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DRYING PHENOMENON IN A FIXED BED UNDER THE BIO FUEL MULTI STAGE DRYING

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ABSTRACT

Drying of solid particles is generally divided into three periods: a short initial heating period, a constant drying rate period, and a falling drying rate period. Drying in a fixed under various drying conditions was experimentally studied. The main goals of the measurements were to calculate the heat transfer coefficient based on the measured values, and to evaluate the validity of the constant drying rate period. The exact duration of the constant drying period concerning the entire bed is short. The calculated heat transfer coefficients seem to be somewhat reasonable.

NOMENCLATURE

A	surface (m^2)
c_p	specific heat (J/kgK)
D_{eff}	effective diameter of particle (m)
h	specific enthalpy (J/kg)
l_v	vaporisation heat of water (J/kg)
Le	Lewis number (-)
$\dot{m}$	mass flow rate (kg/s)
M	molar mass (kg/mol)
p_o	atmospheric pressure (=101.3 kPa)
p_v	partial pressure of vapour (Pa)
p'_v	saturation partial pressure (Pa)
R	gas constant (=8,314 J/molK)
t,T	temperature (^{o}C), (K)
u	moisture of sample (kg/kg$_{d.m}$)
V	Volume (m^3)
x	absolute moisture of air (kg/kg$_{d.a}$)
Greek letters	
α	heat transfer coefficient (W/m^{2}K)
ρ	density (kg/m^3)
τ	time (s)
Subscript	
0	initial
a	air
d.a	dry air
d.m	*dry mass*
e	evaporation
in	inlet
out	outlet
s	particle surface
v	vapour
w	water
wb	wet bulb

INTRODUCTION

In Finland bio fuel represents approximately 19 % of the total primary fuel consumed in forest industry. The most common bio fuels are bark, forest residues, and different kinds of waste wood. The moisture content of bio fuel typically varies between 50 and 63 w-% (water per total mass), considerably reducing the power production of the power plant. Compared to completely dry fuel the theoretical decrease in power production is 12 -21 % for the initial moistures 50 - 63 w-%, respectively.

A multi stage drying system consists of several drying stages between which the same air flow is heated. Figure 1 illustrates in a Mollier diagram the change of the air moisture in multi stage drying having three drying stages compared to one stage drying. The change of air moisture is the same in both cases but the maximum air temperatures are lower in multi stage drying than in one stage drying. Because of the lower drying temperatures, it is possible to utilise lower temperature energy sources such as secondary heat flows in the heating of the drying air than in one stage drying. This increases the electricity production of the combined heat and power plant.

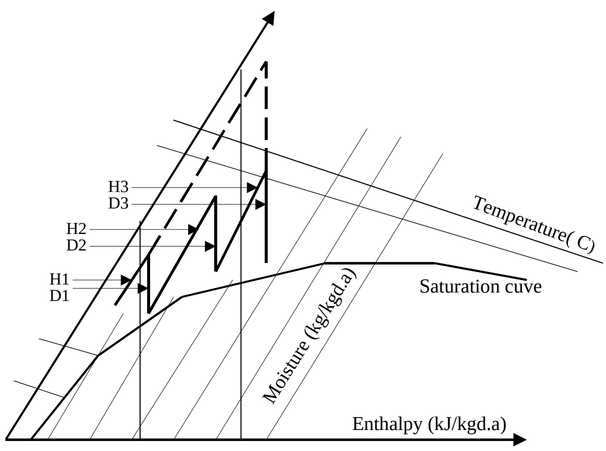

Figure 1- The change of air moisture in multi stage drying with three drying stages compared to one stage drying. D = drying period , H = heating period. The broken line represents one stage drying.

Drying phases of solid particles are usually divided into three periods: a short initial heating period, a constant drying rate period, and a falling drying rate period /2/. During the constant drying rate period, the particle temperature and moisture distributions are assumed uniform, and the heat and mass transfer is in equilibrium at the particle surface:

$$\dot{m}_e l_v = \alpha A(t_a - t_s)\,, \qquad (1)$$

where t_s is the surface temperature. During the constant drying rate period, the surface temperature remains constant, and it is theoretically the same as the wet bulb temperature of air. The theoretical wet bulb temperature can be derived from the Equation (1) based on the analogy of the heat and mass transfer /1/ :

$$t - t_{wb} = \frac{M_v}{\rho c_p}\frac{p_o}{RT} Le^{1-n} l_v \ln \frac{p_o - p_v}{p_o - p_v'(t_{wb})} \qquad (2)$$

If all particles in the fixed bed have reached the local wet bulb temperature, the temperature difference in Equation (1) can be replaced as a logarithmic mean value temperature:

$$\Delta t_{\ln} = \frac{(t_{ain} - t_{s1}) - (t_{aout} - t_{s2})}{In \dfrac{(t_{ain} - t_{s1})}{(t_{aout} - t_{s2})}}\,, \qquad (3)$$

where t_{ain} is inlet air temperature, t_{aout} outlet air temperature, t_{s1} surface temperature in the bottom part of the bed, and t_{s2} surface temperature in the upper part of the bed.

