydroxide were conducted using much smaller water contents ($x = 0.4$ and $x = 0.6$ kg_{H_2O}/kg_{sus}) so that in these runs the feed material was a very dense paste. Each suspension was characterized by the water content, density and particle size distribution, while each dried sample was characterized by the residual water content and particle size analysis. Additionally, when the feed slurry was cineb the active matter content was also analyzed before and after drying.

For the runs where the feed suspension was water, the minimum bed temperature was about 35 ºC (at a constant air flowrate and inlet air temperature). A further increase in the feed water flowrate, *i.e.*, decrease in the bed temperature, lead to agglomeration of the inert particles, causing very poor fluidization. For the runs were the feed material was cineb fungicide and copper hydroxide, the bed temperature was varied between 65 and 115 ºC. Decreasing the bed temperature below about 60 ºC increased the residual water content in the dried product, while bed temperatures above 115 ºC may lead to damage of the active matter.

The specific water evaporation rate (kg_{H_2O}/m^2 h) (feed - pure water) as a function of the temperature difference ($T_{gi} – T_{ge}$), where T_{gi} and T_{ge} are the inlet and outlet air temperature, respectively, are given in Fig. 5. It can be seen that the evaporation for a fixed gas veloxity is directly proportional to the temperature difference. The highest evaporation rate in the performed runs was 647 kg_{H_2O}/m^{2-} h with a superficial air velocity (calculated at 20 ºC) of $U_0 = 2.03$ m/s, at inlet air temperature of $T_{gi} = 245$ ºC and at

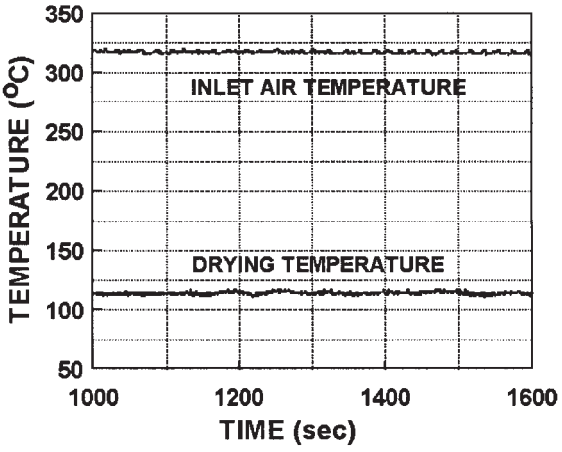

Fig.. 4. Temperature profile during the drying process (feed – cineb).

exit air temperature of $T_{ge} = 67$ °C. The specific heat consumption, calculated on the basis of both the temperature differences $\Delta T_1 = T_{gi} - T_{ge}$ and $\Delta T_2 = T_{gi} - T_0$, where T_0 represents the ambient temperature, is given in Fig. 6. As can be seen, the specific heat consumption q (based on ΔT_1) is approximately independent of the drying conditions and it is slightly above the latent heat of water evaporation. The specific consumption q' (based on ΔT_2) decreases as the temperature difference increases, indicating that overall system efficiency increases with increasing inlet air temperature. The specific air consumption (kg/kg_{H_2O}) decreases with increasing temperature difference ΔT_1, as can be seen from Fig. 7. It can also be seen that the data from 51 runs follow the same line.

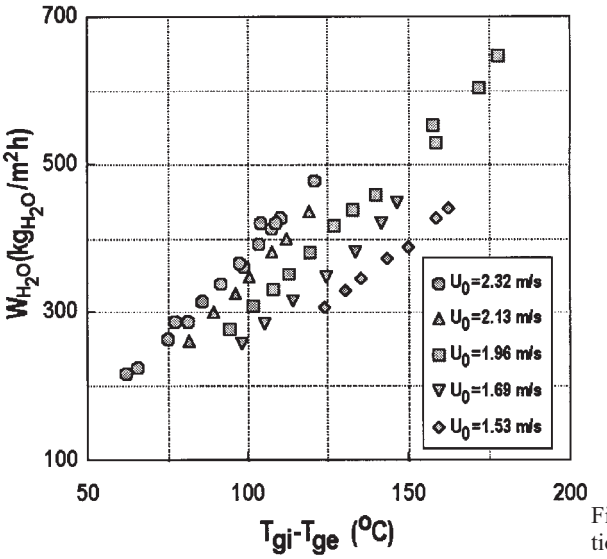

Fig. 5. Specific rate of water evaporation (feed – water).

