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Studio Dei Fenomeni Di Trasporto Nei Processi Di Essiccazione Dei Prodotti Ortofrutticoli

Modelling Of The Transport Phœnomena During Drying Process Of Vegetables

by

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General Introduction

"*Asparagos siccabis, rursum in calidam submittes: callosiores reddes*": as reported by Apicio in *De Re Coquinaria*, food drying was already well known by Romans. Perishability of food is a problem that points back to the appearance of the first social forms, with the fixing of nomad people and the start of earlier forms of agricultural activity. Food preservation can be realized in different ways all with the same purpose: allow the consumption of food not linked with logistic limitations and not only during the harvest time. These factors became very important in modern society due to the globalization, that require the food to cover widely long distance, and to the modern life that has provoked for most people a contraction of the time expendable in domestic affairs, increasing the demand for alimentary preserves and prepared foodstuff. The common objective of all the food preservation methods is to reduce the degradation reactions caused by the activity of microorganisms living inside the food. Drying methods preserve the foods decreasing water activity that stops micro-organisms growth. In a very short time alimentary preserves production grew up from a domestic to an industrial scale. This growing production rate, however, could compromise the level of safety of the products and consequently consumer's health. During drying, the level of water content could be not sufficiently low to stop the micro-organisms activity that will continue to proliferate. As a form of human control for this amount of production is problematic, an automatic control should be devised. The starting point of the automatic control is based on the availability of a predictive model of the process under study. Another important aim of food process modelling is the process optimization: domestic scale production did not present optimization problems, on the contrary, for obvious economic reasons, optimization is a key factor in industrial production.

The aim of this work is developing a predictive model for the most widely diffused drying method: the convective drying. In convective drier, where hot dry air flows about the food sample, different transport phenomena take place: heat, mass and momentum transfer. Moreover, a volume variation of food occurs due to the water leaving the food known as Shrinkage.

Food process modelling involves the resolution of balance equations containing food properties (chemical, physical and transport properties). These properties are difficult to evaluate since they depend on food structure and composition that change during processing; at the same time the potential of any rigorous predictive model is based also on the correct definition of the material properties. Moreover, a standard procedure aimed at the evaluation of most of them does not exist.

In the first part (Part A) of the present work, the transport phenomena that take place during drying have been physically and mathematically modelled to formulate a very high rigorous and versatile drying model that can be easily used for all kind of materials submitted to drying. The model accounts for the most important transport phenomena of the process assumed negligible by other authors as the contemporarily modelling of heat, mass and momentum transport and shrinkage. It does not need the specification of heat and mass transfer coefficients, rather it can be used to evaluate them for irregular geometries. Accounting for the phenomena previously neglected, a very high level of precision and versatility has been reached by the model as confirmed by the experimental validation.

In the second part (Part B) of the present work, an innovative, cheap and versatile experimental procedure for the evaluation of the most important food parameters to model mass transfer during drying has been developed. Moreover, an experimental study of food drying process has been carried out at TOP b.v., a commercial innovation company that provides services for the Food Industry and its technology suppliers, located in Wageningen (The Netherlands), where the author has spent three months as guest scientific researcher. In this last work the effect of drying conditions on food quality has been verified to demonstrate that the optimization process is not a mere economic optimization, but it should be coupled with the optimization of food quality parameters.

Both the first and the second part of this thesis are a collection of papers, some of them already published, some others submitted for publication. All the papers have been co-authored; the author of this work, however, gave the main contribution, whereas the other authors cooperated as supervisors. A more detailed description of the papers will be done in the following introductions to Part A and Part B.

Introduzione Generale

"*Asparagos siccabis, rursum in calidam submittes: callosiores reddes*": come dimostra Apicio nel suo *De Re Coquinaria*, l'essiccazione degli alimenti era tecnica ampiamente diffusa già presso i Romani. Riuscire a sconfiggere la deperibilità delle derrate alimentari, infatti, è da sempre stata una necessità che risale addirittura ai tempi in cui i primi esseri umani diedero vita ad un popolo che da nomade si trasformava in stanziale ed iniziava l'attività agricola.

La conservazione di un alimento può essere realizzata in diversi modi aventi uno scopo comune: la possibilità di consumare un alimento svincolandosi da limitazioni logistiche e temporali. Questi fattori hanno, nella società moderna, un ruolo basilare conferito loro sia dalla globalizzazione sia dai ritmi della vita quotidiana: la prima ha incrementato la richiesta di cibi in luoghi molto lontani da quelli di coltivazione, i secondi hanno provocato un aumento della domanda di cibi pronti e delle conserve alimentari che richiedono brevi tempi di preparazione prima del consumo.

La causa principale dell'alterazione degli alimenti è la degradazione dovuta a microrganismi contaminanti oppure alla catalisi delle reazioni di degradazione promosse dagli enzimi in essi naturalmente presenti. Le tecniche di conservazione hanno lo scopo di evitare l'insorgere di tali reazioni. L'essiccazione, attraverso la diminuzione dell'attività dell'acqua (reagente indispensabile per il metabolismo dei microrganismi), rallenta l'alterazione e quindi aumenta la "shelf-life" dell'alimento.

Il grande sviluppo che l'industria agro-alimentare ha conosciuto in questi anni, ha posto fin da subito il problema della sicurezza relativa ai prodotti commercializzati. Nell'ambito del controllo della qualità degli alimenti, la valutazione del rischio microbiologico è fondamentale; infatti, la legislazione sulla "sicurezza alimentare" impone criteri molto stringenti sui trattamenti di sanificazione il cui controllo non può più essere affidato alla soggettività umana ma che necessita di un controllo automatico. Il punto di partenza di un controllo automatico è la creazione di un modello predittivo del processo in esame che ha un altro importante utilizzo: l'ottimizzazione di processo. Le modalità secondo le quali l' essiccazione viene condotto influenza direttamente il consumo di energia, la qualità dell'alimento essiccato e la produttività dell'azienda; questa influenza suggerisce i notevoli benefici che si potrebbero ottenere attraverso l' ottimizzazione del processo stesso.

L' obiettivo principale del presente lavoro di ricerca è lo sviluppo di un modello predittivo per il metodo di essiccazione più diffuso: l'essiccazione convettiva. L'essiccazione convettiva è generalmente realizzata in dei forni in cui circola aria calda e secca il cui contatto con l'alimento umido provoca il trasporto dell'acqua dall'alimento verso la fase fluida. I fenomeni di trasporto coinvolti nel processo di essiccazione sono vari e complessi ed in prima analisi riassumibili come trasporto di calore, materia e quantità di moto nell'aria essiccante e trasporto di calore e materia all'interno dell'alimento, che durante l'essiccazione subisce un fenomeno di variazione volumetrica noto come *shrinkage*.
La modellazione dei fenomeni di trasporto nell'alimento coinvolge la scrittura di equazioni di bilancio che contengono le proprietà chimiche, fisiche e di trasporto. Tali proprietà sono difficili da valutare perché dipendono da fattori che variano durante la lavorazione dell'alimento come struttura e composizione; d'altro canto, una conoscenza quanto più precisa possibile di tali proprietà è indispensabile per sviluppare un modello valido. Questo, è un punto fondamentale visto che ad oggi non esiste una procedura standard per il calcolo di tali proprietà.

Nella prima parte del presente lavoro di ricerca (Parte A) sono illustrati i risultati ottenuti dalla modellazione fisica e matematica dei fenomeni di trasporto che avvengono durante il processo di essiccazione, allo scopo di creare un modello rigoroso, versatile ed estendibile a qualsiasi tipo di materiale. La modellazione proposta ha progressivamente tenuto conto dei più importanti fenomeni di trasporto coinvolti nel processo, alcuni dei quali in precedenza assunti trascurabili da altri autori, come la modellazione contemporanea del trasporto di calore, materia, quantità di moto e dello *shrinkage*. Il modello ottenuto alla fine del lavoro di modellazione, inoltre, non necessita di informazioni relative ai coefficienti di scambio termico e di materia, anzi, rappresenta un utile strumento per la loro valutazione. La validità dei vari modelli proposti in questa prima parte è stata confermata da un'adeguata verifica sperimentale.

Nella seconda parte (Parte B) del presente lavoro di ricerca è stata proposta una procedura sperimentale innovativa, economica e versatile per la valutazione dei più importanti parametri materiali necessari per la modellazione del trasporto di materia durante l'essiccazione. Inoltre è stato condotto anche uno studio sperimentale del processo in esame presso i laboratori della TOP b.v., una società privata di ricerca e sviluppo operante nel settore agro-alimentare sita a Wageningen (Paesi Bassi), dove l'autore del presente lavoro ha trascorso tre mesi come ricercatore ospite. In quest'ultimo lavoro è stato studiato l'effetto delle condizioni di essiccazione sulla qualità degli alimenti essiccati per sottolineare che l'ottimizzazione del processo non può limitarsi all'aspetto economico ma deve prevedere anche un'ottimizzazione dei parametri indicativi della qualità dell'alimento.

Sia la prima che la seconda parte del presente lavoro sono presentati sotto forma di una raccolta di articoli, alcuni pubblicati altri sottoposti ad approvazione. Tutti gli articoli sono stati realizzati in collaborazione con altri autori i quali hanno dato un contributo soprattutto di supervisione, mentre la parte predominante del lavoro è stata realizzata dall'autore della presente tesi. Nelle seguenti introduzioni alle parti A e B è riportata una descrizione più dettagliata degli articoli presenti in ciascuna sezione.

Part A

Drying Modelling

Besides predictive capacity, versatility is one of the most appreciated features in process modelling. Programmed or unexpected variations can affect a process during its execution: the potentiality of a predictive model is ascertained also by how easily it adapts itself to the occurrence of evolving conditions. Therefore, the purpose of the first part of this thesis is to find out a predictive and versatile model for the drying process of vegetables.

Table 1 summarizes the results obtained by the authors who have previously modelled the drying process.

	Water transport (Food)	Heat transport (Food)	Water transport (Air)	Heat transport (Air)	Momentum transport (Air)	Properties as a function of temperature and water content	Shrinkage (Food)
A	X					Yes/No	
B	X	X				Yes/No	
C	X					Yes/No	X
D	X	X				Yes/No	X
E	X		X		X	Yes/No	
F		X		X	X	Yes/No	
Aim of the present work	X	X	X	X	X	Yes	X

Tab. 1: State of the art in food drying modelling.

The different models described in literature – analyzed in greater detail later on in the present work – are sketched in the A-F groups; they do not include the contemporary modelling of all the transport phenomena occurring during the process, because from this modelling originates a mathematically complex system of partial differential equations. In category A are included the works in which only the transport of water inside the food has been modelled, for the most part assuming a diffusive mechanism of mass transfer and an isothermal process. The works included in the category B ignore the latter assumption, assuming for the most part a conductive mechanism for the heat transfer. In categories C and D are included models similar to those of categories A and B, respectively, with an added phenomenon: the shrinkage. The models in category E account for the water transport in both the domains involved in the drying process (air and food), assuming that the process occurs at a constant temperature. Category F includes those models which do not consider the transport of water, but exclusively the heat transport (so, they are not really drying processing models, related as they are only to the heating process of a food in contact with high temperature air). In all the listed models, authors used properties of air and food as functions of the temperature and/or the water content, as well as constant properties.

The aim of the first part of the present work has been to improve the modelling with the inclusion of the contemporary transport of mass, heat and momentum in both the concerned domains (food and air), so to obtain a complete and versatile predictive model. The first part of the present work is structured in four papers. The first paper presents the modelling of the transport phenomena in the food domain during the drying process. As a difference to the already existing literature, the utilization of a powerful commercial package based on the finite elements method (Comsol Multiphysics®), useful in solving problems with high mathematical complexity, has been introduced. The utilization of the aforementioned software allows the contemporary modelling of the mass and heat transport in a bi-dimensional food domain; furthermore, the food properties, necessary for the balance equations, have been considered as functions of the temperature and of the food moisture content. In defining the boundary conditions, appropriate correlations expressing the thermodynamic equilibrium and the transfer coefficients at the food/air interfaces have been used.

A whole series of important considerations about the process arose from this first paper. One of the most relevant among them is the difficulty to find, in existing literature, food properties and the heat and mass transfer coefficients. Both modelling and experimental approaches have been used to solve these problems. The experimental approach has been selected in order to estimate the properties of food and is the topic of the second part of this work, which will be described later. The modelling approach was conceived to improve the predictive model, in order to freeing it from the problem represented by the knowledge of transfer coefficients that are available only for regular shapes. In the case of foods, on the contrary, initial shape and its variation during drying (due to shrinkage) are unpredictable and irregular. To solve this problem the mathematical modelling of the transport phenomena across the food air interface have been developed. The high number of equations produced by the modelling of both the air and the food domains considerably increased the complexity of the problem; the great effort of this mathematical modelling has been dealt with the utilization of Comsol Multiphysics®.

In the second paper of part A, the model that estimates heat and mass transfer coefficients and the model that utilizes the literature correlations have been compared. The two models have shown a good agreement, and so we used the first of them, in the third paper of part A, for a typical example of real drying process: a flux of drying air flowing parallel to the symmetry axis of a cylindrical carrot. The subsequent experimental validation of the model further confirmed its potential for industrial application.

In the fourth paper of part A shrinkage has been physically and mathematically modelled. Given the complexity of the phenomenon, most authors neglected this effect in their modelling and preferred an experimental approach. However, the shrinkage influences the transport phenomena occurring in both air and food domains, requiring, as it will be explained in the following, a more rigorous modelling approach.

Introduzione Parte A

Una delle caratteristiche maggiormente apprezzate in un modello, dopo la capacità predittiva, è la versatilità. La conduzione di un processo può subire cambiamenti (programmati oppure inaspettati) delle sue variabili caratteristiche; la potenzialità di un modello predittivo è stabilita anche in base alla facilità con cui esso si adatta alle varie condizioni che si possono presentare. La formulazione di un modello predittivo e versatile, per il processo di essiccazione dei vegetali, è l'obiettivo che ci si propone in questa prima parte del presente lavoro di tesi.

I risultati della ricerca, ottenuti dagli autori che hanno in precedenza approcciato lo stesso argomento, sono riassunti nella Tabella 1. I vari modelli presenti in letteratura, di cui si troverà riferimento e maggiore descrizione nel prosieguo, sono schematizzati nei gruppi A-F; nessuno di essi comprende la modellazione contemporanea di tutti i fenomeni di trasporto che avvengono durante il processo perché da essa scaturisce un sistema di equazioni differenziali di notevole complessità matematica.

	Trasporto di acqua (Alimento)	**Trasporto di calore (Alimento)**	**Trasporto di acqua (Aria)**	**Trasporto di calore (Aria)**	**Trasporto di quantità di moto (Aria)**	**Proprietà funzione di Temperatura ed umidità**	**Shrinkage (Alimento)**
A	X					S/N	
B	X	X				S/N	
C	X					S/N	X
D	X	X				S/N	X
E	X		X		X	S/N	
F		X		X	X	S/N	
Scopo del presente lavoro	X	X	X	X	X	S	X

Tab. 1:Stato dell'arte nella modellazione del processo di essiccazione.

Esiste un gruppo di lavori, inclusi nella categoria A, in cui è stato modellato solo il trasporto di acqua all'interno dell'alimento, nella maggior parte dei casi assumendo un meccanismo di tipo diffusivo ed ipotizzando un processo isotermo. Nei lavori inclusi nella categoria B quest'ultima ipotesi è stata rimossa ed il trasporto di calore è stato modellato assumendo, nella maggior parte dei casi, un meccanismo conduttivo. Nelle categorie C e D sono compresi i modelli simili a quelli nelle categorie A e B rispettivamente, ma con un elemento in più: la modellazione dello *shrinkage*. I modelli appartenenti alla categoria E sono quelli che comprendono la modellazione dei fenomeni di trasporto in entrambi i domini coinvolti nell'essiccazione (aria ed alimento) assumendo il processo isotermo. Nella categoria F sono compresi quei modelli che non considerano il trasporto di acqua ma solo quello di calore, infatti, non sono veri e propri modelli per l'essiccazione ma piuttosto riguardano il riscaldamento di un alimento a contatto con aria ad elevata temperatura. I modelli proposti hanno presentato sia versioni che considerano le proprietà di aria ed alimento funzione della temperatura e/o dell'umidità, sia versioni che utilizzano proprietà costanti. Scopo della prima parte del presente lavoro è stato quello di migliorare la modellazione considerando il contemporaneo trasporto di materia, calore e quantità di moto in entrambi i domini coinvolti (alimento ed aria) allo scopo di creare un modello predittivo completo e versatile utilizzabile in fase di ottimizzazione e di controllo del processo.

La prima parte del lavoro di tesi si compone di quattro articoli. Il primo di essi presenta la modellazione dei fenomeni di trasporto nel dominio alimento durante il processo di essiccazione, con le adeguate condizioni al contorno. Rispetto alla letteratura esistente si è voluto introdurre l'utilizzo di un potente software basato sul metodo degli elementi finiti (Comsol Multiphysics®) che ha consentito di risolvere i problemi legati alla complessità matematica della modellazione. L'utilizzo del suddetto software ha consentito la modellazione contemporanea, in un dominio bidimensionale, del trasporto di materia e calore; inoltre, le proprietà dell'alimento, necessarie nelle equazioni di bilancio, sono state considerate funzioni di temperatura e contenuto d'acqua. Per la definizione delle condizioni al contorno sono state utilizzate opportune correlazioni di equilibrio termodinamico ed opportuni coefficienti di scambio termico e di materia.

Da questo primo articolo sono scaturite una serie di importanti considerazioni sulla modellazione del processo. Le principali hanno riguardato la difficoltà di reperire, in letteratura, correlazioni di importanza fondamentale per il modello, in quanto facenti parte delle stesse equazioni di bilancio: le proprietà chimico-fisiche e di trasporto dell'alimento ed i coefficienti di scambio di materia ed energia. Queste esigenze sono state le linee guida per il lavoro successivo, esse hanno dato vita a due linee di ricerca: una di tipo modellistico ed una di tipo sperimentale. L'approccio sperimentale è stato scelto per la stima delle proprietà dell'alimento e caratterizza la seconda parte del presente lavoro di tesi (come sarà descritto in seguito). L'approccio modellistico è stato scelto per l'ulteriore miglioramento del modello predittivo al fine di renderlo capace di affrancarsi dal problema della conoscenza a priori dei coefficienti di scambio. Questi ultimi, infatti, sono disponibili in letteratura solo per geometrie particolarmente semplici: nel caso degli alimenti, le geometrie sono imprevedibili ed irregolari. Inoltre, puntando alla modellazione dello *shrinkage* è stato necessario prevede che i coefficienti di scambio si adattassero in ogni istante alla situazione geometrica attuale dell'alimento in continua evoluzione. A tale scopo è stata realizzata la modellazione dei fenomeni di trasporto che stanno alla base della definizione dei coefficienti di scambio: la modellazione del trasporto di quantità di moto, materia e calore all'interfaccia aria alimento. Tale modellazione ha richiesto dunque di includere nel modello anche i fenomeni di trasporto nel dominio aria. La modellazione dell'essiccazione ha così raggiunto un ulteriore obbiettivo: la possibilità di considerare qualunque tipo di geometria senza doversi preoccupare di reperire o stimare sperimentalmente i coefficienti di scambio. L'elevato numero di equazioni, scaturito dalla modellazione dei fenomeni di trasporto in entrambi i domini coinvolti nell'essiccazione, ha provocato un notevole aumento della complessità matematica del problema. Per la risoluzione del sistema di equazioni differenziali è stato utilizzato il software commerciale Comsol Multiphysics®. Nel secondo articolo della parte A del presente lavoro di tesi, i risultati del modello capace di valutare i coefficienti di scambio sono stati confrontati con i risultati del modello che necessita della loro conoscenza. Il buon accordo tra i due modelli ha dimostrato la validità della modellazione realizzata che nel terzo articolo, della parte A, è stata applicata all'essiccazione di una carota cilindrica investita, parallelamente al suo asse, dall'aria essiccante. La validazione sperimentale del modello ha, in seguito, confermato le sue potenzialità di applicazione a livello industriale. Nel quarto articolo della parte A, è stata realizzata la modellazione fisica e matematica dello *shrinkage*. Data la complessità del fenomeno, molti autori lo hanno trascurato in fase di modellazione optando per la sua caratterizzazione attraverso un approccio sperimentale. Ad ogni modo la notevole influenza che ha lo *shrinkage* , sui fenomeni di trasporto che scaturiscono sia nell'alimento che nell'aria, richiede un più rigoroso approccio modellistico.

An analysis of the transport phenomena occurring during food drying process

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Abstract

This research presented a theoretical model describing the transport phenomena involved in food drying. Aim of the study was to determine the influence of some of the most important operating variables, especially velocity, humidity and temperature of drying air on the performance of drying process of vegetables. The main connotation of this study regarded the possibility of increasing the accuracy of drying process modelling by the use of Finite Elements Method (FEM). With respect to most of the works published in literature, the main innovation introduced in this study was represented with the model formulation. This, in fact, simulated the simultaneous bidimensional heat and moisture transfer accounting for the variation of both air and food physical properties as functions of local values of temperature and moisture content. The resulting system of unsteady-state partial differential equations has been solved by a commercial FEM package, called FEMLAB. The proposed model was general since it was capable of describing both the transport of free and bounded water within the food and the evaporation/condensation phenomena that, depending of the actual driving forces, may occur at the air/food interface. Simulations, carried out in different drying conditions, showed that air velocity and its relative humidity are more effective than air temperature to determine the drying rate. A comparison between the theoretical predictions and a set of experimental results reported in the literature has shown a remarkable agreement especially during the first 2 hours when experimental data and theoretical predictions overlapped and relative errors never exceeded 5%. Later on, a slight deviation can be observed, particularly for the drying test performed with higher air temperature and smaller food thickness, i.e. when a significant volume reduction (shrinkage) was possibly achieved. The observed deviation can be, therefore, ascribed to the lack of accounting for shrinkage effects in the model.

Keywords: Drying, Transport phenomena, modelling, Finite Elements Method.

Nomenclature

a_w	*Water activity in food*	*[/]*
Cp	*Specific heat of food*	*J/(Kg K)*
C	*Water concentration in food*	mol/m^3
Cg	*Water concentration in gas phase*	mol/m^3
D	*Water effective diffusivity in food*	m^2/s
h	*Heat transfer coefficient*	$W/(m^2\ K)$
k	*Thermal conductivity of food*	*W/(m K)*
k_g	*Thermal conductivity of gas phase*	*W/(m K)*
k_c	*Mass transfer coefficient*	*m/s*
N	*Diffusive water flux*	$mol/(m^2\ s)$
P	*System pressure*	*Pa*
P^0	*Vapour pressure of water*	*Pa*
T	*Temperature of food*	*K*
t	*Time*	*s*
T_g	*Temperature of gas phase*	*K*
T_{wb}	*Wet bulb temperature*	*K*
U	*Moisture content in food (wet basis)*	$Kg_{water}/\ Kg_{wet\ carrot}$
Ur	*Relative humidity of air*	*[/]*
v	*Gas velocity*	*m/s*
X	*Moisture content in food (dry basis)*	$Kg_{water}/\ Kg_{dry\ carrot}$
y	*Water molar fraction in food*	*[/]*
y_g	*Water molar fraction in gas phase*	*[/]*

Greek symbol

δ	*Food slice thickness*	*m*
λ	*Water molar latent heat of vaporization*	*J/mol*
ρ	*density of food*	Kg/m^3

subscripts

s	*Food surface*
b	*Bulk condition*
0	*Initial condition (t=0)*

Introduction

Forced convection by hot and dry air is the most common industrial technique to perform food drying. This process is generally realized in ovens where air flows with specific velocity profiles through the food positioned on suitable supports. Typical values of air temperature range between 40°C and 80°C, while air velocity normally changes from 0.5 to 5 m/sec, reaching, in some cases, the value of 10 m/sec; drying time depends on these and other parameters and it can last up to nearly 20 hours. Different physical, mathematical and numerical methods have been proposed to describe the drying processes. Nevertheless, many different aspects can be still improved. These regard the theoretical formulation of the transport model, , the numerical procedures used to solve the governing equations, a proper definition of the transport properties of both air and food material and the evaluation of heat and mass transfer coefficients performed on the basis of literature data or *ad hoc* experiments. An exhaustive analysis is often too onerous in terms of computational time needed to properly analyze the complex transport phenomena involved in food drying. Therefore, simplified approaches have been proposed in the literature to model the drying process of vegetables. Among these, the so-called *thin-layer equation* (Ertekin & Yaldiz, 2004, Doymaz, 2004, Karathanos, 1999, Krokida, Karathanos, Maroulis & Marinos-Kouris, 2003) is quite common, and it is based on the assumption that, in conditions of falling rate, moisture decrease is proportional to the instantaneous difference between material moisture content (assumed uniform within the food) and the moisture content in equilibrium with drying air. The proportionality constant, known as the drying constant, is a function of material moisture content, size of product and its uniform temperature distribution and characteristics of air, i.e. its humidity, temperature and velocity. The above assumptions allow, therefore, to consider only one equation, to represent the *drying kinetics* and to describe the transport phenomena involved in the drying process.

Hernàndez, Pavòn, Garcìa (2000) considered the drying process as isothermal assuming drying temperature equal to air temperature and accounting for mass transfer only in fruits drying. This approach was chosen also by Simal, Femenia, Llull & Rossello (2000) on *Aloe Vera* and by Ben-Yoseph, Hartel & Howling (2000) for sugar films. Wu & Irudayaraj (1996) experimentally verified that drying can be actually considered an isothermal process only if the Biot number is very low. Whereas, when Biot number is high, internal transport resistances are also to be considered.

Ikediala, Correira, Fenton, & Abdallah (1996) developed more precise and rigorous models that take into account the simultaneous heat and mass transfer within the food formulating a bidimensional model for the cooking of chicken patties; Ahmad, Morgan & Okos (2001) modelled the transport phenomena involved in biscuit drying in microwaves ovens accounting for both the transfers. Wang & Brennan (1995) developed a monodimensional model for the simultaneous heat and mass transfer within potatoes slices. The hypothesis of monodimensional transport was experimentally verified in the same study and used also by other authors (Kalbasi, Mehraban, 2000, Rovedo, Suarez, Viollaz,1995, Migliori, Gabriele, de Cindio, Pollini, 2005).

Thorvaldsson & Janestad (1999) analysed drying of porous media, like bread, considering water evaporation on all the exposed surfaces. The authors found that, due to the high internal porosity of the product, vapour inner generation is not negligible; therefore, also vapour transfer must be considered into food structure together with water and heat transfer.

In most of the previous works the materials physical properties depend on local temperature and/or moisture content (Migliori, Gabriele, de Cindio, Pollini, 2005, Ruiz-López , Córdova, Rodríguez-Jimenes, García-Alvarado, 2004, Kalbasi, Mehraban, 2000, Wang & Brennan, 1995). Ruiz-López , Córdova, Rodríguez-Jimenes, García-Alvarado (2004) compared, by experimental temperature and moisture content tests, the models with variable and constant physical properties respectively, suggesting that variable properties should be used only for modelling research purpose, because of the significant increase of numerical computation, while constant properties can be used for balance equation applied to industrial oven.

Since air and food physical properties may be expressed as functions of temperature and moisture content, being heat and mass fluxes coupled together in the boundary conditions, the equations describing heat and mass transfer form a set of non linear, partial differential equations that can be solved only by numerical procedures. Both finite difference (FD) and finite element (FEM) methods have been widely used for this purpose. FEM is particularly suitable for investigation on domains characterized by irregular geometries in presence of complex boundary conditions and for heterogeneous materials. Nevertheless, it is more complex and computationally expensive than FD (Wang & Sun, (2003)). FEM implies the discretisation of a large domain into a large number of small elements, the development of element equations and the assemblage of their contribution to the whole domain, and the solution of the resulting system of equations.

The finite elements discretisation of the governing differential equations is based on the use of interpolating polynomials to describe the variation of a field variable within an element. Although the spatial discretisation is different for the FEM as compared with FD method, the latter is usually employed for the time progression in a transient problem.

In the present work, a theoretical model describing the transport phenomena involved in food drying was presented. Aim of the study was to determine the influence of some process parameters, especially velocity, humidity and temperature of drying air, on food drying processes performed by air flowing in forced convection towards a rectangular food slice(a carrot slab). Actually, either Doymaz (2004) or Ertekin and Yaldiz, (2004) already presented similar approaches to this problem, but resorting to the already-described simplified method of *thin-layer* equation

With respect to most of the works published in literature, the main innovation introduced by this study was represented by the model formulation. This, in fact, simulated the simultaneous bidimensional heat and moisture transfer accounting for the variation of both air and food physical properties as functions of local values of temperature and moisture content. The resulting system of unsteady-state partial differential equations has been solved by FEM. The main objective was to develop an accurate transport model that could be used to simulate the behaviour of real drying processes and to define, over a wide range of drying conditions and of different types of foods, the "optimal" set of operating conditions in each particular situation. In this way, it might be possible to minimize expensive pilot test-runs and to have good indications on the characteristics and the quality of final products.

Theoretical Background

In a moist and cold food put in contact with dry and warm air, two different transport mechanisms simultaneously occur: the heat transfer from air to the material and the transfer of water from food to air . Within the solid material, heat transfer by conduction and water transfer by diffusion take place. In the air, however, also the convective contribution determined by air circulation velocity should be accounted for. Actually, water vapour transport by diffusion within dehydrated material to the external food surface should also be considered. However, this mechanism is significant especially for highly porous media, whereas for vegetables typically characterized by void fraction lower than 0.3 (May & Perré, 2002) can be neglected.

In this study paper, with reference to the drying process of carrot slice, water and heat transfers were modelled by transient mass and energy balances, respectively, whereas evaporation at air-food interface was considered by defining proper boundary conditions expressed in terms of heat and mass transfer coefficients estimated by the well-known empirical correlations (Perry, Green, 1984).

The model is based on the following hypotheses:

- Heat transfer in the product was by conduction;
- Mass transfer in the product was by diffusion;
- Transport of water vapour within dehydrated material was negligible;
- No food shrinkage occurred during drying;
- A bidimensional rectangular domain (0.6 cm long with a thickness ranging from 0.3 to 1.5 cm) was considered;
- The drying air was supplied continuously to the product, and its flow was parallel to its major surfaces;
- Heat and mass transfer resistance were negligible across the net on which food was placed (Thorvaldsson, Janestad 1999, Viollaz, Rovedo, 2002).

Mathematical Model

The energy balance in the solid, based on Fourier's law, leads to

$$\rho C_p \, \partial T / \partial t = \underline{\nabla} \bullet \left(k \underline{\nabla} T \right) \qquad 1$$

where ρ is density (Kg/m^3), C_p is heat capacity (J/Kg K), T is the temperature(K), t is time (s), and k is thermal conductivity (W/m K).

Mass balance, based on Fick's law, leads to:

$$\partial C / \partial t = \underline{\nabla} \bullet \left(D \underline{\nabla} C \right) \qquad 2$$

where C is water concentration in food (mol/m^3), D is the effective diffusion coefficient of water in food (m^2/s).

Supposing that the above material parameters (ρ, C_p, k, D), in the most general case, depend on local water concentration and on food temperature, Eqs. 1 and 2 form a system of non-linear partial differential equations.

Initial conditions are straightforward since it is assumed that, at t = 0, food concentration and its temperature are equal to their initial values, i.e. C_0 and T_0, respectively.

The boundary conditions relative to Eq. 1 applied to food external surfaces where no accumulation occurs actually state that heat transported by convection from air to food is partially used to raise sample temperature by conduction and partially to allow free water evaporation:

$$h\left(T_{gb} - T_s\right) = -\underline{n} \bullet \left(-k \underline{\nabla} T\right) - \lambda \cdot N_s \qquad 3$$

where λ is water latent heat of vaporization (J/mol), N_s is water diffusive flux on food surface (mol/m^2 s), h is heat transfer coefficient (W/m^2 K), T_{gb} is gas temperature in the bulk (K), T_s is food surface temperature(K).

The boundary conditions relative to Eq. 2 applied to food external surfaces express the balance between the diffusive flux of liquid water came from the core of the product and the flux of vapour that left the food surface and is transferred to the drying air:

$$- \underline{n} \bullet (D\underline{\nabla}C) = k_c \left(C_s - C_{gb}\right) \qquad 4$$

where k_c is mass transfer coefficient (m/s), C_{gb} is water bulk concentration in the air (mol/m^3), C_s is water concentration (mol/m^3) evaluated in gaseous phase at food/air interface.

To calculate C_s values, a thermodynamic equilibrium relationship between water concentration in the gas phase and water concentration on food external surfaces has been used:

$$P \cdot y_s = P_s^0 \cdot a_{ws} \qquad 5$$

where P is air pressure (Pa), P_s^0, y_s and a_{ws} are the vapour pressure (Pa), the molar fraction, and the activity of water on external food surfaces , respectively. Their definitions are reported in the following Table 1

$P_s^0 = (1/735.559) \cdot exp\left(A - \frac{B}{(T_s - 273.15) + C}\right)$	(a)
$a_{ws} = 1 - \exp\left[-(389258.9179) \cdot T_s^{-2.058} \cdot X_s^{-10.38+0.0751T_s-0.000128T_s^2}\right]$	(b)
$y_s = c_s$ / gas phase molar density	

a=(Perry, Green, 1984); b= (Ruiz-Lòpez, 2004)

Tab. 1: Definitions of P_s^0 and a_{ws}.

where X_s is food moisture content on a dry base

Equations 3, 4, 5 were essential to properly describe the complex steps involved in drying process. Equation 5, in fact, was expressed in terms of water activity a_{ws} that decreased as food moisture content decreased so that a unity value of a_{ws} meant only the free water evaporation. Whereas when

a_{ws}< 1 bounded water was also removed. Water activity is a distinctive parameter of each product that, being characterized by its own structure, determines how strong are the bonds between food structure and water. Moreover, X_s and T_s, through eq. 5, determine the value of food surface water concentration, C_s, and then the time evolution of drying process. In particular, if X_s and T_s are high enough to have $C_s > C_{gb}$, food drying starts immediately (eq. 4), whereas if their values is such to have $C_s < C_{gb}$, then a preliminary food humidification is observed. This, according to eq. 3, determines a steep food temperature raise owing either to the heat transfer by convection from air to food or to the latent heat of condensation due to vapour condensing on food surface. The above phenomenon continues until food temperature is high enough to make $C_s > C_g$; at this point, according to Eq.4, drying starts and heat is transferred by convection from air to food whereas latent heat of vaporization is transported from food to water thus allowing its evaporation (Eq. 3).

The physical properties of carrot used in the model drying process are reported in the following Table 2

$\rho = 440{,}001 + 90 \cdot X$	[Kg/m^3]
$C_p = 1750 + 2345 \cdot \left(X \big/ 1+X \right)$	[J/(Kg K)]
$k = 0.49 - 0.443 \cdot \exp(-0.206 \cdot X)$	[W/(m K)]
$D = 2.8527 \cdot 10^{-10} \exp(0.2283369 \cdot X)$	[m^2/s]

Tab. 2: Carrot physical properties (Ruiz-Lòpez, 2004).

where X is food moisture content on a dry base. Heat and mass transfer coefficients are calculated on the basis of the well-known semi-empirical correlations expressing the dependence of Nusselt number upon Reynolds and Prandtl numbers and of Sherwood number on Reynolds and Schmidt numbers, respectively. Perry, Green, (1984). The Chilton-Colburn analogy holds.