During the falling drying rate period, the particle surface temperature rises, and the evaporation rate decreases. The drying rate decreases as a result of internal heat and mass transfer resistances inside the particle. Particularly below the fiber saturation point, which is around 30 % (water per dry basis) for Finnish wood, the drying rate significantly decreases.

To improve the heating value of the bio fuel with reasonable investment costs, it is not necessary to dry it below fiber saturation point. If the initial moisture of the bio fuel is 60 w-% (water per total mass) the relative improvement of the heating value ($MJ/kg_{d.m}$) compared to initial moisture is 18 % and 21,5 % for the final moistures 25 w-% and 10 w- %, respectively. The drying time is, however, considerably longer if the final moisture is 10 w-% instead of 25 w-%

If the average final moisture of the fuel particles is around 25 - 30 w-%, one can speculate about the accuracy of the constant drying rate model used to calculate the evaporation rate from wood particles.

These measurements have two main goals: to define the value of the external heat transfer coefficient between particles and air based on the measured data, and to evaluate the validity of the constant drying rate model by comparing real drying results to results calculated using a theoretical constant drying rate model.

Wood particles of different sizes were dried in a fixed bed reactor under drying conditions corresponding as near as possible to real process conditions.

TEST RIG

The flow sheet of the test rig with measurement points is presented in figure 2. The drying reactor consisted of a glass tube and an aluminium shutter grate through which the drying air was led into the glass tube. The inner diameter of the glass tube was 100 mm, height 400 mm and wall thickness 5 mm. The diameter of the shutter grate holes was 1,5 mm. The air mass flow was adjusted to a desired one by a mass flow control unit.

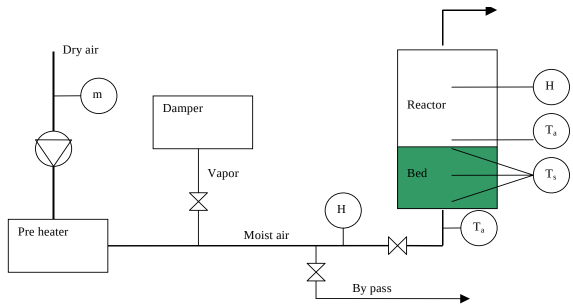

Figure 2 - Flow sheet of test rig with measurement points. H = humidity, T_a = air temperature, T_s = particle surface temperature, m = mss flow control

The air temperatures before and after the bed were measured using thermocouples. There were four temperature measurement points after the bed to take into account the horizontal temperature gradient of the outlet air due to unequal drying rates in the radial direction . The surface temperature of the particle was determined by fixing a thermocouple on the particle surface with studs. The particles with thermocouples were roughly placed in the bottom, mid, and upper parts of the bed. The absolute humidity ($g/kg_{d.a}$) of the air before and after the bed was measured using humidity transmitters.

INITIAL VALUES

Although real fuel particles have no regular shape, it is approximately possible to convert particles of different shapes to a spherical shape by using an effective diameter. The effective diameter is defined as follows/2/:

$$D_{eff} = \frac{6V}{A} \qquad (4)$$

where A is the total evaporation surface of the particle and V the volume of the particle.

Four different effective diameters 5, 10, 15, and 20 mm calculated using the equation (4) were used in the tests. All particles were of regular shape; they were ideal spruce particles specifically made for the laboratory tests. The bed heights were chosen that the particles composed a proper bed in a reactor but the outlet air was not yet saturated after the bed. The real dimensions of the particles and the bed heights used in the tests are presented in table 1.

To equalise the moisture distribution of the test particles they were kept in water for at least one week before the tests. The moisture of the particles was around 40 - 45 w-% before the wetting; after the wetting period, the average moisture varied from 55 w-% to 65 w-% depending on the particle size and the duration of the wetting period. Small particles were usually moister than the large ones.

