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Study of Heat and Mass Transfer in Porous Media: Application to Packed-Bed Drying

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Abstract: This work focuses on tyipical heat and mass transfer phenomena during the processing of products in the context of the packed-bed drying method (products arranged in thick layers into dryers working in forced convection mode). The dryers are modeled as porous media at the macroscopic level. The simulations are carried out using the mass, momentum (written in the framework of the Darcy's law approximation) and energy equations applied for the different components. A diffusion model based on Fick's law is also used to take into account the drying kinetics. This approach allows monitoring of the variations of humidity and temperature (of the medium and the heated air) in time and space. A non-homogeneous drying of the different layers is observed. It is found that the process is strongly affected by external conditions (in particular, the temperature of the heated air).

Keyword: Forced convection, External conditions, Drying kinetics, Local thermal equilibrium (L.T.E.).

1 Introduction

Convective transfer in porous media is currently the object of intensive research activities, because of its important practical applications, such as chemical reactors, heat exchangers, thermal insulation, electronic cooling, food industries and many other industrial processes.

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Generally, two approaches are used to study convective transfers in porous media. The first approach studies the media at the scale of pores (microscopic scale); heat and mass equations are written for the solid, fluid and gas phases. This approach has been adopted by Whitaker (1980), Quintard, Kaviany and Whitaker (1997), Figus, Le Bray, Bories and Prat (1999), Quintard and Whitaker (2000), Altevogt, Roltson and Whitaker (2003), Duval, Fichot and Quintard (2004).

Plumb and Prat (1992), Amir, Le Palec and Daguenet (1987), Masmoudi, Prat (1991) and Prat (2002, 1995) used this strategy to simulate drying process. Boukadida and Ben Nasrallah (2002) used the same method and presented a numerical simulation of convective and convective-radiative drying of a clay brick. The results were used to illustrate the effect of variability of heat and mass transfer coefficients with temperature, gas pressure and vapor concentration. Another work presented by Mhimid, Ben Nasrallah and Fohr (2000) has proven that the variability of the porosity and the coefficient of heat transfer has no influence on the total relative moisture content. Also, a comparison between the local thermal non-equilibrium (LTNE) and the local thermal equilibrium (LTE) hypotheses has been considered. A small difference, leading to the same values at the end of the process, has been observed.

The second approach is macroscopic. Mass and energy balances are enforced for the product and the air. This approach has been largely used in the literature to study the behavior of both the product filled in thick layers and the dryer. It allows knowing the variations of humidity and temperature of the heated air, and also the moisture content and temperature of the dried product (Arnaud and Fohr, 1988; Chauhan, Choudhury and

Garg, 1996; Ben Nasrallah, Amara and Du Peuty, 1997; Srivastava and John, 2002; Johnner and Sumardiono, 2003; Jain and Jain, 2004; Aregba, Sebastian and Nadeau, 2006; Bihercz and Beke, 2006; Aregba and Nadeau, 2007). These works, as the major papers dealing with deep-bed drying of foodstuffs, have considered grain products, such as rice, cereals, corn and wheat, and have supposed the velocity of the heated air to be constant.

Few works can be found in the literature where the drying of thick layers of fruits and vegetables has been studied (Wang and Chen, 1999 and Ratti and Mujumdar, 1995).

In this paper and according to the results obtained by Ben Nasrallah, Amara and Du Peuty (1997), the second approach and the local thermal equilibrium hypothesis is chosen to simulate the process.

Two main objectives are considered. The first is to study the behavior of a packed bed of grapes fruit dried in a forced convection dryer. The air velocity is not considered constant and is calculated using Darcy' law. The second objective is to study the influence of the external conditions which are temperature, velocity and humidity, on the behavior of the different dried layers.

2 Mathematical formulation

The problem is considered to be non-stationary and bi-dimensional, with known characteristics (temperature, flow rate and humidity) of the heated air at the entrance of the dryer. Nevertheless, before writing the governing equations, some simplifying assumptions are made. So, we neglect:

- Layers compression.
- The concentration and temperature gradients between surface and core of the product.
- The walls of the dryer are supposed to be adiabatic and impermeable.

The porous medium is divided into two parts; the heated air (fluid phase) and the dried product (solid phase). The basic conservation laws are used.