The above system of non-linear partial differential equations has been solved by the finite elements method developed by the commercial package called FEMLAB® 3.1, Comsol Sweden. Each food sample has been discretised into 1374 triangular finite elements (Fig.1) with a thicker mesh close to each boundary. On a 1.0 GHz Pentium III PC running under Windows 2000, food drying behaviour over a time horizon of 6 hours was simulated, on average, in about 15 minutes, using the time-dependent iterative nonlinear solver already implemented in the Chemical Engineering Module of Femlab 3.1. It should be remarked that the proposed model does not need any parameters adjustment, but only the specification of a set of input variables that can be varied within a specific range of physical significance to simulate food drying behaviour at different operating conditions.

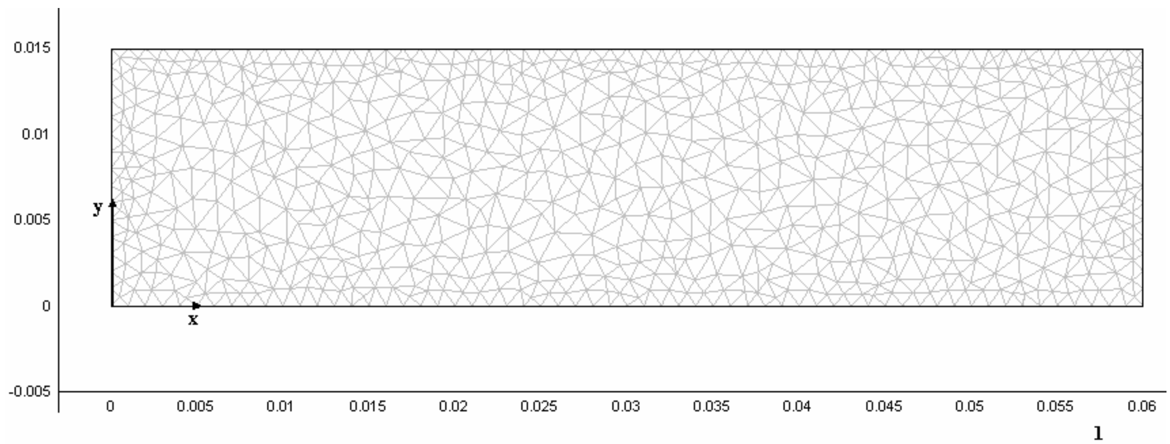

Fig. 1: FEM mesh used to model carrot slab drying process (length expressed in meters).

Results and Discussion

The simulation results expressed as the time evolution of sample average temperature and average moisture content were presented. Both temperature and moisture profiles within the food were also shown as functions of drying time.

Figs. 2 and 3 showed a typical trend of carrot temperature and moisture content during drying process performed at an initial product temperature of 303 K and an initial moisture content of 64%. Air properties were constant and equal to 333 K and 45% for temperature and moisture content, respectively.

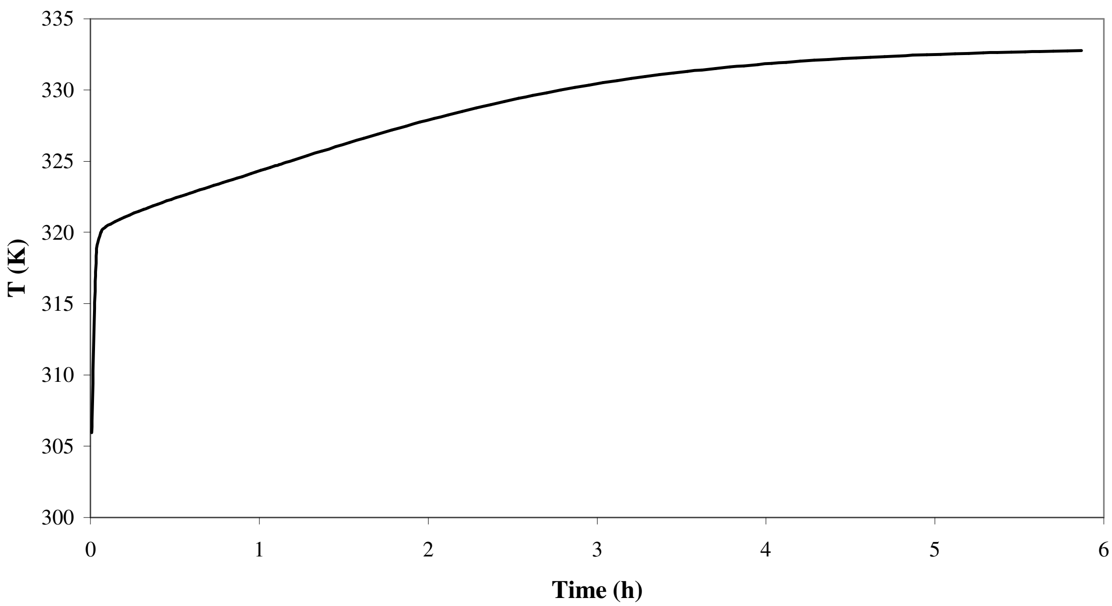

Fig. 2: Food temperature during drying (T_g=333K, U_r=45%, v=0.3m/s, δ=0.003 m, T_0=303K, U_0=0.64, T_{wb}= 320 K).

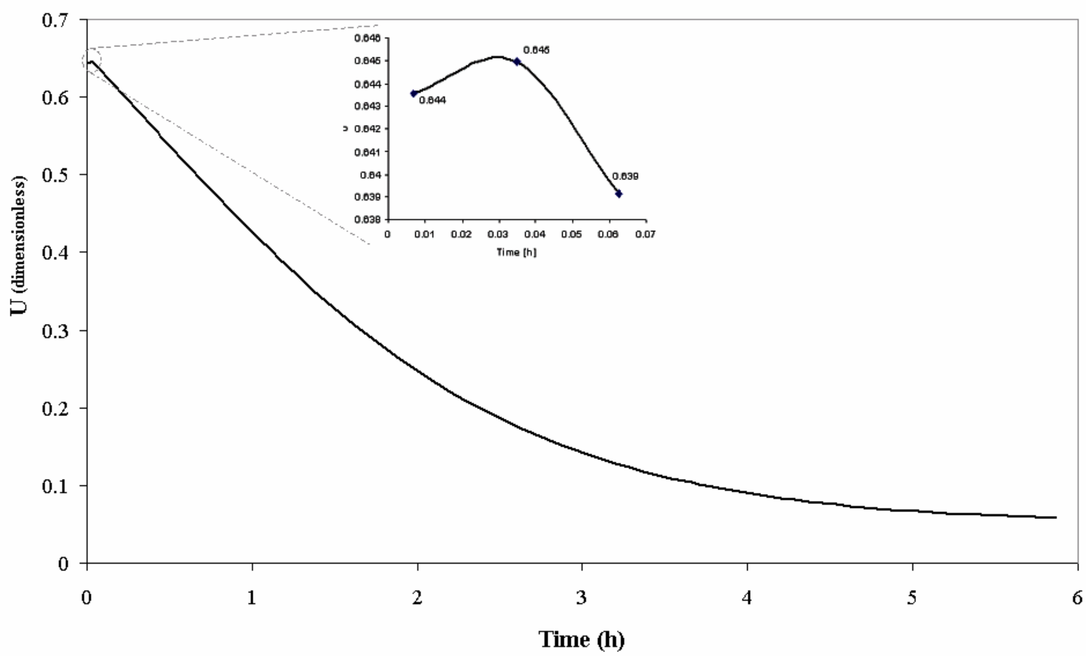

Fig. 3: Food moisture content during drying (T_g=333K, U_r=45%, v=0.3m/s, δ =0.003 m, T_0=303K, U_0=0.64, T_{wb}= 320 K).

Based on Fig. 2, four different regions can be defined; the initial steep temperature rise is, actually the result of two different contributions. In fact, heat transfer depends on the convection, due to the temperature difference between air and solid, and, as soon as Ts is lower than the wet bulb temperature (in the present case 320 K), on the latent heat, due to the condensation of the vapour on food surface. Vapour condensation is also responsible for the corresponding initial slight increase of total food moisture shown in Fig. 3. Once wet bulb temperature was attained C_s exceeded the value of C_{gb} that, depending on air temperature and its relative humidity only, was assumed constant throughout all the process; water evaporation, therefore, prevailed over vapour condensation and heating rate decreases. Hence, while food temperature slowly increased towards air temperature value, food moisture content diminished at decreasing drying rate. Fig.3 showed that initially, when the process was controlled by external heat transfer and the water leaving the sample was not bounded to food structure *(free water)*, drying rate attained its maximum value. Afterwards, when the process rate was controlled by mass transfer and the water leaving the solid was bounded to the food structure, a progressively decrease of drying curve slope was observed. Figs. 4 and 5 gave a detailed microscopic description of food moisture content profiles during the process. They showed very clearly the above-mentioned variation of drying rate; in fact at the beginning of the process, the profiles appeared very well separated, whilst they tended to overlap as soon as the drying rate decreased. It should be noted that the asymmetry shown in Fig. 5 is to be ascribed to the flow conditions that, on each side of the sample, determined a variation of both heat and mass transfer coefficients, becoming sooner dry the side corresponding to the surface subject to dry air impact. The distributions of both temperature and moisture content, corresponding to the above-described drying conditions, were shown in Fig. 6 at a drying time of 3 hours. Figures from 7 to 11 are a representation of drying behaviour as a function of some of the most significant parameters, i.e. carrot initial temperature (T_0) and its thickness (δ), air velocity, temperature and humidity, v, U_r and T_g, respectively. Each parameter has been varied in the range normally used in industrial drying of fresh vegetables.

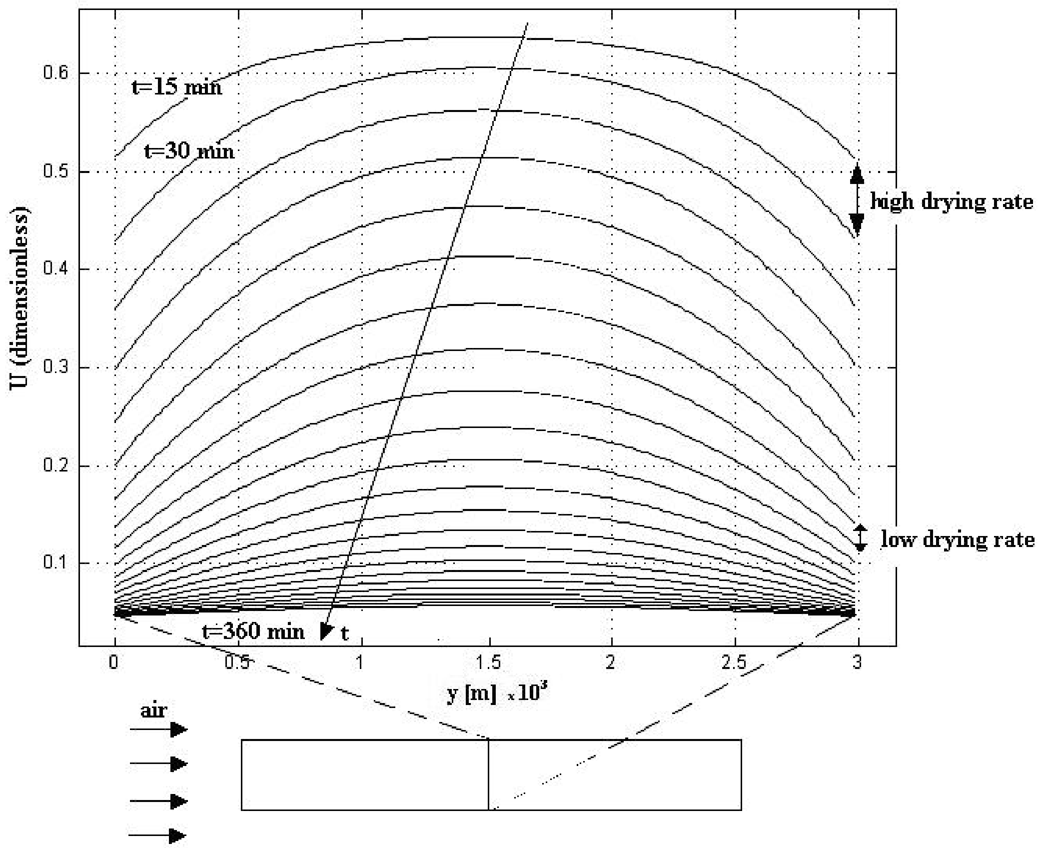

Fig. 4:Dynamic evolution of food moisture content profile at different drying times in a vertical plane (T_g=333K, U_r=45%, v=0.3m/s, δ =0.003 m, T_0=303K, U_0=0.64, T_{wb}= 320 K).

Fig. 7 showed the results obtained at different initial food temperatures. A significant delay in actual drying start-up was noticed when the lowest value of product initial temperature was chosen. In fact, a lower initial temperature corresponded to an increase of the amount of water that condensed on food surface and, therefore, to a delay in reaching the wet bulb temperature. On the contrary, when food initial temperature was already higher than wet bulb temperature, no condensation occurred and food immediately underwent a progressive moisture loss.

As shown in Fig. 8 a reduction of sample thickness determined a decrease of the time necessary to achieve the final target moisture content; this was also due to an earlier release of bounded water in thinner food sample. A decrease of about 34% in food moisture content was observed after 5 hours when food thickness was diminished from 1.5 to 0.3 cm. It should be remarked, however, that the proposed model did not account for shrinkage effects that might affect the results also when the

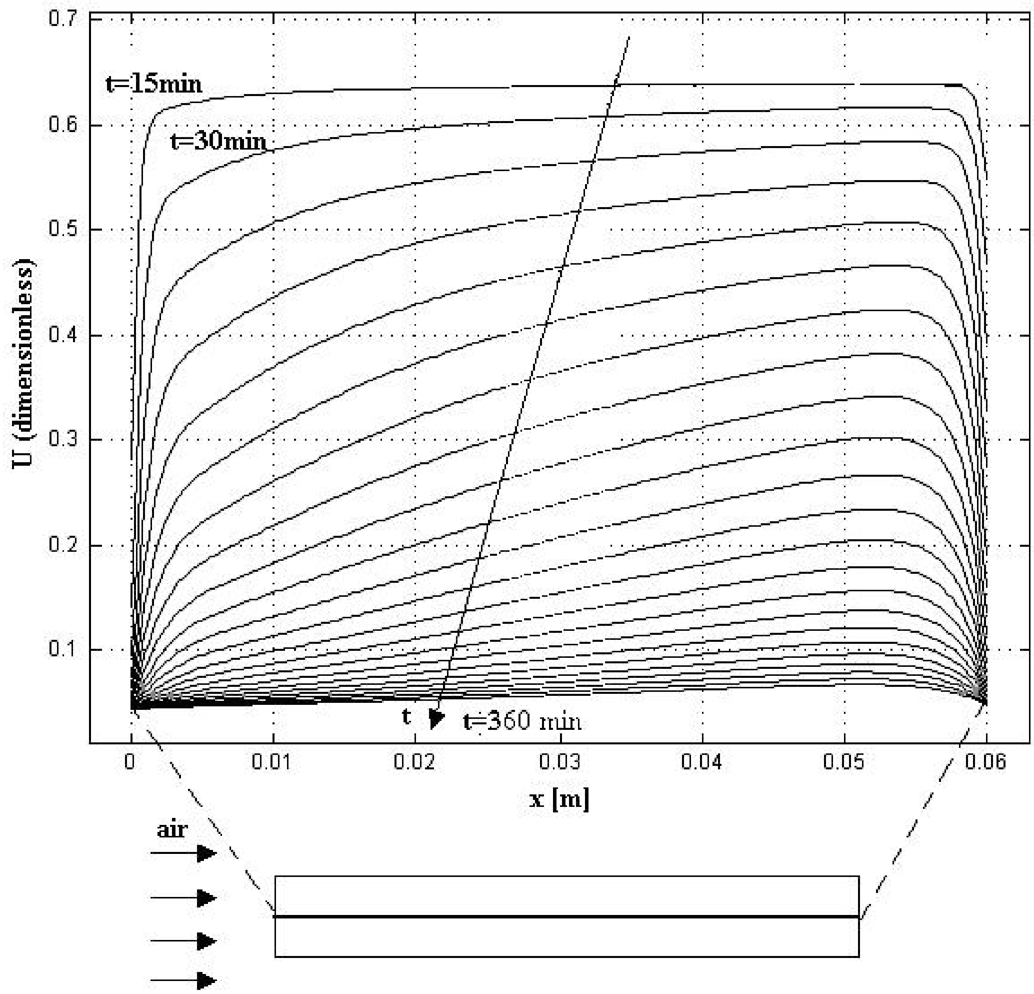

Fig. 5: Dynamic evolution of food moisture content profile at different drying times in the horizontal symmetry plane (T_g=333K, U_r=45%, v=0.3m/s, δ =0.003 m, T_0=303K, U_0=0.64, T_{wb}= 320 K).

drying conditions chosen to perform the numerical simulations (U_r=75%, T_g=323K, v=0.3m/sec) were not particularly severe.

Actually, food characteristics depend on the processes that precede drying and, normally, they do not strictly represent variable parameters even though it was interesting to observe drying behaviour when food initial conditions changed. Drying process is, however, certainly dependent on air characteristics that have to be properly regulated to attain the desired final moisture of the product.

Fig. 9 shows that an increase of air velocity was responsible for an increase of drying rate due to the improvement of both heat and mass transfer coefficients, in fact a 92% reduction of air velocity implied a 9% increase in food moisture content after 5 drying hours. A similar increase of drying rate was observed when a drier air was used (Fig. 10). A decrease of air relative humidity determined an increase of concentration difference that is one of the two driving forces promoting

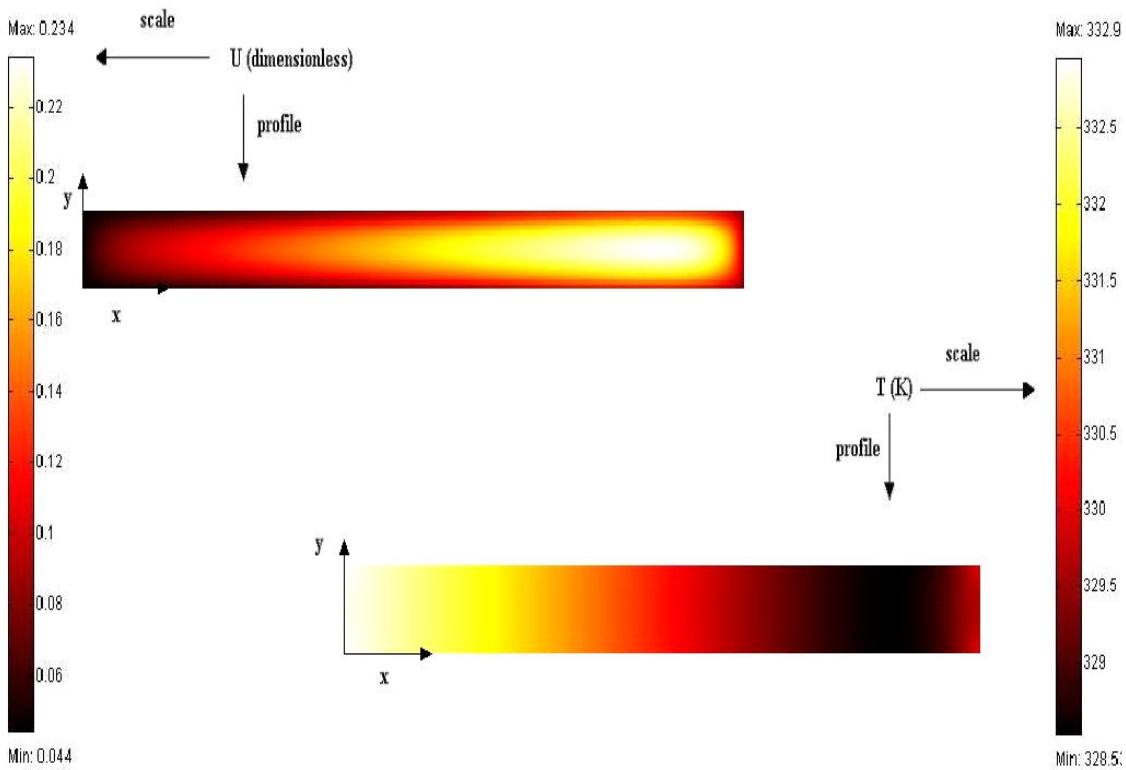

Fig. 6: Food moisture content and food temperature profiles after 3 hours of drying (T_g=333K, U_r=45%, v=0.3m/s, δ =0.003 m, T_0=303K, U_0=0.64, T_{wb}= 320K).

drying process. Fig 11 showed, instead, the effect of the other process driving force, i.e. the temperature difference between air and the sample, on drying behaviour. In this case, the increase of drying rate was lower than that corresponding to air humidity and air velocity variations. A decrease of 66% in air humidity, instead, determined a decrease of 22.6% in food moisture content after 5 hours. An increase of 6% of air temperature corresponded, after 5 hours, to a decrease of 0.6 % in food moisture content. From a physical point of view, this result was a direct consequence of wet bulb temperature rise that drives the system far from equilibrium conditions; in fact, food initial humidification, determined by vapour condensation from air towards food surface, continues for a longer time until wet bulb temperature is reached and condensation stops. Therefore, a trade-off condition on air temperature between the contemporary increase of both the drying rate and the initial moisture content is to be sought in order to choose an air temperature profile that changes during drying process thus minimizing the initial humidification effects.

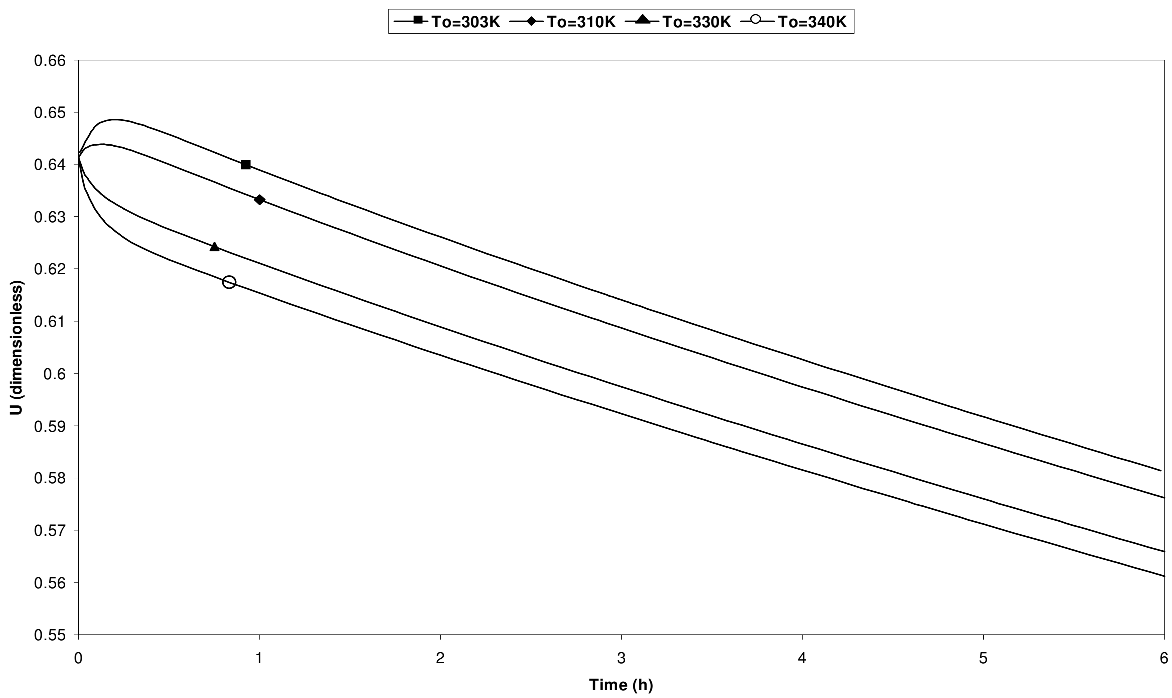

Fig. 7: Dynamic evolution of food moisture content as a function of food initial temperature (T_g=323K, U_r=75%, v=0.3m/s, δ =0.015 m, U_0=0.64, T_{wb}= 318 K).

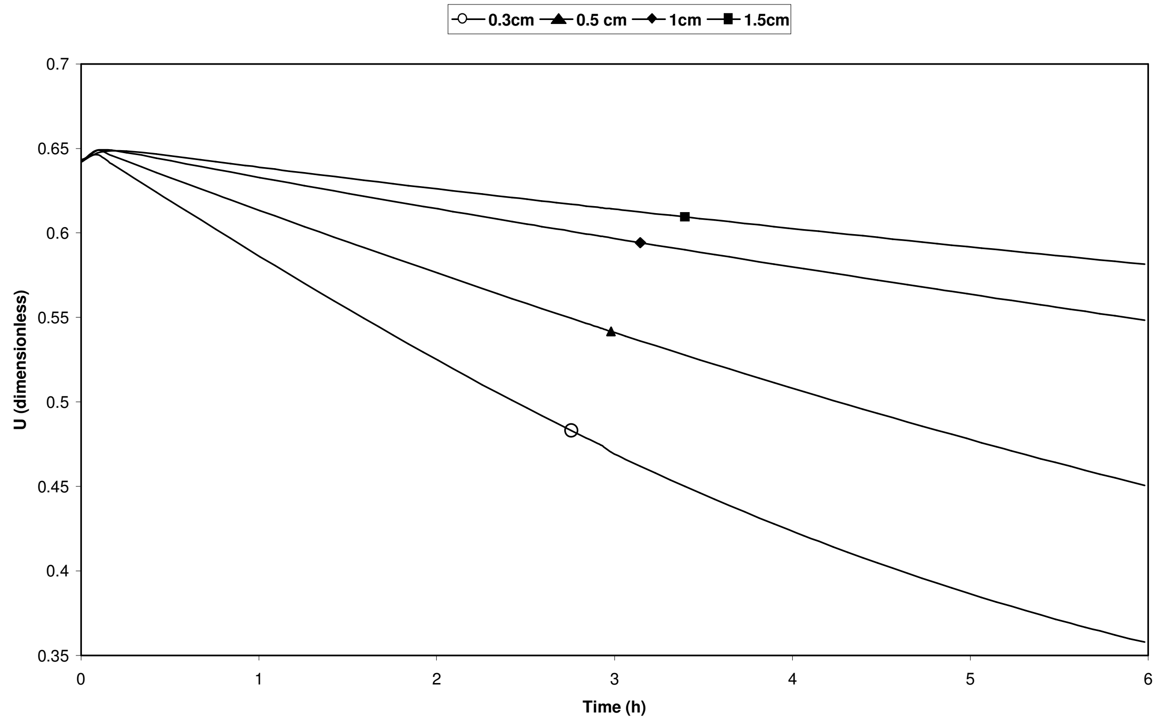

Fig. 8: Dynamic evolution of food moisture content as a function of food slab thickness (T_g=323K, U_r=75%, v=0.3m/s, T_0=303K, U_0=0.64, T_{wb}= 318 K).

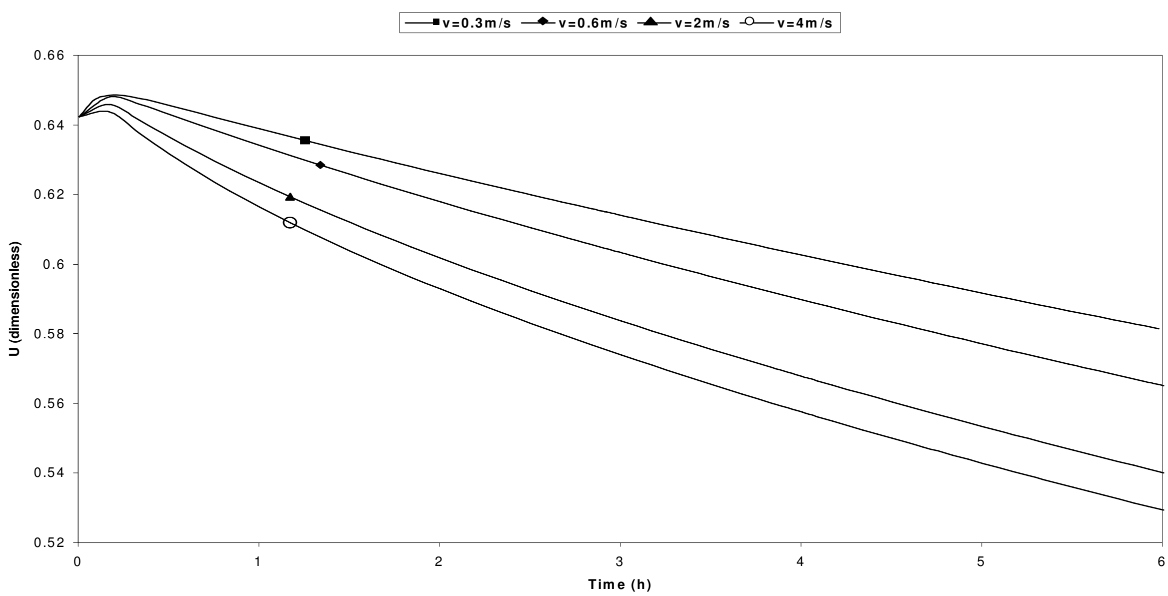

Fig. 9: Dynamic evolution of food moisture content as a function of air velocity (T_g=323K, U_r=75%, δ =0.015 m, T_0=303K, U_0=0.64, T_{wb}= 318 K).

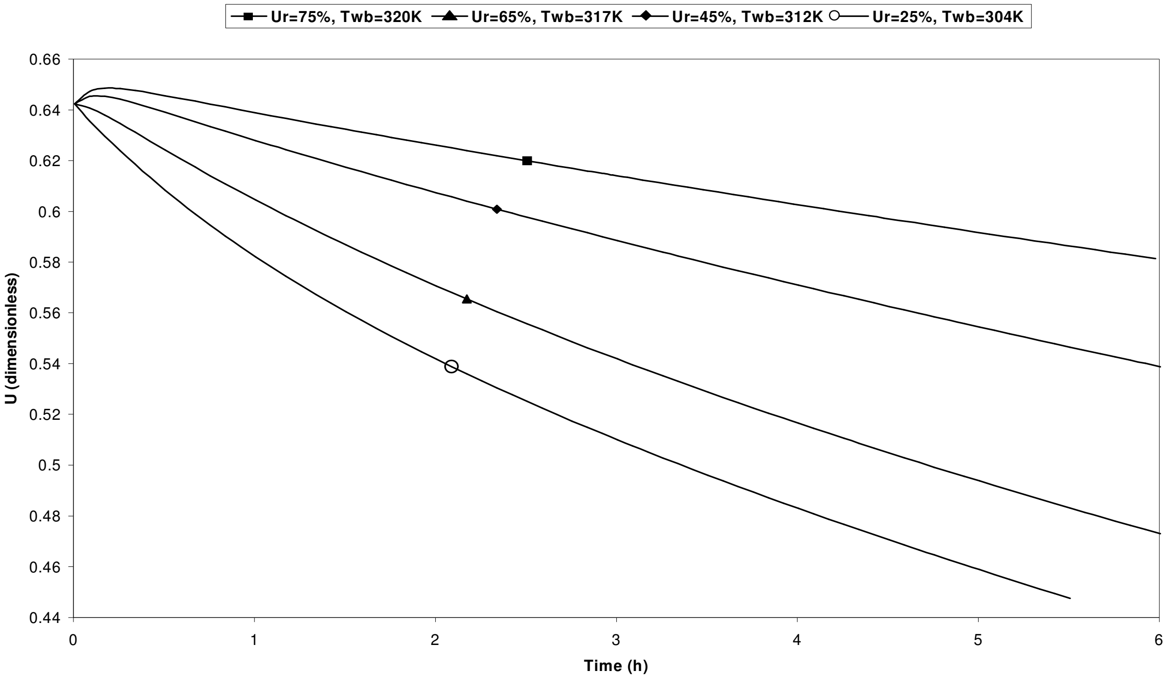

Fig. 10: Dynamic evolution of food moisture content as a function of air humidity (T_g=323K, v=0.3m/s, δ =0.015 m, T_0=303K, U_0=0.64).

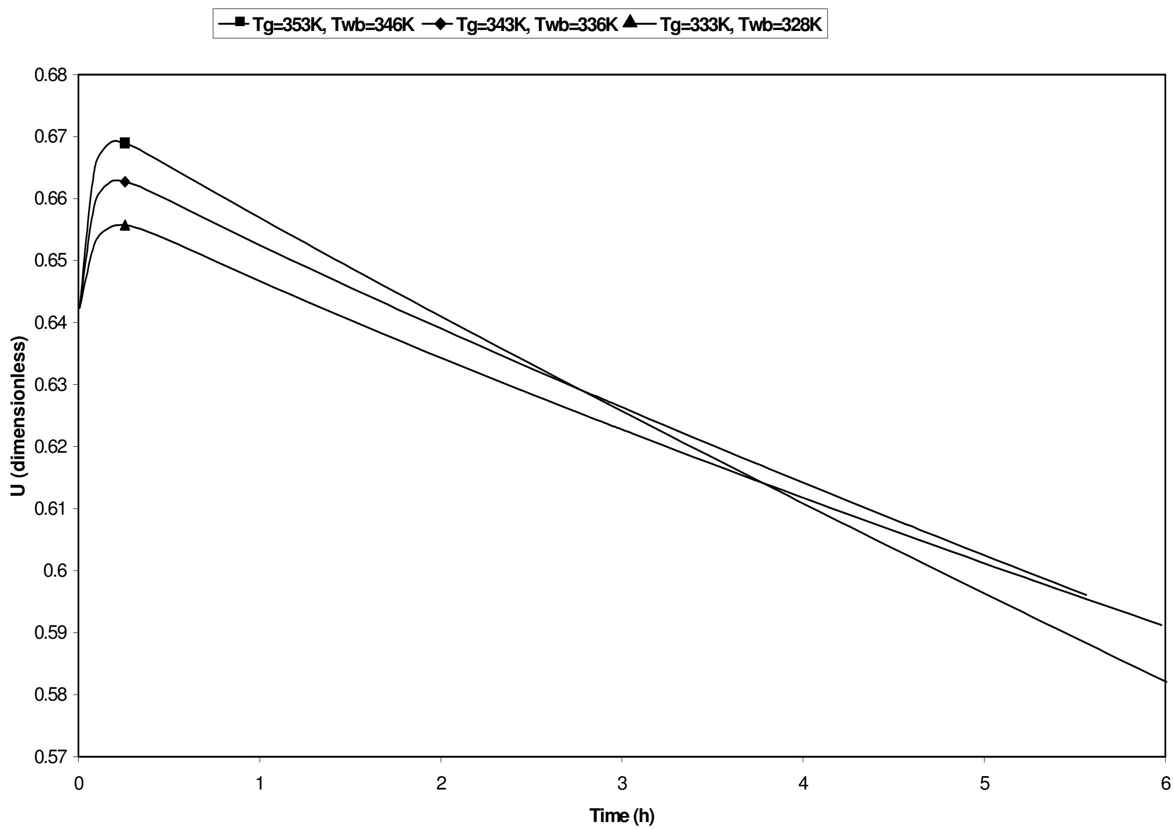

Fig.11: Dynamic evolution of food moisture content as a function of air temperature (U_r=75%, v=0.3m/s, δ =0.015 m, T_0=303K, U_0=0.64).