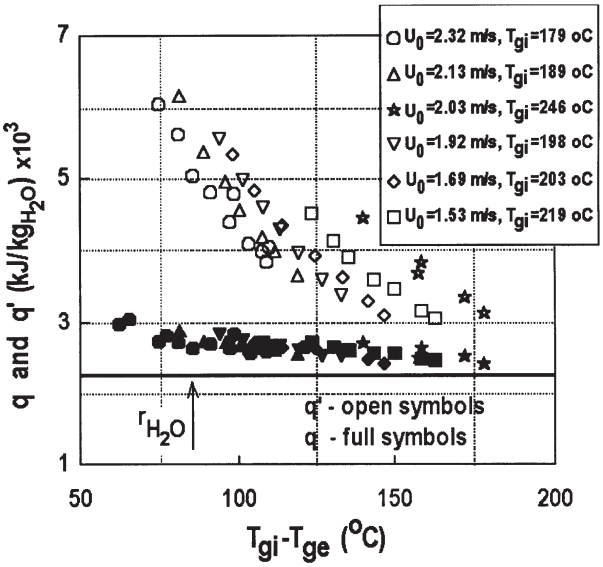

Fig. 6. Specific heat consumption (feed – water).

The corresponding diagrams for the runs where the suspension was cineb fungicide are shown in Figs. 8, 9 and 10. The data scattering in these runs was higher due to the fact that the suspension flowrate was not absolutely constant, since the suspension had a tendency of settling in the feed line. However, in principal, the same conclusion can be drawn from these plots.

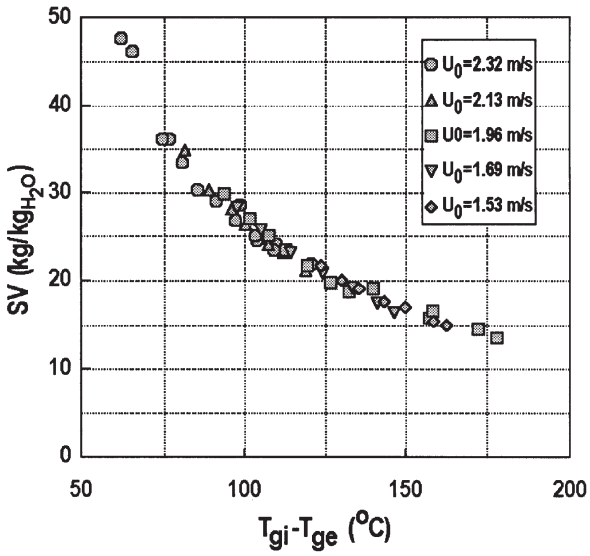

Fig. 7. Specific air consumption (feed – water).

The axial temperature variation in the bed during the drying of copper hydroxide is given in Fig. 11. The thermocouples were placed at the distance of 10 mm from the bed wall. As can be seen, the bed temperature rapidly decreases above bed distributor

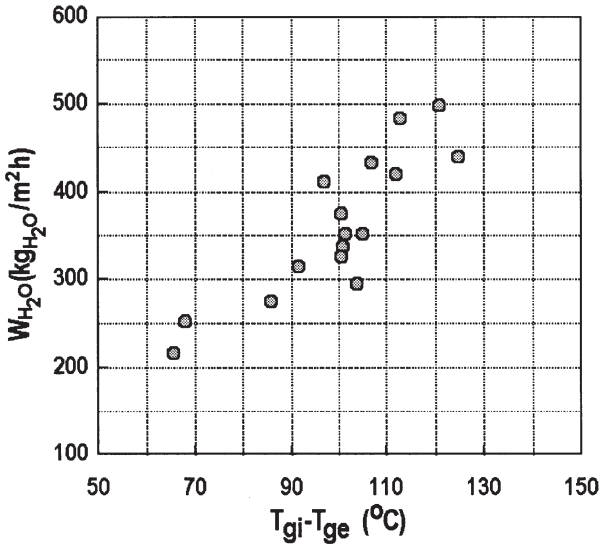

Fig. 8. Specific rate of water evaporation (feed – cineb).

approaching a nearly constant temperature over the major portion of the bed. In this run the inlet air temperature was T_{gi} = 227 ºC. In all the experiments the product had the same particle size as the raw material, due to the effect of collision and friction between the fluidized inert particles. Consequently, for the final formulation of this fungicide, an additional grinding step is unnecessary. Suspension and product hold-up in the bed varies between 6 and 8 % by mass.