Equation of continuity:

$$\frac{\partial U}{\partial x}+\frac{\partial V}{\partial y}=0 \tag{1}$$

Equations of momentum are based on Darcy's Law. They read:

$$U=-\frac{K}{\mu}\frac{\partial p}{\partial x} \tag{2}$$

And

$$V=-\frac{K}{\mu}\frac{\partial p}{\partial y} \tag{3}$$

The mass transfer equation, in the air, is expressed as:

$$\varepsilon\frac{\partial w}{\partial t}+\left[U\frac{\partial w}{\partial x}+V\frac{\partial w}{\partial y}\right]= D\left(\frac{\partial^2 w}{\partial x^2}+\frac{\partial^2 w}{\partial y^2}\right)-(1-\varepsilon)\frac{\rho_s}{\rho_f}\frac{\partial \overline{C}}{\partial t} \tag{4}$$

The heat balance in the fluid phase is defined as:

$$\left[(\rho Cp)_{app}\right]_f\left(\varepsilon\frac{\partial T_f}{\partial t}+U\frac{\partial T_f}{\partial x}+V\frac{\partial T_f}{\partial y}\right)= k\left(\frac{\partial^2 T_f}{\partial x^2}+\frac{\partial^2 T_f}{\partial y^2}\right)-hA\left(T_f-T_s\right) \tag{5}$$

The heat balance in the solid phase is written as:

$$\left[(\rho Cp)_{app}\right]_s\left(\frac{\partial T_s}{\partial t}\right)= hA\left(T_s-T_f\right)+Lv\left(1-\varepsilon\right)\rho_s\frac{\partial \overline{C}}{\partial t} \tag{6}$$

In this work, the local thermal equilibrium (L.T.E.) hypothesis is used. It supposes that there is no difference between the temperature of the fluid and the solid. In this context, Eq. (5) and Eq. (6) are assembled into a single heat transfer equation written as (Haddad, Al nimr and Al Khateeb, 2004; Quintard and Whitaker, 1995; Thevenin and Sadaoui, 1995 and Thevenin, 1995):

$$(\rho Cp)_{eff}\frac{\partial T}{\partial t}+(\rho Cp)_f\left(U\frac{\partial T}{\partial x}+V\frac{\partial T}{\partial y}\right)= k_{eff}\left(\frac{\partial^2 T_f}{\partial x^2}+\frac{\partial^2 T_f}{\partial y^2}\right)-Lv\rho_s\frac{\partial \overline{C}}{\partial t} \tag{7}$$

Where (Thevenin and Sadaoui, 1995; Thevenin, 1995; Hsu, 2000 and Kimura, kiwata, Okajima and Pop, 1997):

$$(\rho Cp)_{eff} = \varepsilon\,(\rho Cp)_f + (1-\varepsilon)\,(\rho Cp)_s \tag{8}$$

And (Thevenin and Sadaoui, 1995 and Nield, Kuznetsov and Xiong, 2002):

$$k_{eff} = \varepsilon k_f + (1-\varepsilon)\,k_s \tag{9}$$

The following initial and boundary equations are used:

At $t = 0$:

$$T = T_0 \tag{10}$$

$$w = w_0 \tag{11}$$

For $y = 0$:

$$w = w_{\text{inf}} \tag{12}$$

$$p = p_{atm} + \frac{\rho V_{\text{inf}}^2}{2} \tag{13}$$

For $y = L$:

$$\left(\frac{\partial T}{\partial y}\right) = 0 \tag{14}$$

$$\left(\frac{\partial w}{\partial y}\right) = 0 \tag{15}$$

$$p = p_{atm} \tag{16}$$

Finally, for $x = 0$ and $x = L$:

$$\left(\frac{\partial T}{\partial x}\right) = 0 \tag{17}$$

(Condition for adiabatic walls)

$$\left(\frac{\partial w}{\partial x}\right) = 0 \tag{18}$$

(Condition for non-condensation)

$$\left(\frac{\partial p}{\partial x}\right) = 0 \tag{19}$$

(Condition for impermeable walls)

The combination of the equation of continuity (Eq.1) and the momentum equations (Eq.2 and Eq.3) leads to another equation, which allows having the distribution of the pressure and thus the air velocity, inside the dryer. The obtained equation is:

$$\left(\frac{\partial^2 p}{\partial x^2}\right) + \left(\frac{\partial^2 p}{\partial y^2}\right) = 0 \tag{20}$$

Eq. (4) and Eq. (7) show that knowing the average moisture content of the dried product is necessary to complete the calculus and thus drying kinetics are needed.