Model Validation

Many authors performed drying experiments, differing from each other as far as food type, geometrical shape and range of air temperature, air humidity and air velocity were concerned. In this paper, in addition to the numerical simulations carried out on drying process in some typical situations and described in previous sections, some others were performed with the specific aim of verifying the model validity. This, was attained by checking the agreement between the model theoretical predictions and a set of experimental data found in the open literature and regarding the drying of carrots slabs (Ruiz-López , Córdova, Rodríguez-Jimenes, García-Alvarado, 2004). In the last paper, the authors showed the results of some experimental tests performed with carrot slabs at fixed initial moisture content and temperature, dried with air characterized by fixed values of both velocity and total humidity. In particular, they reported two experiments at different values of air temperature and food thickness. One of the experiment was carried out in mild drying conditions (T_g=323K, food thickness = 1.5cm) while the other one in more severe drying conditions (T_g=343K, food thickness = 1cm). On performing the numerical simulation aimed at model validation, either the same food properties (ρ, C_p, k, D, a_{ws}) taken from Ruiz-López et al.'s (2004) paper or the same process conditions exploited were used. Fig. 12 shows the comparison between the experimental results and the model predictions, obtained without any parameter adjustment.

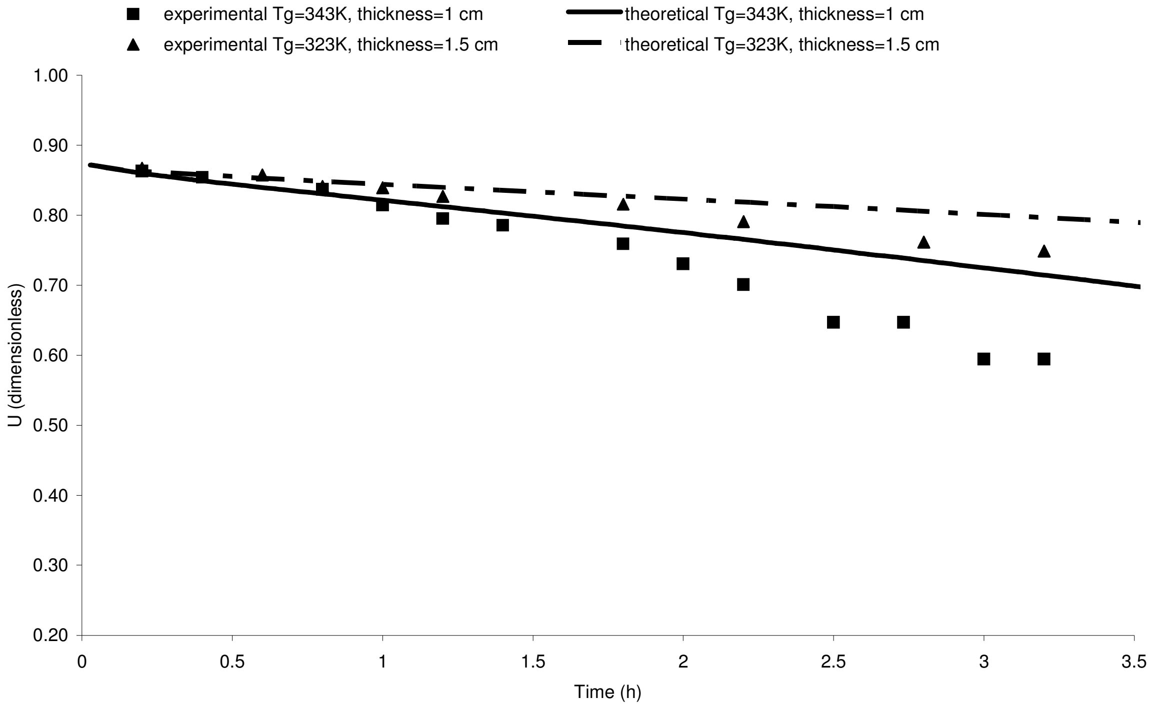

Fig.12: Comparison between model predictions and experimental data - Dynamic evolution of food moisture content (air humidity=0.018 Kg water vapour/Kg dry air, v=2m/s, T_0=300K, U_0=0.87).

A remarkable agreement is observed especially during the first 2 hours when experimental data and theoretical predictions overlap and relative errors never exceed 5%. Later on, a slight deviation can be observed, particularly for the drying test performed with higher air temperature and smaller food thickness, i.e. when a significant volume reduction (shrinkage) was possibly achieved. The observed deviation can be, therefore, ascribed to the lack of accounting for shrinkage effects in the model.

Conclusions

A theoretical analysis of vegetables drying process was presented. With respect to most of the works published in literature, the main innovation introduced by this study was represented by the model formulation. This, in fact, simulated the simultaneous bidimensional heat and moisture transfer accounting for the variation of both air and food physical properties as functions of local values of temperature and moisture content. The numerical solution of unsteady-state heat and mass balance equations, performed by the Finite Elements Method, gave the following results:

* if initial food temperature is lower than wet bulb temperature a preliminary condensation of water on food surface is observed. This phenomenon causes an initial delay in actual drying beginning; on the contrary, when food initial temperature is already higher than wet bulb temperature, no condensation occurs and food immediately undergoes a progressive moisture loss;
* a trade-off condition on air temperature between the contemporary increase of both the drying rate and the initial moisture content is to be sought in order to choose an air temperature profile that changes during drying process thus minimizing the initial humidification effects;
* the agreement between experimental results and model predictions was rather good, particularly during the first 2 hours of operation; when significant shrinkage effects occurred a considerable deviation between theoretical predictions and experimental results can be observed.

The model that contains no adjustable parameter, represents a powerful tool for analyzing the behaviour of a industrial drying ovens as it could be used to individuate, over a wide range of drying conditions and of different types of foods, the “optimal” set of operating conditions in each particular situation. In this way, it might be possible to minimize expensive pilot test-runs and to have good indications on the characteristics and the quality of final products.

The proposed model is general since it is capable of describing both the transport of free and bounded water within the food and the evaporation/condensation phenomena that, as a function the actual driving forces, may occur at the air/food interface. It is very flexible and can be easily adapted to different geometries with no *a priori* limitation for irregular or complex-shaped systems.

Work is, however, in progress for the model features improvement, in particular by seeking an independent and more reliable determination of heat and mass transfer coefficients and accounting for food sample shrinkage during drying process.

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Mathematical modelling of food drying process

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Abstract

Mathematical modeling represents a very important and effective tool for a proper design and control of industrial processes. A model reliably predicting a particular transformation process can, in fact, be used to investigate how the outputs may change with time under the influence of changes of the external disturbances and manipulated variables. In this way, it is possible to optimize the process, thus improving the quality and the safety of the final product. A mathematical model is, generally, based on a relationship, expressed in form of an equation (or a system of equations), whose solution yields the dynamic or static behaviour of the process under examination. Both finite element method (FEM) and finite difference (FD) modelling have been utilized in food engineering research. A detailed analysis of the literature in the last 15 years shows that FEM applied to heat transfer dominates the publications, followed by diffusion calculations and drying process simulations. It is to be remarked, however, that a widespread use of mathematical modelling in food engineering is far from being well assessed.

The main aim of the present chapter is to analyze the transport phenomena involved in food drying process, performed in a convective drier. The formulation of two different theoretical models will be presented, focusing the attention on the differences that may be obtained if a simplified or a more complete analysis of the same transport problem is adopted. Both the models describe the simultaneous transfer of heat and moisture occurring during drying process. Actually, the first approach is much simpler, even though it could be considered a very important advance with respect to the theoretical analyses that are currently available in the literature. On developing the "simple" model, only food domain has been taken into consideration; moreover, heat and mass transfers occurring at the food surface exposed to the drying air have been estimated on the basis of a set of semi-empirical correlations available in the literature. The second model, instead, takes into account also the behaviour of the drying air flowing, in turbulent conditions, about the food sample. This approach is, therefore, more general since it describes the simultaneous transfer of momentum

(for air only), of heat and mass (for both air and food) occurring in a convective drier and does not need the specification of any heat and mass transfer coefficient at the food-air interface that, indeed, is one of the results of the proposed model.

Both the models receive - as inputs - only the initial conditions, the geometrical characteristics of both food and drying chamber and the relationships expressing physical and transport properties of food and air in terms of the local values of temperature and moisture content. The resulting system of non-linear, unsteady-state partial differential equations has been solved by means of the Finite Elements Method. The comparison between two possible different approaches may suggest if a significant increase of computation effort is actually required or, instead, if the utilization of a much simpler and faster method is capable of giving a proper description of the process under consideration. The main objective of the present work is to show how an accurate transport model can be used to determine the influence of operating conditions on drying process. In this way, it might be possible to minimize expensive pilot test-runs and have good indications on the characteristics and the quality of dried products.

Introduction

General Considerations on Drying Process

Drying, or dehydration, is one of the most common unit operation utilized for food preservation. The words drying and dehydration are often reported as synonyms even though it is more appropriate to define a food as dehydrated when its water content is rather low, typically less than 2.5% (Barbosa-Cànovas and Vega-Mercado, 1996). The decrease of water content in food allows reducing both microbial spoilage and deterioration reactions since water is essential for the metabolism of several bacteria and microorganisms. Moreover, food drying can reduce packaging, handling, storage and transport costs because a decrease of food weight and, in most cases, also of its volume is achieved. Several different techniques are used to perform water removal. Among these, freeze-drying, osmotic drying, vacuum drying, microwaves or radio-frequency drying are very well known and are adopted in several industrial applications. Nevertheless, drying by dry and hot air flowing about food samples is, definitely, the most common industrial method to promote water transfer from food, thus reducing its moisture content. On analyzing drying process, it is necessary to characterize the mechanisms determining the water transport within, but also, outside the food. Actually, water transport may be due to a combination of different phenomena: i.e. inner diffusion of water due to concentration gradients, diffusion of water, as a vapor, in the pores filled with air, capillary forces, convective flow determined by pressure differences and flow of water due to vaporization/condensation.

A proper analysis of drying process is usually performed monitoring the variation of food weight with respect to time. The collected experimental data allow calculating the time evolution of food moisture content, X, defined as the ratio (on a mass basis) between the actual amount of water in the food and the amount of dry solids. It is, however, more significant to introduce another variable, i.e. the so called free moisture content, X_f, defined as the difference between X and the moisture content that can be measured when equilibrium conditions are attained, X_{eq}. A drying curve, shown in a typical case in Fig. 1, is obtained plotting free moisture content versus time.

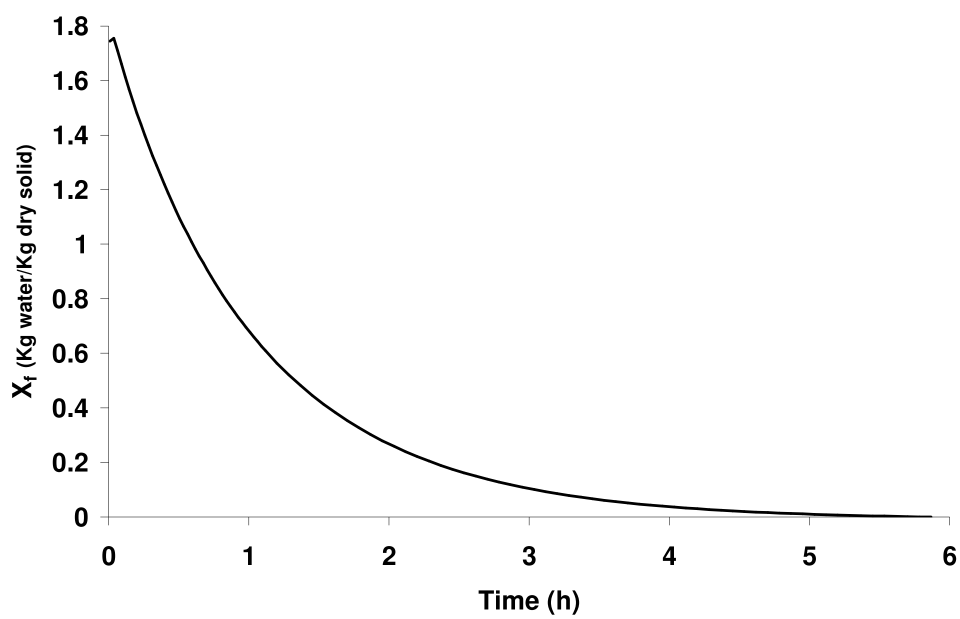


Fig. 1: Drying curve in a typical situation.

The drying rate, R, is proportional to the time derivative of X_f; therefore, the slopes of the drying curve, evaluated for each point, may be used either to estimate the time evolution of drying rate or to determine, as shown in Fig. 2 from a qualitative point of view, the relationship existing between drying rate and free moisture content.

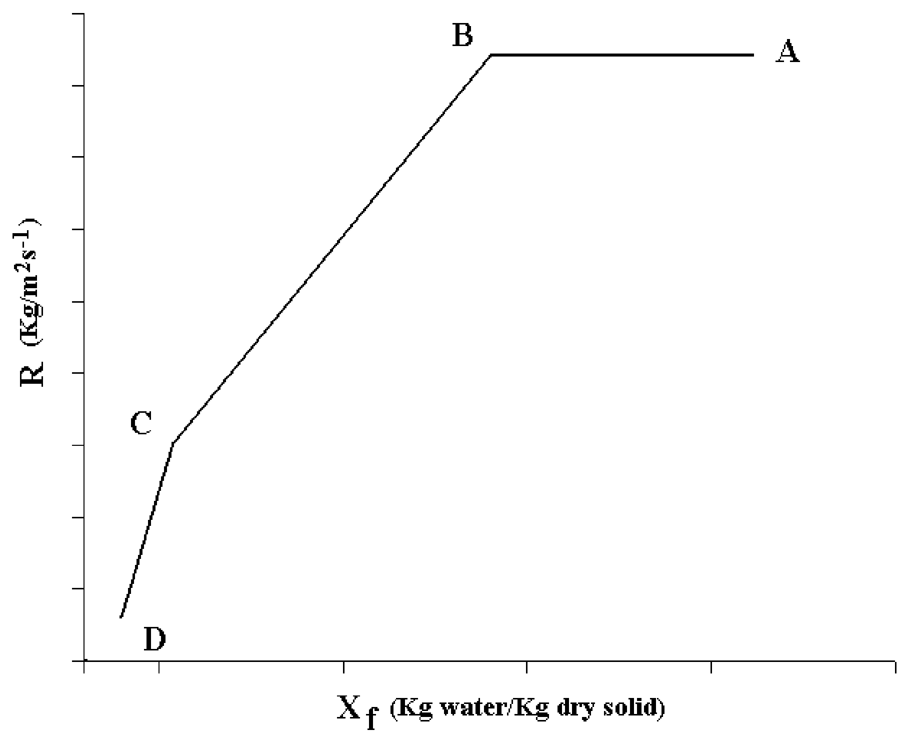


Fig. 2: Relationship between drying rate and free moisture content in a typical situation.

An examination of the above figure puts in evidence that drying usually takes place in different stages, each characterized by a different rate. The initial stage, where drying rate does not change with respect to X_f (line A-B), is known as the constant drying rate period. During this phase, vaporization occurs at the product surface and only unbound water, i.e. the water characterized by unity activity coefficient, is removed from the product. The transport phenomena mostly involved in this stage are the mass transfer of water, as a vapor, from the food surface into a boundary layer developing in the drying air and the heat transfer in the material. The rate at which water is transported from the inner parts of the food is very fast and is capable to compensate the amount of water evaporating at the surface that, therefore, maintains saturation conditions. The temperature of food surface is approximately equal to the wet bulb temperature.

The second stage, known as the falling rate period, starts when free water is no longer available at the surface; water activity on the surface tends to decrease and assumes values lower than unity. The falling rate period may be subdivided in two following stages, recognizable by different slopes in R vs. X_f curve. The first period takes place until food surface is entirely dry (line B-C). The second one, instead, is characterized by a displacement of evaporation front from the surface to food internal regions (line C-D); water vapor has to diffuse through a thicker and thicker layer of already dried material before reaching the surface where it mixes with the air flowing about the sample. During falling rate period, the amount of water removed from the food may be rather small and, consequently, the time necessary to complete this stage is, generally, long. This is mainly due to the fact that drying rate is controlled by diffusion of water toward the surface. Actually, the transport of water within the food structure is a rather complicated phenomenon that can be explained with reference to several mechanisms that include the diffusion of liquid promoted by concentration gradients, the diffusion of water, as a vapor, due to partial vapor pressure differences, the transport of liquid water caused by either capillary or gravity forces, the surface diffusion. In spite of the complexity of the above-described phenomena, a simple diffusion equation based on Fick's law has been widely used in the literature to describe the dependence of food moisture content upon time and position. Nevertheless, it is suggested to make use of an effective diffusion coefficient (D_{eff}), a parameter estimated from the experimental data collected during drying process that accounts for the interactions existing between the water to be transported and the food structure.

Convective Food Drying

When dry and warm air flows about a moist and cold food sample, a simultaneous transfer of both heat and water occurs. In particular, heat is transferred from air to the material; water is transported from the food core, then, to its surface and, eventually, to air. The rates of heat and mass transfer depend on both temperature and concentration differences, but also on the air velocity field that strongly affects the transfer rate and, therefore, has to be properly evaluated. Typical values of air temperature range between 35°C and 80°C, air relative humidity varies between 10% and 70%, while air normally flows at a velocity ranging from 0.5 to 5 m/sec, reaching - in some cases - the value of 10 m/sec. The main aims of convective drying are a decrease of food water content and an increase of its temperature: both the above effects improve food preservation since microbial spoilage is favoured by low temperature and high moisture content.

Convective drying is definitely the most common method for food preservation, so several different approaches have been proposed to model this process (Wang & Sun, 2003; Datta 2007 Part I). The models available in the open literature may be subdivided in two categories: simplified and complex approaches. The latter are often regarded as too onerous and time consuming for practical purposes; the former, instead, are quite simple but are based either on simplification hypotheses, not applicable in several real cases, or on the utilization of empirical correlations, necessary to estimate - by means of a set of transport coefficients – the heat and water fluxes at food-air interface. In simplified models the utilization of the so-called *thin-layer equation* (Ertekin & Yaldiz, 2004; Doymaz, 2004; Karathanos, 1999; Krokida, Karathanos, Maroulis & Marinos-Kouris, 2003) is quite common. It is based on the hypothesis that food moisture reduction is proportional to the instantaneous difference between the material moisture content (assumed uniform within the food) and the water concentration calculated under the assumption that equilibrium conditions with drying air do actually exist. A considerable simplification is, therefore, attained since the description of all the transport phenomena involved in drying process can be reduced to the solution of a single ordinary differential equation. The above-mentioned proportionality is expressed in terms of a constant, known as the drying constant, that is a function of several parameters such as the material moisture content, the size of product and its temperature distribution (assumed uniform), and the main characteristics of drying air, i.e. its humidity, temperature and velocity. Also other simplified models, based on a single equation but supported by fundamental considerations, were proposed; they actually model the water transport in food by means of Fick's diffusion equation and assuming that drying can be considered as an isothermal process (Hernàndez, Pavòn, Garcìa, 2000; Simal, Femenia, Llull, Rossello, 2000). Nevertheless, a more rigorous approach should also take into account for the heat transfer in food (Wang, Brennan ,1995; Kalbasi, Mehraban, 2000; Migliori, Gabriele, de Cindio, Pollini, 2005) by means of Fourier's equation of conduction or, in a simpler way, assuming a uniform temperature profile in food, thus considering a macroscopic heat balance (Rovedo, Suarez, Viollaz, 1995). It should be observed, however, that all the above-mentioned models are based on the solution of a set of transport equations that are defined in terms of both fundamental transport properties, like the diffusion coefficient of water in food and the effective thermal conductivity of food, and physical parameters, like the heat capacity and the density of the material to be dried. Their values actually depend on food structure, on the local values of both temperature and moisture content and are very difficult to estimate due to the complex chemical and physical changes that take place during food drying. The importance of considering the variability of physical and transport properties with food temperature and moisture content was investigated by several authors (Datta, 2007 Part II, Panagiotou, Krokida, Maroulis, Saravacos, 2004, Lewicki, 2004). In particular, Datta described the different possible procedures that can be adopted for the evaluation of food properties; the author remarked the need to utilize a more rigorous methodology since every single complex phenomenon involved in food processing plays a crucial role and might strongly affect parameters estimation. Saravacos & Maroulis (2001) confirmed this necessity, showing that huge differences are found in the literature as far as the estimation of the same property, such as the diffusion coefficient of water in the food and the food thermal conductivity, is concerned.

As it will be shown more in detail in the following section, the transport phenomena involved in food drying process are rather complex since they involve a simultaneous transfer of momentum, heat and mass. The governing equations form a set of un-steady, non-linear, partial differential equations that can be solved only by means of numerical procedures. Both finite difference (FD) and finite element (FEM) methods have been widely used for this purpose. It was reported that FEM is particularly suitable to perform investigations on domains characterized by irregular geometries, in presence of complex boundary conditions and for heterogeneous materials. Nevertheless, it is more complex and computationally expensive than FD (Wang & Sun, 2003). To solve drying balance equations, suitable boundary conditions, especially at food-air interface, are required. They could be either Neuman-type or Dirichlet-type; the former, however, gives better results than the latter (Markowski, 1997). Dirichlet-type boundary conditions are defined specifying at food-air interface both water concentration and food temperature. Neuman-type boundary conditions are defined with reference to a set of heat and mass transfer coefficients that are estimated from the experimental data collected during drying process. Most of available literature data are expressed in terms of semi-empirical correlations and refer to the estimation of heat transfer coefficient (Saravacos, Maroulis, 2001); the Chilton-Colburn analogy is, however, usually adopted to estimate mass transfer coefficient once heat transfer coefficient has been determined (Bird, Stewart, Lightfoot, 1960). These correlations are parametric equations expressed in terms of characteristic dimensionless numbers. In particular, Nusselt number, *Nu*, is related to both Reynolds, Re, and Prandtl, Pr, numbers by a general equation like:

$$Nu = a \mathrm{Re}^b \mathrm{Pr}^c \qquad 1$$

in which a, b and c are model parameters.

In the case of quite common geometries (i.e. slabs, cylinders, spheres), the model parameters are very well known in a wide range of fluid-dynamic and process conditions. In other practical cases, experimental test-runs have be performed to estimate model parameters as a result of best fitting procedures. The available data, however, put in evidence that large variations do, actually, exist (Saravacos, Maroulis, 2001). This is to be ascribed to the fact that even small differences in fluid-dynamic conditions or in food characteristics cause relevant deviations in the estimated values of transfer coefficients. As a matter of fact, Kondjoyan and Boisson, 1997) and Verboven et al. (1997 and 2001) verified that minor errors in the estimation of transfer coefficients, can lead to huge variations in the calculated time evolutions of both food temperature and its moisture content, thus leading to an unfair design of drying equipments or to severe problems during food processing. In principle, it should be observed that a very good prevision model might fail if heat and mass transfer coefficients are estimated by means of unsuitable literature correlations. In spite of this real difficulty, many authors proposed transport models based on the utilization of semi-empirical literature correlations to evaluate heat and mass fluxes at the food-air interface (Pavón-Melendez, Hernández, Salgado, García, 2002; Karim & Hawlader, 2005; Aversa, Curcio, Calabrò, Iorio, 2007; Wang & Brennan, 1995). To overcome the problems related to the utilization of a proper set of heat and mass transfer coefficients, it would be, however, advisable to formulate a more general drying model. This is to be based on the resolution of the fundamental transport equations (momentum, heat and mass) in both air and food domains, achieved by the utilization of a proper set of boundary conditions at the food-air interfaces. In this way, no literature correlation is actually required and the model might fit for all possible food shapes and all the operating and fluid-dynamic conditions. Nevertheless, a so exhaustive analysis has been often regarded in the past years as too onerous in terms of computational effort and calculation time.

In the following, two different approaches to the modelling of convective food drying process will be presented. The attention will be focused on the differences that can be obtained if either a simplified or a more complete analysis of the same transport problem is adopted. Both the models describe the simultaneous transfer of heat and moisture occurring during drying process. Actually, the first approach is much simpler, even though it could be considered a very important advance with respect to the theoretical analyses that are currently available in the literature. On developing the “simple” model, only food domain has been taken into consideration; moreover, heat and mass transfers occurring at the food surface exposed to the drying air have been calculated on the basis of semi-empirical correlations available in the literature.

The second model, instead, takes into account also the behaviour of the drying air flowing, in turbulent conditions, about the food sample. This approach is, therefore, more general since it describes the simultaneous transfer of momentum (for air only), of heat and mass (for both air and food) occurring in a convective oven.

The main objective of the present chapter is to present a rather detailed analysis of the transport phenomena occurring during drying process in order to develop a proper transport model that can be used to simulate the behaviour of real dryers. In particular, the comparison between two possible different approaches may suggest if a significant increase of computation effort is actually required or, instead, if the utilization of a much simpler and faster method is capable of giving a proper description of the process under consideration.

Theoretical

In the following, the main assumptions and the mathematical equations used to model the unsteady-state behaviour of a convective drying oven will be shown for both the models that are to be developed.

It is supposed that drying air is continuously supplied to the oven inlet section and flows about the food, actually a vegetable, in the axial direction (y), parallel to its main dimension (Fig. 3). Heat and mass transfer resistance are assumed negligible across the net on which the vegetable is placed (Thorvaldsson & Janestad, 1999; Viollaz & Rovedo, 2002). On the basis of the above hypothesis, the system under investigation can be considered as a classical symmetric one; a symmetry axis, in fact, may be individuated and only half of the original domains will be henceforth considered. Moreover, any possible variation occurring with reference to the spatial coordinate z is assumed not relevant for the present study. This allows analyzing a 2D geometry so that food sample can be considered as a slab and the oven as a rectangular chamber, characterized by a length of 25 cm and by a height of 10 cm. Food sample is placed 8 cm far from the oven inlet section and it is 6 cm long and 5 mm thick. Each dependent variable is expressed as a function of two spatial coordinates, *x* and *y*, and of time, *t*.

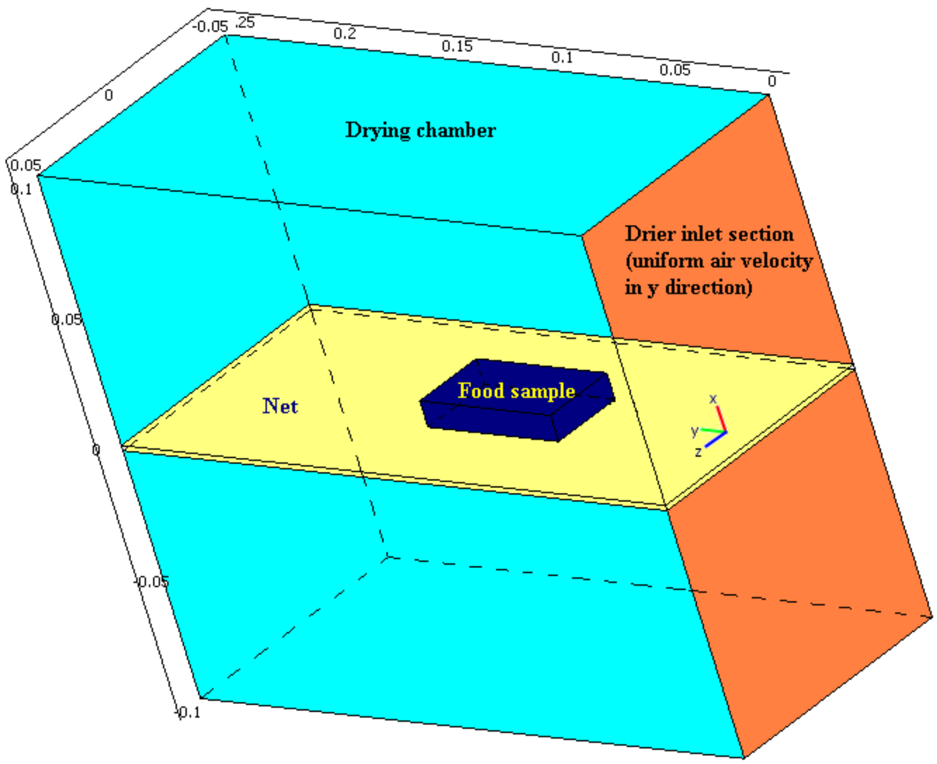


Fig. 3: Schematic representation of the drying chamber under consideration.

Both the proposed models account for the variation of air and food physical properties, defined in terms of the local values of temperature and moisture content. As far as the food sample is concerned, a continuum approach has been chosen; it has been supposed, in fact, that the actual multiphase hygroscopic porous medium may be replaced by a fictitious continuum, any point of which has variables and parameters that are continuous functions of point spatial coordinates and of time. Convective contributions in the transport equations written for the food have been neglected, assuming weak inner evaporation. Also the transport of water vapor by diffusion within the dehydrated material toward the external food surface has not been considered since this mechanism is significant especially for highly porous media, whereas for the vegetables under consideration - typically characterized by void fraction lower than 0.3 (May & Perré, 2002) – can be neglected. Mass transfer in the product occurs, therefore, only by diffusion; heat transfer in the product occurs only by conduction. Shrinkage effects were assumed negligible in the range of average moisture content taken in consideration.

Simplified Approach

On the basis of the above discussion, water and heat transfers occurring during food drying process were modelled by the unsteady state mass and energy balances, respectively, whereas evaporation, occurring at air-food interfaces only, was considered by defining a proper set of boundary conditions expressed in terms of heat and mass transfer coefficients estimated by empirical correlations available in the literature (Perry & Green, 1984).

The energy balance in the solid, based on Fourier's law, leads to

$$\rho C_p \,{}^{\partial T}\!/_{\partial t} = \underline{\nabla} \bullet \left(k_{eff}\, \underline{\nabla} T\right) \tag{2}$$

where ρ is the food density, C_p is its heat capacity, T is the temperature, t is time and k_{eff} is the effective thermal conductivity of food.

The mass balance, based on Fick's law, leads to:

$${}^{\partial C}\!/_{\partial t} = \underline{\nabla} \bullet \left(D_{eff}\, \underline{\nabla} C\right) \tag{3}$$

where C is water concentration in food and D_{eff} is the effective diffusion coefficient of water in food.

The subscript *eff* considers that food transport properties are evaluated as effective indexes, thus accounting for a possible combination of different transport mechanisms.

Supposing that the above material parameters and transport properties (ρ, C_p, k_{eff}, D_{eff}), in the most general case, depend on the local values of food moisture content and of temperature, Eqs. 2-3 form a system of unsteady, non-linear partial differential equations (PDEs). The expressions of ρ, C_p, k_{eff}, D_{eff}, derived by Ruiz-López, Córdova, Rodríguez-Jimenes, García-Alvarado (2004) in the case of carrot slices drying, have been used.

The initial conditions are straightforward since it has been assumed that before drying process actually begins (t = 0), food moisture content and its temperature have definite values, i.e. C_0 and T_0, that are to be specified before performing each simulation.

The boundary conditions relative to eq. 2 apply to each external surface of the food sample, where no accumulation occurs; they actually state that the heat transported by convection from air to food is partially used to raise sample temperature by conduction and partially to allow free water evaporation:

$$h\left(T_{gb} - T_s\right) = -\underline{n} \bullet \left(-k_{eff}\, \underline{\nabla} T\right) - \lambda \cdot N_s \qquad 4$$

where λ is the latent heat of vaporization for water, N_s is the diffusive flux of water at the food surface, h is the heat transfer coefficient, T_{gb} is the bulk temperature of air, T_s is the temperature at the food surface; $\underline{n}$ is a generic unity vector normal to the surface.

The boundary conditions relative to Eq. 3, apply to each external surface of food sample, where no accumulation occurs; they express the balance between the diffusive flux of liquid water transported from the product core to the surface and the flux of vapor that leaves the food surface and is transferred to the drying air:

$$-\ \underline{n} \bullet \left(-D_{eff}\, \underline{\nabla} C\right) = k_c\left(C_i - C_{gb}\right) \qquad 5$$

where k_c is the mass transfer coefficient, C_{gb} is the bulk concentration of water in air, C_i is water concentration evaluated in gaseous phase at the food/air interface.

An equilibrium relationship between the water concentration in the air and the water concentration, on the food surfaces actually exposed to the drying air, can be also formulated (Smith, Van Ness, Abbot, 1987):

$$\gamma_w \, x_w \, f_w = \hat{\phi}_w \, y_w \, p \qquad 6$$

The superscripts *v* and *l* (vapour and liquid, respectively) were omitted with the understanding that γ_w, the activity coefficient of water and f_w , the fugacity of water, refer to the liquid phase, $\hat{\phi}_w$, the fugacity coefficient of water, refers – instead - to the vapor phase; x_w and y_w are the molar fractions of water in food and in air, respectively, p is the pressure within the drying chamber. The fugacity of water can be expressed by:

$$f_w = \phi_w^{sat} \, P_w^{sat} \exp\left[\frac{V_w\left(p - P_w^{sat}\right)}{RT}\right] \qquad 7$$

where P_w^{sat} is the vapor pressure of water, expressed in terms of the local values of temperature, and the exponential term is known as the Poynting factor.

If the quantity Φ_w is introduced:

$$\Phi_w = \frac{\hat{\phi}_w}{\phi_w^{sat}} \exp\left[-\frac{V_w\left(p - P_w^{sat}\right)}{RT}\right] \qquad 8$$

The following relationship is obtained:

$$\gamma_w \, x_w \, P_w^{sat} = \Phi_w \, y_w \, p \qquad 9$$

At low pressures (up to at least 1 bar), vapor phase usually approximates ideal gases ($\hat{\phi}_w = \phi_w^{sat} = 1$) and the Poynting factor differs from unity by only a few parts per thousand; moreover, values of $\hat{\phi}_w$ and ϕ_w^{sat} differ significantly less from each other than from unity and their influence in Eq. 8 tends to cancel. Thus, the assumption that $\Phi_w = 1$ introduces a little error for low-pressure vapor-liquid-equilibrium (VLE) and allows simplifying Eq. 9 to:

$$\gamma_w \, x_w \, P_w^{sat} = y_w \, p \qquad 10$$

It should be observed that in hygroscopic materials, like most of the foods, the parameter γ_w accounts for the effects related to the amount of physically bound water so it is usually expressed as a function of both food moisture content and of its temperature (Datta, 2007 Part II, Ruiz-López, Córdova, Rodríguez-Jimenes, García-Alvarado, 2004). Once the activity coefficient is known for the particular food under examination, Eq. 10 permits calculating the molar fraction of water in the vapor phase and, therefore, the value of C_i that appears at the right-hand side of the above Eq. 5.

Eqs. 4, 5 and 10 are essential to properly describe the complex steps involved in drying process. Eq. 10, in fact, is expressed in terms of water activity coefficient γ_w that decreases as food moisture content decreases so that a unity value of γ_w means that only unbound (free) water actually evaporates. When $\gamma_w < 1$ also bound water is removed from the food sample. Water activity coefficient is a distinctive parameter of each product that, being characterized by its own structure, determines how strong the bonds between food structure and water are. Moreover, it should be observed that both the moisture content and the temperature on food external surfaces determine, through Eq. 10, the value of food surface water concentration and then the time evolution of drying process. In particular, if on the food surface either moisture content or temperature are high enough to have $(C_i - C_{gb}) > 0$, then, on the basis of Eq. 5, food drying starts immediately; whereas if their values are such to have $(C_i - C_{gb}) < 0$, then a preliminary humidification of food surface is observed.

This, according to Eq. 4, causes a steep food temperature raise owing to both the heat transfer by convection from air to food and the latent heat of condensation due to vapor condensing on food surface. The above phenomenon continues until food temperature is high enough to make (C_i - C_{gb}) > 0; at this point, according to eq.5, drying begins and heat is transferred by convection from air to food whereas latent heat of vaporization is transported from food to water, thus allowing its evaporation (eq. 4).