Table 1 - The real dimensions of the particles and the corresponding bed heights

D_{eff} mm	length mm	width mm	thickness mm	bed height mm
5	10	10	2,5	~30
10	20	20	5	~60
15	30	30	7,5	~100
20	40	40	10	~150

Two different inlet air temperatures, 70 and 120 °C were used. With inlet temperature 70 °C, the air was mostly dry but a series of tests with particle size 10 mm was carried out by moisturizing the air to around 8 $g/kg_{d.a}$ before the drying chamber. With inlet temperature 120 °C the inlet moisture content was around 20 $\pm$ 4 $g/kg_{d.a.}$

Air velocities used in the tests were 0,3 m/s, 0,45 m/s (only with inlet temperature 70 °C), 0,6 m/s, 0,9 m/s and 1,2 m/s. All velocities were calculated per free sectional area of the grate. As air density becomes smaller when the air temperature rises, the mass flows were smaller with inlet temperature 120 °C.

RESULTS

Figure 3 shows how the measured temperatures and moistures change during one hour of drying when particle size is 10 mm, air velocity is 0,45 m/s, inlet air temperature 70 °C and inlet air moisture 0 $g/kg_{d.a}$. It can be seen there is a clear constant drying rate period but its duration is quite short, only a few hundred seconds. The surface temperature in the bottom part of the bed rises relatively rapidly quite soon after the beginning of the drying while the surface temperatures in the mid and upper part of the bed remain constant longer. The moisture content of the outlet air also decreases quite soon after the beginning of the drying and accordingly the temperature of the outlet air increases. Regardless of the drying conditions and the particle sizes, the changes of the measured values were relatively uniform when the Figure (3) is scaled by dividing the drying time by the drying time of each test. Drying conditions and particle size have a considerable effect on the absolute drying time. For example, with an inlet air temperature of 120 °C and air velocity 0,6 m/s, the drying times to moisture content 30 % (water per dry basis) are: 840 s (u_o=2,00), 2840 s (u_o=1,74), 3890 s (u_o=1,70), 4480 s (u_o=1,38) for particle sizes 5, 10, 15, and 20 mm, respectively.

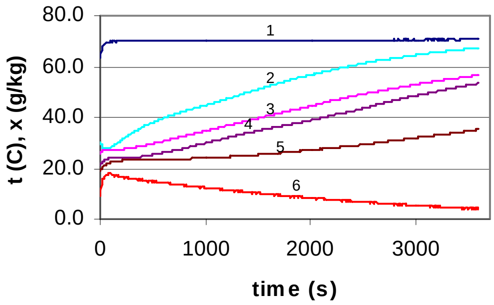

Figure 3 - Changes of the measured temperatures and moisture contents during the drying session. Particle size 10 mm, air velocity 0,45 m/s and the initial moisture content of the sample 1,38. 1 temperature of the inlet air, 2 surface temperature in bottom part of the bed, 3 temperature of the outlet air, 4 surface temperature in middle part of the bed, 5 surface temperature in upper part of the bed, 6 moisture content of the outlet air.

Heat transfer coefficients were calculated based on the measurement data by applying Equations (1) and (3). All temperatures required to define the logarithmic mean value temperature are obtained from the measurement data. The evaporation rate of the bed (kg/s) can be calculated as follows:

$$\dot{m}_e = \dot{m}_{d.a}(x_{out} - x_{in}) \quad (5)$$

The number of particles in the bed was manually calculated. Because of the regular shape of the particles, it was possible to precisely calculate the external evaporation surface.

The temperatures and moisture contents obtained from the measurement data are the mean values for the period when the entire bed is simultaneously or almost simultaneously in the period of constant drying rate. In some cases (mainly with maximum air velocities) the constant drying period in the bottom part was so short that the particles in the upper part had not yet reached the wet bulb temperature which slightly skews the principle of the definition of the heat transfer coefficient. Nevertheless, the heat transfer coefficients have been calculated for each air velocity based on Equations (1) and (3). Figure 4 shows an example of the period when the whole bed is simultaneously in the period of constant drying.

The calculated heat transfer coefficients based on the measurement data are presented in Figures 5 and 6 as a function of air velocity per free sectional area of the grate.