For all the runs, the residual water content in the product as well as the content of the active matter were at the desired level. It is important to note that the content of ac-

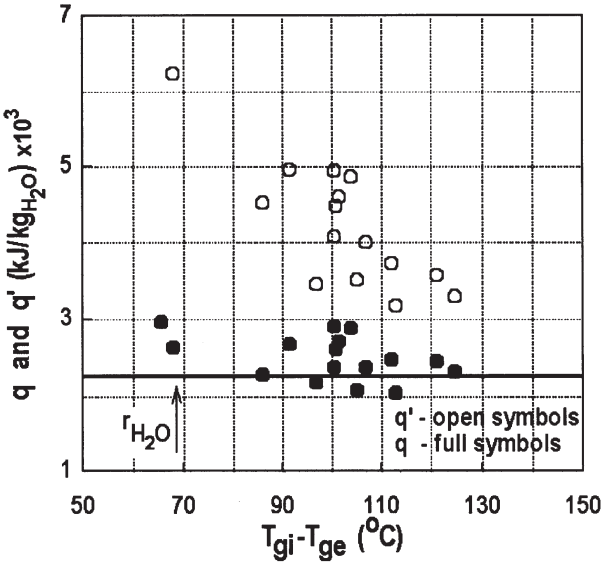

Fig. 9. Specific heat consumption (feed – cineb).

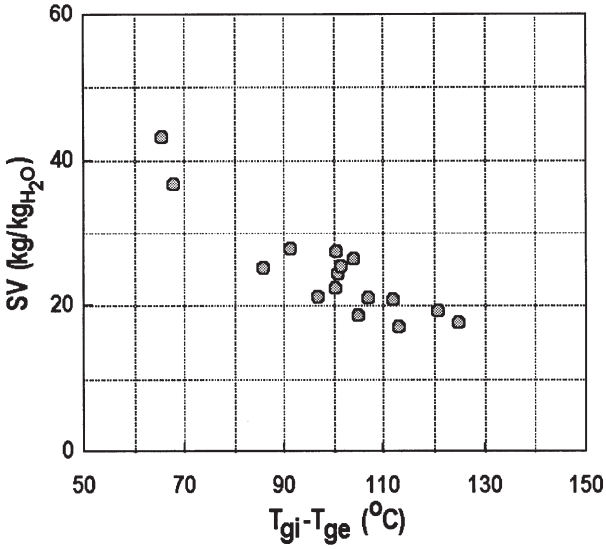

Fig. 10. Specific air consumption (feed – cineb).

tive matter for the cineb fungicide was at least 3 % higher than the corresponding content obtained by drying in an ordinary tunnel dryer. The drying time in a tunnel dryer is 48 h at 80 ºC, while the nominal residence time of the drying particles in a fluidized bed is about 2 min. Table I gives the basic drying parameters for a typical drying run of cineb fungicide (x = 0.7 kg_{H_2O}/kg_{sus}).

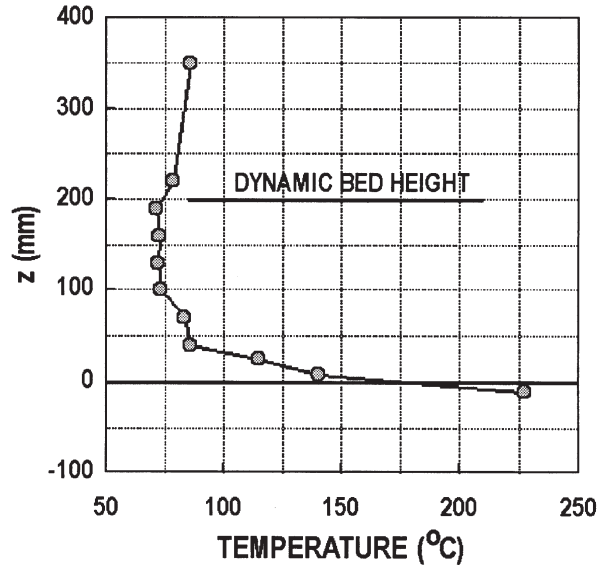

Fig. 11. Axial temperature variation (feed – copper hydroxide, T_{gi} = 227 ºC, T_{ge} = 70 ºC).

