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MODELLING OF TWO-COMPONENT FLUID BED DRIERS: AN APPROACH TO THE EVALUATION OF THE DRYING TIME

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ABSTRACT

Experimental heat transfer data measured in a laboratory scale fluidised bed have been worked out as reference drying times, in order to predict the performance of two-component fluid bed driers. The effects of the size of the solid to be processed, of bed temperature and of fluidization velocity have been investigated and discussed, and a correlation for the reference drying time is proposed.

INTRODUCTION

Fluidization techniques are widely used in processing particulate solids in the chemical as well as in the pharmaceutical and food industry, due to the high efficiency of the gas-solid contact they achieve. Accordingly, all processes controlled by heat and mass transfer rates, like drying, can be successfully carried out by means of fluid bed contactors.

Fluid bed driers are usually employed to process particulate solids of relatively fine size, because of their low fluidization velocities, but also coarse materials can be efficiently treated in the so called two-component fluidized bed driers, as recently suggested (1,2). Two-component fluidization consists in suspending the relatively coarse solids to be processed in a fluidised bed of fine particles. This technique allows operating the drier with a relatively low flow rate and obtaining very high transfer rates, but undergoes a peculiar fluid dynamic regime with complex interactions between the fine particles bed and the suspended solids.

While the modelling of conventional fluid bed driers has received much attention, so that reliable design criteria have been set up, only a limited effort has been produced on the modelling of two-component driers, whose design is based so far on empirical scale-up evaluations. In this case, very few reliable experimental data on heat and mass transfer coefficients are reported (3,4).

This work aims at discussing the heat transfer properties of fluid bed driers operated under two-component regime and at defining a simple design criterion based on the generalisation of experimental heat transfer data in the form of model parameters. This approach allows to perform a preliminary assessment of the performance of such driers and to evaluate approximately the size of the industrial-scale apparatus for a given set of process requirements.

MODEL DEFINITION

A mathematical model of a two-component fluid bed drier suitable for design purposes should be able to obtain the following results:

a) Evaluation of the drying time for the solid to be processed in the prevailing conditions of the bed;

b) Calculation of the volume of the fluid bed for a given feed flow rate, in batch and in continuous operation.

The approach followed has been as simple as possible, with the objective of utilizing directly experimental heat transfer data measured on laboratory scale, and of embodying them in the model predicting the behaviour of large scale driers.

The following hypotheses have been made:

- The fine solid in the bed is completely mixed, i.e. its temperature is constant throughout the bed.
- The gas in the bed is completely mixed, i.e. its temperature and humidity are constant throughout the bed.
- The drying process follows a pattern which can be represented by a drying rate curve including a constant-rate period followed by a falling-rate period.

In these conditions, drying time can be evaluated from the formula:

$$t_D = - \frac{W_s \, X_0}{A \, R_c} \int_{X_0}^{X_1} \frac{dX/X_0}{R/R_c} = \tau H \qquad (1)$$

where W_s is the mass of the body to be dried, A its external surface, X_0 the initial moisture content and R_c the constant drying rate, while X and R are the current values of moisture content and drying rate respectively.

This expression separates the effect on the drying time of the geometry of the processed body and external transport phenomena from one side (the term τ), and of the internal transport phenomena from the other (the term H).

On this basis, it is possible to evaluate the first term, which can be defined as a reference drying time, as a function of the transport properties of the fine particles bed and of the geometry of the solid to be processed, irrespective of the internal drying characteristics of the solid.

If we express the rate in the constant rate period as a function of heat transfer coefficient, we obtain:

$$R_c = h\ (T - T_{wb}) / \lambda \qquad (2)$$

and thus:

$$\tau = \frac{W_s\ X_0}{A\ h\ (T - T_{wb}) / \lambda} \qquad (3)$$

where T and T_{wb} are dry and wet bulb temperature of the gas respectively, and λ is the latent heat of vaporisation of water.

The total drying time can be thus evaluated by multiplying the term τ and the term H which can be separately obtained from simple laboratory experiments. This method does not imply performing laboratory scale fluid bed drying experiments, but requires heat transfer measurements between a fluid bed of fine particles and a model body of the same shape and size of the processed body and drying experiments on the real body in a forced air drier, with the further assumption that in these conditions the drying curve keeps the same shape as in the fluid bed drier.