2.1 Drying kinetics

The simulated product is seedless grape and a diffusion model, based on Fick's law, is used for modeling the variations of the moisture content of such a product. As it is assumed to have a spherical shape, the model equation reads:

$$\frac{\partial C}{\partial t} = D\left(\frac{\partial^2 C}{\partial r^2} + \frac{2}{r}\frac{\partial C}{\partial r}\right) \tag{21}$$

The most common difficulty in using the diffusion model lies in the determination of the coefficient of diffusion. Bennamoun and Belhamri (2006a), based on the experimental work of Berna, Rosselo, Cañellas and mullet (1991), have expressed the coefficient of diffusion as a function of the heated air temperature and its velocity. It is found that increasing air temperature leads to an increase in the coefficient of diffusion. In a same way, increasing the air velocity leads to an increase in the coefficient. However, at high velocities the influence is less important.

An average value of the moisture content of the product is used for Eq. (4) and Eq. (7).

3 Results and discussion

The finite difference method with an implicit scheme has been used to solve the obtained system of differential equations. The obtained system is converted, using discretization, into a system of equations, which can be rewritten into a matrix form.

As, the matrix contains sparse coefficients (equal to zero), an iterative method is more rapid and more economical in memory requirement for the computer. Also the method offers the advantage of self correcting [Gerold and Wheatly (1989)]. Accordingly, the Gauss-Seidel iterative method has been used.

We have considered a regular mesh and we have used 10 nodes in the x-direction and 10 nodes in the y-direction. The calculation has been performed in one-hour time steps. The simulation code has been developed and written in FORTRAN and the results have been calculated with a relative error of 10^{-4} %.

The presented results have been obtained using a flow rate equal to 99 kg/h, initial air absolute humidity w_0 = 0.01 kg/kg. At the entrance of the dryer, the air has a humidity of about w_{inf} = 0.00837 kg/kg (which represents a relative humidity equal to 13%) and a temperature of about 50°C.

The following three figures show the variation of the air humidity, with time and for different heights of the dryer.

Fig. 1 shows the variation of the air humidity for first dried layers at $y^* = 0.1$.

Ten hours are sufficient to dry the first layers and, as shown in the figure, the maximum of evaporation (around $w^* = 1.39$) has already reached its highest value. After that, the air humidity starts decreasing. It is deduced that the layers have received a sufficient quantity of energy.

At the beginning of drying process, the obtained energy serves only to evaporate water from the product. This is characterising the first period of drying, called period of constant drying rate. After that, the obtained energy serves, in the one hand, to evaporate the water of the product and, on the other hand, to increase the temperature of the media. Hence, it is found that, after 10 hours of drying, the first layers are in the second phase of the drying process and the temperature increases with time. However, this is not the case for the last layers, where the obtained energy only serves to water evaporation. These results are confirmed by simulation in a work presented by Bennamoun and Belhamri (2006b).

Fig. 2 and Fig. 3 show that drying proceeds in a non- homogeneous manner. It is seen in Fig.

