Computational modelling of the transport phenomena occurring during
convective drying of prunes

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Abstract
A mathematical model was developed to describe the simultaneously coupled heat, mass and momentum transfer processes occurring in convective drying of prunes. In this model, the fruit is viewed as a composite ellipsoidal body comprising of two materials (flesh and stone) having different properties. The model accounts for the variation of both air and food properties expressed as a function of temperature and moisture content. The resulting systems of transient non-linear partial differential equations (PDEs) in the space-time domain together with the set of initial and boundary conditions were numerically solved by the finite element method (FEM) coupled to the Arbitrary Lagrange-Eulerian (ALE) procedure to account for the shrinkage phenomenon using a commercial package (COMSOL Multi-physics v3.5a, Comsol AB, Sweden).

A series of laboratory drying experiments were undertaken using a computer-controlled dehydration system developed to obtain quality drying kinetics data under varying operating conditions for validation of the results from the proposed model. Comparison of the model predictions against the experimental results has shown a very good agreement, implying that the proposed numerical model can be used with
confidence as a tool in optimising the design and operation of the prune drying system. A parametric study performed using the modelling tool has demonstrated the impact of key operational parameters on the drying kinetics. It is expected that the model can be applied for other food products and processes involving similar phenomena.

Keywords:
Mathematical modelling, tunnel drying, prunes, dried plums

1. Introduction
Prune is one of the important products for the dried fruit industry with worldwide production in excess of 257,000 metric tons (IPA, 2010). It is widely consumed for its nutritional value including high dietary fibre, antioxidants and its effectiveness in enhancement of digestive function. The industrial production of prunes usually relies on drying of high-moisture fresh plums in long dehydration tunnels to 20-24% final moisture content (Sabarez and Price, 1999). The drying process of prunes is energy-intensive and takes a long time to complete (up to 35 hours depending on drying conditions). In Australia, the present cost of prune dehydration accounts for a major portion of the total production costs (up to 50%) depending on the dehydrator design, drying conditions, type of fuel used, fruit size, etc (Sabarez, 2010). There is a significant interest in the industry to improve the energy efficiency in dehydration to maintain its profitability, competitiveness and sustainability due to the prospect of a continuously increasing trend in fuel prices and the need for eco-friendly processes to mitigate environmental impact coupled with the rising consumer demand for high quality products.
The work presented here formed part of a larger research project, which was initiated that aims to provide strategies to improve the cost efficiency in dehydrating prunes to help the industry remain globally competitive through improved design and operation of the existing industrial drying system to reduce energy use and improve production throughput with consistent product quality. In particular, this work was underpinned by the development of mathematical models for the drying process of prunes. This enables to perform parametric studies to determine the impact of various operating and design scenarios and test options for improvement without indulging to lengthy and costly industrial scale experimentations.

Several studies have sought to model the drying process of prunes (Bertin et al., 1983; Techasena et al., 1991; Price et al., 1997; Sabarez et al., 1997). In previous studies (Sabarez and Price, 1999; Sabarez and Price, 2001), moving boundary diffusion models based on numerical solution with constant and variable diffusivities were successfully developed to describe the kinetics of drying plums. Whilst these models fitted the experimental data very well but their validity appeared to be limited as these models mainly account the internal resistance to moisture transfer. It should be noted that at the air/food interface the external heat and mass transfer rates could be the controlling step in the drying process depending on the drying conditions, which are strongly dependent on the drying air velocity field. It is therefore important to account the momentum transport dynamics to improve the predictive precision of the model. This work was therefore aimed to develop and validate a computational model accounting for the simultaneous transfer of heat, mass and momentum in air as well as within the food capable of predicting the drying behaviour of prunes with moving boundaries over a very wide range of operating and fluid-dynamic conditions.
2. Experimental

2.1. Materials

Fresh plums of d’Agen variety obtained from Cavaso Farming P/L (Darlington Point, New South Wales, Australia) were used in the laboratory drying experiments. Fruit weighing between 10 and 20 g, which represent the average size, were utilised. Two harvest seasons of fruit samples (2007 and 2008) were used in this study. Samples were kept in sealed bag purged with nitrogen and then stored in a cool store at 4°C prior to the drying experiments.