As it will be specified in the following, heat and mass transfer coefficients have been estimated on the basis of the well-known semi-empirical correlations expressing the dependence of Nusselt number upon Reynolds and Prandtl numbers and of Sherwood number on Reynolds and Schmidt numbers, respectively (Perry & Green, 1984). It has been assumed that the Chilton-Colburn analogy holds.

The above system of non-linear partial differential equations has been solved by the finite elements method developed by the commercial package Comsol Multiphysics 3.3, Comsol Sweden. The domain corresponding to food sample has been discretized into 3550 triangular finite elements (Fig. 4) with a thicker mesh close to the boundaries actually exposed to the drying air.

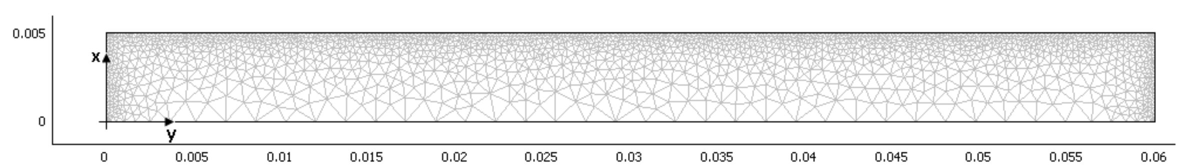


Fig. 4: Discretization of food sample into triangular finite elements.

The total number of degrees of freedom was equal to about 15000. On a 1.4 GHz Pentium IV computer running under Linux, a typical drying process duration of 5 hours was simulated, on average, in about 5 minutes, using the time-dependent nonlinear direct solver already implemented in the Comsol package. It should be remarked that the proposed model does not need any parameters adjustment, but only the specification of a set of input variables that can be varied within a specific range of physical significance to simulate food drying behaviour at different operating conditions.

Complete Model

The development of the previous model is based on the utilization of a proper set of heat and mass transfer coefficients that are to be provided by literature correlation or by experimental test-runs. In both the above cases, there are certainly some problems. Literature data about heat and mass transfer coefficients do actually cover a limited number of possible situations as far as both the geometry and the operating conditions are concerned; experimental tests are time consuming and not general since they are performed with reference to a specific vegetable, dried using a rather restricted range of operating conditions. A better way to analyze food drying process might be the solution of the transport equations in both the domains (food and air) and the definition of a completely new set of boundary conditions at the food-air interfaces. In this way, heat and mass transfer coefficients can be, actually, considered as the final result of a complex calculation and not as the essential tool necessary to describe the transport phenomena involved in drying process.

Therefore, the simultaneous transport of momentum, heat and mass and momentum in the drying air is to be combined, to the previous PDEs 2-3 describing, on the basis of the same simplification hypotheses already formulated, heat and moisture transport within the food sample.

In the air, heat transfer occurs by convection and conduction whereas water transfer takes place by convection and diffusion. The convective contributions to heat and mass transfer have to be considered due to air circulation whose velocity field has to be determined solving the non-isothermal momentum transport equations in turbulent conditions, coupled to the continuity equation. Evaporation/condensation effects at the air-food interfaces are, also in this case, taken into account by a set of boundary conditions that, however, are completely different with respect to the previous simplified approach. As it will be described in the following, these boundary conditions express the continuity of temperatures, and of heat and mass fluxes at the food/air interfaces.

Moreover, the same relationship between water concentration in air and food moisture content does exist and the thermodynamic equilibrium expressed by Eq. 10 holds.

The non-isothermal turbulent flow of air within the drying chamber has been modelled by means of the well-known k-ε model that is based on two additional semi-empirical transport equations for the variables k and ε, i.e. the turbulent kinetic energy and the turbulent energy dissipation rate, respectively.

The unsteady-state momentum balance coupled to the continuity equation written for the drying air lead to (Bird, Stewart, Lightfoot, 1960; Verboven et al., 1997;Verboven et al. 2001):

$$\partial \rho_a / \partial t + \underline{\nabla} \bullet (\rho_a \underline{u}) = 0 \qquad 11$$

$$\rho_a \, \partial \underline{u} / \partial t + \rho_a \underline{u} \bullet \underline{\nabla} \underline{u} = \underline{\nabla} \bullet \left[-p\underline{I} + (\eta + \eta_t)\left(\underline{\nabla} \underline{u} + (\underline{\nabla} \underline{u})^T - (2/3)(\underline{\nabla} \bullet \underline{u})\underline{I}\right)\right] \qquad 12$$

$$\rho_a \, \partial k / \partial t + \rho_a \underline{u} \bullet \underline{\nabla} k = \underline{\nabla} \bullet \left[(\eta + \eta_t / \sigma_k)(\underline{\nabla} k)\right] + \eta_t P(\underline{u}) - (2\rho_a k/3)(\underline{\nabla} \bullet \underline{u}) - \rho_a \varepsilon \qquad 13$$

$$\rho_a \, \partial \varepsilon / \partial t + \rho_a \underline{u} \bullet \underline{\nabla} \varepsilon = \underline{\nabla} \bullet \left[(\eta + \eta_t / \sigma_\varepsilon)(\underline{\nabla} \varepsilon)\right] + (c_{1\varepsilon} \varepsilon / k)[\eta_t P(\underline{u}) - (2\rho_a k/3)(\underline{\nabla} \bullet \underline{u})] - c_{2\varepsilon} \rho_a \varepsilon^2 / k \qquad 14$$

where ρ_a is the air density, η is its viscosity, both expressed in terms of the local values of temperature and water content, p is the pressure within the drying chamber, $\underline{u}$ is the velocity vector. c_μ ,σ_k, σ_ε, $c_{1\varepsilon}$ and $c_{2\varepsilon}$ are constants whose value depends on the particular k-ε turbulence model that is used; in the case of the present analysis, the standard k-ε model has been adopted (Verboven, 2000).

The term $P(\underline{u})$, finally, contains the contribution of the shear stresses:

$$P(\underline{u}) = \underline{\nabla} \underline{u} : \left(\underline{\nabla} \underline{u} + (\underline{\nabla} \underline{u})^T\right) - (2/3)(\underline{\nabla} \bullet \underline{u})^2 \qquad 15$$

The following definition for η_t, i.e. the turbulent viscosity, in Eqs. 12-14, holds:

$$\eta_t = \rho_a \cdot c_\mu k^2 / \varepsilon \qquad 16$$

The energy balance in the drying air, accounting for both convective and conductive contributions, leads to (Bird, Stewart, Lightfoot, 1960):

$$\rho_a C_{pa} \, {\partial T_2}/{\partial t} - \underline{\nabla} \bullet \left(k_a \underline{\nabla T_2} \right) + \rho_a C_{pa} \underline{u} \bullet \underline{\nabla T_2} = 0 \qquad 17$$

where T_2 is the temperature of air, C_{pa} is its specific heat and k_a its thermal conductivity.

The mass balance in the drying air, referred to the water and accounting for both convective and diffusive contributions, leads to (Bird, Stewart, Lightfoot, 1960):

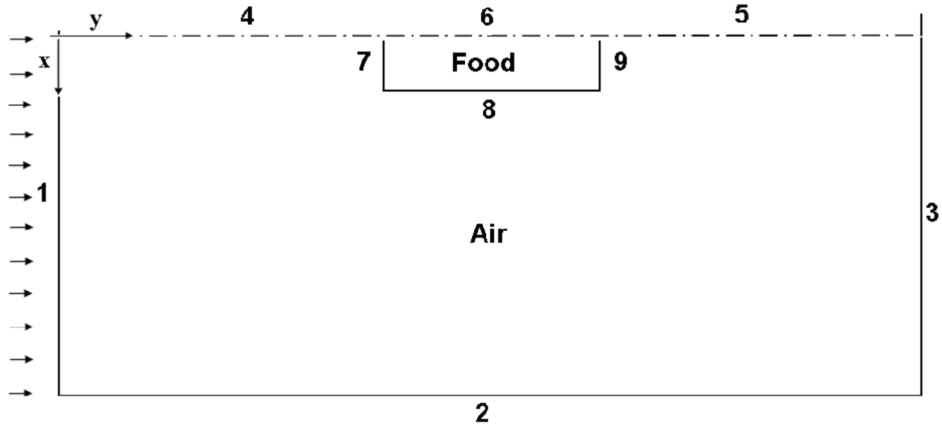
$$\partial C_2 / \partial t + \underline{\nabla} \bullet \left(- D_a \underline{\nabla C_2} \right) + \underline{u} \bullet \underline{\nabla C_2} = 0 \qquad 18$$

where C_2 is the water concentration in air and D_a is the diffusion coefficient of water in air.

Since physical and transport properties of both air and food are expressed in terms of the local values of temperature and moisture content, Eqs. 7-18, together with Eqs. 2-3 describing heat and moisture transport in the food, represent a system of unsteady, non-linear partial differential equations that can be solved only by means of a numerical method.

A set of proper initial conditions is necessary to perform the numerical simulations. The initial conditions are referred to both air and the food. In particular, water concentration, C_{20}, temperature, T_{20}, the pressure and the velocity field in the drying chamber, $p_0 = p_{atm}$ and $\underline{u} = \underline{0}$, have been fixed for air before drying process takes place, whereas the initial values of moisture content, C_0, and temperature,T_0 have been used for food. The above values will be, however, specified on presenting the simulation results in the following “results and discussion” section.

As far as the boundary conditions are concerned, they are reported, for sake of brevity, in the fig. 5 where each different boundary has been identified by a number ranging from 1 to 9.



Boundary	Type of condition
1	$u_x = 0$, $u_y = u_0$ $T_2 =$ Tair $C_2 =$ Cair
2	$\mathbf{u}$ =Logarithmic wall function $T_2 =$ Tair $C_2 =$ Cair
3	$p = p_0$ (Atmospheric pressure) $\mathbf{n}\cdot(-k_a\nabla T_2)=0$ (Convection prevailing over conduction) $\mathbf{n}\cdot(-D_a\nabla C_2)=0$ (Convection prevailing over diffusion)
4, 5, 6	Axyal symmetry (except for momentum on boundary 6)
7, 8, 9	$\mathbf{u}$ =Logarithmic wall function $T = T_2$ (Temperature continuity) $\mathbf{n}\cdot(-k_a\nabla T_2+\rho_a C_{pa} T_2\mathbf{u})=\mathbf{n}\cdot(-k_{eff}\nabla T)-\lambda\mathbf{n}\cdot(-D_{eff}\nabla C)$ (Heat flux continuity) Equilibrium : $\gamma_w\, x_w\, P_w^{sat} = y_w\, p$ $\mathbf{n}\cdot(-D_{eff}\nabla C)=\mathbf{n}\cdot(-D_a\nabla C_2+C_2\mathbf{u})$ (Mass flux continuity)

Fig. 5: Boundary conditions used to formulate *mod2*.

Some of the above boundary conditions are straightforward; some others need a more detailed explanation. At the food-air interfaces (boundaries 7 - 9), where no accumulation occurs, the continuity of heat fluxes actually accounts for the fact that the heat transported by convection and conduction from air to food is partially used to raise sample temperature by conduction and partially to allow water evaporation. The latter effect is described by considering water latent heat of vaporization (λ) that is expressed in terms of the interfacial food temperature and of its moisture content (Ruiz-López, Córdova, Rodríguez-Jimenes, García-Alvarado, 2004):

$$\underline{n}\bullet\left(-k_a\underline{\nabla}T_2+\rho_a C_{p_a} T_2\underline{u}\right)=\underline{n}\bullet\left(-k_{eff}\underline{\nabla}T\right)-\lambda\underline{n}\bullet\left(-D_{eff}\underline{\nabla}C\right) \qquad 24$$

On the same boundaries, a balance between the diffusive flux of liquid water coming from the product core and the flux of vapor that leaves the food surface and is transported, by both diffusion and convection, into the drying air applies:

$$\underline{n} \bullet \left(- D_a \underline{\nabla} C_2 + C_2 \underline{u} \right) = \underline{n} \bullet \left(- D_{eff} \underline{\nabla} C \right) \qquad 25$$

Also in this case, an equilibrium relationship between the water concentration in the air and the water concentration, on the food surfaces actually exposed to the drying air (Eq. 10), applies.

On boundary 2, it has been assumed that air temperature and its concentration are the same as those measured at the drier inlet section. These two conditions are valid under the hypothesis that the penetration of both temperature and concentration profiles is, actually, confined in two very thin regions that develop close to the food-air interface. In the following, the above assumptions will be validated (or falsified) on discussing the simulations results.

At the oven outlet section, conduction and diffusion phenomena can be neglected with respect to convection (Danckwerts conditions). As far as the boundary conditions referred to the momentum balance on the solid surfaces are concerned, a two-velocity scale wall function was used, as reported by Lacasse, Turgeon and Pelletier, 2004.

The above system of non-linear Partial Differential Equations, together with the above-described sets of initial and boundary conditions, has been solved by Finite Elements Method. Both the domains were discretized into a total number of 5846 triangular finite elements, leading to about 60000 degrees of freedom. Fig. 6 shows a detail of the considered mesh. On a Pentium IV computer running under Linux, a typical drying process duration of 5 hours was simulated, on average, in about 1 hour also in this case by the commercial package Comsol Multiphysics 3.3, using a time-dependent nonlinear direct solver already implemented in the same package.

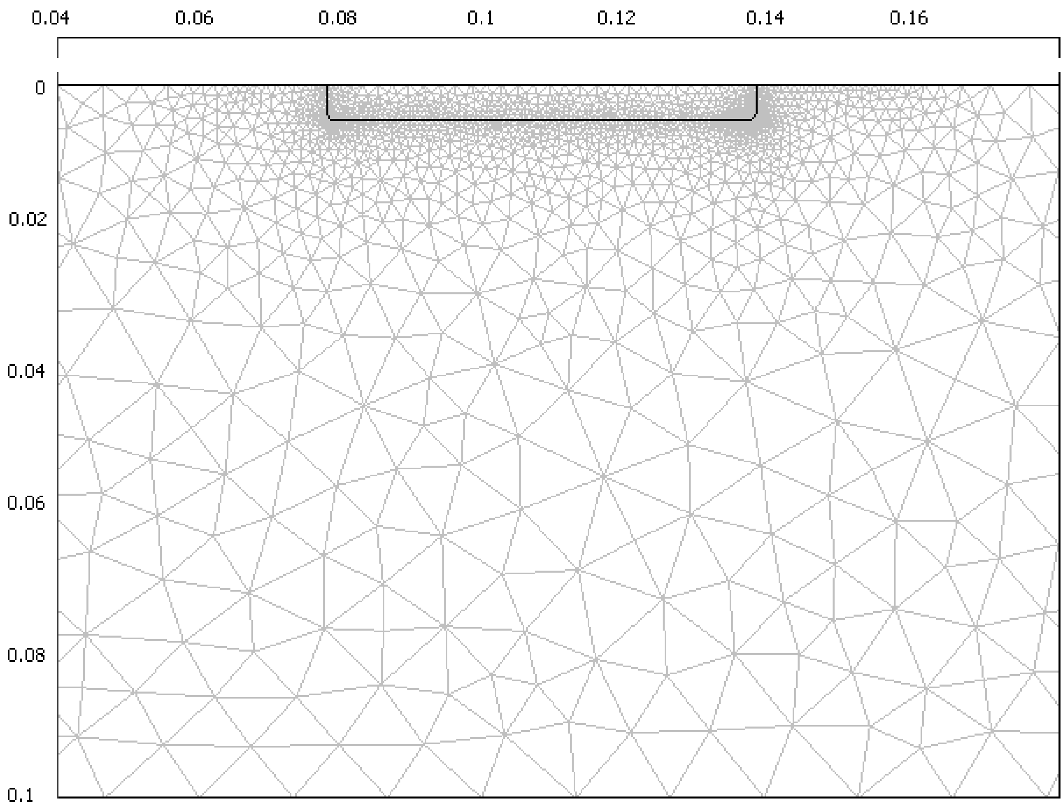


Fig. 6: Discretization of food and air domains into triangular finite elements (detail of the mesh).

Results and Discussion

The analysis of transport phenomena henceforth presented will be based on the examination either of some characteristic average variables or of temperature and concentration profiles. Simulations at different drying conditions have been performed to show how air characteristics affect food drying process and, hence, the drying time. In order to properly explain the physical meaning of the obtained results, it will be necessary to bear in mind the general considerations on drying process already described at the beginning of the present chapter. In particular, fundamental parameters like the wet bulb temperature, the difference existing between free and bound water and their relationship with the activity coefficient, the drying rate, the driving force, etc. will be often mentioned so to precisely interpret the behaviour of drying process. Starting from a reference case, different simulations have been performed varying every single air property (i.e. the inlet temperature, its relative humidity and the feed velocity) in a specific range, typical of food drying process. To define the reference case, it has been assumed that the food sample, i.e. a 6-cm-long carrot slice with a half-thickness of 0.5 cm, had an initial temperature equal to 283 K and an average initial moisture content (on a wet basis), $\overline{U_0}$, of 0.88 Kg/Kg, corresponding to an average initial moisture content (on a dry basis), $\overline{X_0}$, of 7.3 Kg/Kg. The drying air entering the oven had a uniform feed velocity of 5 m/s; in the tested range of operating conditions, turbulent conditions are attained with Reynolds numbers, within the drier, well higher than 10^4. Finally, it was supposed that at the drier inlet section air had a dry bulb temperature, T_{gb}, of 343 K and a relative humidity, U_r, of 30%.

In the following, it will be sketched a comparison among the predictions that can be obtained simulating the drying process behaviour by means of the simplified model, based on the utilization of literature transport coefficient and henceforth identified as *mod1*, and of the more complete approach that accounts also for the transport phenomena occurring in the drying air and that will be henceforth indicated as *mod2*. As far as *mod1* is concerned, it should be again observed that the utilization of a particular set of semi-empirical correlations for heat and mass transfer coefficients evaluation is critical to ascertain the model capability to predict the time evolutions of temperature and moisture content within the food. The model predictions, in fact, may significantly differ if two distinct choices of semi-empirical correlations, equally acceptable for the same drying problem, are performed. This aspect has, generally, deep repercussions on food processing. Wrong predictions of real system behaviour might lead to stop drying process despite, for instance, of local moisture content that is still so high to promote microbial spoilage or might suggest to adopt operating conditions that, actually, produce local damages in the food structure, thus determining a poor product quality. In the present analysis, *mod1* makes use of three different correlations, one for each of the surfaces exposed to drying air, to evaluate the local values of heat transfer coefficients from the Nusselt number, *Nu*, expressed in terms of characteristic Reynolds, Re, Prandtl, Pr, and Grashof, *Gr*, numbers (Perry & Green, 1984, Welty, Wicks, Wilson, Rorrer, 2001):

Boundary	Semi-empirical relationship	Definition of dimensionless numbers
Impact surface (identified as boundary 7)	$Nu_x = 0.25 \mathrm{Re}_x^{0.588} \mathrm{Pr}^{1/3}$	$Nu_x = h_{x\|7} \cdot x / k_a$; $\mathrm{Re}_x = \rho_a \cdot u_0 \cdot x / \eta$
Surface parallel to air flow (identified as boundary 8)	$Nu_y = 0.648 \mathrm{Re}_y^{0.5} \mathrm{Pr}^{1/3}$	$Nu_y = h_{y\|8} \cdot y / k_a$; $\mathrm{Re}_y = \rho_a \cdot u_0 \cdot y / \eta$
Rear surface (identified as boundary 9)	$Nu_x = 0.59 (\mathrm{Pr} \cdot Gr_x)^{1/4}$	$Nu_x = h_{x\|9} \cdot x / k_a$; $Gr_x = x^3 \cdot \rho_a \cdot g \cdot \Delta\rho_a / \eta^2$

Tab. 1: Semi-empirical correlations used to estimate heat transfer coefficients on each food surface.

The Prandtl number is defined as $\mathrm{Pr} = C_{pa} \cdot \eta / k_a$. In the previous relationships, *x* and *y* define the local position on each of the food surfaces; $h_{(x,y)|i}$, characterized by a numeric subscript (*i*) corresponding to each boundary, is the local value of heat transfer coefficient; u_0 is air velocity at the inlet section of the drier; g is the acceleration gravity. All the physical and transport properties defining each dimensionless numbers have been evaluated at film conditions (the subscript *f* has been omitted).

It should be observed that, as far as the rear face of food sample is concerned, the Nusselt number has been actually expressed in terms of Grashof number since it can be supposed that air circulation in proximity to boundary 9 is so limited that free convection prevails over forced convection.

Once the heat transfer coefficients have been estimated, the Chilton-Colburn analogy is used to calculate the mass transfer coefficients referred to each corresponding boundary (Perry & Green, 1984).

The following Fig. 7 shows a comparison of food moisture content profiles (on a dry basis) that are obtained by both *mod1* and *mod2* after 15 minutes, 45 minutes and 3 hours from drying beginning. It has been assumed that drying air had a temperature of 343 K, a relative humidity of 10% and an inlet velocity of 5m/s. In these conditions the wet bulb temperature is equal to 307.6 K (Perry & Green, 1984). The food surface upon which air impinges is heated more rapidly than the other surfaces and its temperature suddenly rises above the wet bulb temperature, thus initiating the water removal. The impact surface is interested by two different simultaneous phenomena: 1) the heat transferred from air provides both the latent heat of vaporization necessary for water evaporation and the heat necessary to increase, by conduction, the temperature of food dry matter; 2) the drying rate is very high and involves the free water that, therefore, vaporizes and is immediately transported into the air. Both *mod1* and *mod2* are capable to predict the different transport rates of water that are achieved on each exposed surface. The water removal is responsible for the movement of dry/moist interface toward the food core. This displacement is more rapid from the surface where drying air impinges and is rather slow, especially in the very initial stage of the process, from the opposite side. The rear face of food, in fact, is interested by a rather wide "segregation" region that forms due to an ineffective air circulation, as it is confirmed by the velocity field reported in Fig. 8, and that determines a significant delay on reaching the wet bulb temperature necessary to initiate free water removal. This result, found over the whole range of tested velocities, confirms that heat transfer near the back surface of food sample occurs mainly by free convection. It should be also observed that if a unique semi-empirical correlation had been used to estimate the Nusselt number, an identical behaviour of all the exposed surfaces would have been observed. This would have determined an equal displacement of the so-called dried front from the outer regions to the food core; a result that, obviously, conflicts with the physical experience about the phenomenon under consideration.

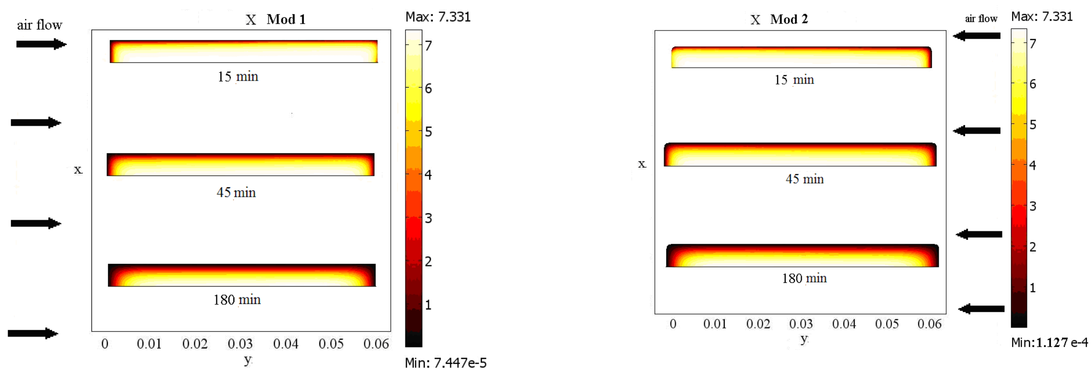


Fig. 7: Profiles of food moisture content (on a dry basis) obtained by both the models.

It should be preliminary observed that several authors, as reported by Saravacos and Maroulis, (2001), made use of the same semi-empirical correlation to estimate, for all the exposed surfaces, the Nusselt number as a function of a unique average Reynolds number. This choice might be responsible for very inaccurate predictions of the real transfer rates at the food/air interfaces since each of the surfaces or, actually, each point of a surface is characterized by different fluid-dynamic conditions that have a strong influence on the external resistances to both heat and mass transfer. The time evolutions of food surfaces temperature, averaged over each of the surfaces and evaluated as a function of drying time ($\overline{T_s}(t)$) by *mod2*, are reported in Fig. 9.

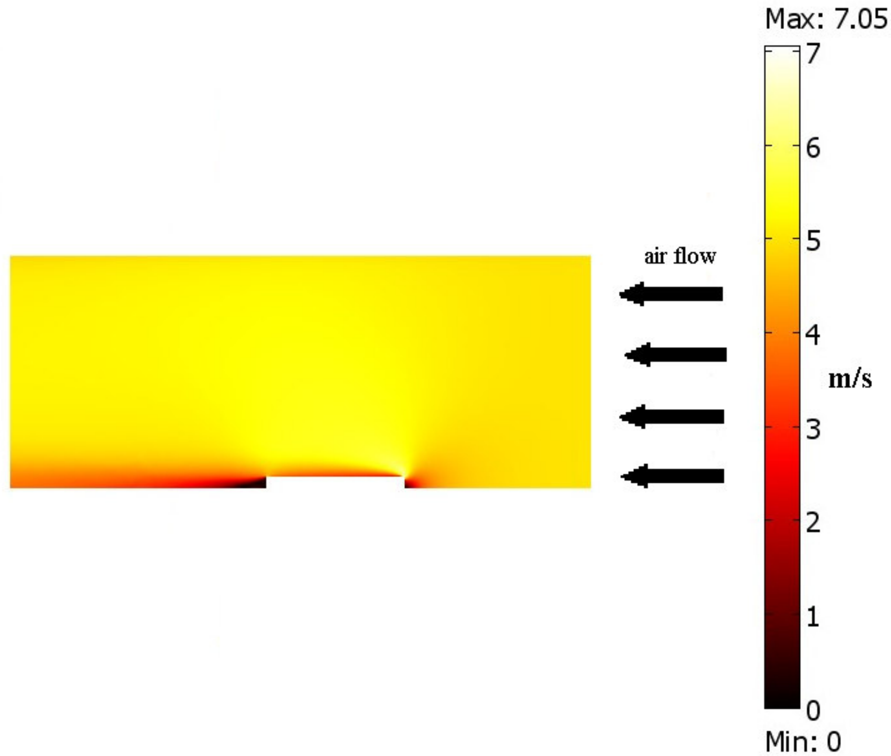


Fig. 8: Air velocity field in the drying chamber.

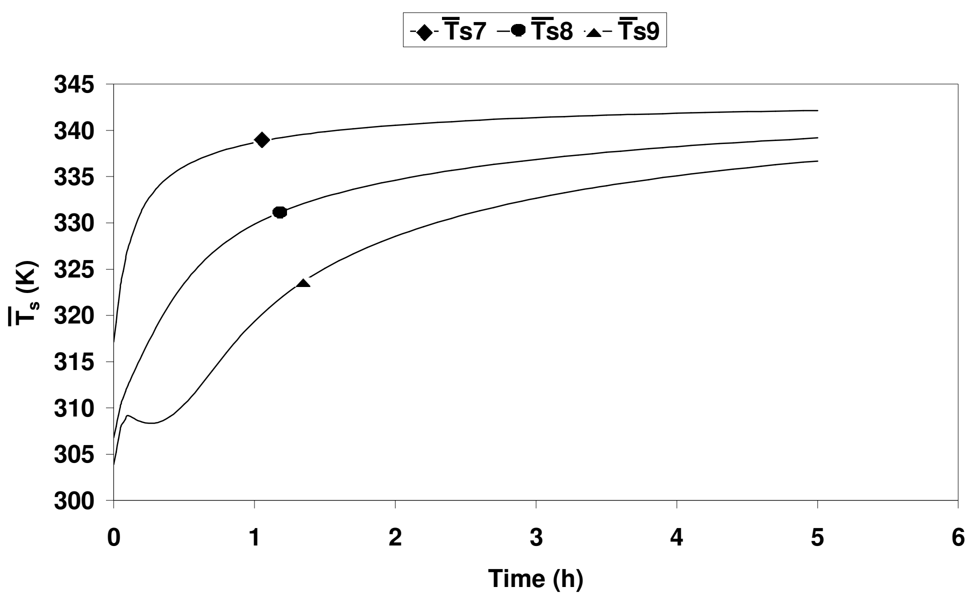


Fig. 9: Time evolutions of food average temperature on each of the exposed surfaces (air temperature 343 K, relative humidity 10%, inlet air velocity 5 m/s).

The food surface upon which air impinges (boundary 7) is heated more rapidly than the other surfaces and its temperature rises, quickly, above the wet bulb temperature. This behavior is definitely to be ascribed to the flow conditions. The impact surface is interested by two different simultaneous phenomena: 1) the heat transferred from air provides both the latent heat of vaporization necessary for water evaporation and the heat needed to increase, by conduction, the temperature of food dry matter; 2) the drying rate is very high and involves the free water that, therefore, vaporizes and is immediately transported into the air. A completely different behavior can be observed on the opposite site, i.e. on boundary 9, where, at the very beginning of drying process, the heat transported from air increases food temperature up to the wet bulb temperature; later on, the average temperature profile has a nearly constant value. During this stage, whose duration depends on the operating conditions, the heat transferred from air provides only the latent heat of vaporization to free water. When also bound water begins to be removed, the heat transported by convection and conduction from air to food is partially used to raise sample temperature by conduction and partially to allow water evaporation. Average temperature now increases up to a plateau value, with a slope depending on the operating conditions. The behaviour of the food surface parallel to the main direction of air flow can be considered as "intermediate" between the other two boundaries. Temperature profiles are characterized by a nearly constant value, but its duration is shorter or does not actually exist when, like in the present case, velocity field is so high to determine a significant reduction of transport resistance external to food sample.

In the following the effects of air characteristics on the drying rates predicted by both the models will be shown. As a general consideration, it should be remarked that, among air characteristics, velocity has a strong influence on the transport resistances outside the food sample and, therefore, on the transfer rates of both heat and water at food-air interface. Temperature and humidity of air entering the dryer mainly affect the available driving forces promoting transport phenomena between food and air.

Actually, an increase of air velocity corresponds to a decrease of external resistances to heat and mass transfer. This effect is however significant only when internal resistances do not control the overall rate of the process. Therefore, the improvement related to higher values of feed velocity can be observed only when free water evaporates and leaves the food. As soon as bound water is also removed, internal resistance to mass transfer predominates and air velocity has a weak influence on process rate; during this stage the value of effective diffusivity of water in food and its dependence on both temperature and food moisture content actually control the drying rate. Figs. 10 and 11 show the predictions obtained, by both the models, when drying process is performed by an air temperature of 343 K, a relative humidity of 30% and two different feed velocities equal to 5 and 10 m/s, respectively. In the above conditions the wet bulb temperature is 321 K (Perry & Green, 1984).

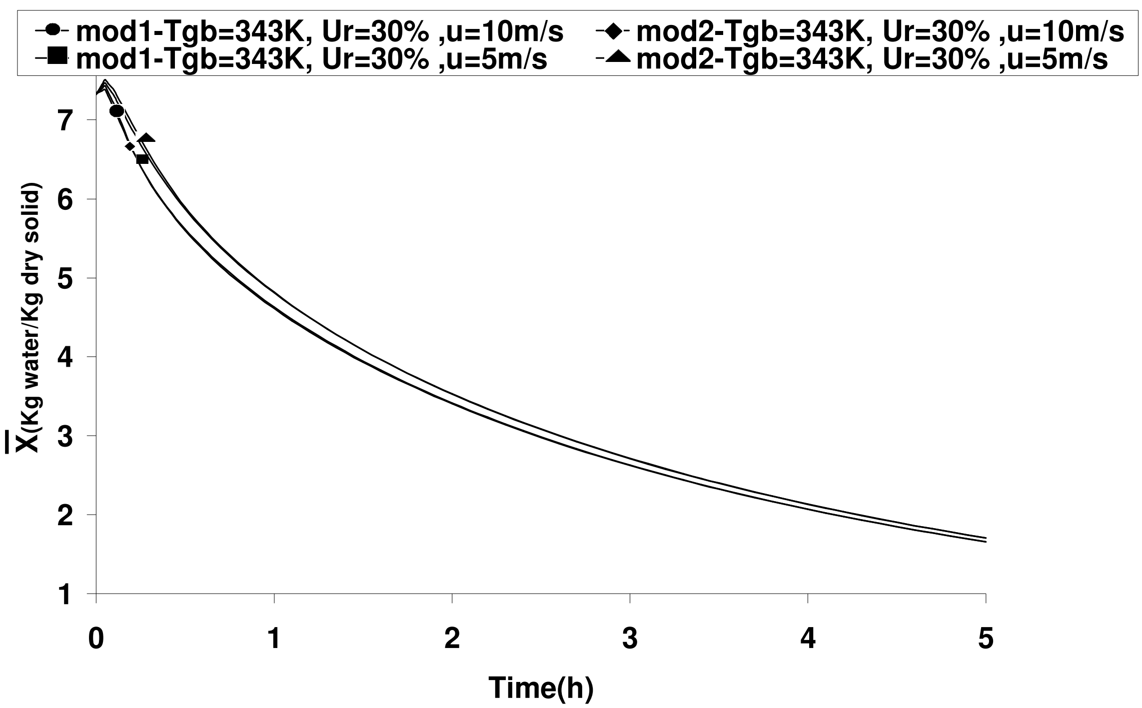


Fig. 10: Average food moisture content as predicted by both the models (effect of air inlet velocity).

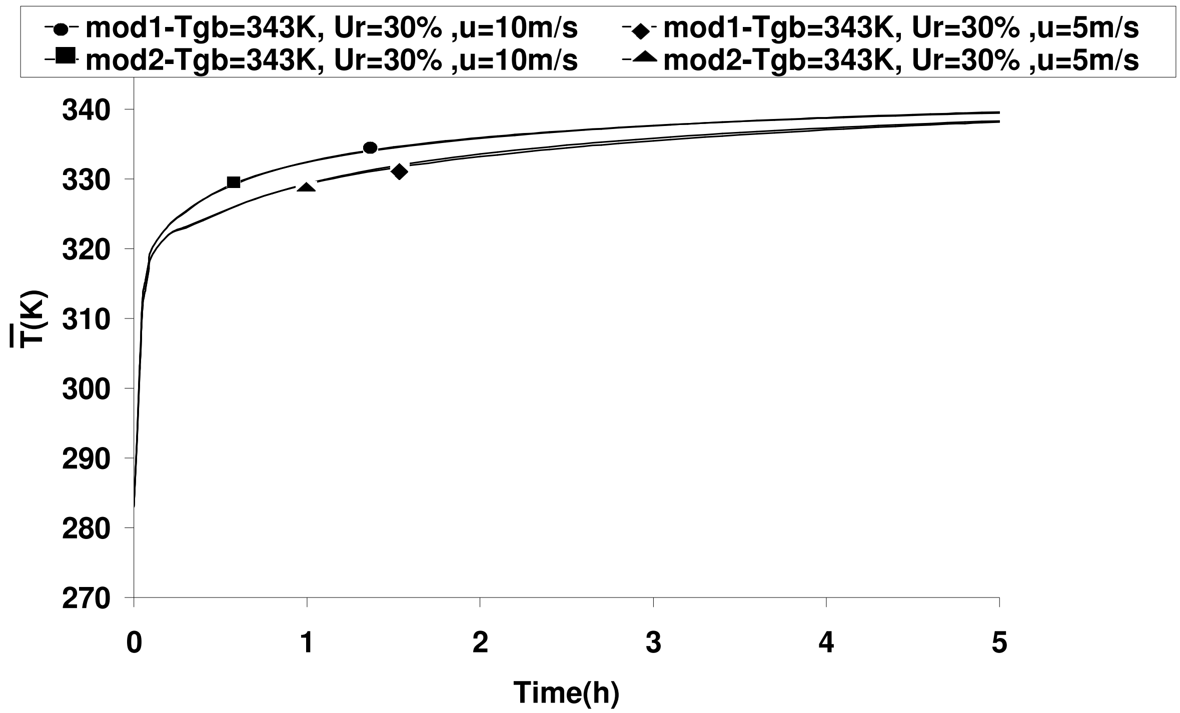


Fig. 11: Average food temperature as predicted by both the models (effect of air inlet velocity).