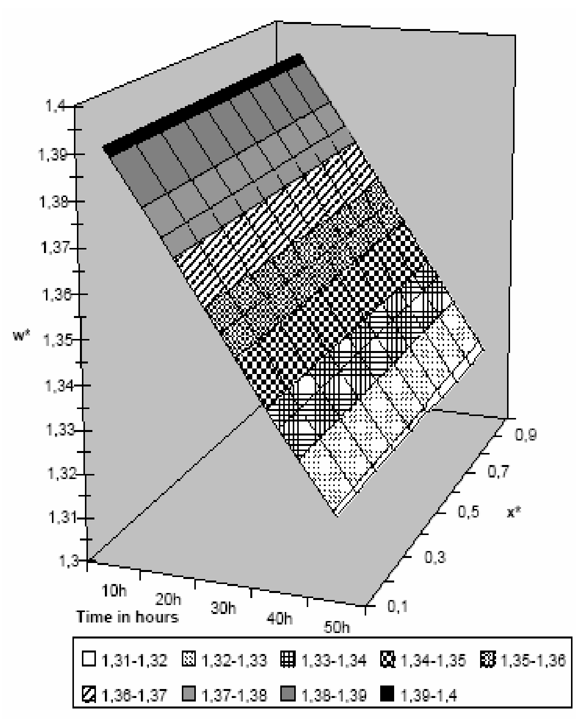

Figure 1: Variation of the air humidity at $y^* = 0.1$

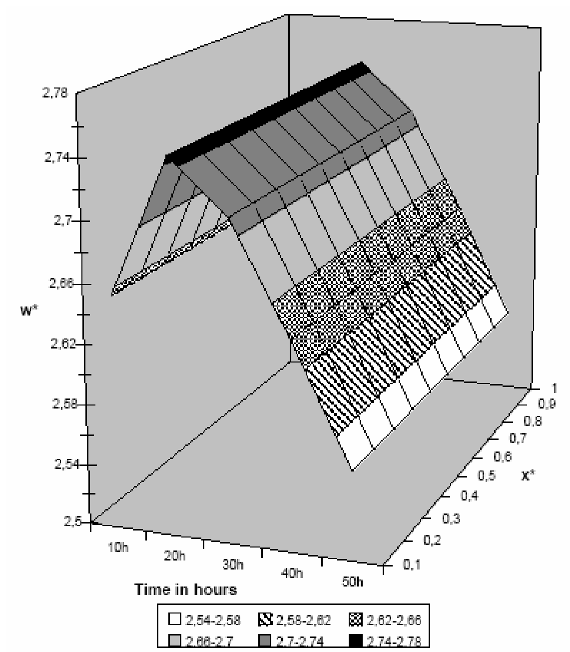

Figure 2: Variation of the air humidity in the middle of the drying chamber ($y^* = 0.5$)

2 that the maximum of evaporation is reached in the middle of the dryer after 20 hours ($w^* = 2.74$) while it is reached after 40 hours ($w^* = 3.98$) for the last layers, as shown in Fig. 3. It can be deduced that the received energy is absorbed essentially by the first layers and as these are dried, more energy is delivered to the last layers. It is observed in these three figures that the quantity of the evaporated water increases (as it is shown by the maximum values). This is not caused by the important evaporated quantities from the corresponding layers but it is an accumulation of the evaporated water from the preceding layers. These results are in agreement with those obtained by Bennamoun and Belhamri (2006b).

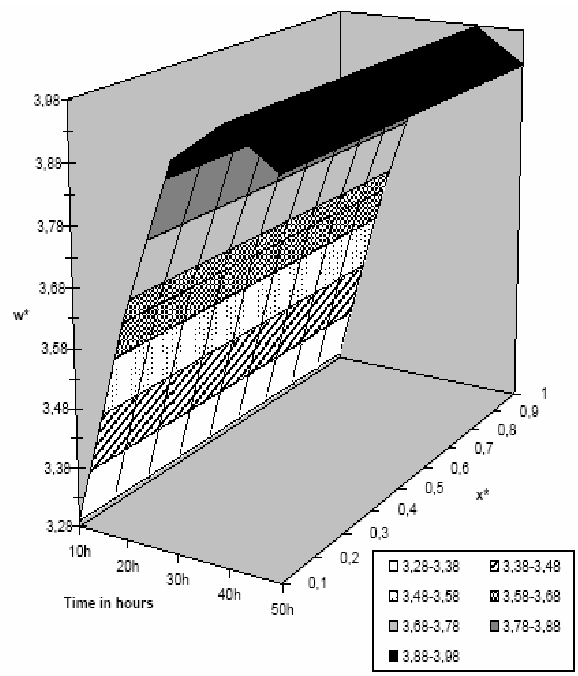

Figure 3: Variation of the air humidity at the exit of the drying chamber ($y^* = 1.0$)

These researchers monitored the gradient of air humidity ($\partial w^*/\partial y^*$). The gradient of air humidity is important at the beginning of the drying process (in the first 10 hours), for the first layers and tends to vanish for the last dried layers. This observation is no longer valid after 50 hours. The last layers now obtain more energy and more water quantities are evaporated. Consequently the