2.2. Moisture content determination

The initial moisture content of the fruit samples was determined using a standard AOAC method (AOAC, 1995) by vacuum drying over magnesium perchlorate desiccant (Sabarez and Price, 1999). In this method, 5 to 10 g of ground sample was spread in an aluminium moisture dish and dried in a vacuum oven at 70°C for 6 hours under reduced pressure (<100 mm Hg). This was repeated at least three times to obtain a representative average. The initial moisture content was expressed on a wet basis (kg H$_2$O/kg wet sample).

2.3. Experimental drying setup

Laboratory drying experiments were performed using a computer-controlled drying system illustrated in Fig.1. The purpose-built test drying facility incorporates a number of special features including a fully programmable cyclic control of process conditions (i.e. temperature, humidity and airflow) and a dedicated weighing system. It is equipped with controllers for controlling the process variables. A number of additional sensors (i.e. thermocouples, Infrared non-contact temperature sensors, air
velocity sensors, etc.) were interfaced to a computer-based data acquisition and control system of the dryer for further online monitoring and recording of the various processing conditions.

Fig. 1. Schematic diagram of the laboratory test drying setup.

The temperature of the air in the drying chamber can be regulated between 10°C below ambient to 250°C (accuracy of ±0.5°C). The dryer can be operated with a relative humidity of the air in the range of 10-95% (±1.1%). The temperature and relative humidity of the drying air were measured using a Rotronic H290 transmitter connected to a microprocessor controller (Eurotherm 2604) and controlled using a PID algorithm. Two centrifugal fans driven by variable speed motors were used to allow variation in air velocity. The air velocity can be adjusted of up to 5 m/s either in
horizontal or vertical directions. The airflow rate was measured using a hot film anemometer (E+E Electronik GmbH, Austria) with a sensitivity of ±0.2 m/s.

A Precisa AD30000 electronic balance with a capacity of up to 30 kg (resolution ±0.1g) is used for measurement of the sample mass. The balance is mounted on the outside top of the dryer with stainless steel rods directly suspended from it inside the drying chamber to enable continuous monitoring and recording of the sample mass throughout the drying process at pre-determined time intervals without interruptions. In order to obtain a stable mass reading, the fans are automatically switched off for a short period of time (just enough to allow any movement in the sample tray to cease) during the weighing process to minimise the impact of airflow draft and vibrations on each recorded mass. The operation of the integrated drying facility is automatically done using a PC with a specific application software developed in CitectSCADA version 6.0 (Schneider Electric, France) to enable monitoring, supervision and control of all process variables. This also allows easy access of the process data (both real-time and historical) for decision making and further analysis.

2.4. Drying experiments

Prior to each laboratory drying experiment, fruit samples were allowed to equilibrate under room conditions. Approximately 1 kg of fruit samples for each run were spread uniformly in a single layer on the tray and then loaded into the drying chamber after the desired drying conditions had stabilised. Each drying run was carried out until the moisture content (in wet basis) of the fruit reached about 20% (consistent with commercial dried prunes). All drying experiments were replicated at least twice.
3. Mathematical modelling

3.1. Model development

A 2-dimensional numerical model was developed to describe the simultaneous transfer of momentum (air only), heat and mass (air and food) occurring in convective air drying of prunes. In this model, prune is viewed as a composite ellipsoidal body, comprising of two materials (flesh and stone) having different properties. In order to simulate the drying process under these conditions, the following assumptions were made:

- convection and conduction heat transfer in the air
- convection and diffusion water transfer in the air
- heat transfer by conduction within the solid interior for both flesh and stone
- mass transfer in the solid interior by diffusion (for flesh component only)
- moisture evaporation at the air/food interface
- deformation shrinkage of the flesh component only due to moisture loss
- air flows under turbulent conditions around the food material

3.2. Mathematical formulation

In convective air drying of food products, two distinct transport mechanisms occur simultaneously, involving heat transfer from the drying air to the food material and water transport from the interior of the solid product to its surface and eventually to the air through evaporation. The heat and mass transfer rates depend on both temperature and concentration differences as well as on the air velocity field.