TABLE I Typical drying run of cineb fungicide (x = 0.7 kg_{H2O}/kg_{sus}).

Air flowrate (20 ºC)/m^3/h	V_0	270
Superficial air velocity (20 ºC)/m/s	U_0	2.1
Inlet air temperature/ºC	T_{gi}	195
Exit air temperature/ºC	T_{ge}	67
Bed pressure drop/Pa	ΔP	1980
Bed expansion/%	H/H_0	250
Suspension flowrate/kg_{sus}/h	G_{sus}	25.5
Water flowrate (in suspension)/kg_{H2O}/h	G_{H2O}	17.8
Dry matter production/kg_{dm}/h	G_{dm}	7.64
Specific rate of water evaporation/kg_{H2O}/m^2 h	W_{H2O}	491.4
Specific air consumption/kg/kg_{H2O}	SV	18.1
Specific heat consmption (based on $T_{gi} - T_{ge}$)/kJ/kg_{H2O}	q	2405
Specific heat consumption (based on $T_{gi} - T_0$)/kJ/kg_{H2O}	q'	3250
Water content in the product/%	s	0.78
Suspension and product hold-up in the bed/%	h	7.1

MATHEMATICAL MODEL

Theoretically, the water evaporation capacity can be determined from the overall heat balance:

$$G_v c_v(T_{gi}-T_{ge}) = G_{dm}c_{dm}(T_{ge}-T_0) + G_{H_2O}[c_{H_2O}(T_{ge}-T_0) + r_{H_2O}]+Q_g \quad (1)$$

where G_v –air mass flowrate, G_{dm} – mass flowrate of dry matter, G_{H_2O} – water mass flowrate, r_{H_2O}– latent heat of water evaporation and Q_g – heat losses. Since

$$G_{sus} = G_{dm} + G_{H_2O} \quad (2)$$

and if the water content is defined as $x = G_{H_2O}/G_{sus}$, it follows that $G_{dm} = (1-x)G_{sus} = [(1-x)/x]G_{H_2O}$. Using these relationships, Eq. (1) becomes

$$W_{H_2O} = \frac{G_{H_2O}}{A_c} = \frac{1}{A_c} \cdot \frac{G_v c_v (T_{gi} - T_{ge}) - Q_g}{\left[(1-x)/x\right] c_{dm} (T_{ge} - T_0) + c_{H_2O} (T_{ge} - T_0) + r_{H_2O}} \quad (3)$$

where W_{H_2O} is the specific rate of water evaporation and A_c is column cross-sectional area. For a fixed geometry of the fludized bed (A_c), the air flowrate, *i.e.*, superficial air velocity, follows from the fluid bed mechanics[8] and usually should be (3 ÷ 4) times higher than minimum fluidization velocity (U_{mF}). Since the drying temperature (T_{ge}) is selected according to the thermal stability of the drying material, Eq. (3) gives the simple relationship between inlet air temperature (T_{gi}) and evaporation capacity (G_{H_2O}).

A comparison between experimental and calculated values of W_{H_2O}, using an estimated value of $c_{dm} \approx 1200$ kJ/kg ºC, is shown in Fig. 12. As can be seen, the agreement is very good, although the calculated values are systematically slightly higher than the experimental ones (4 % on average) due to the fact that heat losses were neglected in the calculations.

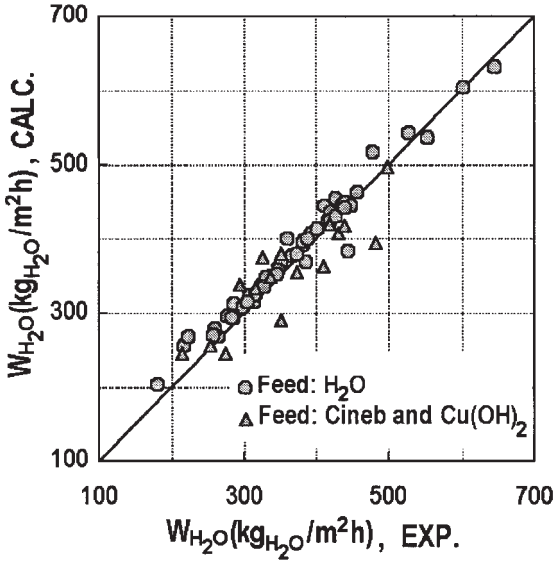

Fig. 12. Comparison of the experimental and calculated values of the specific rate of water evaporation, W_{H2O}.