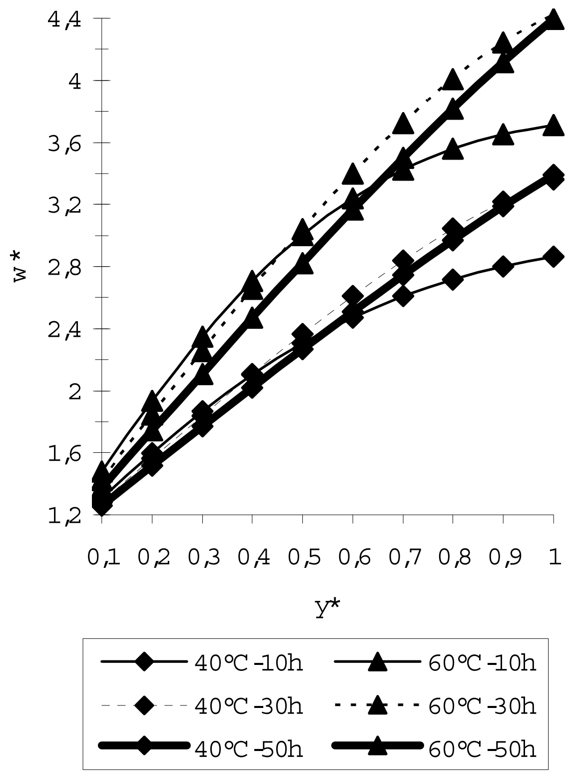

Figure 4: Influence of the heated air temperature on its humidity at several times of drying

gradient of air humidity, for these last layers, is increasing.

The influence of the temperature of the heated air is illustrated in Fig. 4.

The figure shows that temperature is a significant parameter. Increasing the temperature of the heated air allows giving more energy. However, it has no effect on the first layers, as shown in Fig. 4. This can be understood by considering that the energy given at 40°C is sufficient for the first layers. On the other hand, this increase is not sufficient for last layers, because the air is saturated for these layers (the air gradient, $\partial w^*/\partial y^*$, tends to zero). These observations are made after 10 hours of drying. After that, the energy given reaches the last layers and air saturation vanishes. The simulation with air at 60°C shows that the evaporation is more important at $t = 10$h than at $t = 30$h which in turn is more important than at $t = 50$h. Thus, it can be concluded that after 50 hours, the entire product is dried (except $y^* = 0.9$ and $y^* = 1.0$).

Fig. 5 illustrates the variation of the medium temperature with the heated air temperature, for several drying times.

The figure shows that although the energy increases as temperature increases, this effect is not strong enough to be visible for the last layers, as shown for $t = 10$h. This especially profits to the first layers, where it is seen that the temperature of the medium has reached that of the heated air temperature, after just 10 hours. For the first layers, the received energy serves to evaporate water from the product and increase its temperature. However, for the last layers, as this energy is not sufficient to evaporate all the water in the product, it does not serve to increase its temperature

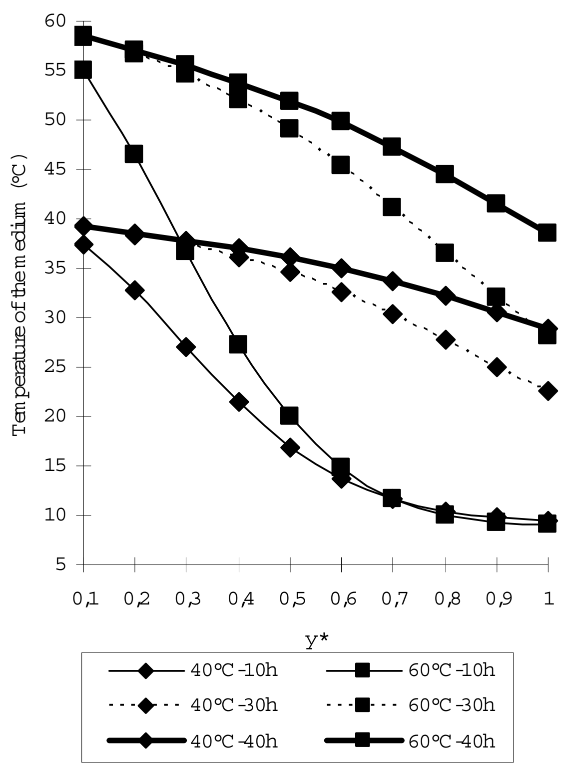

Figure 5: Influence of the heated-air temperature on the media temperature

After the first layers are dried, more energy is left to the profit of higher layers, and, everywhere in the medium, the temperature increases gradually, as seen in the curves at $t = 30$h and $t = 40$h, until it reaches the temperature of the heated air.

The influence of the flow rate and humidity of the heated air on the air humidity is illustrated in Fig. 6.

The increase of the flow rate (calculated at 20% relative humidity) leads to an increase in the heat and mass transfer represented by the increase in the concentration of the air and the temperature of the medium. Here, only air humidity is illustrated. Bennamoun and Belhamri (2006b) have shown an increase of heat transfer when the flow rate increases. It is important to note that the influence of the flow rate is less significant with respect to the effect of temperature.