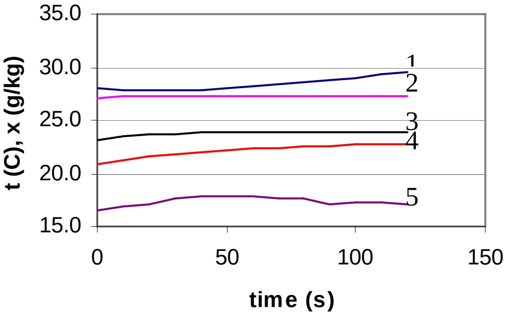

Figure 4 - Period of constant drying rate. Inlet air temperature 70 ^{o}C, inlet air moisture 0 g/kg$_{da}$, air velocity 0,45 m/s and particle size 10 mm. 1 surface temperature in the bottom, 2 outlet air temperature, 3 surface temperature in the middle, 4 surface temperature in the upper, 5 absolute moisture content of the air

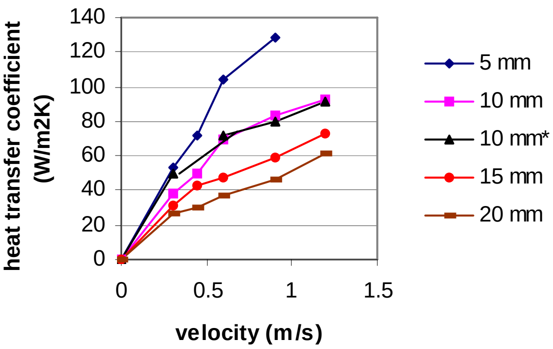

Figure 5 - Calculated heat transfer coefficients for inlet air temperature 70 ^{o}C. Values on the 10 mm* line are the calculated heat transfer coefficients with moist air.

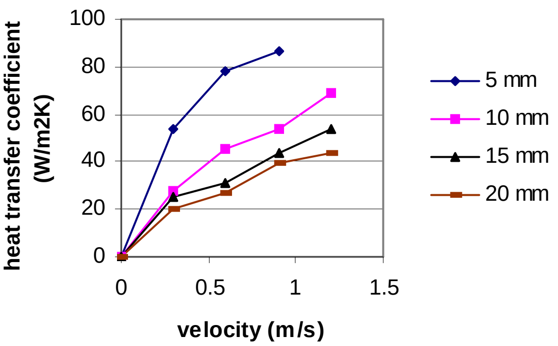

Figure 6 - Calculated heat transfer coefficient for inlet air temperature 120 ^{o}C.

THE EVALUATION OF THE VALIDITY OF THE CONSTANT DRYING RATE MODEL

The validity of the constant drying rate model was evaluated by comparing the theoretical change of the sample moisture to actual change of the sample moisture. To calculate the theoretical drying rate of the bed, it was divided into equal drying layers according to Figure 7. Evaporation rate is calculated for each layer based on the constant drying rate model. The evaporation rate of the entire bed is the sum of the evaporation rates of the layers.

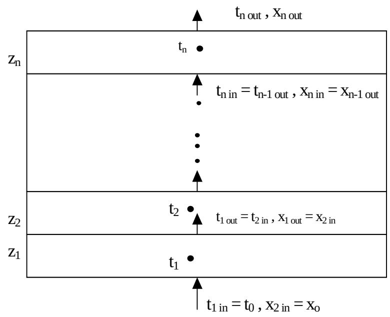

Figure 7 - The principle of the model for calculating the theoretical evaporation rate of the bed.

The following assumptions were used in the calculations:

- The evaporation surface is equal in each layer.
- The surface temperature of the particles in each layer is the same as the theoretical wet bulb temperature of the inlet air
- Surface temperatures remain constant during the entire drying period
- Drying conditions are constant in each layer. They change only between the layers (see Fig. 7)
- Drying rate does not change in the radial direction of the bed
- Drying process is adiabatic
- All particles have an equal initial moisture
- The heat transfer coefficient is the same in each layer
- A short initial heating period is neglected

With these assumptions, the evaporation rate in each layer can be calculated using Equation (1). As the drying process is assumed to be adiabatic, the air temperature after the layer can be calculated by applying the energy balance of the adiabatic moisturising:

$$\dot{m}_{d.a} h_{in} + \dot{m}_e c_{pw} t_s = \dot{m}_{d.a} h_{out} \qquad (6)$$

where h is an enthalpy of the air per kilograms of the dry air. The enthalpy of the wet air (kJ/kg$_{d.a}$) at the particular temperature (^{o}C) in equation (6) is expressed as follows:

$$h = c_{pa} t + x(c_{pv} t + 2501) \qquad (7)$$

The reference point in Equation (7) is water at zero degrees.