CONCLUSIONS

The drying of solutions, suspensions and pastes in a fluidized bed of inert particles is a simple and very effective technique for all materials that do not adhere to the inert particles. This drying concept has some important advantages compared to other drying systems, such as: higher capacity per unit volume of the dryer, lower energy consumption and lower specific air consumption. The high drying efficiency results from the large contact area and from the large temperature difference between the inlet and outlet air. The rapid mixing of the particles, due to the aggregative fluidization and mechanical agitation, leads to nearly isothermal conditions throughout the bed. In our experiments suspensions and very dense pastes were successfully treated. The suspension and product hold-up in the bed varies between 6 and 8 % by mass and the dried product has the same particle size as raw material.

Acknowledgment: The financial support of the Research Council of Serbia is gratefully acknowledged.

LIST OF SYMBOLS

A_c – cross-sectional area of the column, m^2
c_{dm} – specific heat of dry matter, kJ/kg ºC
c_{H2O} – specific heat of water, kJ/kg ºC
c_v – specific heat of air, kJ/kg ºC
d_p – inert particle diameter, m
D_c – column diameter, m
G_{dm} – mass flowrate of dry matter, kg/s
G_{H_2O} – water mass flowrate, kg/s
G_{sus} – suspension flowrate, kg/s

G_v – air mass flowrate, kg/s
h – suspension and product hold-up in the bed, %
H_0 – static bed height, m
H – dynamic bed height, m
ΔP – bed pressure drop, Pa
q – specific heat consumption, based on $T_{gi} - T$ge, kJ/kg$_{H_2O}$
q' – specific heat consumption, based on $T_{gi} - T_0$, kJ/kg$_{H_2O}$
Q_g – heat losses, kJ/s
r_{H_2O} – latent heat of water evaporization, kJ/kg$_{H_2O}$
s – product moisture content, %
SV – specific air consumption (G_v/G_{H_2O}), kg/kg$_{H_2O}$
T_{gi} – inlet air temperature, ºC
T_{ge} – outlet air temperature, ºC
T_0 – ambient temperature, ºC
U_0 – superficial fluid velocity (at T_0), m/s
V_0 – air flowrate (at T_0), m^3/s
W_{H_2O} – specific evaporation rate of water (=G_{H_2O}/A_c), kg/m^{2}s
x – water content in the suspension (G_{H_2O}/G_{sus}), kg/kg
z – vertical coordinate, m

ИЗВОД

СУШЕЊЕ СУСПЕНЗИЈА И ПАСТА У ФЛУИДИЗОВАНОМ СЛОЈУ ИНЕРТНОГ МАТЕРИЈАЛА

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Извршени су огледи сушења суспензија и паста у флуидизованом слоју инертног материјала. Испитиван је утицај оперативних услова на капацитет сушионика и квалитет продукта. Експерименти су извршени у цилиндричној колони пречника 215 mm и висине 1200 mm. Као инертне флуидизоване честице коришћене су стаклене сфере пречника 0.925 mm. Третиране су суспензије цинеб фунгицида, бакар хидроксида као и чиста вода. У односу на основне критеријуме ефикасности, тј. у погледу специфичне испарљивости, специфичне потрошње топлоте и специфичне потрошње ваздуха систем сушења у флуидизованом слоју инертог материјала представља врло атрактивну алтернативу осталим техникама сушења. Висока ефикасност сушења последица је велике контактне површине и велике разлике температура између улазног и излазног ваздуха. Интензивно мешање инертних честица услед агрегативне флуидизације и механичког мешања има за последицу приближно равномерну температуру по целом волумену слоја. У нашим огледима успешно су третиране ретке суспензије и врло густе пасте. Hold-up материјала који се суши у слоју креће се од 6 до 8 мас. %, а у свим огледима гранулометријски састав продукта је практично исти са гранулометријским саставом полазне суспензије односно пасте.

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