Generally, research on drying neglects the effect of air humidity (Karanoudis, Maroulis and Marinos-Kouris, 1992). Only few scientists studied this parameter (Inazu, Iwasaki and Furuta, 2002, Ratti and Crapiste, 1992). These scientists were concerned with the effect of air humidity on the drying kinetics of a single sample or a thin layer. They found that increasing the humidity results in longer drying time. Thus, it is interesting to study the effect of this parameter on deep-bed drying. Fig. 6 shows that the effect is important and that it is negative. Increasing the humidity limits the heated air capacity. In consequence, it will take more time for all the product layers to be dried.

4 Conclusion

Tyipical heat and mass transfer phenomena related to the processing of products in the context of the packed-bed drying method (products arranged in thick layers into dryers working in forced convection mode) have been investigated in the framework of a numerical approach.

The results show that there is a non-homogenous drying of the different layers of the dryer. It has been found that, for the considered conditions, after 10 hours the first layers only are dried. An important part of the received energy is consumed by these layers. It is used both to evaporate water from the product and to increase the product temperature. whereas the energy that reaches the last layers is only used to evaporate water from

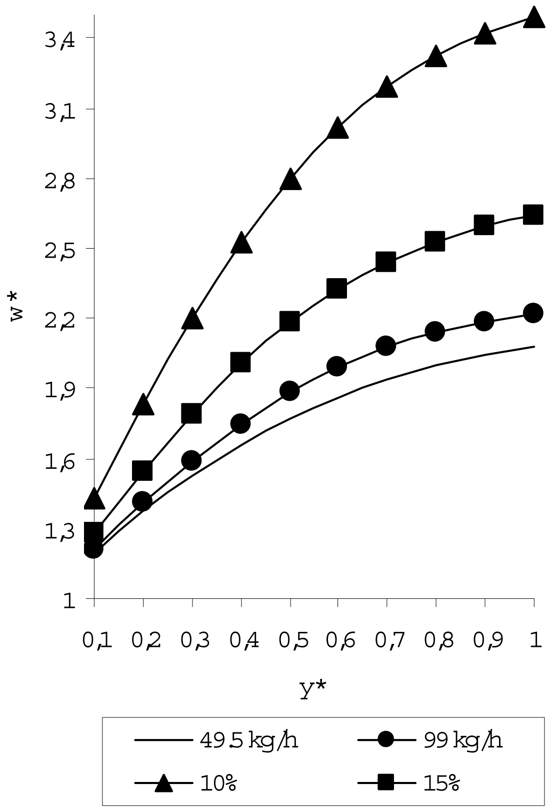

Figure 6: Influence of the flow rate and the humidity of the heated air on the concentration of the air

the product.

A parametric study has been carried out to clarify the effect of the external conditions. It has been found that increasing the heated air temperature leads to an increase in the evaporated quantities with a possibility of air saturation for the last layers at the beginning of the process.

The study has also shown that the flow rate of the heated air is less significant than the air temperature. However, increasing the flow rate increases heat and mass transfer and thus humidity of the air and the temperature of the medium.

The heated air humidity is, generally, neglected for thin layers drying. However, it has an important effect on deep-bed drying. This effect is negative: increasing the air humidity limits its power of evaporation.

This means, the effective external conditions must be rigorously chosen before starting the operation; also, effective control of the industrial tools used (fans, heaters . . .) may be useful for achieving an efficient drying process.

It will be interesting to extend this study to solar drying where no control on the external conditions is possible.

Nomenclature

A	Contact surface (m^2)
C	Moisture content ($kg.kg^{-1}$)
$\overline{C}$	Mean value of the moisture content ($kg.kg^{-1}$)
Cp	Specific heat ($J.kg^{-1}.°C^{-1}$)
D	diffusion coefficient ($m^2.s^{-1}$)
h	coefficient of transfer by convection ($W.m^{-2}.°C^{-1}$)
K	Permeability (m^2)
k	adapted coefficient of transfer by conduction ($W.m^{-2}.°C^{-1}$)
L	Height of the drying chamber (m)
Lv	Latent heat of vaporization ($J.kg^{-1}$)
p	pressure (Pa)
r	radius (m)
T	Temperature (K or °C)
T^*	dimensionless temperature ($T.T_{inf}^{-1}$)
t	time (s)
U,V	Velocity components ($m.s^{-1}$)
w	Absolute humidity ($kg.kg^{-1}$)
w^*	Dimensionless humidity ($w.w_{inf}^{-1}$)
x,y	Spatial coordinates
y^*	Dimensionless coordinate ($y.L^{-1}$)