The process time for a batch apparatus is the drying time t_D, while for a continuous process the calculated drying time must be taken as the residence time of the processed solid, and accordingly the bed volume is:

$$V = \frac{t_D\ F_s}{\rho_s\ C_s} \qquad (4)$$

being F_s the solid flow rate, ρ_s its density and C_s the allowable concentration of coarse solids in the fluid bed. For most two-component apparatus, due to the difference in size between coarse and fine solids, the residence time of the processed solid can be taken as constant, being controlled by external devices, such as perforated transport ribbons. For fluid bed systems in which the solid discharge is obtained by overflow, allowance must be made for the residence time distribution, on the basis of models well established in process engineering, and bed volume must be larger.

EXPERIMENTAL

Experiments have been planned in order to obtain reliable data on the reference drying time defined in equation (1). A simple geometry of immersed bodies has been selected at first: spheres of different size, fixed in the bed at a definite position.

The experimental apparatus consists of a fluidization vessel 150mm ID which can be heated up to a prefixed temperature through an external oven and the parallel preheating of the fluidizing air.

Heat transfer coefficients are measured by working out the unsteady-state temperature rise curve of a copper sphere immersed in the bed, on the hypothesis that the thermal conductivity of the sphere is so high that the transfer is completely controlled by the external heat transfer.

The bed consists of sand particles, 200-400 μm size, with a height of 300 mm. The sphere is suspended at a distance from the bottom of 200 mm. The location of the sphere has been selected on the basis of previous experiments as the position whose heat transfer properties are closest to average conditions of the bed.

Fluidization velocity is set at different levels, as well as bed temperature in the range of practical interest for drying processes.

EXPERIMENTAL RESULTS

Experiments have been carried out at different levels of the following variables:

- Temperature of the bed: 90, 200 and 300 °C;

- Size of the immersed body: 8, 12, 16 and 20 mm diameter;

- Fluidization velocity: 0.8, 2.6 and 6 cm/s above minimum fluidization velocity evaluated at the bed temperature.

Heat transfer coefficients are evaluated as described above in all the different experimental conditions listed. Their dependence on sphere size for different bed temperatures is represented in Figure 1. It appears that heat transfer is enhanced by temperature rise, also in a range in which radiation has a negligible contribution, while the effect of increasing the size of the immersed body is that of reducing heat transfer coefficients, as already observed in previous experiments (2).

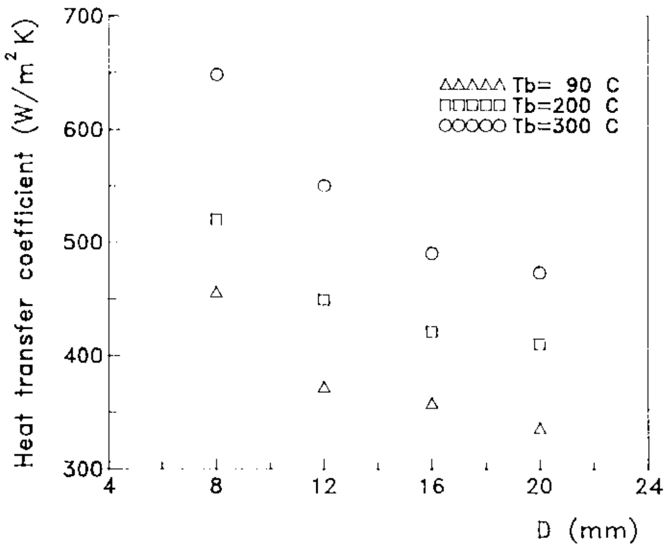

FIGURE 1. Heat tranfer coefficient between immersed spheres and fluidised bed as a function of spheres size, at different bed temperatures. $U-U_{mf}$= 2.6 cm/s.

Heat transfer data have been worked out as reference drying times , with reference values of $X_0 = 2$ kg_w/kg_{ds} and $(T - T_{wb}) = 30\,°C$. Figures 2, 3 and 4 report the values of the drying times as a function of the different experimental variables.

The effect of the size of the immersed body on the reference drying time is well defined for a given fluidization velocity (Fig.2). In accordance with the theory of fluidization (5), the effect of increasing the flow rate in excess of minimum fluidization is that of increasing heat transfer coefficients at first, while at higher velocities coefficients tend to level off as a consequence of the increasing fraction of bubbles in the bed, whose contribution to heat transfer is very poor (Fig.3).