The governing partial differential equations (PDEs) describing the simultaneous transfer of heat, mass and momentum in two distinct sub-domains (air and food)
during drying of prunes are presented. The non-isothermal turbulent flow of air in the
drying chamber is described according to the standard $ \kappa \varepsilon $ model. The equations for
the momentum transport and continuity are the following:

\[ \rho \frac{\partial u}{\partial t} - \nabla \left[ \left( \eta + \rho \frac{C_{\mu} \kappa^2}{\sigma_{\varepsilon}} \right) \left( \nabla u + \left( \nabla u \right)^T \right) \right] + \rho u \cdot \nabla u + \nabla P = 0 \]  
(3.1)

\[ \frac{\partial \rho}{\partial t} + \nabla (\rho u) = 0 \]  
(3.2)

The turbulence energy equation is given by

\[ \rho \frac{\partial \kappa}{\partial t} - \nabla \left[ \left( \eta + \rho C_{\mu} \frac{\kappa^2}{\varepsilon} \right) \nabla \kappa \right] + \rho u \cdot \nabla \kappa = \frac{1}{2} \rho C_{\mu} \frac{\kappa^2}{\varepsilon} \left( \nabla u + \left( \nabla u \right)^T \right)^2 - \rho \varepsilon \]  
(3.3)

and the dissipation equation by

\[ \rho \frac{\partial \varepsilon}{\partial t} - \nabla \left[ \left( \eta + \rho C_{\varepsilon} \frac{\varepsilon^2}{\kappa} \right) \varepsilon \right] + \rho u \cdot \nabla \varepsilon = \frac{1}{2} \rho C_{\varepsilon} \kappa \left( \nabla u + \left( \nabla u \right)^T \right)^2 - \rho C_{\varepsilon} \frac{\varepsilon^2}{\kappa} \]  
(3.4)

The energy balance in the food material leads to the transient heat transfer equation
according to Fourier’s law of heat conduction (equation. 3.5), whilst the energy
balance in the drying air, taking into account for both convective and conductive
contributions is given in equation 3.6.

\[ \rho C_v \left( \frac{\partial T}{\partial t} \right) + \nabla (-k \nabla T) = 0 \]  
(3.5)
The transient moisture transport within the food matrix was modelled using the basic law governing the movement of moisture according to Fick’s law of diffusion (equation 3.7), whilst the water mass balance in the drying air, taking into account for both convective and diffusive contributions is given in equation 3.8.

\[
\rho_a C_{pa} \left( \frac{\partial T_2}{\partial t} \right) + \nabla (- k_a \nabla T_2) + \rho_a C_{pa} \mu \nabla T_2 = 0 \quad (3.6)
\]

\[
\left( \frac{\partial c}{\partial t} \right) + \nabla (- D \nabla c) = 0 \quad (3.7)
\]

\[
\left( \frac{\partial c_2}{\partial t} \right) + \nabla (- D \nabla c_2) + \mu \nabla c_2 = 0 \quad (3.8)
\]

The boundary conditions used to formulate the mathematical model are summarised in Fig. 2. In particular, the boundary condition at the air/food interface (at \( t>0 \)) for heat transfer, considering the mass transfer at the air/food interface, thus coupling the heat and mass transfer equations simultaneously is given in equation 3.9. This means that the heat transported by convection and conduction from the drying air to the food is partly used to raise the food temperature by conduction and partly for water evaporation at the food surface.

\[
-n(-k \nabla T) = \lambda T_c \left( c_2 - c_s \right) + h_c \left( T_2 - T_s \right) \quad (3.9)
\]

Whilst the boundary condition at the air/food interface for mass transfer is given in equation 3.10, which accounts the balance between the diffusive flux of liquid water
coming from the interior of the product and the flux of vapour from the food surface to the drying air.

\[-n(DVc) = k \left( c_2 - c_x \right) \]  

(3.10)

<table>
<thead>
<tr>
<th>Boundary Type</th>
<th>Condition</th>
</tr>
</thead>
<tbody>
<tr>
<td>1, 2, 3, 4, 5</td>
<td>Axial symmetry</td>
</tr>
</tbody>
</table>
| 6             | Momentum: Inlet (velocity)  
                 Heat transfer (air): Temperature  
                 Mass transfer (air): Concentration |
| 7             | Momentum: Outlet (pressure, no viscous stress)  
                 Heat transfer (air): Convective flux  
                 Mass transfer (air): Convective flux |
| 8             | Insulation/Symmetry |
| 9, 10         | Momentum: Wall (logarithmic wall function)  
                 Heat transfer (air/product): Heat flux  
                 Mass transfer (air/product): Mass Flux |
| 11, 12        | Continuity |