The outlet moisture of the air after the layer can be written as follows:

$$x_{out} = x_{in} + \frac{\dot{m}_e}{\dot{m}_{d.a}} \qquad (8)$$

The air temperature in Equation (1) is the mean value of the inlet and outlet air temperature of the layer. When the evaporation rate in Equation (8) is expressed by means of Equation (1) and the Equation (8) is placed in the energy balance of the adiabatic moisturising, where the enthalpy is expressed by the Equation (7), the only unknown term in the equation (6) is the outlet temperature of the air which can be solved. When the outlet temperature of the air after the layer is known the evaporation rate and the outlet moisture of the air can be calculated using Equations (1) and (8). The calculated outlet temperature and moisture are then the new inlet values of the air in the next layer and so on.

As long as the constant drying rate assumption is valid in the entire bed the theoretical outlet moisture of the sample can be calculated as follows:

$$u_{out} = u_0 - \frac{\Delta\tau \sum_{i=1}^{n} \alpha A_{ei}(t_{ai} - t_{si})}{m_{d.m}} \qquad (9)$$

where u_o is the initial moisture of the sample, $\Delta\tau$ the duration of the constant drying rate period, and n the number of layers.

Since the moisture in the first drying layer has decreased to zero there are nine drying layers left and the inlet air temperature into the second drying layer is the same as in the first layer, and so on. It is, of course, not correct to apply the constant drying rate model particularly below the fiber saturation point. Because the aim of this study is to evaluate the accuracy of the constant drying model, the decrease in the evaporation rate is not taken into account in the calculations. The theoretical evaporation rate of the bed merely decreases when the number of the dry layers increases in the bed.

The actual outlet moisture of the sample based on the measurement data was calculated as follows:

$$u_{out} = u_0 - \frac{\Delta\tau \dot{m}_{d.a} \sum_{j=1}^{m} \frac{(\Delta x_j + \Delta x_{j-1})}{2}}{m_{d.m}} \qquad (10)$$

where $\Delta\tau$ is the time step, Δx the difference between outlet, and inlet air moisture and m the number of scans.

The wet bulb temperature of the air was calculated using the Equation (2). The densities of dry air and vapour in Equation (2) were calculated by applying the equation of ideal gas. The pressure of the saturated vapour as a function of temperature can be approximately calculated using the correlation/1/:

$$p_v' = p_0 \exp\left(\frac{11{,}78 \cdot (t\text{-}99{,}64)}{t + 230}\right) \qquad (11)$$

where t is the wet bulb temperature in Celsius Degrees.

The diffusion coefficient between water and air needed to define the Lewis number depended on the temperature according to proportion $D \sim T^{3/2}$ /4/. The value of the exponent n in Equation (2) was 0,4.The bed was divided into ten layers in all cases by assuming the evaporation surface is equal in each layer.

Figures 8 and 9 show how air velocity and particle size affect the validity of the constant drying rate model.

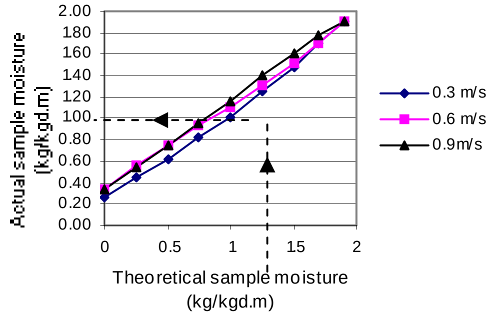

Figure 8 - The effect of air velocity on the validity of the constant drying rate model for particle size 5 mm and inlet air temperature 120 °C. Moisture content in x-axis is the theoretical moisture of the sample calculated by the model and moisture content in y-axis is the actual moisture of the sample determined by the measurement.

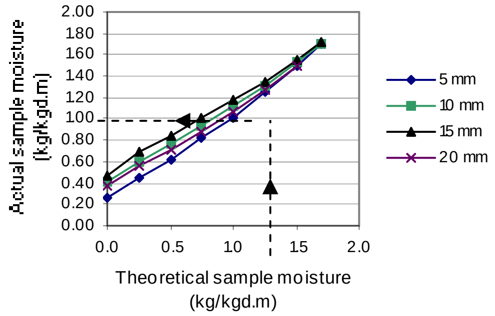

Figure 9 - The effect of particle size on the validity of the constant drying rate model with air velocity 0,3 m/s and inlet temperature 120 °C. Moisture content in x-axis is the theoretical moisture of the sample calculated by the model and moisture content in y-axis is the actual moisture of the sample determined by the measurement.