The effect of temperature on drying time (Fig.4) is coherent with the trend of heat transfer coefficient reported in Figure 1. Some discrepancies in experimental data are probably due to local fluctuactions of bubble pattern in the bed.

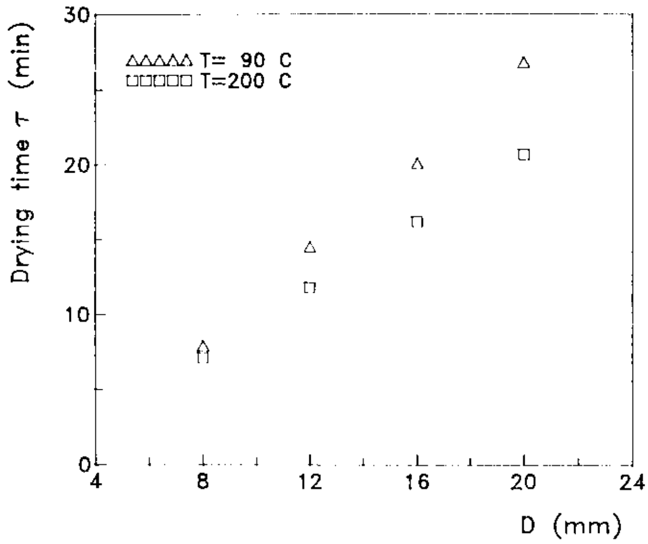

FIGURE 2. Reference drying time as a function of sphere size at different bed temperatures. $U-U_{mf}$= 2.6 cm/s.

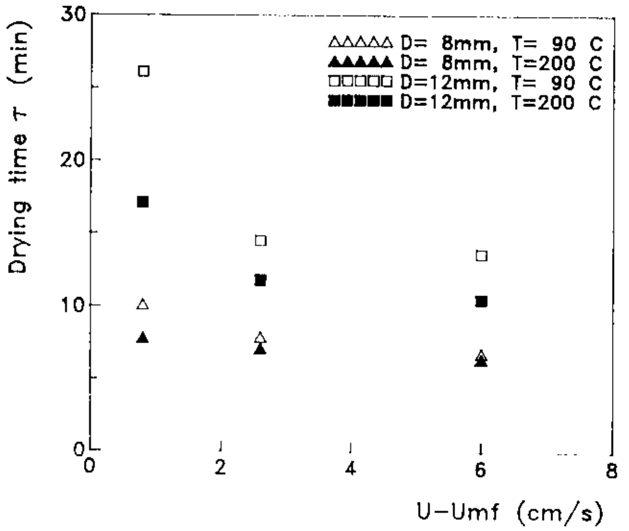

FIGURE 3. Reference drying time as a function of excess fluidization velocity for two different sphere sizes and two different bed temperatures.

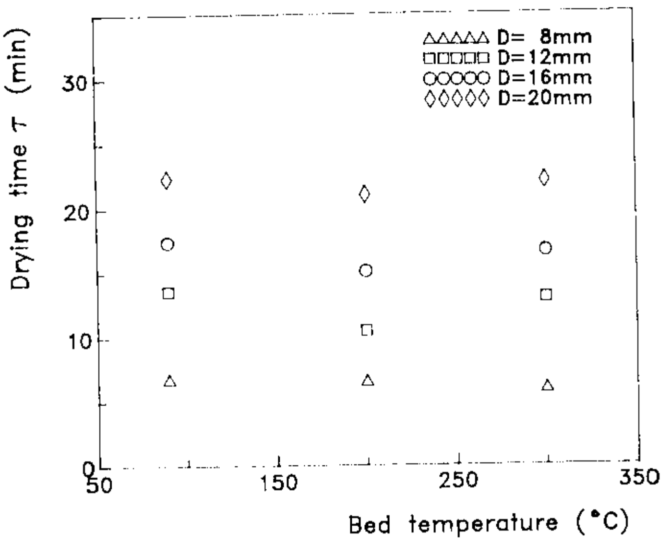

FIGURE 4. Reference drying time as a function of bed temperature for different sphere sizes $U-U_{mf}$= 6 cm/s.