The results are presented in terms of food moisture content (on a dry basis), averaged over the whole food sample and evaluated as a function of drying time, $\overline{X}(t)$, and of food temperature, averaged over the whole food sample and evaluated as a function of drying time, $\overline{T}(t)$. In the tested cases both the models give very close predictions and a slight difference can be noted only during the very initial stage of drying process. An improvement of drying rate is observed as air velocity is increased, even though the curves tend to approach each other as soon as the internal resistances prevail over the external ones. The initial steep temperature rise shown in Fig. 11 is, actually, the result of two different contributions. In fact, heat transfer depends on both the convection (due to the temperature difference between air and the solid surface), and, as long as food surface temperature is lower than the wet bulb temperature, the latent heat (due to the condensation of the vapor on food surface). Vapor condensation is also responsible for the corresponding initial slight increase of total food moisture shown in Fig. 10. Once wet bulb temperature is attained, water concentration at food-air interface, C_i, exceeds the value of bulk concentration of water in air that, depending on air temperature and its relative humidity only, can be assumed as constant throughout all the process; water evaporation, therefore, prevails over vapor condensation and heating rate decreases. Hence, while food temperature slowly increases towards air temperature value, food moisture content diminishes at a decreasing drying rate. Fig. 10 confirms that initially, when the process is controlled by external heat transfer and the water leaving the sample is not bound to food structure *(free water)* drying rate attains its maximum value. Afterwards, when the process rate is controlled by internal mass transfer and the water leaving the solid is bound to the food structure, a progressively decrease of drying curve slope is observed. A comparison between the results obtained by *mod1* and *mod2* shows, in the tested cases, that both the models predict a similar behavior. This could suggest that the proposed simplified approach is capable of giving as accurate predictions as the more complete model, as long as a proper set of transport coefficients is used.

Figs.12 and 13 show the predictions obtained by both the models when drying process is performed by an air temperature of 343 K that is fed to the drier at a velocity of 5 m/s; two different values of relative humidity, equal to 30% and 10%, have been chosen to test the models. In the above conditions the wet bulb temperature is equal to 321 K and to 307.6 K, respectively (Perry & Green, 1984). Also in this case, the results are presented in terms of $\overline{X}(t)$ and of $\overline{T}(t)$. When air relative humidity is changed from 10% to 30% a simultaneous increase of both the wet bulb temperature and of bulk water concentration, by a factor of about 3, is observed. An increase of relative humidity does, therefore, determine a lower drying rate and is also responsible for a longer time to initiate water removal. This behaviour can be explained observing that, at the very beginning of the process, i.e. when free water is actually subtracted from the food, an increase of air relative humidity significantly lowers the drying rate owing to a reduction of the driving force promoting water transport.

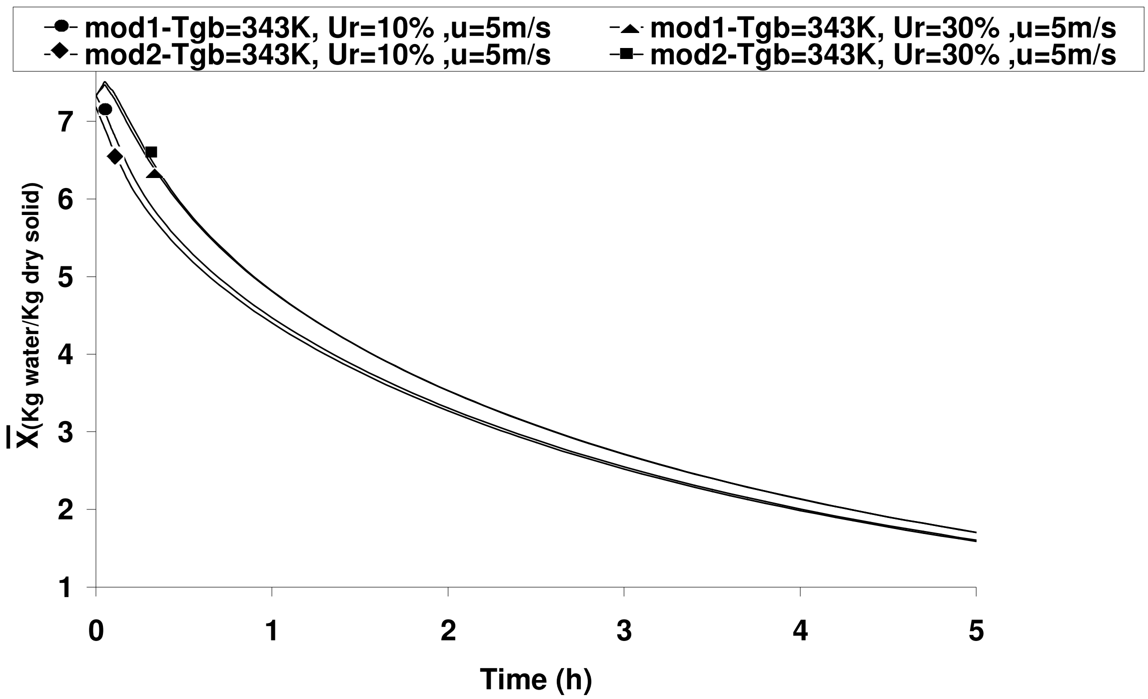


Fig. 12: Average food moisture content as predicted by both the models (effect of air relative humidity).

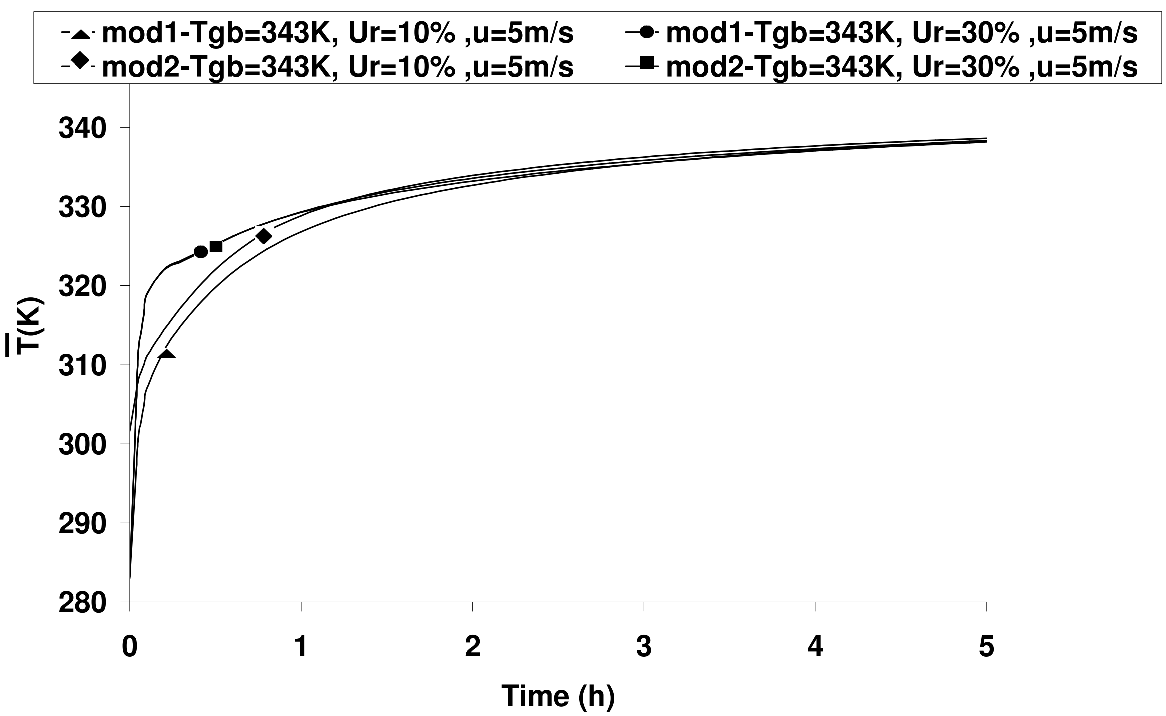


Fig. 13: Average food temperature as predicted by both the models (effect of air air relative humidity).

Moreover, it can be noted that vapor condensation does not occur appreciably when air relative humidity is equal to 10% (Fig. 12). When bound water is progressively involved in drying, internal resistance to mass transfer represents the actual limiting step and the concentration difference outside food sample affects the process rate only to some extent. A comparison between model predictions shows, in this case, that the two curves corresponding to the more drastic operating conditions differ, especially during the first two hours of drying process, for both $\overline{X}(t)$ and $\overline{T}(t)$. When food drying is performed under milder conditions, both *mod1* and *mod2* predict a similar behavior.

Figs.14 and 15 show the simulation results obtained by the two models when drying process is performed by a drying air having a relative humidity of 30 % that is fed to the drier at a velocity of 5 m/s; two different values of dry bulb temperature, equal to 308 K and 343 K, have been used to test the models.

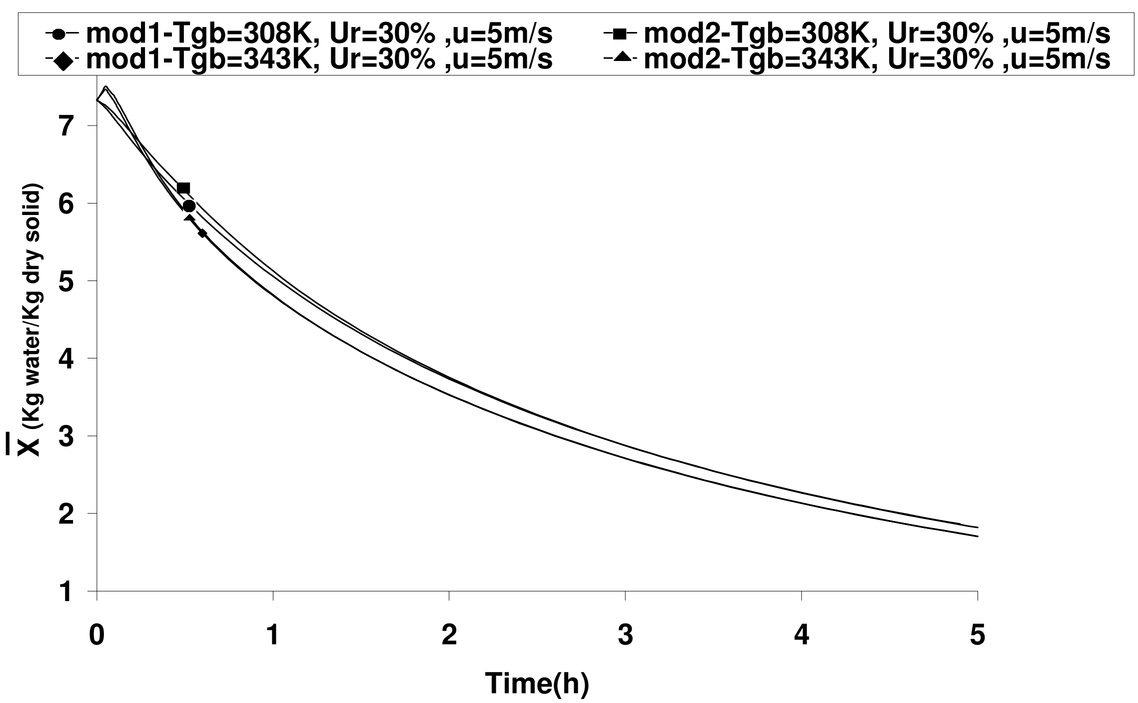


Fig. 14: Average food moisture content as predicted by both the models (effect of air temperature).

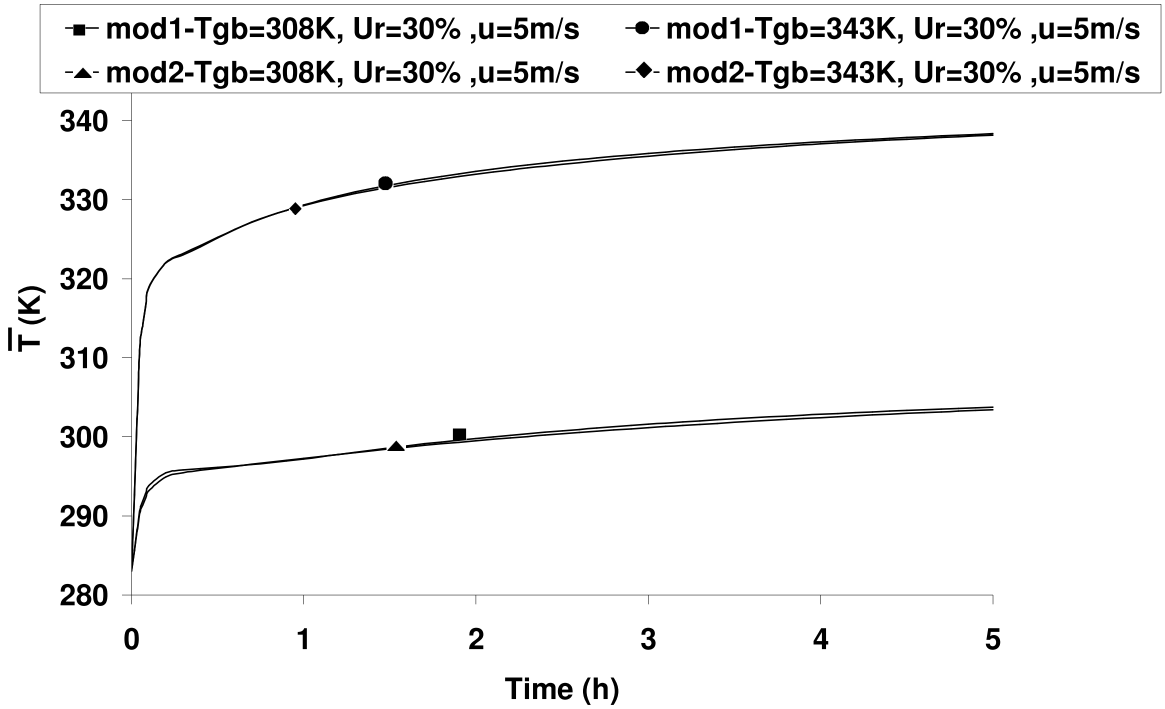


Fig. 15: Average food temperature as predicted by both the models (effect of air temperature).

An increase of dry bulb temperature from 308 K to 343 K determines an initial delay of drying process beginning, due to the increase of wet bulb temperature from 295 K to 321 K, (Perry & Green, 1984). Also in this case, vapor condensation does not occur appreciably when dry bulb temperature is equal to 308 K (Fig. 14). Moreover, under the same conditions of relative humidity, an increase of air temperature causes both a significant reduction of the driving force to mass transfer from food to air and an improvement of heat transfer from air to food. The observed rate of drying process can be, therefore, considered as a trade-off between different physical phenomena that determine a dependence on air inlet temperature mainly during the initial step of the process, where a higher operating temperature is actually responsible for a faster drying. The effect of temperature on drying performance is, instead, less important in the last part of the process, when internal resistance to mass transport prevails and represents, actually, the limiting step. During this stage, it can be observed from Fig. 14 that drying rate is much slower and the slopes of each of the curves are almost the same. A comparison among model predictions shows, also in this case, that both the models predict a similar behavior, except during the initial stage of drying process performed at a temperature of 308 K, where *mod1* tends to slightly overestimate the process rate with respect to *mod2*.

It is now interesting to analyze the following Figs. 16 and 17 that show, respectively, the distributions of water concentration and temperature profiles in the air after 15 minutes, obtained in the so-called reference case.

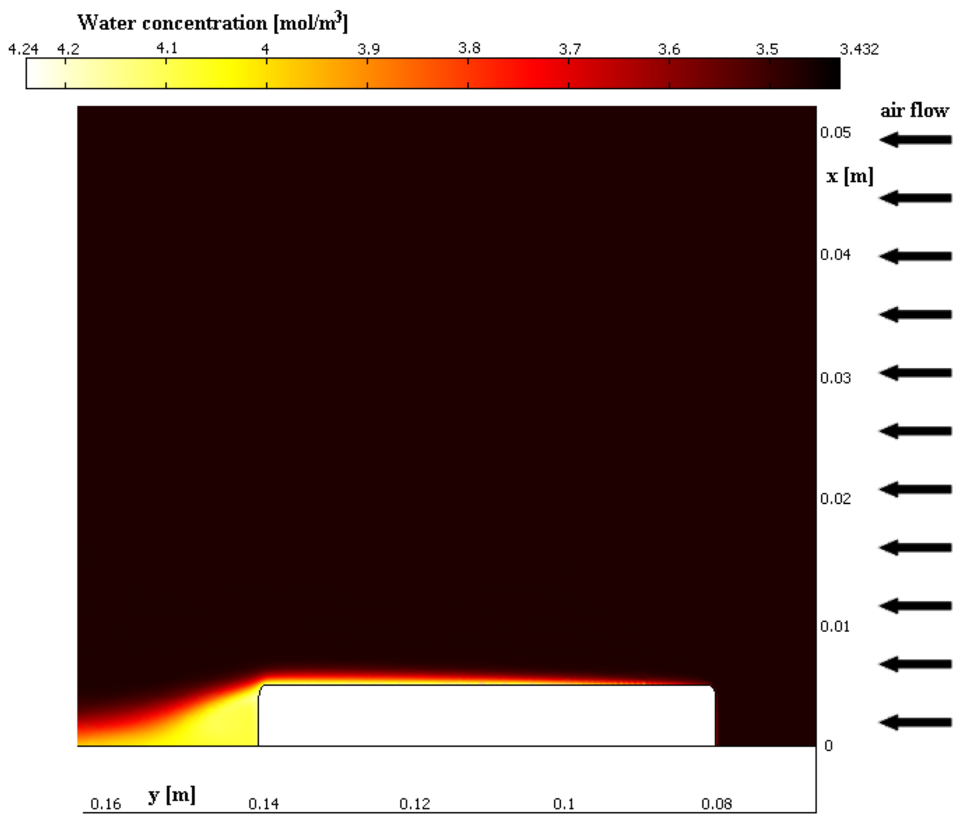


Fig. 16: Water concentration profile in the air outside the food sample (t = 15 min).

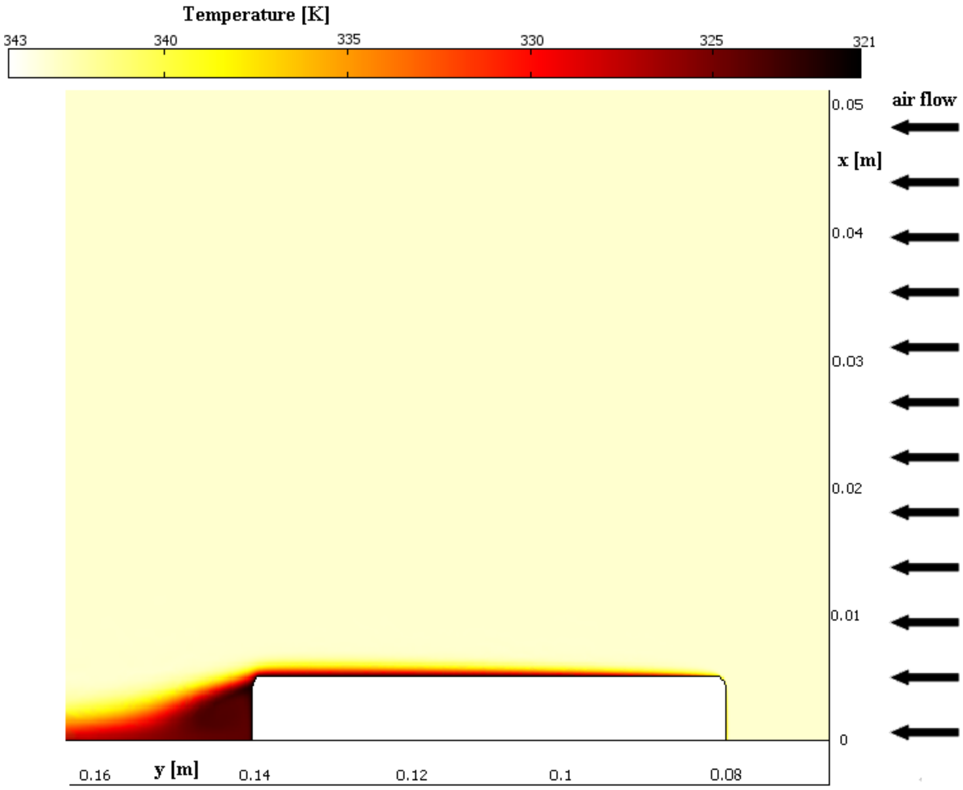


Fig. 17: Temperature profile in the air outside the food sample (t = 15 min).

It can be observed that both concentration and temperature gradients develop in two very thin regions close to the food surface whose thickness is very similar to that predicted, for instance, by the boundary layer theory (Schlichting, 1960). Figs. 16-17 confirm that some of the boundary conditions described in Fig.5 may be considered as a true representation of real system behavior, since both concentration and temperature gradients vanish at a very short distance from the food surface equal to about 3 mm compared to the dryer transverse dimension of 10 cm.

Before dealing with the last part of the present analysis that will regard the estimation of heat and mass transfer coefficients by means of *mod2*, it is now necessary to suggest some guidelines that might help deciding which of the two approaches is to be preferred. Actually, *mod1* is much simpler, it is capable of simulating a complete drying operation in just a few minutes and its predictions are in a very good agreement with the more complete model, i.e. *mod2*. Nevertheless, *mod1*is strongly dependent on heat and mass transfer coefficients that, generally, are known only for very simple geometries and for drying operations performed in a restricted range of process conditions. Moreover, it should be observed that *mod1* made use of different semi-empirical correlations for each of the surfaces actually exposed to the drying air. This practice is not so common in the literature, since it is usually preferred to utilize the same semi-empirical correlation to estimate, for all the exposed surfaces, the Nusselt number as a function of a unique average Reynolds number. One of these correlations was proposed, for example, by Ratti & Crapiste (1995):

$$Nu_L = 0.249\,\mathrm{Re}_L^{\ 0.64} \qquad 26$$

the subscript L indicates that the characteristic length utilized to define the Reynolds number is the whole slab length, but it could be also the diameter (in the case of both a cylindrical-shaped and a sliced vegetable). It should be observed that the utilization a unique semi-empirical correlation is often dictated by the objective difficulties related to the experiments. These have to be performed to derive a proper set of transport coefficients, in a as wide as possible range of controlled conditions. In addition, a considerable number of the papers currently available in the literature do not account for the variation of heat and mass transfer coefficient with the local position. At this point of the analysis, it is interesting to compare, in some typical situations, the predictions obtained by the same model, i.e. *mod1*, when either Eq. 26 or the semi-empirical correlations reported in Tab. 1 are adopted to simulate drying process. The Chilton-Colburn analogy has been used to calculate the mass transfer coefficient starting from the estimated values of heat transfer coefficients.

Figs. 18 and 19 show this comparison when two different values of dry bulb temperature and of air velocity are chosen to perform the simulations. Significant differences can be observed in all the tested conditions. The utilization of a unique correlation is responsible for a major overestimation of operating conditions effects on the drying rate. The general trends are not significantly affected by the values of semi-empirical correlations, but a rather lower value of food moisture content is calculated when the unique correlation is adopted, for both the considered velocities. The observed differences are actually more evident during the first two hours of the process, but persist, to a lesser extent, also till the end of simulations. This behaviour definitely does not depend on the model formulation and is to be ascribed only to an incorrect choice of semi-empirical correlation. Especially for the rear food surface, Eq. 26 estimates a heat transport coefficient and, on the basis of Chilton-Colburn analogy, also a mass transfer coefficient that is not as high as that predicted. The results shown in Figs. 18-19 are meaningful since they confirm that the choice of a proper set of transfer coefficients is, actually, critical when drying process is to be simulated.

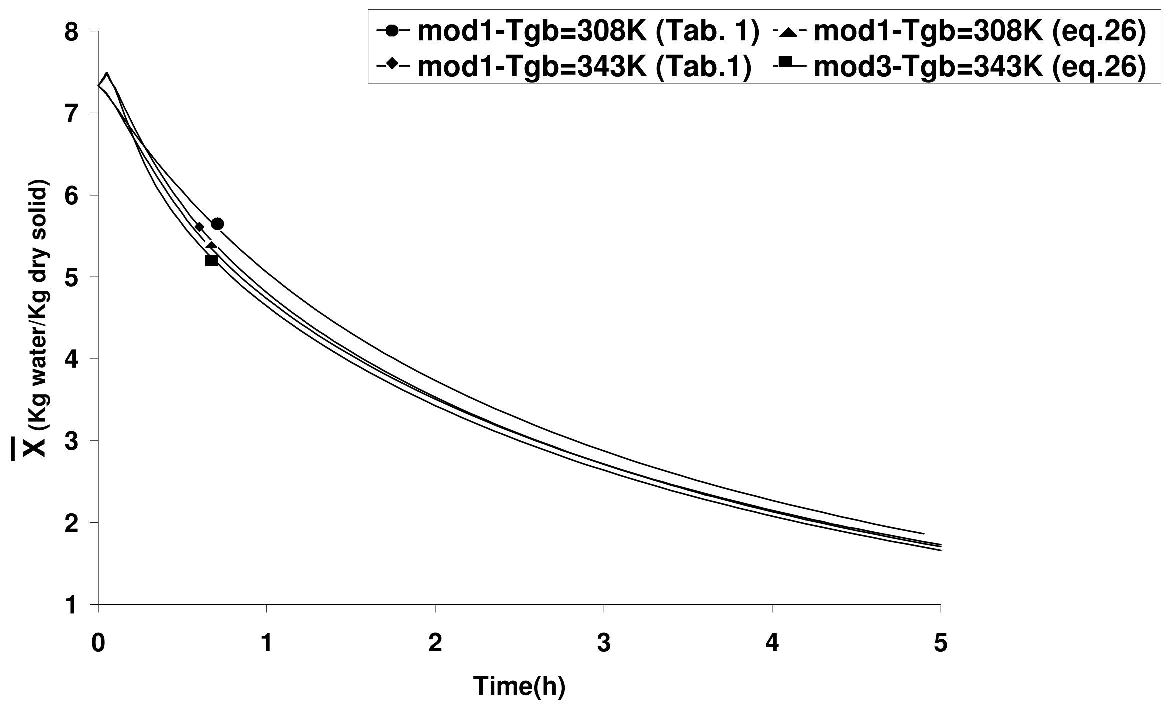


Fig. 18: Effect of semi-empirical correlations on average food moisture content in two different cases (U_r = 30%, u = 5 m/s).

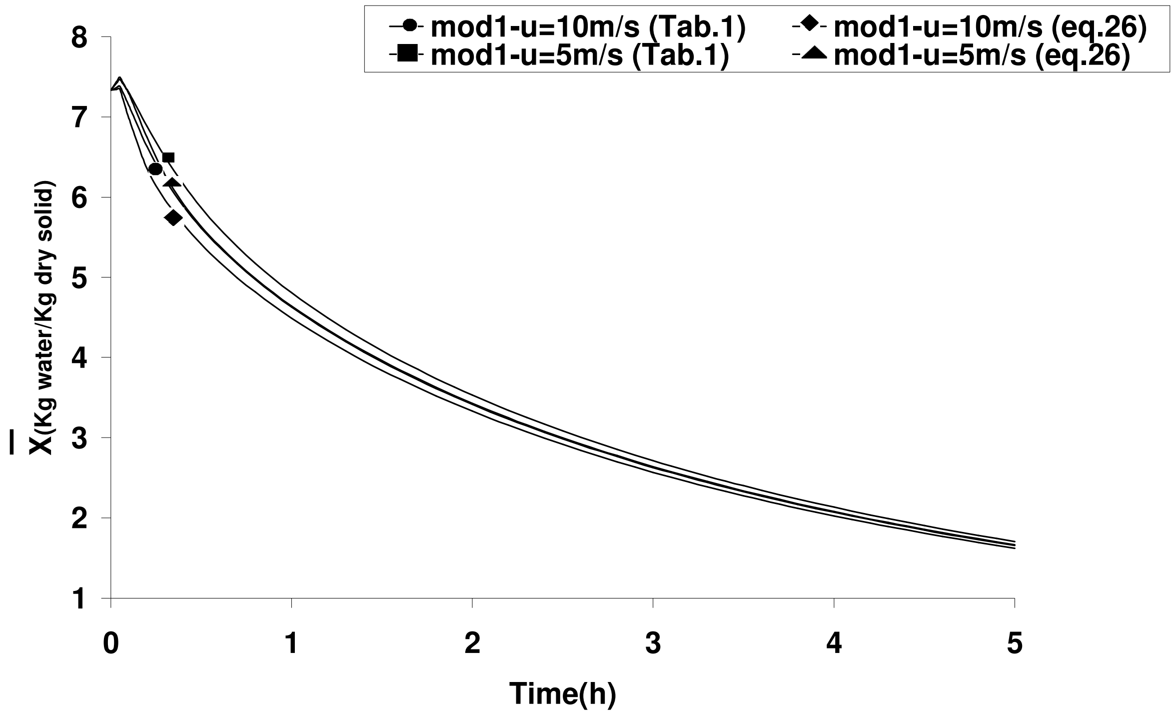


Fig. 19: Effect of semi-empirical correlations on average food temperature in two different cases (T_{gb} = 343 K, U_r = 30%).

On the basis of the above discussion, it seemed that *mod1* - provided that a proper set of semi-empirical correlation is actually available for the process under investigation - is to be preferred to *mod2* due to its simplicity, to the accuracy of its predictions (in most of the tested cases) and to the low computational effort that is needed to perform numerical simulations. *Mod2*, however, was developed to analyze drying process from a more general point of view that could be particularly useful in those conditions (irregular food shapes, operating variables changing with time, etc.) for which it might be rather difficult either to find literature correlations or to estimate them as a result of suitable and accurate experiments. *Mod2* has certainly a wider range of validity and, although more complicated and computationally more onerous, provides a more precise description of the actual system behavior than *mod1*.

One of the potentialities of *mod2* definitely concerns its "ability" to calculate the transport coefficients at food/air interfaces. This calculation is performed by means of fundamental considerations only and without resorting to any empirical correlation. For example, the local mass transfer coefficients at each surface can be calculated once both the molar diffusive flux of water and a characteristic concentration difference are obtained as a result of the above-described systems of PDEs (Eqs. 2-3 and Eqs.11-18). The following definition can be used to estimate the local values of mass transfer coefficient, $k_c(y)$, at the food surface parallel to the main flow of drying air (Bird, Stewart, Lightfoot, 1960):

$$k_c(y) = \frac{-D_a \left.\dfrac{\partial C_2(x,y)}{\partial x}\right|_{x=\text{food surface}}}{C_i(y) - C_{gb}} \tag{27}$$

Both the numerator and the denominator of Eq. 27 are obtained as a result of each numerical simulation, so they directly account for the actual values of the operating variables chosen to analyze drying process behaviour in each situation.

Similar considerations can be done to estimate the mass transfer coefficients at the other two exposed surfaces, but also to calculate heat transfer coefficients, h, once both heat flux and a characteristic temperature difference are known at each position of food surfaces.

Figs 20 and 21 show the local values of heat transfer coefficient evaluated, after one hour of drying, on the food surface where air impinges and on the surface parallel to air flow, respectively.

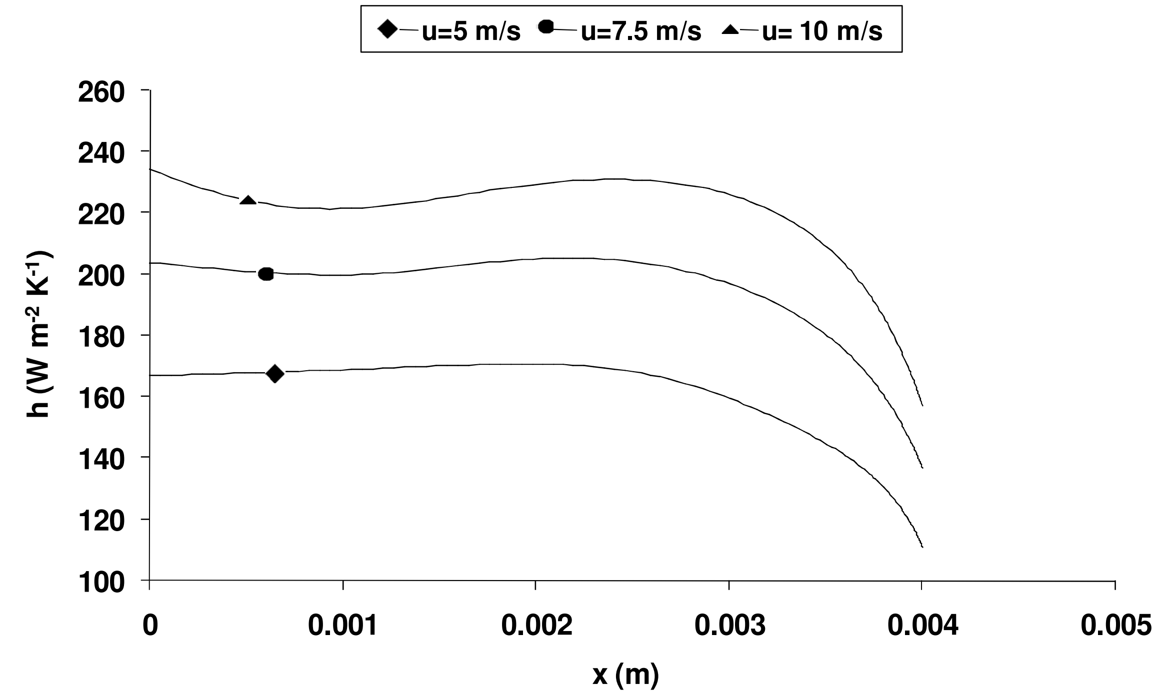


Fig. 20: Local values of heat transfer coefficient on the food surface where air impinges (effect of inlet velocity).