Greek Symbols

ε	Porosity
μ	Dynamic viscosity ($kg.m^{-1}.s^{-1}$)
ρ	Mass density ($kg.m^{-3}$)

Subscripts

0	Initial value
app	Apparent
atm	Atmosphere
eff	Effective
f	Fluid
inf	Entrance to the drying chamber
s	Solid

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References

Altevogt A.S.; Rolston D.E.; Whitaker S. (2003): New equations for binary gas transport in porous media, Part2: experimental validation. *Advances in Water Resources*, Vol. 26, pp. 717-723.

Amir H.; Le Palec G.; Daguenet M. (1987): Séchage superficiel d'un matériau poreux humide par convection forcée d'air chaud : couplage entre les équations de transferts dans le matériau et celles de la couche limite. *International Journal of Heat and Mass Transfer*, vol. 46, pp. 1149-1158.

Aregba A.W.; Nadeau J.P. (2007): Comparison of two non-equilibrium models for static grain deep-bed drying by numerical simulations. *Journal of Food Engineering*, vol. 78, pp. 1174-1187.

Aregba A.W.; Sebastian P.; Nadeau J.P. (2006): Stationary deep-bed drying: A comparison study between logarithmic model and a non-equilibrium model. *Journal of Food Engineering*, vol. 77, pp. 27-40.

Arnaud G.; Fohr J.P. (1988): Slow drying simulation in thick layers of granular products. *International Journal of Heat and Mass Transfer*, vol. 31, pp. 2517-2526.

Bennamoun L.; Belhamri A. (2006a): Numerical simulation of drying under variable external conditions: Application to solar drying of seedless grapes. *Journal of Food Engineering*, vol. 76, pp. 179-187.

Bennamoun L.; Belhamri A. (2006b): Study of convective heat and mass transfer in a porous media: application to packed bed drying. In: A.S. Mujumdar (Ed.), *Drying 2006*, Szent Istvàn University Publisher, Gödöllö, Hungary, pp. 532-538.

Ben Nasrallah S.; Amara T.; Du Peuty M.A. (1997): Convection neturelle instationnaire dans un cylindre rempli de grains ouverts à ses extrémités et dont la paroi est chauffée par un flux de chaleur constant: validité de l'hypothèse de l'équilibre thermique local. *International Journal of Heat and Mass Transfer*, vol. 40, pp. 1155-1168.

Berna A.; Rosselo C.; Cañellas J. Mulet A. (1991): Drying kinetics of a Majorcan seedless grape variety. In: A.S. Mujumdar (Ed.) *Drying 91*, Elsevier, Amsterdam, pp. 455-462.

Bihercz G.; Beke J. (2006): Semi-Empirical model of convective drying with wide range layer depth validity. *Drying Technology*, vol. 24, pp. 1165-1172.

Boukadida N.; Ben Nasrallah S. (2002): Effect of the variability of heat and mass transfer coefficient on convective and convective-radiative drying of porous media. *Drying Technology*, vol. 20, pp. 67-91.

Chauhan P.M.; Choudhury C.; Garg H.P. (1996): Comparative performance of coriander dryer coupled to solar air and solar air-heater-cum-rock bed storage. *Applied Thermal Engineering*, vol. 16, pp. 475-486.

Duval F.; Fichot F.; Quintard M. (2004): A local thermal non-equilibrium model for two-phase flows with phase-change in porous media. *International Journal of Heat and Mass Transfer*, vol. 47, pp. 613-639.

Figus C.; Le Bray Y.; Bories S.; Prat M. (1999): Heat and mass transfer with phase change in a porous structure partially heated. Continuum model and pore network simulations. *International Journal of Heat and Mass Transfer*, vol. 42, pp. 2557-2569.

Gerold C.F.,; Wheatly P.O. (1989): *Applied numerical analysis*, 4th ed., Canada; USA, Addison. Wesley.

Haddad O.M.; Al-Nimr M.A.; Al Khatteb A.N. (2004): Validation of the local equilibrium assumption in natural convection from vertical plate embedded in porous medium: non-Darcian model. *International Journal of Heat and Mass Transfer*, vol. 47, pp. 2037-2042.

Hsu C.T. (2000): Heat conduction in porous media. In: K. Vafai (Ed.) *Handbook of porous media*, Marcel Dekker, New York, pp. 171-200.