Fig. 2. Geometry, discretisation of the domains and boundary conditions.
3.3. Model parameters

The solution of the governing partial differential equations requires knowledge of the thermo-physical and transport properties of the product and air. The thermo-physical properties of the product (i.e. thermal conductivity, specific heat, and density) were assumed to be dependent on product composition (i.e. water, protein, fat, carbohydrate, and ash) expressed as a function of the local temperature (ASHRAE, 1995). The typical proximate composition of prunes is reported in the literature (Di Matteo et al, 2008).

The convective heat transfer coefficient required in the boundary condition in equation 3.9 was estimated from Nusselt-Reynolds-Prandtl correlation for local convective heat transfer for solid sphere (forced convection) given by Heldman and Lund (2007). The overall mass transfer resistance is comprised of two contributions from the external mass transfer resistance of the external boundary layer and the mass transfer resistance of the skin of the food product (van der Sman, 2003). The air film mass transfer coefficient which describes the convective mass transfer at the surface of the product was obtained from Sherwood-Reynolds-Schmidt correlation for average convective mass transfer for a solid sphere (forced convection) (Heldman and Lund, 2007). The water vapour concentration of the air in equilibrium with the surface of the fruit exposed to convection was estimated from the sorption isotherms using Henderson’s equation (Henderson and Perry, 1955). For prunes, these authors reported the values for empirical constants \( c \) and \( n \) to be \( 1.25 \times 10^{-4} \) and 0.865, respectively. The effective diffusion coefficients were estimated by a least squares fit between the predicted and experimental drying data, assuming the dependence on both moisture concentration and temperature of the product given by the relationship
reported elsewhere (Sabarez and Price, 1999; Sabarez and Price, 2001). Thus, the influence of temperature and moisture content on the effective diffusivity could be established from a single experiment, as the change in temperature and moisture content of the product during drying depends on the operating conditions. In order to account for the shrinkage factor, the dimensional change of the prune is assumed to occur only in the flesh section and that the volume of the stone component remains constant throughout the drying process. The change in volume during drying was assumed to be equivalent to the amount of moisture removed (Raghavan et al., 1995). The latent heat of vaporisation of water in the air is given by Lydersen (1983) as a function of temperature.

3.4. Numerical solution

A numerical approach is required to solve the set of time-dependent PDEs for heat, mass and momentum transfer processes as these equations cannot be solved analytically. The resulting systems of highly coupled non-linear PDEs in the space-time domain together with the set of initial and boundary conditions were numerically solved by the finite element method (FEM) coupled to the Arbitrary Lagrange-Eulerian (ALE) procedure to account for the shrinkage phenomenon using a commercial package (COMSOL Multi-physics v3.5a, Comsol AB, Sweden). Both the air and food domains were discretised into a number of triangular finite elements (Fig. 2). A preliminary grid independency test was carried out to ensure that the solution is independent of grid size with further grid refinements. The time-dependent problem was solved by an implicit time-stepping method. The resulting systems of non-linear equations were solved using Newton’s method whilst a direct linear system solver was adopted to solve the resulting systems of linear equations. The relative and
absolute tolerances were set to 0.001 and 0.0001, respectively to control the error at each integration step.

4. Results and discussion

4.1. Model validation

The validity of the proposed models to describe real systems is usually verified by determining the mean relative percentage deviation (%P) between experimental and the predicted values using the expression described elsewhere (Lomauro et al., 1985; Madamba et al., 1996; Palipane and Driscol, 1994). Fig. 3 shows the predicted drying curves by the model for the experimental data during drying of prunes for different relative humidity levels at 80°C. As can be seen from this figure, the simulated results agreed well with the experimental data. Generally, the experimental data banded around the straight line which indicates the suitability of the model in describing the drying behaviour of prunes (Fig. 4). The mean relative percentage deviation (%P) was found to be in the range of 0.3 to 7.8%. According to Kaymak-Ertekin and Gedik (2005), and McLaughlin and Magee (1998) a model is acceptable, or a good fit, when P<10%. These results confirm the validity of the model and demonstrate that the parameters used in the model are reasonable, indicating the suitability of the model to describe the drying process under various conditions.