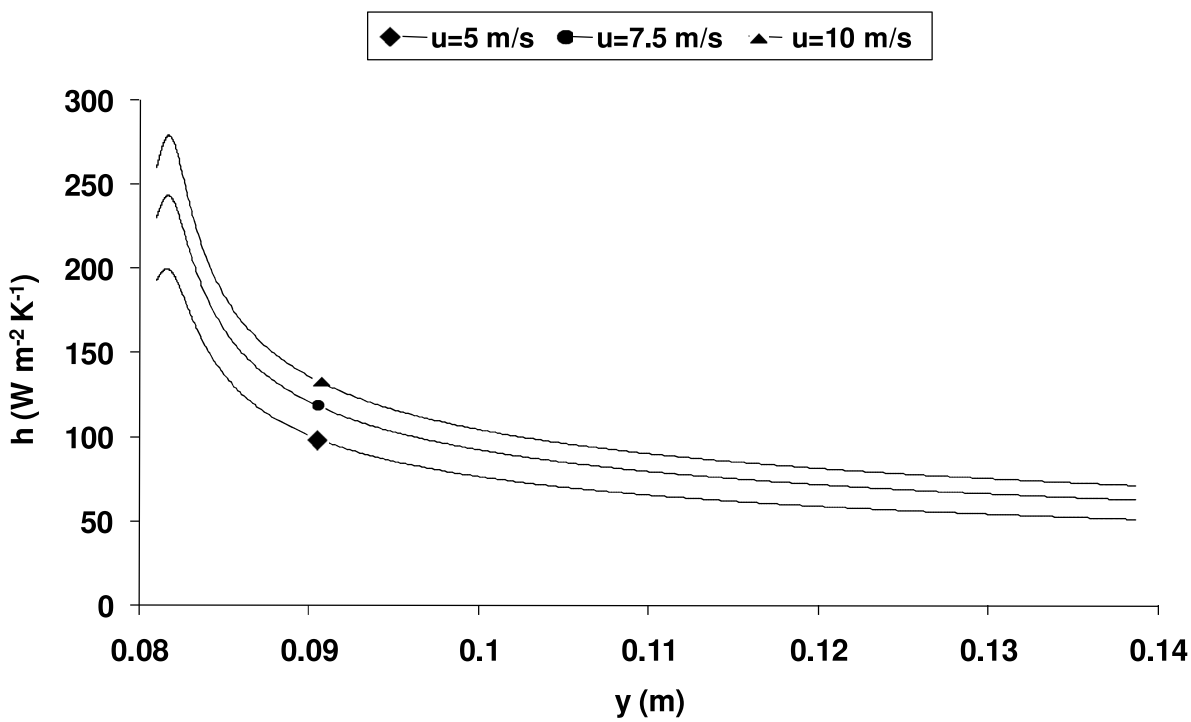


Fig. 21: Local values of heat transfer coefficient on the food surface parallel to air flow (effect of inlet velocity).

Three different values of air velocity have been tested (5, 7.5, 10 m/s) with a dry bulb temperature and an air relative humidity equal to 343 K and 30%, respectively. It should be again remarked that heat and mass transfer coefficients play a significant role when *free water* is actually removed from food and the controlling mechanism is the external transport of both heat and water. When external transport limits drying rate any variation of either air characteristics or fluid-dynamic conditions can lead to a great improvement or, conversely, to a significant worsening of both drying and heating rates. As it is shown in Figs. 20-21, impact surface has much higher values of transfer coefficients than the food surface parallel to air flow. Moreover, air velocity is responsible, as expected, for a significant variation of heat transfer coefficient that is, on average, 38% higher when feed velocity increases from 5 to 10 m/s. Similar considerations can be done about the mass transfer coefficient that is shown, for the food surface parallel to air flow only, in Fig. 22. It should be observed that the local values of *k* presented in Fig. 22 are not the result of the application of Chilton-Colburn analogy, but come directly from the solution of PDEs 7-18 coupled to PDEs 2-3. Nevertheless, by a comparison of the numerical results shown in Figs. 21-22, it can be verified that Chilton-Colburn analogy actually applies since the following relationship (Bird, Stewart, Lightfoot, 1960) is satisfied:

$$\frac{h}{k_c} = \frac{\rho_a C_{pa}}{C_{tot}} \left(\frac{Sc}{\Pr} \right)^{2/3} \qquad 28$$

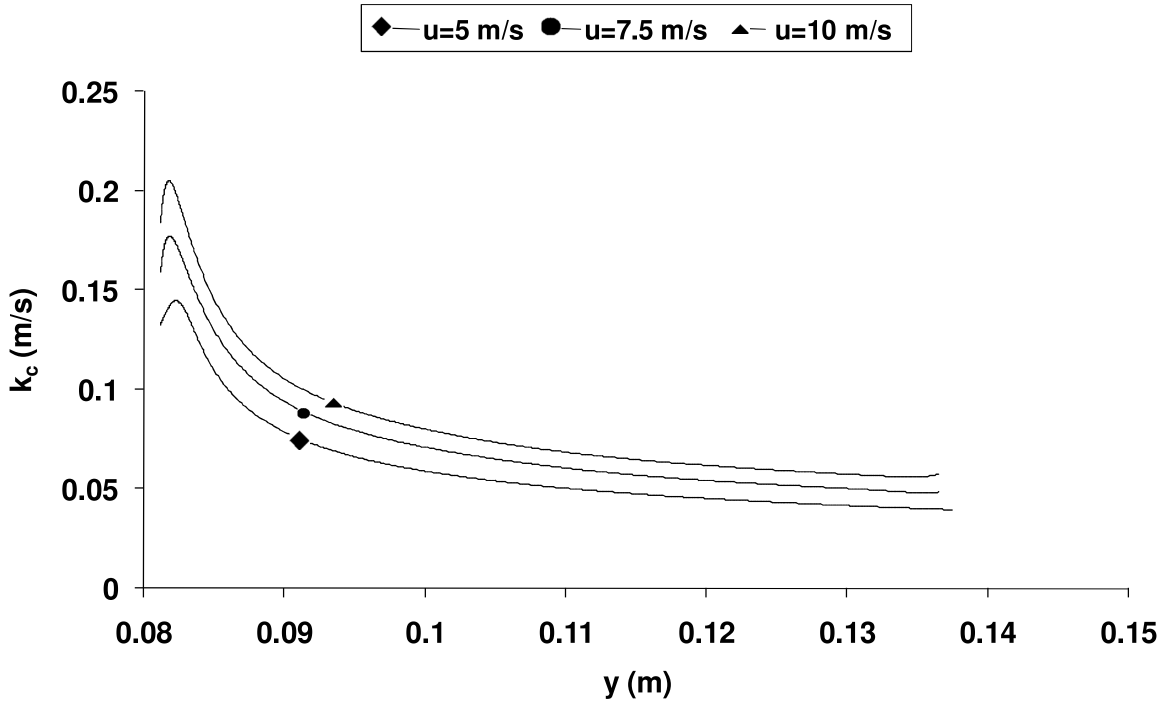


Fig. 22: Local values of mass transfer coefficient on the food surface parallel to air flow (effect of inlet velocity).

Model Validation

In addition to the numerical simulations carried out on a bidimensional flat system and described in the previous sections, some others were performed with the specific aim of verifying the model validity. This has been achieved checking the agreement between the experimental results obtained with reference to the drying, by air, of cylindrical-shaped carrots and the model theoretical predictions attained by means of *mod2*. The general transport equations (Equations 2–3 and 7-18) and their corresponding boundary conditions have been converted (Bird, Stewart, Lightfoot, 1960) to account for the cylindrical-shaped geometry of both food sample and lab-scale drying apparatus. Fig. 23 shows, in two different situations, the comparison between experimental and predicted food moisture content during two drying tests performed at the same value of air velocity (1.11 m/s) but changing inlet air temperature and its humidity. In the milder condition, dry bulb temperature was maintained at 308 K and air relative humidity was equal to 22%; in the more severe case, air temperature was of 320 K and air relative humidity was equal to 16%. The comparison shows that, in the first case, the relative error never exceeds 4% while, in the second case, it is even lower and never larger than 2.6%.

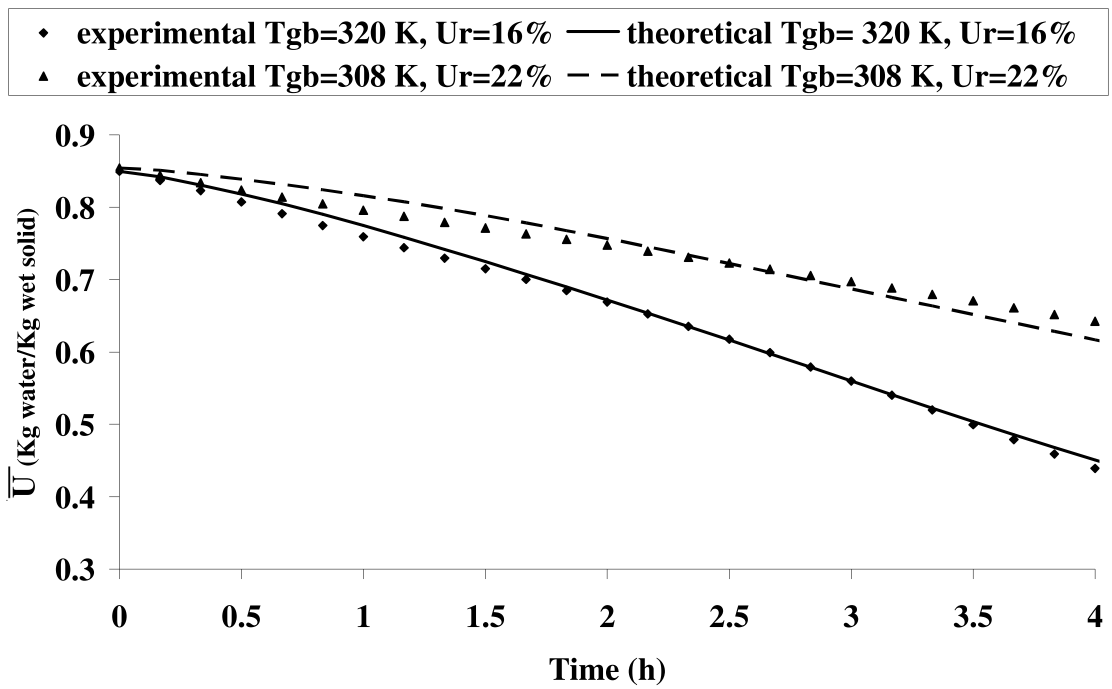


Fig. 23: Comparison between model predictions and experimental data - Dynamic evolution of food moisture content (u= 1.11 m/s, T_0= 16°C).

Conclusion

In the present contribution it has been shown that mathematical modelling represents a very effective tool for the analysis of industrial transformations, such as food drying process. Two different models have been developed with the specific aim of determining if a much simpler approach, focused on the analysis of transport phenomena within the food only, is to be preferred to a more complex analysis in which both the domains (food and air) are taken into account. Certainly, the latter approach is more general since it does not require the knowledge of any heat and mass transfer coefficient. For this reason, it is capable to simulate drying process also in those conditions for which either semi-empirical correlation are not currently available (complex food geometries) or operating conditions are changed during drying process. The simple model is capable of giving very accurate predictions in a quite short time, provided that a correct set of semi-empirical correlations is used to estimate heat and mass fluxes at the food/air interfaces. Nevertheless, the same model, strongly dependent on heat and mass transfer coefficients, fails when a wrong choice of semi-empirical correlations is performed since the actual transfers at the food/air interfaces are not correctly estimated.

Both the models can be improved to account for some phenomena that, in the present analysis, have been neglected. Both shrinkage effects and transport of water vapor, by diffusion, within the dehydrated material are definitely very important and, especially in some practical situations, may affect drying rate and, therefore, the quality of final product.

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Simulation of food drying: FEM analysis and experimental validation

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Abstract

The aim of the present work is the formulation of a theoretical model describing the simultaneous transfer of momentum, heat and mass occurring in a convective drier where hot dry air flows, in turbulent conditions, about a food sample. The proposed model does not rely on the specification of interfacial heat and mass transfer coefficients and, therefore, represents a general tool capable of describing the behavior of real driers over a wide range of process and fluid-dynamic conditions. The system of non-linear unsteady-state partial differential equations modelling the behavior of a cylindrical-shaped vegetable sample in a drier, has been solved by using Finite Elements Method. It has been observed that air characteristics influence drying performance only when external resistance to mass transfer is the rate controlling step. An experimental study was undertaken which shows very good agreement between model predictions and experimental results.

Keywords: Food drying, Transport phenomena, Finite Elements Method, Process Modelling.

List of Simbols

C	Water concentration in food	mol/m^3
C^+	Scalar variable	-
C_2	Water concentration in air	mol/m^3
C_{2i}	Water concentration in gaseous phase at the food/air interface	mol/m^3
C_{pa}	Air specific heat	J/(kg K)
C_{ps}	Food specific heat	J/(kg K)
c_μ	Model parameter	-
$c_{1\varepsilon}$	Model parameter	-
$c_{2\varepsilon}$	Model parameter	-
D_a	Diffusion coefficient of water in air	m^2/s
D_{eff}	Effective diffusion coefficient of water in food	m^2/s
f_w^0	Fugacity of water (liquid phase)	Pa
$\underline{I}$	Identity matrix	-
k	Turbulent kinetic energy	m^2/s^2
k_a	Air thermal conductivity	W/(m K)
k_c	Karman's constant	-
k_{eff}	Effective thermal conductivity of food	W/(m K)
$\underline{n}$	Unity vector normal to the surface	-
p	Pressure within the drying chamber	Pa
P_w^{sat}	Vapor pressure of water	Pa
R	Gas constant	J/(mol K)
r	Spatial coordinate	m
T	Food temperature	K
t	Time	s
T_2	Air temperature	K
T_{air}	Air temperature at the drier inlet	K
$\overline{T_b}$	Average temperature at the food surface	K
U	Moisture content on a wet basis	kg water/kg wet solid
$\overline{U}$	Average moisture content on a wet basis	kg water/kg wet solid

$\underline{u}$	Velocity vector	m/s
u_0	Air velocity at the drier inlet	m/s
U_r	Relative humidity of air	-
$\overline{X}$	Average moisture content-dry basis	kg water/kg dry solid
X_e	Equilibrium moisture content on a dry basis	kg water/kg dry solid
x_w	Molar fraction of water in the food	-
y_w	Molar fraction of water in air	-
z	Spatial coordinate	m

Greek Simbols

γ_w	Activity coefficient of water (liquid phase)	-
δ_w	Distance from the wall	m
δ_w^+	Dimensionless distance from the wall	-
ε	Turbulent energy dissipation rate	m^2/s^3
η_a	Air viscosity	Pa*s
η_t	Air turbulent viscosity	Pa*s
λ	Water latent heat of vaporization	J/kg
ρ_a	Air density	kg/m^3
ρ_s	Food Density	kg/m^3
σ_ε	Model parameter	-
σ_κ	Model parameter	-
ϕ	Dimensionless average moisture content of food	-
$\hat{\phi}_w$	Fugacity coefficient of water-vapor in solution	-
ϕ_w^{sat}	Fugacity coefficient of water-vapor at saturation conditions	-

subscripts

0	Initial condition (t=0)
atm	Atmospheric conditions
db	On a dry basis
wb	On a wet basis

Introduction

When warm dry air flows about a cold moist food sample, simultaneous heat and mass (water) transfer occurs, leading to, both, a decrease in food water content and an increase in its temperature. The above effects enhance food preservation, since microbial spoilage is generally promoted by low temperature and high moisture content. Heat and mass transfer rates depend on both temperature and concentration differences, but also on the air velocity field which strongly influences the transfer rate at food-air interfaces and, therefore, has to be properly evaluated. An exhaustive analysis of all the complex transport phenomena involved in drying process has often been regarded as being too onerous and time consuming for practical purposes. For this reason, many simplified approaches have been proposed, and are widely used by industrial drier designers (Mujumdar Arun, 2006). These approaches are based either on simplifying hypotheses, which may not be applicable in practice, or on the use of semi-empirical correlations for estimating the heat and water fluxes at food-air interfaces (Saravacos and Maroulis, 2001).

Hernàndez et al. (2000) assumed fruit drying as an isothermal process occurring at a fixed air temperature; this simplification restricted the analysis to mass transfer only. Wu and Irudayaraj, (1996) experimentally verified that drying is an isothermal process only if Biot number is very low. When Biot number is significantly greater than unity, the internal transport resistances are not negligible. Wang and Brennan (1995) developed a one-dimensional model for the simultaneous heat and mass transfer within potato slices. The hypothesis of one-dimensional transport was experimentally verified in the same study and also adopted by other authors (Kalbasi, Mehraban, 2000, Rovedo et al., 1995, Migliori et al., 2005). In a recent paper, Datta (2007 Part I) showed the different approaches that have to be used to model heat and mass transfer in food drying process. In some cases, the size of the pores as well as vapor generation within the sample have to be taken into consideration, wheras in some others cases, they can be neglected. A transport model describing the simultaneous two-dimensional heat and moisture transfer was formulated to analyze the influence of some of the most important operating variables on the drying rate of carrots (Aversa et al., 2007). The authors solved the system of unsteady-state partial differential equations, modelling the transport phenomena occurring within the food, by means of the Finite Elements Method. The dependence of physical and transport properties on food temperature and moisture content has been investigated by several authors (Datta, 2007 Part II, Panagiotou et al., 2004, Lewicki 2004, Saravacos, Maroulis, 2001). The large number of available studies dealing with drying process suggests, however, that more general and versatile mathematical models must be formulated. These models can potentially overcome some problems regarding, for instance, the use of the most appropriate semi-empirical correlations for evaluating heat and mass transfer coefficients at the food/air interfaces. Kondjoyan and Boisson (1997) and Verboven et al. (1997), in fact, found that even small errors in the estimation of transfer coefficients could lead to large deviations between estimated and real values of temperature and moisture content, leading to inappropriate equipment design or severe processing problems. It is therefore desirable to formulate a transport model accounting for the simultaneous transfer of momentum, heat and mass in air as well as within the food. Moreover, with a proper set of boundary conditions expressing the continuity of both heat and mass fluxes at the food/air interfaces, it is possible to estimate the actual transport rates without resorting to any empirical correlation. Such a model is therefore capable of predicting the drying behavior of foods available in all shapes, and over a very wide range of operating and fluid-dynamic conditions. The present work is intended to fill such a gap in modelling the drying of foods. The main objective of this paper is to formulate an accurate transport model analyzing the simultaneous transfer of momentum (for air only), and of heat and mass (for both air and food) occurring in a convective drier where hot dry air flows, under turbulent conditions, around a cold wet food sample. The model can be used to study the time evolution of some characteristic parameters, namely the moisture content of food and its temperature, expressed as functions of the operating conditions, i.e. inlet velocity, relative humidity and temperature of the air. The model proposed, minimizes expensive pilot scale tests and satisfactorily indicates the characteristics and safety of dried products.

Theoretical background

In a convective drier, two different transport mechanisms simultaneously occur: heat is transferred from air to the material; water is transported from the core of the material to its surface and, eventually, to air. The rates of heat and mass transfer certainly depend on both temperature and concentration differences, but also on air velocity field which any model has to properly predict. In the following, the main physical assumptions and the mathematical equations used to model the unsteady-state behavior of a convective drier will be discussed. It is assumed that drying air is continuously supplied to the oven inlet section and flows around a cylindrical food sample in the axial direction, parallel to its length (Fig. 1). Heat and mass transfer resistance are assumed negligible across the mesh on which the food is placed (Thorvaldsson and Janestad 1999; Viollaz and Rovedo, 2002). The system under investigation is symmetric; and only half of the original domain will henceforth be considered. Moreover, variation occurring in the θ direction (i.e. angular coordinate) will be neglected, thus restricting the analysis to a 2D geometry. Each dependent variable is, therefore, function of the radial and axial coordinates, r and z, respectively, and of time, t. The drier under investigation has a length of 25 cm and a radius of 10 cm. The food sample, assumed to be a half cylinder 6 cm long and 5 mm radius, is placed 8 cm far from the drier inlet.

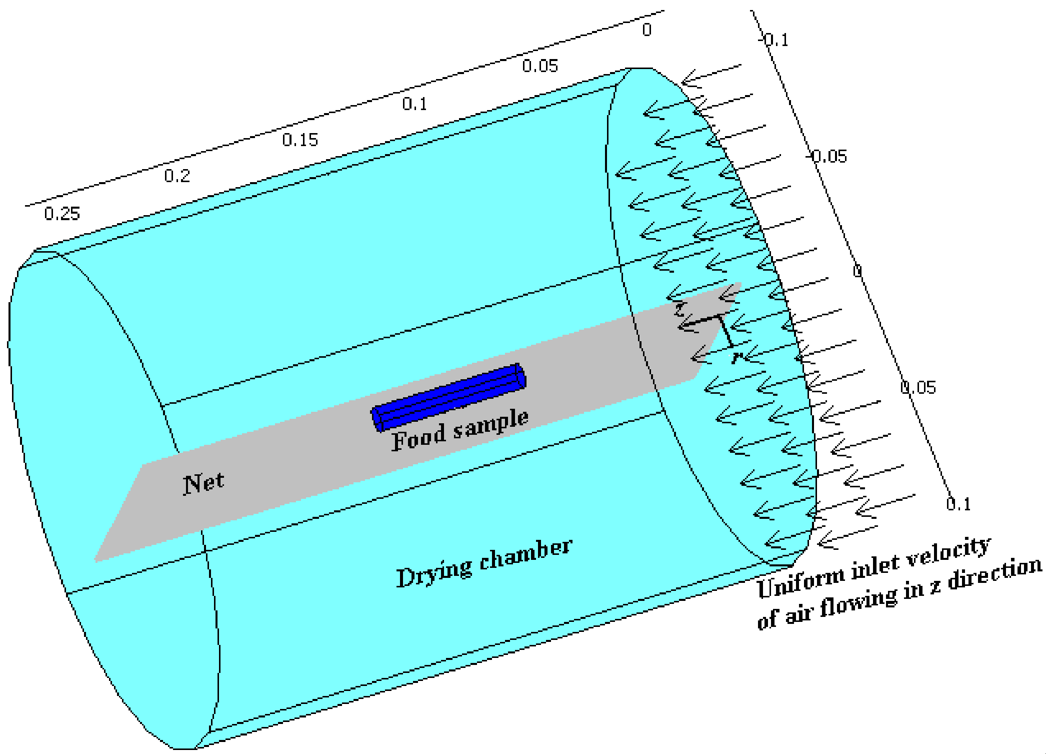

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Fig. 2: Schematic representation of the drying chamber.

The food sample is a multiphase hygroscopic porous medium; yet it is assumed to be a fictitious continuum. The contribution of convection to the transport equations written for the food sample has been neglected, assuming weak internal evaporation. Also the transport of water vapor by diffusion within the dehydrated material towards the external food surface, has not been considered since this mechanism is only significant in the case of highly porous media, whereas it can be neglected in the case of vegetables since typical void fraction values are less than 0.3 (May and Perré, 2002). Mass transfer in the product, therefore, occurs only by diffusion; and heat transfer, only by conduction. Shrinkage effects were assumed to be negligible in the range of moisture content considered here. The proposed model accounts for the variation of food physical and transport properties which are dependent on the local values of temperature and moisture content.

In air, heat transfer occurs by convection and conduction, whereas the water is transferred by convection and diffusion. The convective contributions to heat and mass transfer are related to air circulation, and the velocity field is to be determined by solving the non-isothermal momentum transport equations for turbulent flow, coupled with the continuity equation. It is assumed that evaporation/condensation only occur at the air-food interface where a proper set of boundary conditions has to be specified. These boundary conditions substantially express the continuity of temperatures and heat and mass fluxes; thermodynamic equilibrium is assumed to prevail between the water concentration in air and the moisture content at the food-air interface. Physical and transport properties of air have been also expressed in terms of the local values of temperature and water content.

Following on from the Fick's law, the unsteady-state mass transfer equation for the movement of water within the food sample is (Bird et al. 1960, Welty et al., 2001):

$$\partial C / \partial t + \underline{\nabla} \cdot \left(- D_{eff} \, \underline{\nabla} C \right) = 0 \qquad 1$$

where C is the water concentration in food and D_{eff} is the effective diffusion coefficient of water in the food.

According to Fourier's law, the energy balance in the food material leads to the unsteady-state heat transfer equation (Bird et al., 1960, Welty et al., 2001):

$$\rho_s C_{p_s} \partial T / \partial t - \underline{\nabla} \cdot \left(k_{eff} \, \underline{\nabla} T \right) = 0 \qquad 2$$

where T is temperature; ρ_s is the density of the food sample; C_{ps} its specific heat and k_{eff} is the effective thermal conductivity. The subscript eff refers to food transport properties evaluated as effective, i.e. accounting for a possible combination of different transport mechanisms. The energy balance does not contain any evaporation term since it is assumed that evaporation only occurs at the food surface which is actually exposed to the drying air.

The non-isothermal turbulent flow of air within the drying chamber has been modelled by means of the well-known k-ε model (Verboven et al., 1997, Verboven et al., 2001) that is based on two additional semi-empirical transport equations for the variables k and ε, i.e. the turbulent kinetic energy and the turbulent energy dissipation rate, respectively. The unsteady-state momentum balance coupled to the continuity equation lead to (Bird et al. 1960, Verboven et al., 2001):

$$\partial \rho_a / \partial t + \underline{\nabla} \cdot \rho_a \underline{u} = 0 \qquad 3$$

$$\rho_a \, \partial \underline{u} / \partial t + \rho_a \underline{u} \cdot \underline{\nabla} \underline{u} = \underline{\nabla} \cdot \left[- p \underline{I} + (\eta_a + \eta_t) \left(\underline{\nabla} \underline{u} + (\underline{\nabla} \underline{u})^T - (2/3)(\underline{\nabla} \cdot \underline{u}) \underline{I} \right) \right] \qquad 4$$

$$\rho_a \, \partial k / \partial t + \rho_a \underline{u} \cdot \underline{\nabla} k = \underline{\nabla} \cdot \left[\left(\eta_a + \eta_t / \sigma_k \right) (\underline{\nabla} k) \right] + \eta_t P(\underline{u}) - (2 \rho_a k / 3)(\underline{\nabla} \cdot \underline{u}) - \rho_a \varepsilon \qquad 5$$

$$\rho_a \, \partial \varepsilon / \partial t + \rho_a \underline{u} \cdot \underline{\nabla} \varepsilon = \underline{\nabla} \cdot \left[\left(\eta_a + \eta_t / \sigma_\varepsilon \right) (\underline{\nabla} \varepsilon) \right] + (c_{1\varepsilon} \varepsilon / k)[\eta_t P(\underline{u}) - (2 \rho_a k / 3)(\underline{\nabla} \cdot \underline{u})] - c_{2\varepsilon} \rho_a \varepsilon^2 / k \qquad 6$$

where ρ_a is the air density and ηa is its viscosity, both expressed in terms of the local values of temperature and of water content; p is the pressure within the drying chamber; u is the velocity vector; c_μ ,σ_k, σ_ε, $c_{1\varepsilon}$ and $c_{2\varepsilon}$ are constants whose value depends on the k-ε turbulence model used. In the present analysis, the standard k-ε model has been adopted (Verboven et al., 2000). The term, P(u), contains the contribution of the shear stresses:

$$P(\underline{u}) = \underline{\nabla u} : \left(\underline{\nabla u} + (\underline{\nabla u})^T\right) - \left(2/3\right)\left(\underline{\nabla} \cdot \underline{u}\right)^2 \qquad 7$$

The following definition for η_t, i.e. the turbulent viscosity, in Eqs. 4-6, holds:

$$\eta_t = \rho_a c_\mu k^2 / \varepsilon \qquad 8$$

The energy balance in the drying air, accounting for both convective and conductive contributions, leads to (Bird et al., 1960, Welty et al., 2001):

$$\rho_a C_{pa} \partial T_2 / \partial t - \underline{\nabla} \cdot \left(k_a \underline{\nabla T_2}\right) + \rho_a C_{pa} \underline{u} \cdot \underline{\nabla T_2} = 0 \qquad 9$$

where T_2 is the air temperature; C_{pa} is its specific heat; and k_a is the thermal conductivity.

The water mass balance in the drying air, accounting for both convective and diffusive contributions, leads to (Bird et al., 1960, Welty et al., 2001):

$$\partial C_2 / \partial t + \underline{\nabla} \cdot \left(- D_a \underline{\nabla C_2}\right) + \underline{u} \cdot \underline{\nabla C_2} = 0 \qquad 10$$

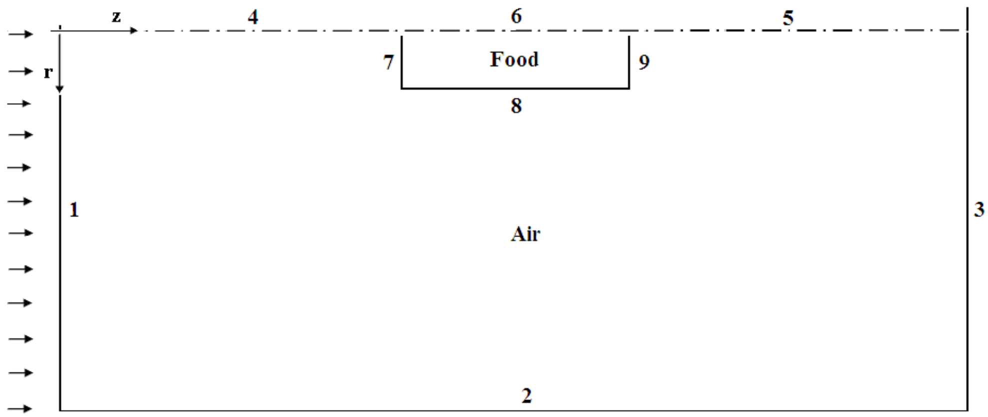
where C_2 is water concentration in the air and D_a is the diffusion coefficient of water in air. Since physical and transport properties of both air and food are expressed in terms of the local values of temperature and moisture content, Eqs. 1-10 represent a system of unsteady, non-linear partial differential equations which can only be solved by means of a numerical method.

A set of initial conditions, typically prevailing in an industrial drying process (Table 1), is necessary to perform the numerical simulations. As far as the air is concerned, it was assumed that: the water concentration, C_{20}, was equal to 0.729 mol/m3 (corresponding to a relative humidity, U_{r0}, of 20%); the temperature, T_{20}, took the value 318 K; and the pressure in the drying chamber, p_0, was 1 atm.

Initial condition	Description	Value
$\boldsymbol{C_{20}}$	Water concentration in air	0.729 mol/m^3
$\boldsymbol{T_{20}}$	Air temperature	308 K
$\boldsymbol{p_0}$	pressure in the drying chamber	1 atm
$\underline{u}$	Air velocity	$\underline{0}$
$\boldsymbol{T_0}$	food temperature	289 K
$\boldsymbol{C_0}$	moisture content in food	44860 mol/m^3 = 0.85 kg H_2O/kg_{wb} = 5.67 kg H_2O/kg_{db}

Tab. 1: Initial conditions used in the present theoretical model

It was also assumed that the air was stationary (i.e $\underline{u} = \underline{0}$). The initial values of the food temperature, T_0, and its moisture content, C_0, were set equal to 289 K and 44860 mol/m3, respectively. The adopted C_0 value refers to an average initial moisture content, typical of fresh carrots corresponding to 0.85 kg H_2O/kg_{wb} (on wet basis) or to 5.67 kg H_2O/kg_{db} (on dry basis). The boundary conditions are reported, for the sake of brevity, in Fig. 2 where each different boundary is identified by an integer ranging from 1 to 9.



Boundary	Type of condition
1	$u_r = 0;\ u_z = u_0$ $T_2 = T_{air}$ $C_2 = C_{air}$
2	Logarithmic wall function: $\left[(\eta_a+\eta_t)\left(\nabla \underline{u}+(\nabla \underline{u})^T-(2/3)(\nabla\cdot\underline{u})\underline{I}\right)\right]\underline{n}=\left[\dfrac{\rho_a c_\mu^{0.25} k^{0.5}}{\left(\dfrac{\ln(\delta_w^+)}{k_c}+C^+\right)}\right]\underline{u};$ $\underline{n}\cdot\nabla k=0;\quad \varepsilon=\dfrac{c_\mu^{0.75}\,k}{k_c\,\delta_w};$ where $\delta_w^+=\dfrac{\delta_w\,\rho_a c_\mu^{0.25}\,k^{0.5}}{\eta_a}$ $T_2 = T_{air}$ $C_2 = C_{air}$
3	$p = p_{atm}$ (Atmospheric pressure) $\underline{n}\cdot(-k_a\nabla T_2)=0$ (Convection prevailing over conduction) $\underline{n}\cdot(-D_a\nabla C_2)=0$ (Convection prevailing over diffusion)
4, 5, 6	Axial symmetry (except for momentum on boundary 6)
7, 8, 9	Logarithmic wall function: $\left[(\eta_a+\eta_t)\left(\nabla \underline{u}+(\nabla \underline{u})^T-(2/3)(\nabla\cdot\underline{u})\underline{I}\right)\right]\underline{n}=\left[\dfrac{\rho_a c_\mu^{0.25} k^{0.5}}{\left(\dfrac{\ln(\delta_w^+)}{k_c}+C^+\right)}\right]\underline{u};$ $\underline{n}\cdot\nabla k=0;\quad \varepsilon=\dfrac{c_\mu^{0.75}\,k}{k_c\,\delta_w};$ where $\delta_w^+=\dfrac{\delta_w\,\rho_a c_\mu^{0.25}\,k^{0.5}}{\eta_a}$ $T = T_2$ (Temperature continuity) $\underline{n}\cdot\left(-k_a\nabla T_2+\rho_a C_{pa}\,\underline{u}\,T_2\right)=\underline{n}\cdot\left(-k_{eff}\nabla T\right)-\lambda\,\underline{n}\cdot\left(-D_{eff}\nabla C\right)$ (Heat flux continuity) $\underline{n}\cdot\left(-D_a\nabla C_2+C_2\,\underline{u}\right)=\underline{n}\cdot\left(-D_{eff}\nabla C\right)$ (Mass flux continuity) $\gamma_w\,x_w\,P_w^{sat}=y_w\,p$ (Thermodynamic equilibrium)

2

Fig. 2: Boundary conditions used to formulate the theoretical model.

At the food-air interface (boundaries 7 - 9), where no accumulation occurs, the continuity of both heat and water fluxes is imposed. In particular, the heat transported by convection and conduction from air to food is partly used to raise the sample temperature by conduction and partly to allow water evaporation described by considering the latent heat of vaporization (λ). Similarly, a balance also applies between the diffusive flux of liquid water coming from the core of the sample, and the flux of vapor leaving the food surface and transported into the drying air. Moreover the thermodynamic equilibrium at the air-food interface can be expressed as (Smith et al., 1987):

$$\gamma_w \, x_w \, f_w^0 = \hat{\phi}_w \, y_w \, p \qquad 11$$

where γ_w, the activity coefficient of water and f_w^0, the fugacity of water, refer to the liquid phase, $\hat{\phi}_w$, the fugacity coefficient of water, refers – instead – to the vapor phase; x_w and y_w are the mole fractions of water in food and in air, respectively; and p is the pressure within the drying chamber. At low pressures, the vapor phase usually approximates to ideal gas behaviour and eq. 11 can be simplified as:

$$\gamma_w \, x_w \, P_w^{sat} = y_w \, p \qquad 12$$

where P_w^{sat} is the vapor pressure of water at the interface temperature. It may be noted that in hygroscopic materials, like most of the foods, the parameter γ_w accounts for the effects of physically bound water; it is, therefore, usually expressed (as in the present study) as a function of both food moisture content and its temperature (Datta 2007 Part II, Ruiz-López et al., 2004). Once this relationship between activity coefficient and moisture content, and temperature is known for the particular food under examination, Eq. 12 permits calculating the mole fraction of water in the vapor phase and, therefore, the value of C_{2i} (i.e. water concentration evaluated in gaseous phase at the food/air interface).

On boundary 2, it is assumed that air temperature and its moisture concentration are equal to the values measured at the drier inlet. These two conditions are valid under the assumption that the temperature and concentration profiles are confined to two very thin regions which develop close to the food-air interface. the validity of the above assumptions will be considered while discussing the simulations results. At the drier outlet (boundary 3), conduction and diffusion can be neglected in favour of convection which dominates (Danckwerts' conditions). Finally, the boundary conditions for momentum balance at the solid surfaces are expressed in terms of a two-velocity scale wall function, as reported by Lacasse et al. (2004).