We can see in Figure 8 that air velocity has some effect on the validity of the constant drying rate model but it is quite small. Figure 10 presents how valid the constant drying rate model is for all particle sizes and temperatures used in the tests when the effect of the air velocity is ignored. Figure 11 shows a mean value of the curves in Figure 10. Regardless of the particle size and drying conditions we can approximately determine the accuracy of the constant drying rate model as a function of the sample moisture in Figure 11.

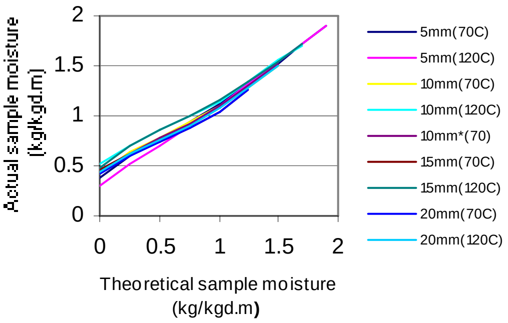

Figure 10 - The effect of particle size and temperature on the validity of the constant drying model.

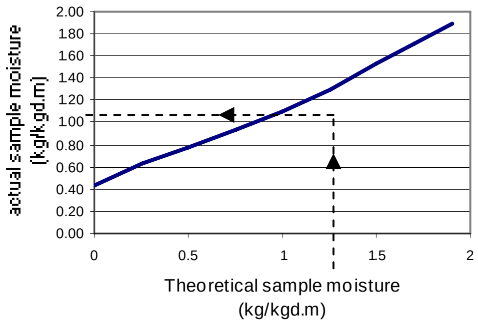

Figure 11 - The average accuracy of the constant drying model.

CONCLUSIONS

Drying of wood particles was experimentally studied in a fixed bed under various drying conditions.

Heat transfer coefficients were calculated based on the measured values. According to the results, the heat transfer coefficients seem to behave relatively logically as a function of velocity and particle size when the factors causing deviations are observed in the results.

Factors mainly causing deviations are: the measurement of surface temperatures, the measurement of the outlet air temperature and the variation of effective evaporation surface. The deviations of the measured values and the calculated heat transfer coefficients were estimated by conducting five tests for the same initial values. Particle size was 10 mm, air velocity 0,45 m/s, inlet air temperature 70 °C and inlet air moisture 0 g/kg$_{d.a}$. The variations of the measured values were the following: surface temperature in the bottom 26,3 - 28,2 °C, surface temperature in the upper 21,6 - 22,1°C, outlet air temperature 27,2 - 29,5 °C and evaporation rate 0,384 - 0,410 g/m^{2}s. The variation of the calculated heat transfer coefficients is 48,3 - 53.7 W/m^{2}K. The mean value is 50,1 W/m^{2}K. It is important to remember the measured temperatures and moistures are not necessarily exact values.

The values of the heat transfer coefficients seem to be higher when the inlet air temperature is 70 °C. This at least partly results from the smaller mass flow the air had when the inlet air temperature was 120 °C. Similarly, The heat conduction from the reactor little improves the absolute values of the heat transfer coefficients. The heat conduction was not wanted to eliminate as there is also heat conduction in a real application.

According to measurement data, the exact duration when the entire bed is in the period of constant drying rate is relatively short. On the basis of the measured surface temperatures, the particles in the bottom are quite soon in the period of the falling drying rate while the constant drying rate period is valid longer in the middle and upper part of the bed (see Fig. 3).

Air velocity and particle size have some influence on the validity of the constant drying rate model. When particle size and air velocity are small (in these measurements 5 mm and under 0,6 m/s) the constant drying model can be used with reasonable accuracy to average final moisture 0,6 - 0,7 kg/kg$_{d.m}$. When particle size and air velocity increase the accuracy of the model is slightly worse. Inlet air temperature under these drying conditions has no discernible effect on the validity of the model.

It is reasonable to conclude that sample moisture is the most influential factor affecting the validity of constant drying rate model. Since the moisture content has decreased below 0,8 -0,9 kg/kg$_{d.m}$ (see. Fig.11) the accuracy of the model deteriorates in most cases regardless of the drying conditions and particle size.

It is, however, possible to compare different drying conditions by using the constant drying rate model because the error compared to actual moisture content seems to be in the order of the same in all cas

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