Fig. 5 shows the measured surface and centre temperatures of the product together with the predicted values. It can be seen from this figure that the agreement between the measured and predicted values is reasonably good. The deviations (%P) between the experimental and predicted values were in the range of 1.5% to 5.5%. This
validates the assumption on the dependency of the thermal properties of the product as a function of both temperature and moisture content. Clearly, the figure indicates differences between the product surface temperature and the drying air temperature (about 5-10°C) particularly in the early stages of drying. This could be explained by the evaporative cooling effect due to the rapid moisture flux on the surface during this period, justifying the use of elevated temperatures in a parallel-flow tunnel drying system. The rapid moisture flux on the surface of the product during this period requires more energy for moisture evaporation and hence less heat received by the product initially. Then as the moisture flux decreases in the later stages of drying, the product temperature gradually increased approaching near equilibrium with the drying air temperature. It was also found that a temperature gradient exists within the product particularly in the initial stages of drying.

Fig.3. Predicted versus experimental drying kinetics of prunes (T=80°C; \(u=7.0\) m/s).
Fig. 4. Predicted versus experimental moisture content (T=80°C; RH=10%).

\[ y = 1.0122x + 0.4915; R^2 = 0.9916 \]

Fig. 5. Predicted versus experimental fruit temperature profile (T=70°C; RH=30%; \( u=2.7 \text{m/s} \)).
The advantage of the proposed numerical model is that the temperature and moisture distributions at any time in both domains (solid food and drying air) can be established. Fig. 6 demonstrates the predicted evolution of local moisture content of the product, temperature and airflow distributions of the drying air. This information is relevant from the food quality point of view as it allows detection of the temperature and moisture distributions within the food matrix as the changes in quality parameters (i.e. bioactive, colour, microbial spoilage, etc.) generally depend on both the temperature and moisture content of the product. In addition, the prediction of the drying air stream conditions flowing across the product surface which would affect the drying behaviour of the solid product at any time and position in the dryer is also of particular importance in simulating the drying process of industrial drying systems where a systematic variation in drying conditions is typical.

In general, the proposed model developed and tested in this research has advanced the capability to predict the drying process of prunes over a wide range of conditions. This would be a useful tool in decision-making strategies in relation to the optimal design and operation of the prune tunnel drying system. The successful validation of the simulation model under various conditions against the experimental data suggests that the predictive tool can be used with confidence for prediction of important processes during tunnel drying of prunes. Any discrepancies between the experimental and predicted values in some cases were probably due to uncertainties in the experimental data and some of the input model parameters (i.e. heat and mass transfer correlations, sorption isotherms, etc.). A parametric sensitivity study presented in the next section further investigates the effects of uncertainties of various
input parameters on the model’s predictions and demonstrates the usefulness of the predictive tool to identify critical operational factors affecting the drying process.

Fig. 6. Predicted product moisture content, temperature and velocity profiles of the drying air during drying of prunes (T=85°C; RH=15%; $u=5$ m/s).

4.2. Parametric sensitivity study

A number of computer simulations were undertaken to study the interplay between the factors in the system and to examine the critical parameters affecting the drying
process of prunes. Figs. 7 through to 11 show the simulated drying kinetics of prunes at various parameters. Clearly, the simulated results depicted a wide range of variations in the drying kinetics when the product properties and operating conditions are changed. For instance, it is evident that sufficient effort should be made particularly into characterising the fruit size (Fig. 7). Obviously, smaller fruit dried faster than larger fruit. Grading the fresh fruit is therefore important in obtaining uniform and efficient drying. It is also worthwhile noting that the initial moisture content of the fresh fruit (or solid content) has significant impact in the drying process (Fig. 8). This exemplifies the need to measure this parameter and regulate the drying conditions to avoid over-drying (i.e. waste of energy and reduce product quality) as the moisture content of the fresh fruit changes from the start of harvest towards the end of the season depending on climatic conditions during harvest.