The above system of unsteady, non-linear PDEs has been solved by Finite Elements Method using a commercial package, i.e. Comsol Multiphysics 3.3. Both food and air domains were discretized into a total number of 5470 triangular finite elements leading to about 57500 degrees of freedom. In particular, the mesh consisted of 1576 and 3894 elements within the food and air domains, respectively. Fig. 3 shows the details of the mesh considered, which provided a satisfactory spatial resolution for the system under study.

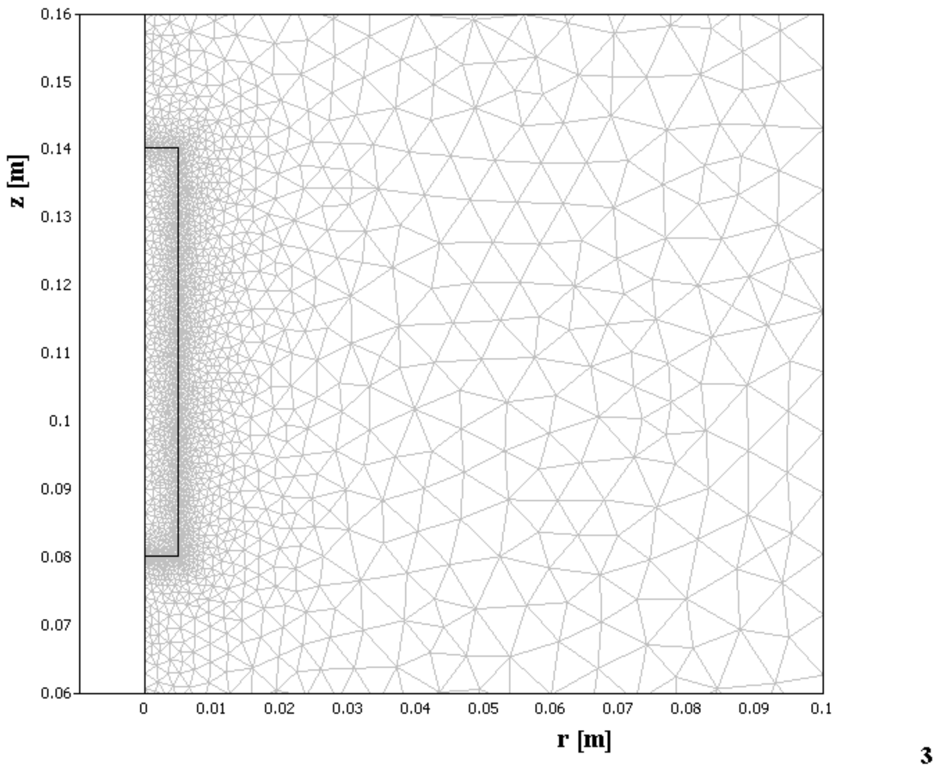


Fig. 3: Discretization of food and air domains into triangular finite elements (detail of the mesh).

It was also found that the solution was independent of the grid size even with further refinements. Lagrange finite elements of order two were chosen for the components, u_r and u_z, of air velocity vector u, for the turbulent kinetic energy and the turbulent energy dissipation rate but also for water concentration and temperature in, both, air and food. A Lagrange finite element of order one was used for variable p, i.e. the pressure distribution within the drying chamber. The time-dependent problem was solved by an implicit time-stepping scheme, leading to a non-linear system of equations for each time step. Newton's method was used to solve each non-linear system of equations, whereas a direct linear system solver was adopted to solve the resulting systems of linear equations. The relative and absolute tolerances were set to 0.005 and 0.0005, respectively. On a Pentium IV computer running under Linux, a typical drying time of 5 hours was simulated in about 45 minutes.

Experimental

The validity of the proposed theoretical model was ascertained by undertaking experiments using cylindrical carrots samples, dried by air of known characteristics. A standard procedure was adopted to perform the experiments which aimed to monitor the time evolution of the average moisture content of carrots and a single value of temperature measured at a point close to the surface where the drying air impacted with the sample . Each experiment was repeated twice to ascertain the reproducibility of the results. Nevertheless, only the average value of each of the two measurements will be reported here, since the calculated standard deviations never exceeded 5%. Carrots were purchased from the local market. They were cut by a proper tool that allowed obtaining regular cylinders having a diameter of 1 ± 0.01 cm and a length of 6 ± 0.01 cm. The above dimensions were measured by a vernier caliper providing a precision of 0.02 mm. Some fragments, collected randomly from the same carrot used to perform the experiments, were used to measure the initial moisture content of each sample by an electronic moisture analyzer (HB43, Mettler Toledo). The cylindrical carrot samples were put into a lab-scale drying cell consisting of a cylinder having a diameter 15 cm and a length 18 cm, equipped with a stainless steel wide-mesh net where each sample was placed. To determine the time evolution of the average moisture content, weight losses were measured at given time intervals by a precision balance, Sartorius CP225D-OCE. To follow the time evolution of local temperature within the sample, a capillary thermocouple (PTFE,K,PK, RS components) was inserted 0.8 mm deep inside the carrot sample, 1 cm from the impact surface. The thermocouple, having a diameter of 0.6 mm, did not interfere with the measurements, and it was connected to a two-channel digital thermometer (KRS52 dual, RS components). Drying air was fed to the cell by an electric fan that allowed selecting air inlet flow rate, its temperature and the relative humidity. The drying air entered in the axial direction and flowed parallel to the main axis of the food sample. Air characteristics, i.e. velocity, temperature and relative humidity were continuously monitored at the inlet section of the drying cell by an Atmos anemometer. Different conditions were chosen to test the theoretical model proposed.

However, the experimental results obtained at three different air temperatures (308 K, 311 K and 320 K), two different values of relative humidity (16% and 22 %) and two different air velocities (1 m/s and 1.5 m/s) will be discussed in this paper. It may be noted that the operating conditions selected were such that shrinkage effects, not accounted for by the present model, occurred only to a limited extent and in the very last part of each experiment.

Results and discussion

The analysis of the transport phenomena presented here is based on the examination of the time evolutions of averaged variables, i.e. the dimensionless moisture content and the temperature on each exposed surface, and water concentration distributions. Simulations at different drying conditions were performed to show how air characteristics (dry bulb temperature, relative humidity and inlet velocity) influenced the drying process and, in particular, the drying rate. It should be remarked that turbulent conditions prevailed in the drying chamber in the range of operating variables tested, since the Reynolds numbers encountered were well in excess of 104.

Figs. 4-6 show the time evolutions of dimensionless average moisture content, $\varphi(t)$ with air inlet temperature, relative humidity and air velocity, respectively, as parameters. $\varphi(t)$ is defined as:

$$\varphi(t) = \frac{\overline{X}(t) - X_e}{\overline{X_0} - X_e} \qquad 13$$

where $\overline{X}(t)$ is the moisture content (on a dry basis), averaged over the sample volume and evaluated as a function of drying time; $\overline{X_0}$ is the average initial moisture content; and X_e is the equilibrium moisture content as defined by Ruiz-López, et al. (2004).

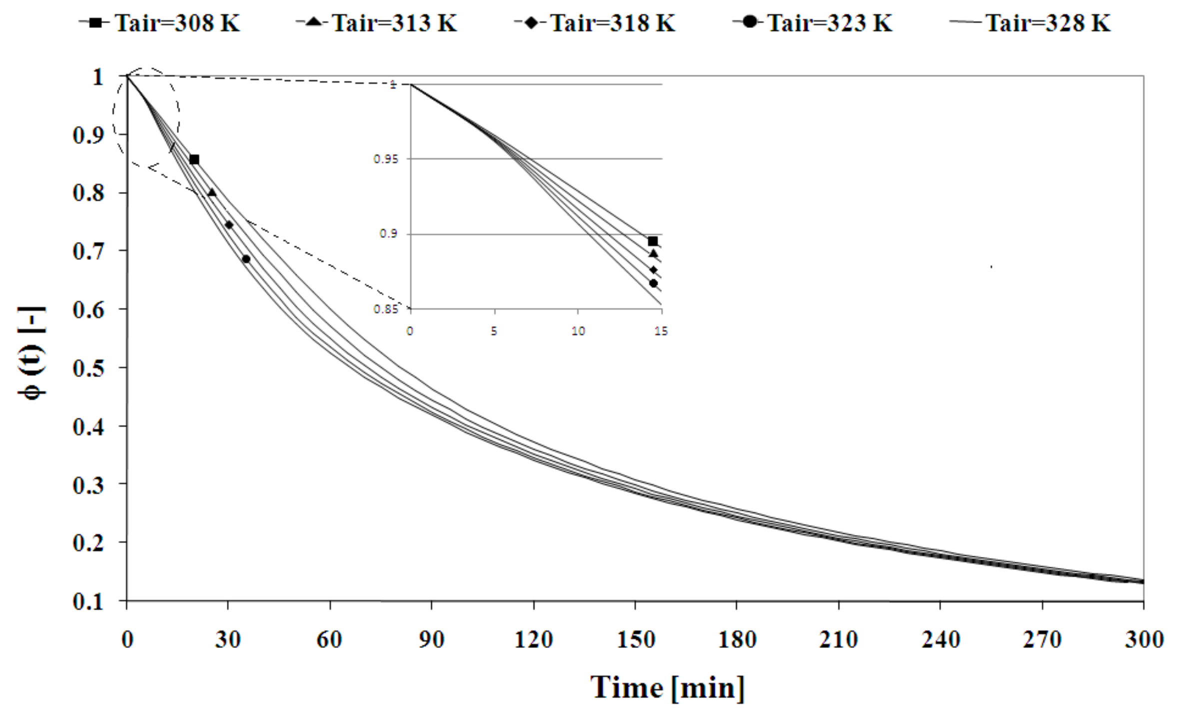


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Fig. 4: Time evolution of dimensionless average moisture content with inlet air temperature as the parameter ($T_0 = 289$ K, $U_0 = 0.85$ kg H_2O/kg$_{wb}$, $u_0 = 1.5$ m/s, $U_r = 20\%$).

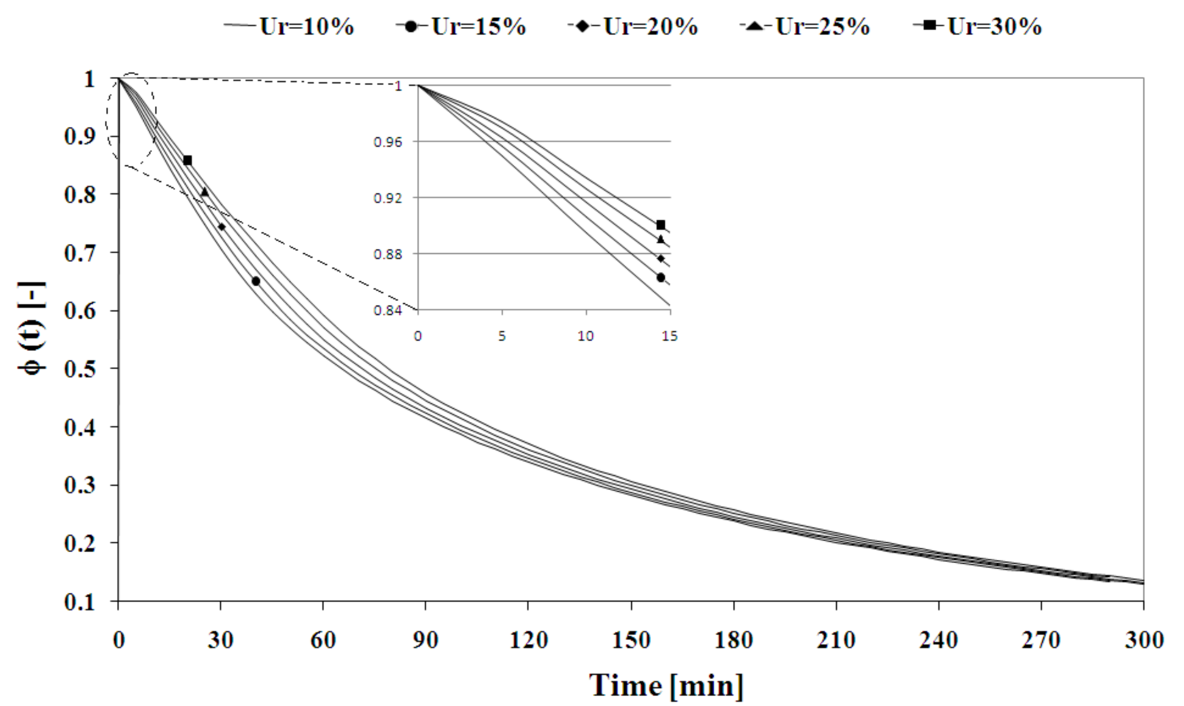

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Fig. 5: Time evolution of dimensionless average moisture content with inlet air relative humidity as the parameter ($T_0 = 289$ K, $U_0 = 0.85$ kg H_2O/kg$_{wb}$, $u_0 = 1.5$ m/s, $T_{air} = 318$ K).

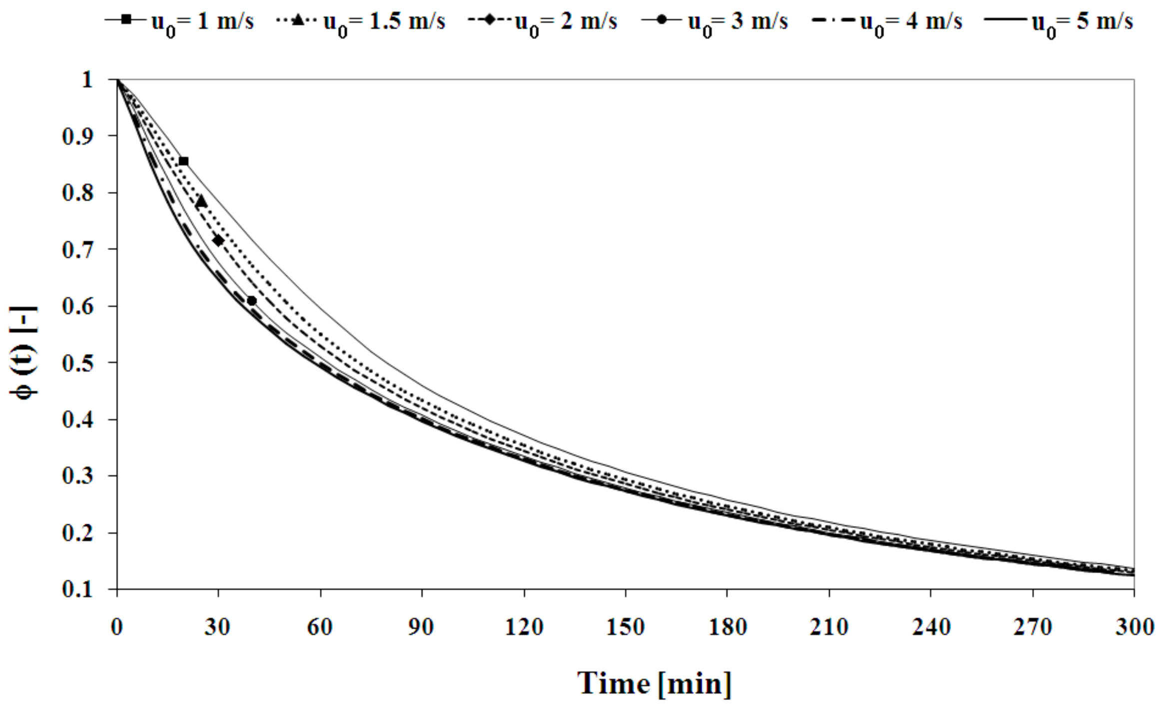

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Fig. 6: Time evolution of dimensionless average moisture content with inlet air velocity as the parameter (T_0 = 289 K, U_0 = 0.85 kg H_2O/kg_{wb}, T_{air} = 318 K, U_r = 20%).

An increase in air inlet temperature from 308 K to 328 K (Fig. 4) results in an increase in wet bulb temperature from 292 K to about 305 K (Perry and Green, 1984) that drives the system away from equilibrium conditions determining an initial delay in the start-up of drying. At the very beginning of the process all the curves overlap showing an almost identical slope; then, they part evidencing a drying rate (proportional to the slopes of φ (t) vs. time curves) that is higher and higher as dry bulb temperature increases. The increase in wet bulb temperature may potentially cause an initial humidification of the exposed surfaces, especially at higher values of air relative humidity. This phenomenon is caused by the condensation of water from the drying air on the cold food surfaces and continues until the surface temperature is lower than wet bulb temperature. Once the wet bulb temperature is attained, evaporation of free water dominates over vapour condensation and the food actually begins to dry. Moreover, at a given relative humidity, an increase in the dry bulb temperature results in an increase in the moisture content of the air which causes a significant reduction in the driving force to mass transfer from food to air; however, this is accompanied by an improvement in heat transfer from air to food. The observed rate of drying therefore depends on air inlet temperature during the initial stage of the process, where higher inlet temperature causes faster drying. On the other hand, the effect of air temperature on drier performance is less important in the last three hours of the process, when bound water is removed and the internal resistance to mass transport prevails and represents the limiting step. During this stage, it can be observed from Fig. 4 that the drying rate is much slower, and φ (t) values and the corresponding drying rates almost coincide regardless of the inlet air temperature.

When the relative humidity of inlet air increases from 10% to 30%, an increase in, both, the wet bulb temperature by about 7.5 K, as well as its moisture content by a factor of about three, is observed. Therefore, an increase in relative humidity results in a lower drying rate and causes a longer initial delay before drying actually commences (Fig. 5). Moreover, when free water is removed from the food, an increase in relative humidity of air from 10% to 30% significantly lowers the drying rate owing to a reduction in the driving force which promotes moisture transfer. Nevertheless, when bound water is involved in the drying process, the internal resistance to mass transfer represents the rate limiting step, and the concentration difference outside the food sample does not significantly influence the drying rate. Also, in this case, the various curves showing φ vs. time tend to coincide during the later stage of the drying process.

Fig. 6 shows the effect of air inlet velocity on moisture removal. An increase in air velocity corresponds to a decrease in external resistances to heat and mass transfer. This results in faster drying which can be clearly observed when free water evaporates and leaves the food, although no significant variation is observed when inlet velocity is increased, for example, from 4 to 5 m/s. As soon as bound water is removed from the food, internal resistance to mass transfer predominates and air velocity has a weak influence on the process rate, since the value of effective diffusivity of water in food, dependent on both temperature and moisture content, controls the drying rate. The present model confirms that changes in dry bulb temperature, inlet humidity and inlet velocity influence the drying rates when free water is involved in the transfer process. On the other hand, when bound water is to be removed from the food and internal resistances control the process rate, the operating variables have little effect on the system behaviour. The proposed model may represent a powerful tool to analyze the behaviour of industrial driers over a wide range of process and fluid-dynamic conditions. It may also be used to find an "optimal" set of operating conditions aimed at improving the final product quality.

The time evolution of food sample temperature, averaged on each surface exposed to the drying air, i.e. boundaries 7-9, are shown in Figs. 7 and 8.

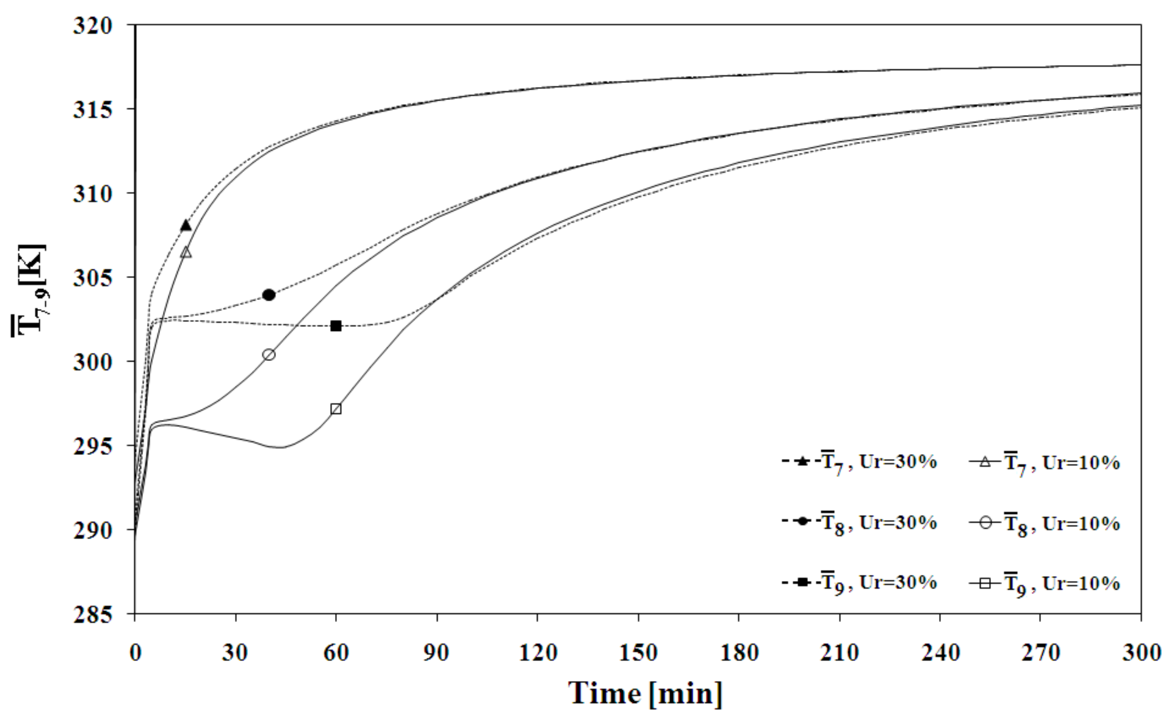

Fig. 7: Time evolution of average temperature on each food surface with inlet air relative humidity as the parameter (T_0 = 289 K, U_0 = 0.85 kg H_2O/kg$_{wb}$, u_0 = 1.5 m/s, T_{air} = 318 K).

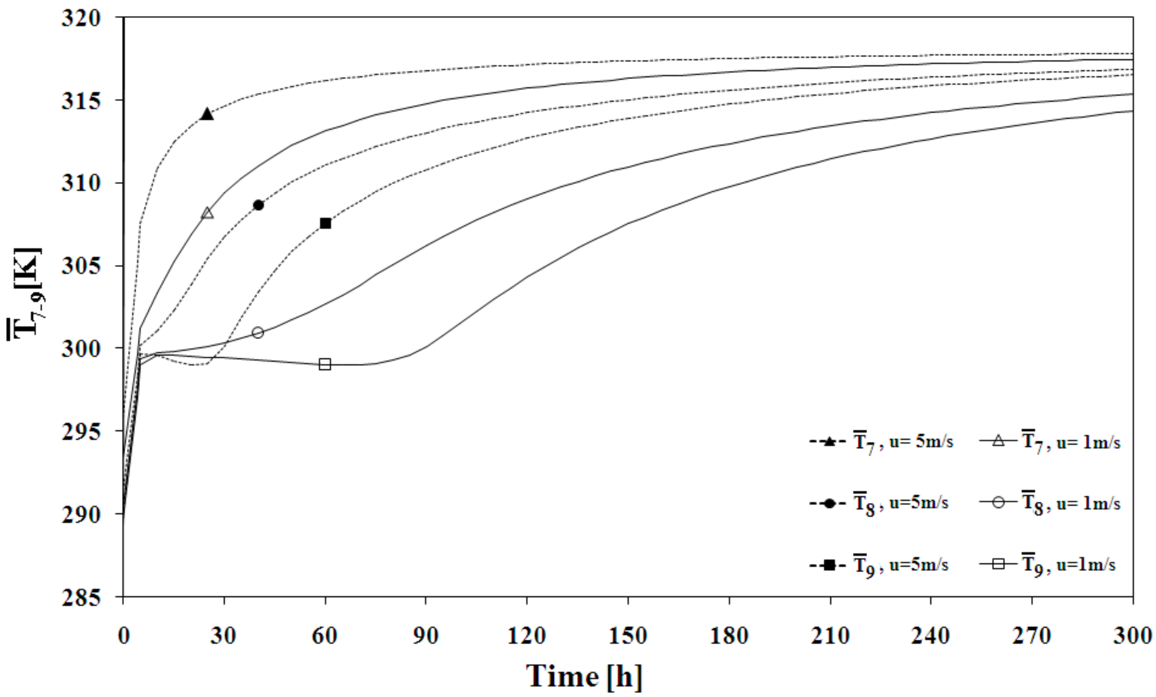

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Fig. 8: Time evolution of average temperature on each food surface with inlet velocity as the parameter (T_0 = 289 K, U_0 = 0.85 kg H_2O/kg$_{wb}$, U_r = 20%, T_{air} = 318 K).

The plots were obtained at two different values of relative humidity and inlet velocity, respectively. The air-impact food surface (boundary 7) is heated more rapidly than the other surfaces and its temperature rises quickly above the wet bulb temperature. This behavior is definitely ascribed to the flow conditions which have a strong influence on the external resistances to both heat and mass transfer. At the impact surface: 1) the heat transferred from air is provides both the latent heat of vaporization necessary for water evaporation and the sensible heat necessary to increase the temperature of the dry matter by conduction; 2) the drying rate is immediately very high and it involves evaporation of free water which is transported into the air. A completely different behavior is observed at the opposite end, i.e. on boundary 9, where a rather wide “segregation” region forms due to inefficient air circulation, as shown by the velocity field shown in Fig 9.

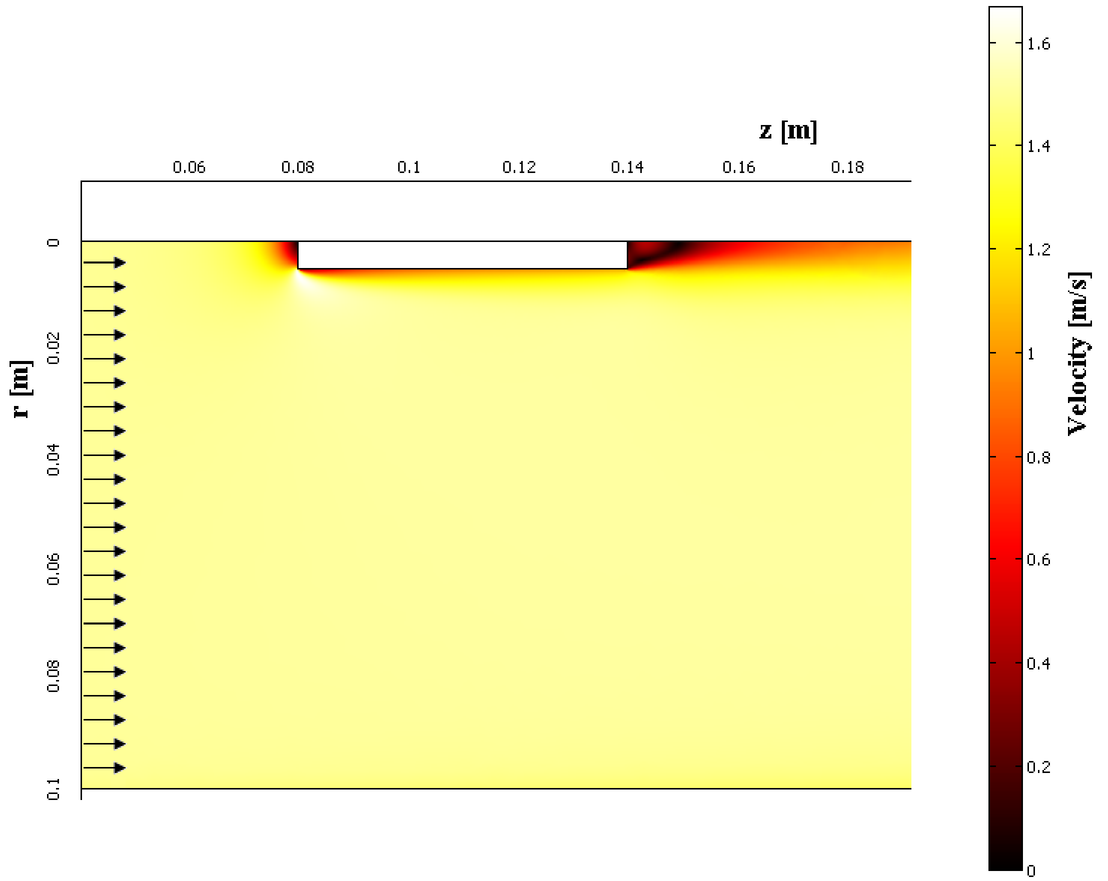

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Fig. 9: Air velocity field developing close to the food sample (t = 60 min, T_0 = 289 K, U_0 = 0.85 kg H_2O/kg$_{wb}$, u_0 = 1.5 m/s, T_{air} = 318 K, U_r = 20%).

The time evolutions of average temperature corresponding to the "rear" end of the food surface show that, at the very beginning of drying process, the heat transported from air increases the food temperature up to the wet bulb temperature; later on, the average temperature is nearly constant. During this stage, the duration of which depends on the operating conditions (shorter when air relative humidity is low or inlet velocity is high), the heat transferred from air provides only the latent heat of evaporation to the free water. When bound water starts to get eliminated, the heat transported by convection and conduction from air to food is partially used to raise the sample temperature by conduction and partially to allow water evaporation. The average temperature continuously increases up to a plateau value, with a slope that depends on the values of air velocity and relative humidity. The behavior of the food surface parallel to the direction of air flow (boundary 8) can be considered to be "intermediate" between that of the other two boundaries.

Temperature profiles do exhibit a nearly constant value whose duration is, however, shorter, when air relative humidity is equal to 10%, or tends to vanish as soon as the air velocity reaches its highest value, thus determining a significant decrease in external transport resistances. A comparison between the values of wet bulb temperature predicted by the model and that obtained from the psychrometric chart (Perry and Green, 1984), under several conditions, is given in Tab.2; a very good agreement is evident with relative errors being at most 0.5%.

Dry bulb Temperature [K]	**Air relative Humidity (%)**	**Wet Bulb Temperature [K] (psychrometric chart)**	**Wet Bulb Temperature [K] (predicted by the model)**	**Relative Error (%)**
318.15	20	298.35	299.75	0.47
308.15	20	292.05	293.55	0.51
313.15	20	295.15	296.65	0.51
323.15	20	301.55	302.85	0.43
328.15	20	304.75	305.95	0.39
318.15	10	294.25	295.25	0.34
318.15	15	296.35	298.05	0.57
318.15	25	300.15	301.35	0.40
318.15	30	301.85	302.55	0.23

Tab 2: Comparison among wet bulb temperatures as predicted by the proposed model and those ones obtained by the psychrometric chart

The proposed model can also describe the temperature and spatial moisture profiles in both air and food at any time. This is highly relevant from the food safety point of view, inasmuch as it allows detection of the regions within the food core where high values of moisture content might determine microbial spoilage. Fig. 10 shows the water concentration in the air, as a function of the radial distance from the food sample at different drying times.

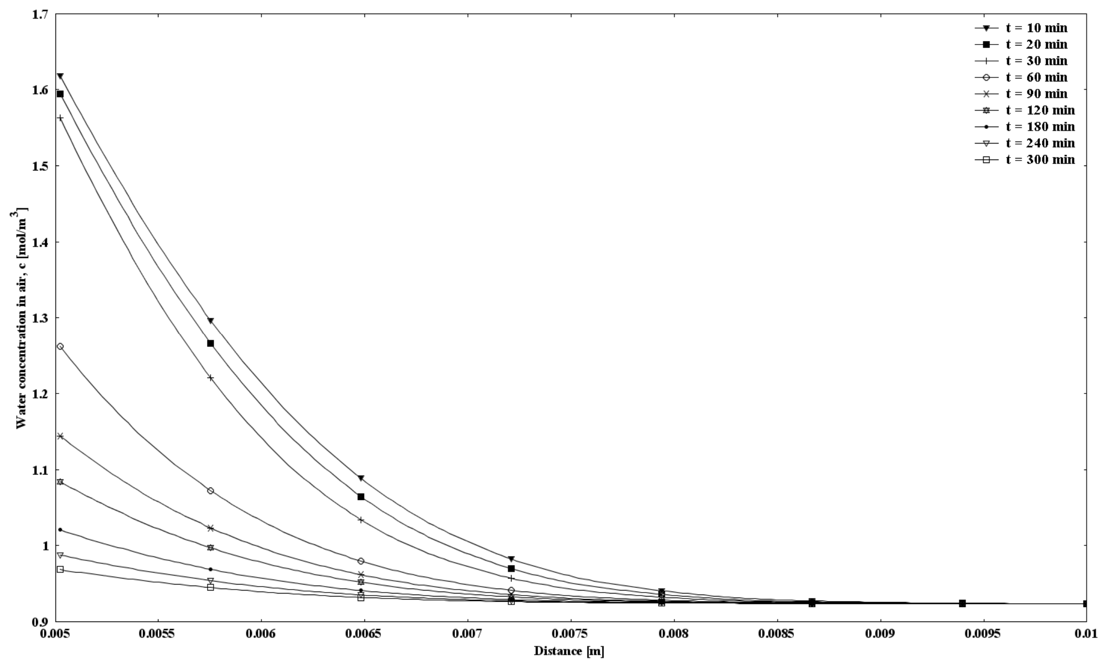

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Fig. 10: Detail of water concentration profiles developing in air in the radial direction at different drying times (Axial position: $z = 0.125$ m (half length of the drier), $T_0 = 289$ K, $U_0 = 0.85$ kg H_2O/kg_{wb}, $u_0 = 1.5$ m/s, $T_{air} = 318$ K, $U_r = 20\%$).

It can be observed that a concentration gradient develops in a very thin region close to the food surface as predicted, for instance, by the boundary layer theory (Schlichting, 1960). Fig. 10 confirms that the above boundary conditions referred to water concentration at boundary 2 (see Fig. 2) may be considered as a true representation of behaviour of a real system. Concentration gradients, in fact, vanish at a very short distance, equal to about 3 mm, from the food surface in comparison to the drier radius of 10 cm. It can be also noted that the external resistances to mass transfer tend to decrease as drying proceeds; this is further evidence that internal resistance to water transport actually controls the final stage of the process. Similar conclusions can also be drawn in respect of air temperature profiles (data not presented). Fig. 11 shows the moisture content profiles developing within the food sample in the radial direction at different times. It can be observed that the internal

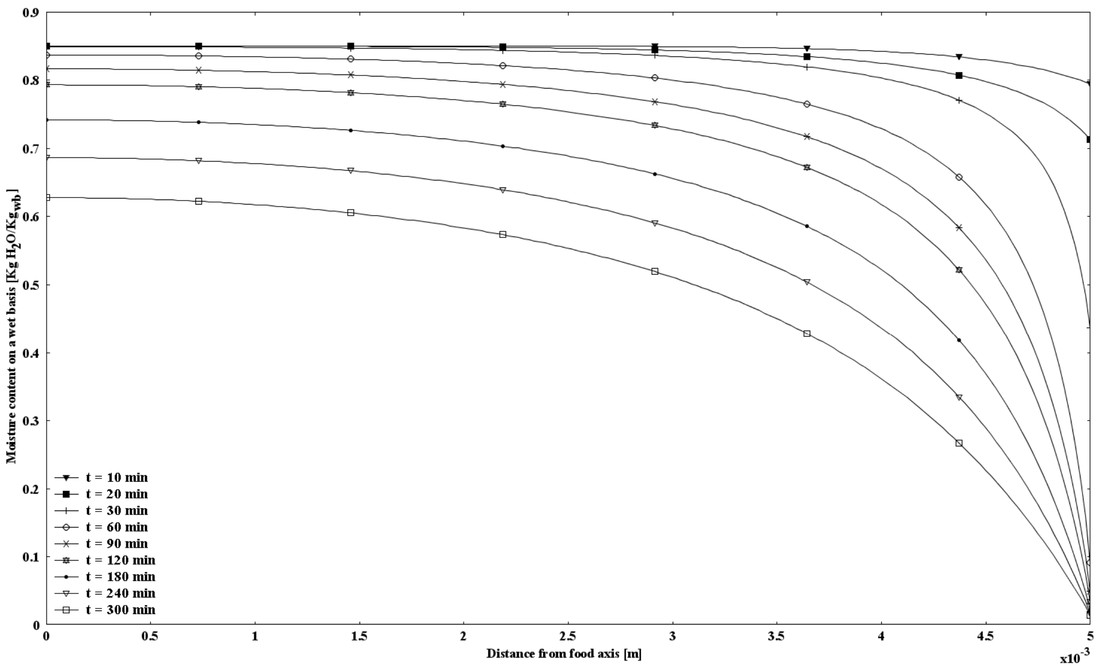

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Fig. 11: Moisture content profiles developing within the food in radial direction at different drying times (Axial position: z = 0.11 m (half length of the food), T_0 = 289 K, U_0 = 0.85 kg H_2O/kg_{wb}, u_0 = 1.5 m/s, T_{air} = 318 K, U_r = 20%).