Inazu T.; Iwasaki K.; Furuta T. (2002): Drying air conditions of fresh Japanese noodle (Udon). In: Mujumdar A.S. (Ed.), *Drying 2002*, pp. 1556-1562.

Jain D.; Jain R.K. (2004): Performance evaluation of an inclined multi-pass solar air heater with in-built storage on deep-bed drying application. *Journal of Food Engineering*, vol. 65, pp. 497-509.

Johnner P.S.I.; Sumardiono S. (2003): Modelling and simulation of momentum, heat and mass transfer in a deep-bed grain dryer. *Drying Technology*, vol. 21, pp. 217-229.

Kimura S.; Kiwata T.; Okajima A.; Pop I. (1997): Conjugate natural convection in porous media. *Advances in Water Resources*, vol. 20, pp. 111-126.

Kiranoudis C.T.; Maroulis Z.B.; Marinos-Kouris D. (1992): Drying kinetics of onion and green paper. *Drying Technology*, vol. 10, pp. 995-1011.

Liu W.; Jiang P.; Shen S. (2005): Numerical and experimental investigation of convective drying in unsaturated porous media with bound water. *Heat and Mass Transfer*, vol. 41, pp. 1103-1111.

Masmoudi W.; Prat M. (1991): Heat and mass transfer between a porous medium and a parallel external flow application to drying of capillary porous material. *International Journal of Heat and Mass Transfer*, vol. 34, pp. 1975-1989.

Mhimid A.; Ben Nasrallah S.; Fohr J.P. (2000): Heat and mass transfer during drying of granular products- simulation with convective and conductive boundary conditions. *International Journal of Heat and Mass Transfer*, vol.43, pp. 2779-2791.

Nield D.A.; Kuznetsov A.V.; Xiong M. (2002): Effect of local thermal non-equilibrium on thermally developing forced convection in a porous medium. *International Journal of Heat and Mass Transfer*, vol. 45, pp. 4949-4955.

Plumb O.A.; Prat M. (1992): Microscopic models for the study of drying of capillary porous media. In: A.S. Mujumdar (Ed.) *Drying 92*, Elsevier, Amsterdam, pp. 397-406.

Prat M. (2002): Recent advances in pore-scale models for drying of porous media. *Chemical Engineering Journal*, vol. 86, pp. 153-164.

Prat M. (1995): Isothermal drying of non-hygroscopic capillary-porous materials as an invasion percolation process. *International Journal of Multiphase Flow*, vol. 21, pp. 786-791.

Quintard M.; Whitaker S. (2000): Theoretical analysis of transport in porous media. In: K. Vafai (Ed.) *Handbook of porous media*, Marcel Dekker, New York, pp. 1-52.

Quintard M.; Kaviany M.; Whitaker S. (1997): Two medium treatment of heat transfer in porous media: numerical results for effective properties. *Advances in Water Resources*, vol. 20, pp. 77-94.

Quintard M.; Whitaker S. (1995): Local thermal equilibrium for transient heat conduction: theory and comparison with numerical experiments. *International Journal of Heat and Mass Transfer*, vol. 38, pp. 2779-2796.

Ratti C.; Mujumdar A.S. (1995)**:** simulation of packed bed drying of foodstuffs with airflow reversal**.** *Journal of Food Engineering*, vol. 26, pp. 259-271.

Ratti C.; Crapiste G.H. (1992): A generalized drying curve fir shrinking food materials. In: Mujumdar A.S. (Ed.), *Drying 92*, Elsevier, Amsterdam, pp. 864-873.

Srivastava V.K.; John J. (2002): Deep bed grain drying modelling. *Energy Conversion & Management*, vol. 43, pp. 1689-1708.

Thevenin J.; Sadaoui D. (1995): About enhancement of heat transfer over a circular cylinder embedded in a porous medium. *International communications in Heat and Mass Transfer*, vol. 22, pp. 295-304.

Thevenin J. (1995): Transient forced convection heat transfer from a circular cylinder embedded in a porous medium. *International communications in Heat and Mass Transfer,* 1995, vol. 22, pp. 507-516.

Wang Z.H.; Chen G. (1999): Heat and mass transfer in fixed-bed drying**.** *Chemical Engineering Science*, vol.54, pp. 4233-4243.

Whitaker S. (1980): Heat and mass transfer in granular porous media. In: Mujumdar A. S. (Ed.) *Advances in Drying I*, pp. 23-61.