Fig. 7. Predicted effect of fruit size on the drying kinetics of prunes (T=80°C; RH=15%; u=5m/s).
Further simulations were also carried out to demonstrate the effect of input data uncertainties on the accuracy of the model. The uncertainty in the drying air temperature was mainly considered as an example as it was shown to have a major influence on the drying kinetics with likely greater variability in the measurement accuracy. Fig. 9 depicts the effect of uncertainties in the drying air temperature measurement on the model accuracy. It indicates that the uncertainties in the measurement of this parameter are likely to represent a greater contribution to the accuracy in the model predictions. This demonstrates that accurate measurement of the drying air temperature and perhaps its sensing location are important in industrial drying operations.
Fig. 9. Predicted effect of air temperature uncertainties on the drying kinetics of prunes (RH=15%; $u=5\text{m/s}$).

Fig. 10. Predicted effect of air velocity on the drying kinetics of prunes (T=85°C; RH=15%).
It is also important to identify critical operational factors that offer significant and measurable opportunities for improvement in the drying process. The conditions of the drying air (i.e. airflow, temperature and relative humidity) are considered to be the main factors influencing the drying performance in tunnel dehydrators. The effect of different air velocity levels was taken as an example to demonstrate the impact of this parameter on the drying kinetics (Fig. 10). In this instance, it appears that the drying time significantly decreases as the air velocity increases but only to a certain point and then beyond this level the air velocity plays a proportionally decreasing role in reducing the drying time. Under these conditions, it can be seen that the optimum air velocity level is around 5 m/s. This implies that the role of the external resistance to moisture transport becomes negligible compared to the internal resistance as the air velocity increases. It should be noted that the air stream mainly facilitates drying by removing moisture reaching the surface. As the fruit dries, internal resistance to mass transfer becomes the governing mechanism and higher temperatures rather than airflow are more efficient in removing the remaining moisture. The result also unveils that further increase in air velocity beyond the optimum level would significantly increase the energy consumption with minimal increase in throughput (i.e. slight reduction in drying time). This is obvious as increases in air volume would result in increased energy requirement to heat the large volume of air to the desired temperature level. If the air velocity is too high, the costs of energy required to heat the excessive air would tend to offset the benefits of slight reduction in drying time. Consequently, there is little to be gained by using a very high air velocity. On the other hand, it appears that lower energy consumptions could be achieved if operating at lower air velocities, however with the expense of longer drying times. Apart from reduced throughput, operating at longer drying times would also increase the labour
costs associated in drying and possibly affect the product quality due to prolonged exposure times. Also, one of the major issues in tunnel drying is the non-uniformity in drying conditions (i.e. temperature and humidity), which is more apparent for large industrial drying systems depending on air velocity field. If the velocity is too low, convection currents and other disturbances will cause wide variations in temperature and relative humidity across the system, resulting in uneven drying.

The mode of operation in a tunnel drying system is one of the key factors that may influence the drying performance. The industrial tunnel dehydrators are currently operated in either counter-flow or parallel-flow mode of operation. In a counter-flow configuration, the drying air is introduced into one end of the tunnel while the trolleys of food products enter at the other end and each moves in opposite directions. This configuration is characterised by having conditions most conducive to intense heating at the end of the drying cycle when the products are nearly dry and less heating in the early stages. The operation of the parallel-flow tunnel is opposite to that of the counter-flow. The trolleys and drying air enter at the same end of the tunnel and progress through the tunnel in the same direction. This configuration is characterised by intense heating in the early stages where the product to be dried is still very wet followed by slow drying as the product approaches the cooler end.

A number of computer simulations were undertaken to mimic the commercial tunnel drying of prunes both in the counter-flow and parallel-flow modes of operation. In simulated parallel-flow drying, the drying air temperature was set initially at 85°C then decreased linearly with drying time until up to 70°C and the drying air relative humidity was set initially at 15% then linearly increased as drying progresses until the
relative humidity reaches 30%. For the simulated counter-flow drying, the drying conditions of temperature and relative humidity were exactly in the opposite profile with that in parallel-flow drying. The air velocity condition was kept constant at 5.0 m/s for both drying modes. The selected conditions are typical in commercial prune tunnel drying operations.