DISCUSSION

Different theories have been developed to represent heat transfer phenomena in fluidised beds. They are mainly concerned about heat transfer between large surfaces and the bed and are based on different hypothesis about the limiting resistance to the transfer. For the case of relatively small bodies surrounded by bed particles, it is likely that heat transfer is controlled by the renewal of the emulsion of fine particles and gas at the surface. This hypothesis, developed through the application of the unsteady-state heat transfer theory, brings to the following expression (6):

$$h = (k_e \rho_e c_{ps}/ \pi t)^{1/2} \qquad (5)$$

where k_e and ρ_e are the thermal conductivity and the density of the emulsion phase of the fluid bed and c_{ps} is the specific heat of the solid, while t is the residence time of the fine particles packets at the surface of the body. The last variable is strongly dependent on fluid dynamic regime of the bed.

By assuming that the process of particles renewal at the surface is an uniform phenomenon, and that all particles contact the surface for the same time, flowing downwards with constant velocity, their residence time on the surface of the sphere has been calculated according to the model description of the fluidised bubbling bed proposed by Kunii and Levenspiel (5):

$$t = b\ D/(U - U_{mf}) \qquad (6)$$

being b a constant depending on fine particles properties and D the sphere diameter.

Inserting this expression into equation (5), and substituting in turn it into equation (3) specified for a sphere, we obtain the following expression for the adimensional drying time :

$$\tau = a \frac{D^{3/2}}{(U - U_{mf})^{1/2}} \qquad (7)$$

where the term a embodies the effect of bed properties and bed temperature, while the dependence on fluidization velocity and on sphere size is expressed explicitely.

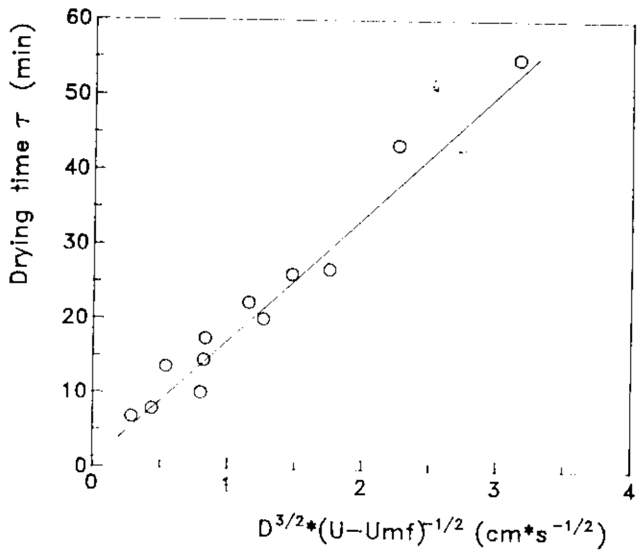

FIGURE 5. Correlation of reference drying time with fluidization excess velocity and sphere diameter according to equation (7). T_b= 90 °C.

According to the interpretation above, for a given temperature and a given set of fine particles bed characteristics, reference drying times should be correlated by a simple power law to the sphere diameter and to the excess of fluidization velocity. Data reported in Figure 5 confirm this finding, being all correlated by a straight line with a small dispersion in a coordinate system based on relationship (7).

CONCLUSIONS

The experiments performed on heat transfer between fluidised beds of fine particles and immersed spheres have elucidated the effect of different experimental variables, such as the size of the immersed body, the bed temperature and and the excess of fluidization velocity in respect to minimum fluidization velocity. Data have been succesfully compared with the theory which considers as controlling step in the heat transfer the mechanism of particle renewal at the surface.

Moreover, heat transfer data have been worked out as a generalized drying time to be used in the modelling of two-component fluid bed driers according to the design approach proposed in this work. This approach allows to convert data measured on small scale apparatus operating on model solids in pure heat transfer conditions, into a transfer parameter easy to use in a design procedure.

LIST OF SYMBOLS

A	surface area, m^2
C	volume concentration, m^3/m^3
c_{ps}	specific heat, joule/kg C
D	diameter of the sphere, m
h	heat transfer coefficient, Watt/m^2 C
H	adimensional variable in eq. (1)
k	thermal conductivity, Watt/m C
R_c	constant drying rate, kg /m^2 s
t_c	residence time, s
t_D	drying time, s
T	temperature, °C
U	flow velocity, m/s
W	mass of the solid, kg
X	moisture content, kg water/kg dry solid
λ	latent heat of vaporization, joule/kg
ρ	density, kg/m^3
τ	reference drying time, s

Subscripts

e emulsion phase of the fluid bed
mf at minimum fluidization conditions
s solid
wb at wet bulb conditions
0 at initial conditions

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