Fig. 11 shows the drying kinetics of prunes in simulated tunnel drying for both counter-flow and parallel-flow operations. Under the drying conditions investigated, the parallel-flow operation would apparently result in shorter overall drying time to reach the desired final moisture content (20%) compared with the counter-flow operation. It took about 18.5 hours to dry the fruit in parallel-flow whilst drying in counter-flow would take around 20.5 hours. The predicted overall drying times were found to be in close agreement with those obtained in the commercial dehydration trials undertaken under similar drying conditions (Sabarez, 2010). The result of these
commercial dehydration trials shows that drying in parallel-flow mode of operation
requires less energy and gives higher production throughput (i.e. shorter drying time)
in comparison with the counter-flow operation. Parson (1968) reported a 33% faster
in drying time with less fuel consumed per tonne of prunes in parallel-flow drying of
prunes as compared to the counter-flow operation.

As can be seen in the figure, drying in parallel-flow was much faster in the initial
stages of drying and tends to significantly slow down in the later stages due to the fact
that as drying progresses the fruit is subjected to less intense drying (low temperature
and high humidity) at the end of the drying process. In counter-flow drying, the fruit
experienced slow drying in the initial stages and then progressively exposed to intense
drying conditions. Also, the time required to dry the fruit with both modes of
operation would tend to converge and cross over in the last stages of drying as
equilibrium is approached. It shows that prunes dried in counter-flow operation could
reach very low moistures compared to that in parallel-flow dried prunes. This implies
that the counter-flow drying operation is more vulnerable to achieving very low
moistures of the dried product if over-dried. This is consistent with the drying
conditions that the fruit would be exposed in the last stages of drying. Also, drying the
fruit below the desired moisture content level (or over-drying) would appear to greatly
affect the economics of drying as the last remaining moisture in the product would
normally take much longer to be removed as evidenced from the curves. This is
particularly more apparent in parallel-flow operation in which the drying curve tends
to flatten towards the end of the drying process. For example, the figure shows that
under the prevailing drying conditions (in parallel-flow), over-drying the fruit by just
2.5% from the desired level of moisture content (20%) would add an extra drying
time of around 2.5 hours. This shows that the determination of the final moisture content of the product is an important avenue to reduce the total drying time and hence costs.

5. Conclusion

The proposed model developed and tested in this research has advanced the capability to predict the drying process of prunes over a wide range of conditions. Predicted results were found to be in good agreement with experimental values. The successful validation of the simulation model under various conditions against the experimental data suggests that the predictive tool can be used with confidence for prediction of important processes during tunnel drying of prunes. It was shown to identify and predict the key operational parameters affecting the drying process. This would be a useful tool in decision-making strategies in relation to the optimal design and operation of the tunnel drying system. The developed model can readily be adapted to account for other physics (e.g. ultrasound, electromagnetic) and can be extended for three-dimensional geometries in complex-shaped systems.

Nomenclature

\( c \) water concentration in food \( (\text{mol/m}^3) \)
\( c_2 \) water concentration in air \( (\text{mol/m}^3) \)
\( C_p \) specific heat \( (\text{J/kgK}) \)
\( C_{\mu} \) model parameter -
\( C_{\varepsilon_1} \) model parameter -
\( C_{\varepsilon_2} \) model parameter -
\( D \) effective water diffusivity in food \( (\text{m}^2/\text{s}) \)
heat transfer coefficient \( h_c \) (W/m\(^2\)K)

thermal conductivity \( k \) (W/mK)

mass transfer coefficient \( k_c \) (m/s)

moisture content \( MC \) (% wet basis)

direction normal to surface \( n \) -

pressure \( P \) (Pa)

relative humidity \( RH \) (%) 

food temperature \( T \) (°C)

air temperature \( T_2 \) (°C)

time \( t \) (sec)

Subscripts:

air \( a \) -

initial \( i \) -

food surface \( s \) -

Greek letters:

dissipation rate \( \varepsilon \) (m\(^2\)/s\(^3\))

latent heat of evaporation \( \lambda \) (J/kg)

dynamic viscosity \( \eta \) (N.s/m\(^2\))

turbulence energy \( \kappa \) (m\(^2\)/s\(^2\))

density \( \rho \) (kg/m\(^3\))

model parameter \( \sigma_e \) -

velocity \( u \) (m/s)
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