The moisture content gradient developing in radial direction is, in fact, more pronounced as process time advances; this is also substantiated by the greater differences between water concentration at the sample core and that at the surface exposed to air.

Model Validation

In addition to the numerical simulations described in the previous section, other simulations were performed with the specific aim of verifying the model validity, by checking the agreement between experimental results, obtained under different fluid-dynamic and operating conditions and model predictions. Model validation has been carried out for the case of air-drying performed on cylindrical-shaped carrot samples. Fig. 12 shows the comparison between experimental and predicted average moisture content during two drying tests performed at the same value of air velocity (1 m/s) but changing inlet air temperature and its relative humidity. In the milder condition, the dry bulb temperature was maintained at 308 K and air relative humidity was equal to 24%; in the more severe case, the air temperature was 323 K and relative humidity was 16%.

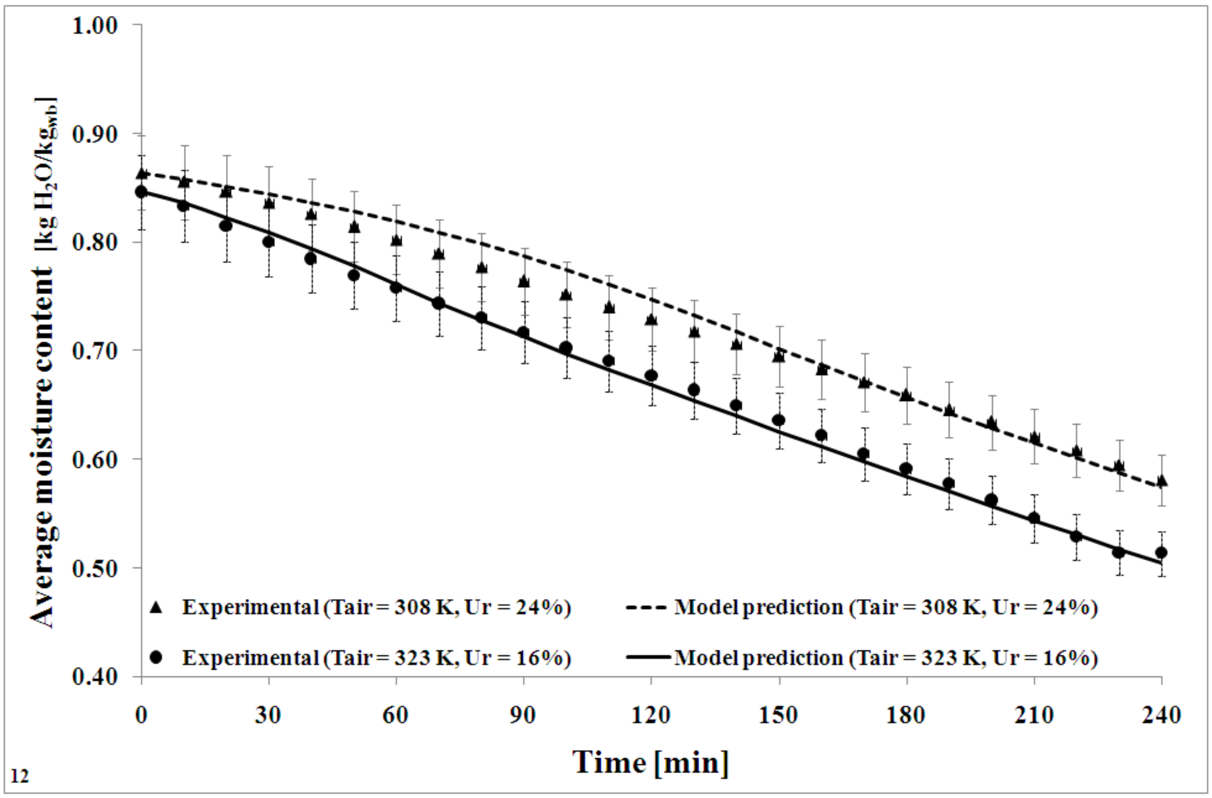

Fig. 12: Comparison between model predictions and experimental data - Time evolution of average moisture content on a wet basis (u_0= 1 m/s).

Experimental data and model predictions show remarkable agreement, with relative errors never exceeding 3% in the second case, and 1.7% in the second case. Fig 13 shows that the proposed model is also capable of giving very good predictions of the actual time evolution of food temperature. As mentioned earlier, the food temperature was experimentally measured by a capillary thermocouple inserted in a specific point located 1 cm away from the impact surface at a depth of 0.8 mm. In this case, the effect of air inlet velocity on drying behavior has been taken into account.

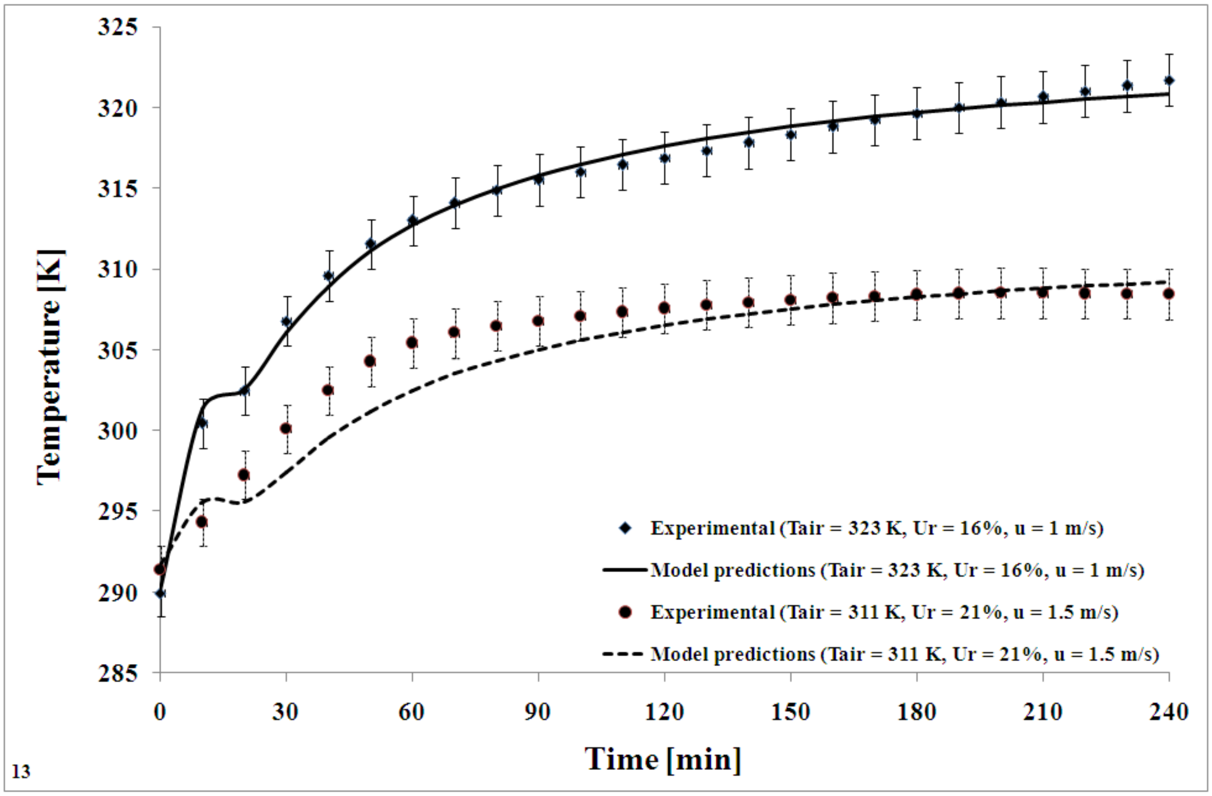

Fig. 13: Comparison between model predictions and experimental data - Time evolution of temperature 1 cm from the impact surface at a depth of 0.8 mm.

The agreement between experimental data and model predictions is rather good with relative errors never exceeding 1%. The proposed model therefore gives a good representation of the behavior of the real system, since it is capable of reproducing experimental data not only at the very beginning of drying process, but also over a the entire time duration.

Conclusions

The transport phenomena involved in food drying have been analyzed. A general predictive model, i.e. not based on any semi-empirical correlation for estimating heat and mass fluxes at food-air interface, has been formulated. The model, which also accounts for momentum transport of air flowing around the food sample, was capable of describing the real drier behavior over a wide range of process and fluid-dynamic conditions. The proposed model is particularly useful for those situations for which either semi-empirical correlations are not currently available (i.e. complex food geometries) or operating conditions are changed during drying process. Moreover, the model can also be used to determine which set of operating conditions would enhance the quality and the safety of the final product. It was observed that air properties (dry bulb temperature, relative humidity, inlet velocity) play a significant role only during the initial stage of the drying process, i.e. when external resistances to heat and mass transfer control the drying rate. On the other hand, when bound water is to removed from the food and internal resistances control the process rate, the operating variables have insignificant effect on the system behavior. The proposed model also predicted the spatial moisture profiles at all times, thus allowing detection of the regions within the food core, where high values of moisture content can promote microbial spoilage. Simulations results were found to be in good agreement with experimental data in terms of time evolutions of food moisture content and temperature. The models can be improved by taking into account phenomena such as product shrinkage and transport of water vapor, by diffusion, within the dehydrated material which can influence the drying rate and, therefore, the quality of final product.

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Modelling of transport phenomena and shrinkage during food drying process

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Abstract

The aim of the present work is the formulation of a theoretical model describing the transport phenomena involved in food drying process. The attention has been focused on the simultaneous transfer of momentum, heat and mass occurring in a convective drier where hot dry air flows, in turbulent conditions, about the food sample. Water loss produces changes in shape and dimensions of the food sample; shrinkage as well as the other transport phenomena occurring in both air and food domains, have been described formulating an exhaustive model aimed at drying process optimization. The system of non-linear unsteady-state partial differential equations modelling the process has been solved by means of Finite Elements Method coupled to the ALE (Arbitrary Lagrangian Eulerian) procedure that, by a proper modification of integration domain, accounts for shrinkage effects .

Keywords. Food drying, Transport phenomena, Finite Elements Method, Process Modelling, Shrinkage

Introduction

Among the different methods for food preservation currently used, drying is one of the most ancient. Drying by dry and hot air flowing about food samples is, definitely, the most common industrial method to promote water transfer from food, thus reducing its moisture content.

The decrease of water content in food reduce both microbial spoilage and deterioration reactions since water is essential for the metabolism of several bacteria and microorganisms

When dry and warm air flows about a moist and cold food sample, a simultaneous transfer of heat, water and momentum occurs. Moreover a volume variation of food occurs due to the water leaving the food; this phenomenon is known as shrinkage.

Many authors modelled food drying process using different assumptions (Zogzas, Maroulis, Marinos-Kouris 1994, Lozano, Rotstein & Urbicain, 1983, Suzuki, Kubota, Hasegawa, & Hosaka, 1976). In particular, shrinkage is often considered negligible, although a more detailed characterization can improve the control of the quality parameters and re-hydration capacity of the final product (Krokida and Maroulis, 2000). Both fundamental and empirical models have been used to fit experimental results in terms of food volume variations versus moisture content: Mayor and Sereno (2004) summarized the results obtained by different authors. Results obtained by experimental data fitting were used by Chalida Niamnuy, et al (2008) for mathematical modelling of drying process assuming strain displacement as proportional to shrinkage coefficient in food domain. The authors, however, did not consider the transport phenomena in air domain; moreover, heat and mass transfer coefficients, necessary to express the boundary conditions, were taken from the available literature correlations.

Some authors assumed drying as an isothermal process modelling shrinkage and mass transfer in food domain only (Fusco, Avanza, Aguerre, Gabitto, 1991); other authors accounted for heat transfer too (Hui Yang, Noboru Sakai, Manabu Watanabe, 2001, Balaban, and Pigott, 1988). The equations describing the transport phenomena involved in food drying process form a system of nonlinear Partial Differential Equations (PDEs) which can be solved only by a numerical method (Wang, and Sun 2003), as – for example – the finite element method (FEM). It is based on the discretization of food domain into finite elements, interconnected through a finite number of points where problem variables are evaluated.

The aim of the present work is the formulation of a theoretical model describing the transport phenomena involved in food drying process, accounting also for time-dependent volume variation of the sample. The attention has been focused on the simultaneous transfer of heat and water (for both food and air) coupled to momentum transport (in air only) occurring in a convective drier where warm and dry air flows, in turbulent conditions, about the food sample whose shape and dimension change due to shrinkage. The system of non-linear unsteady-state partial differential equations modelling the process has been solved by means of Finite Elements Method coupled to the so-called Arbitrary Lagrangian Eulerian procedure implemented to account for the variation of air and food domains. The proposed model does not need the specification of any heat and mass transfer coefficients at food-air interfaces; for this reason, the present approach can be considered as a general one, suitable for any drying process the may require a preliminary analysis or optimization.

Theoretical and Mathematical Modelling

In a previous paper, the authors of the present work modelled drying process accounting for simultaneous transport of heat, water and momentum (Curcio, Aversa, Calabrò, Iorio, 2008). Diffusion and conduction were considered as the only transport mechanisms occurring in food domain, whereas convective contributions, coupled to momentum transport, were accounted for in air. Food shrinkage was considered negligible assuming that, under mild conditions, volume variations are actually not so relevant. The resulting system of equations was solved by FEM using a commercial package, i.e. Comsol Multiphysics ®.

In the present work also shrinkage has been physically and mathematically modelled. Mathematical modelling of shrinkage has been carried out by an Arbitrary Lagrangian Eulerian (ALE) method that properly accounts for time-dependent domains variation and, therefore, for the so-called moving mesh. The ALE procedure, implemented within the finite element method framework, allows solving all the transport equations at each time step in two deformed domains that consider the actual food shape (and its dimension) and, also, the corresponding change of air domain (Donea, Huerta, Ponthot and Rodríguez-Ferran, 2004). The proposed ALE procedure is based on both the Eulerian and Lagrangian approaches and, therefore, requires the definition of a proper relationship, having a physical foundation, expressing how mesh is displaced. In the present work, it was assumed that the drying shrinkage strain rate vector is proportional, through a constant to be determined, to the local water content variation, i.e. to the water mass flux leaving the food sample. Actually, the drying shrinkage strain rate vector corresponds to the mesh velocity (as defined within Comsol Multiphysics® environment), evaluated locally at the food-air interfaces, thus providing a relationship between mesh displacement and the water flux transported in the air. It should be remarked that this assumption is valid in a very wide range of food moisture content, as it will be also shown in the second part of this thesis, where an experimental analysis of drying process will be presented. The plots of volume variation during drying (shrinkage) vs. weight loss (food moisture content) do indeed show, in most cases, a linear relationship that validates the present assumption.

As far as the proportionality constant is concerned, it was determined, through an iterative scheme, by a global mass balance on the water contained in the food and assuming that food volume variation can be easily related to the amount of water actually evaporated (Janjai, Lamlert, Intawee, Mahayothee, Haewsungcharern, Bala, Müller , 2008). This assumption is valid for several fruits and vegetables and throughout almost all the drying process, at least when food structure presents high mobility (Mayor and Sereno, 2004).

Results and Discussion

In the following, the results obtained in some typical cases will be presented showing the effect of shrinkage on cylindrical carrot samples. Figs. 1 and 2 show water content and temperature profiles developing in the food whose shape and dimensions change with respect to time and as a function of the actual value of local water flux. Figs. 1 and 2 show that food surface where air impinges is characterized by a higher volume variation.

The proposed model can be also used to compare how operating conditions affect food volume variations, as shown in Fig. 3 and 4. Volume variation rate increases with air temperature and with air velocity.

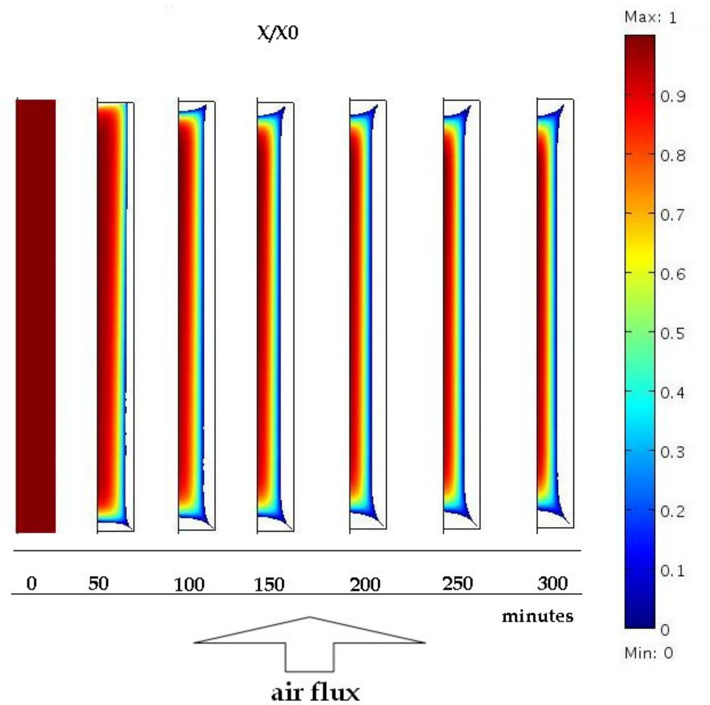

Fig. 1: Shrinkage and water content profiles in a cylindrical carrot sample (initial diameter 1 cm an initial length 6 cm) during drying by air having the following characteristics: velocity=1.5 m/s, relative humidity = 20%, temperature = 323K.
X is the moisture content on dry base (Kg water/Kg dry solid), X_0 is its initial value.

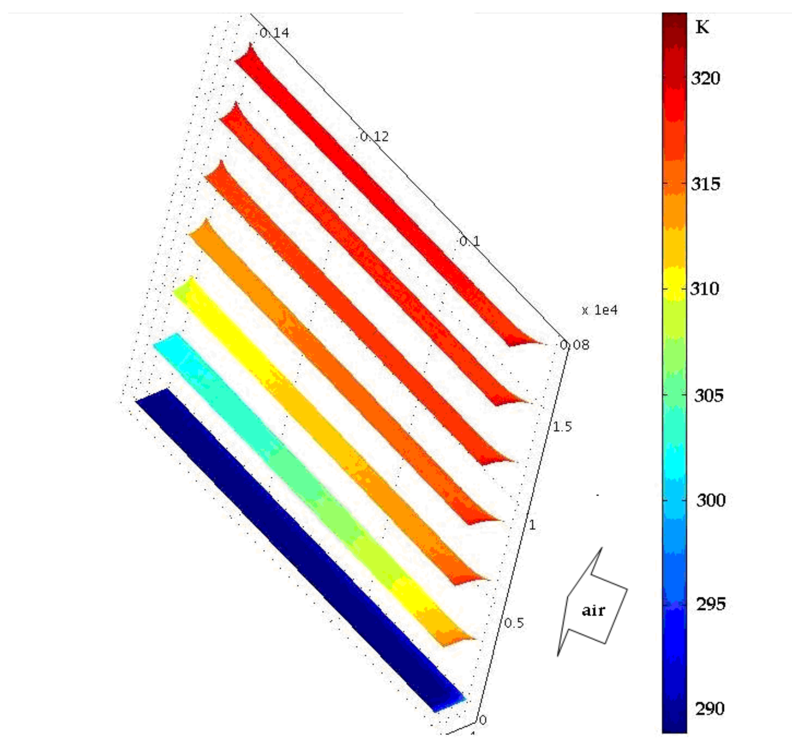

Fig. 2: Shrinkage and temperature profiles in a cylindrical carrot sample (initial diameter 1 cm an initial length 6 cm) during drying using air having the following characteristics: velocity=1.5 m/s, relative humidity = 20%, temperature = 323 K.

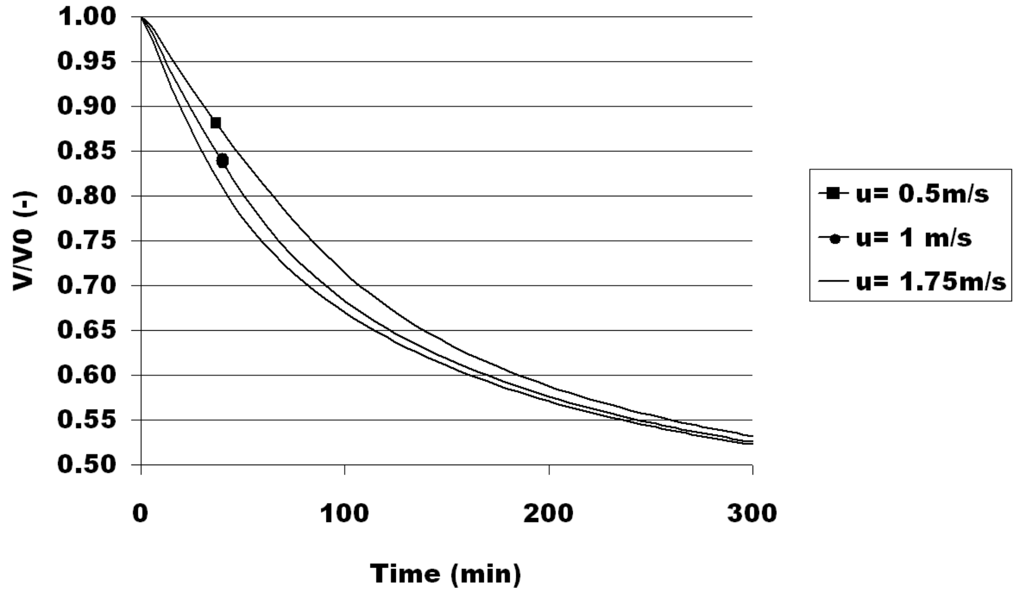

Fig. 3: Shrinkage of a cylindrical carrot sample (initial diameter 1 cm an initial length 6 cm) during drying by air having the following characteristics: relative humidity = 20%, temperature = 323 K. V is the food total volume (m^3), V_0 is its initial value, u is air velocity.

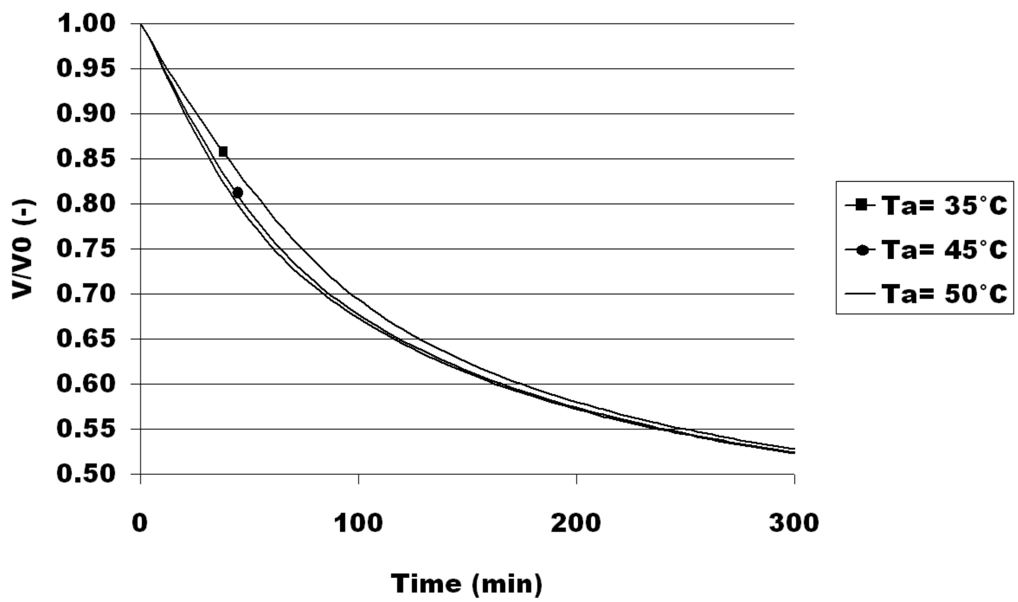

Fig. 4: Shrinkage of a cylindrical carrot sample (initial diameter 1 cm an initial length 6 cm) during drying by air having the following characteristics: air velocity= 1.5 m/s, relative humidity = 20%. V is the food total volume (m^3), V_0 is its initial value, Ta is air temperature.

The above preliminary results demonstrate the potentiality of the present model that may be capable to predict the time evolution of food shape and dimensions so providing helpful information that can be used to asses and to predict food quality as a function of operating conditions chosen to perform drying process.

Conclusions

In the present work a physical and mathematical modelling of shrinkage occurring during drying process has been performed. The obtained results showed the importance of accounting for shrinkage phenomenon that cannot be neglected in many cases of actual interest. It is intended, however, to couple the present model to a proper set of additional equations that may allow evaluating the actual deformations occurring within the food subjected to drying.

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Part B

Food properties

Introduction - Part B

In the second part of the present work, structured in two papers, some of the most important food properties (related to drying modelling) have been estimated. In this case, the experimental approach has been preferred because the food matrix during the drying process undergoes to chemical-physical and structural changes that are difficult to predict and characterize in theory.

In the first paper of part B an experimental method, for the estimation of some food properties relevant in the mass transfer during drying process, has been developed. The work shows an experimental procedure that allows the contemporary estimation of the effective diffusion coefficient of water in food, the shrinkage and the bulk density during drying. The approach has been focused on the research of a versatile, economical and quick estimation method. It should be, however, remarked that, in the literature, a standard method for the evaluation of food properties does not exist, due to the extremely complex structure involved.

In the second paper of part B some important quality parameters have been monitored, during drying of carrot samples, using different operating conditions. The quality parameters used were shrinkage, re-hydration capacity and colour changes. The obtained results showed that milder conditions, during drying, allow a better product quality but longer processing time. This means that economic optimization does not agree with optimization of food quality parameters, however, they both should be considered. The above-mentioned study has been carried out at TOP b.v., a commercial innovation company that provides services for the Food Industry and its technology suppliers, located in Wageningen (The Netherlands), where the author has spent three months as guest scientific researcher.

Introduzione Parte B

La seconda parte del presente lavoro di tesi, si compone di due articoli. In essi è stata realizzata la determinazione sperimentale di alcune proprietà dell'alimento, fondamentali durante il processo di essiccazione e per la sua corretta modellazione. In questa seconda parte del lavoro è stato scelto l'approccio sperimentale in quanto l'alimento subisce, durante l'essiccazione, notevoli cambiamenti fisici, chimici e strutturali i quali sono assolutamente poco prevedibili e caratterizzabili a priori. Nel primo articolo della parte B del presente lavoro è stata sviluppata una tecnica sperimentale per la determinazione di alcune proprietà dell'alimento fondamentali nella modellazione del trasporto di materia durante l'essiccazione. Tali proprietà sono, in particolare, il coefficiente di diffusione dell'acqua nelle matrici vegetali, la densità di *bulk* e lo *shrinkage* durante l'essiccazione. L'attenzione è stata focalizzata sulla ricerca di una metodologia di misura che fosse soprattutto versatile, economica e relativamente rapida. In letteratura non esiste una metodologia standard per la valutazione delle proprietà degli alimenti proprio a causa della loro struttura particolarmente complessa.

Nel secondo articolo della seconda parte (Parte B), è stata realizzato un monitoraggio sperimentale di alcuni importanti indici della qualità dell'alimento durante l'essiccazione. Tali proprietà sono state lo *shrinkage*, la capacità di reidratazione ed il colore di carote sottoposte ad essiccazione in diverse condizioni. I risultati ottenuti dimostrano che l'ottimizzazione di un processo dell'industria agroalimentare non può e non deve limitarsi ad una ottimizzazione di tipo economico ma deve prevedere anche il controllo della qualità finale dell'alimento, anche se spesso queste due esigenze vanno in conflitto. Lo studio suddetto è stato condotto presso i laboratori della TOP b.v., una società privata di ricerca e sviluppo operante nel settore agro-alimentare sita a Wageningen (Paesi Bassi), dove l'autore del presente lavoro ha trascorso tre mesi come ricercatore ospite.

An estimation procedure of water diffusion coefficient, bulk density and shrinkage during drying of food samples

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Abstract.

The aim of this work is to propose an innovative experimental method that, combined to a classical mathematical model describing the transient, mono-dimensional transport of water in food samples, allows estimating the effective diffusion coefficient of water in food. Sample volume variation with respect to its moisture content has also been described in order to characterize eggplants shrinkage. Finally, an estimation of eggplants bulk density and its dependence on sample moisture content has been performed. The present experimental technique, based on the utilization of a Halogen Moisture Analyzer, has several advantages with respect to the traditional methods usually reported in the literature. It is, indeed, very cheap, it does not require high energy consumptions and, in principle, can be applicable to different types of foods, thus providing reliable estimations of some of the most important parameters involved in drying process.

Keywords. Effective diffusion coefficient, shrinkage, density, eggplants, drying.

Introduction

Three main stages can be distinguished in food drying process (Üretir , Özilgen & Katnaş, 1996). The drying rate is initially controlled by external mass transfer and attains its maximum value. Afterwards, the process rate is controlled by internal mass transfer and a progressive decrease of drying curve slope is observed. Drying rate, eventually, approaches a nearly zero value; during this stage, evaporation front moves from food outer surfaces towards its core and the resistance due to internal mass transfer increases since water is to be transported through a progressively thicker layer of dry matter. The above three periods are called the constant rate, the first falling rate and the second falling rate period, respectively.

As drying goes by, the initial mechanical equilibrium, which determines the food size and its shape, may fail due to the water transport from the solid matrix. A field of contracting stresses develops, thus promoting the attainment of a new equilibrium situation that leads to a change in food shape and size. This phenomenon is known as shrinkage (Mayor & Sereno, 2004, del Valle, Cuadros, Aguilera 1998).

The modification of food structure causes a time-dependent variation of physical, chemical and transport properties (Datta, 2007 II, Khraisheh, McMinn & Magee, 2004) that has to be properly accounted for especially on formulating transport models aimed at the prediction of industrial driers performance. In the literature, a large number of papers dealing with the experimental evaluation of food properties is found and great efforts have been made to estimate diffusion coefficient of water within a solid matrix (Saravacos & Maroulis, 2001). Several methods have been actually proposed and applied to regular-shape foods undergoing a transformation under well-defined experimental conditions (steady or transient state).

The permeability method (Boudhrioua, Bona & Daudin, 2003) makes use of a thin layer of food, placed between two environments maintained at constant and uniform water concentration and temperature. A constant, linear concentration gradient develops in the sample and diffusion coefficient (or permeability) can be estimated by means of the first Fick's law. The concentration–distance curve method (Saravacos et al. 2001) is based on the measurement, at a definite time t, of moisture concentration profile developing within a sample for which water transport occurs by diffusion only, along a single direction. Solute concentration profile is actually determined by slicing and weighing the sample, even though some experimental problems, essentially related to slicing itself, are reported (Boudhririoua et al., 2003). The so-called sorption/desorption method (Saravacos et al. 2001), as well as the drying method, is based on the utilization of a regular-shape sample undergoing processing under constant and controlled conditions; the sample weight is monitored at regular times in order to evaluate the actual moisture loss (or the weight increase, in the case of sorption) until the sample mass remains unchanged. Actually, only the period in which drying rate is controlled by water transport within the material is taken in consideration for diffusion coefficient estimation.

None of the above-described procedures can be considered, however, as a standard method. The values of diffusion coefficient available in the literature do, indeed, show a large variability (Panagiotou, Krokida, Maroulis & Saravacos, 2004), essentially due to the time evolution of food structure and composition that may change even if two samples having similar initial characteristics are dried under the same conditions. Saravacos et al. (2001) published a detailed review collecting the experimental results of water diffusion coefficient in food as estimated by different authors. Some types of food were deeply analyzed, some other ones, as in the case of eggplants, lack of useful data and require a proper characterization.

Shrinkage and food density are also essential on describing water transport in food matter. The former defines the shape of the domain in which water is transported; the latter is necessary to evaluate the actual value of water concentration. Several factors affect shrinkage, whose extent depends on the mobility of the biopolymeric matrix constituting the food, on its water content and on its temperature, but also on the drying conditions (Mayor et al., 2004). As far as density is concerned, it should be remarked that two different definitions are usually used: the particle (or true) density and the bulk (or apparent) density. The former is the ratio between the total mass and the sum of the volume occupied by both dry solid and water; the latter is the ratio between the total mass and the total volume, i.e. the sum of the volumes occupied by dry solid, water and air. (Krokida, Oreopoulou & Maroulis, 2000, Maroulis, Krokida & Rahman, 2002, Krokida & Maroulis 1997). Both shrinkage and bulk density are usually estimated by monitoring food sample dimensions and its water content; the collected experimental data are eventually interpolated so to provide a set of functions correlating each measured variables to sample moisture content.

The aim of the present work is to propose an innovative experimental method that, combined to a classical mathematical model describing the transient, mono-dimensional transport of water in eggplants slices, allows estimating the effective diffusion coefficient of water in food. The food sample volume change with respect to its moisture content has been described in order to characterize eggplants shrinkage. Finally, an estimation of eggplants bulk density and its dependence on sample moisture content has been performed.

Theory

Simplified approaches have been proposed to model food drying processes. One of the most common assumptions considers food as a hygroscopic and a capillary porous material for which inner evaporation is so weak that can be neglected. The vapor flow through the porous medium, usually modeled by the Darcy's equation, is, therefore, ignored.

Water, instead, is transported as both a liquid and a vapor by capillary flow and by molecular diffusion, respectively. For the sake of simplicity, it is commonly assumed (Panagiotou et al. 2004, Saravacos et al., 2001) that this combination of different transport mechanisms can be expressed referring to a unique transport property known as effective diffusion coefficient, D_{eff}. For a significantly wet material the prevailing transport mechanism is capillary diffusion of liquid water so that D_{eff} can be considered as a capillary diffusion-like coefficient, whereas, for a very dry material molecular diffusion of water, as a vapor, prevails over capillary flow of liquid and effective diffusion coefficient can be identified as a molecular diffusion coefficient (Datta, 2007 -I).

Un-steady state, mono-dimensional water transport in food can be therefore modeled by a Fick's-like diffusion equation, leading to the following partial differential equation:

$$\frac{\partial C}{\partial t} = \frac{\partial}{\partial z}\left(D_{eff} \frac{\partial C}{\partial z} \right) \qquad 1$$

where C is the water concentration in food, t is the time and z is the spatial coordinate along which water is transported.

An analytic solution of Eq.1 leading to the time evolution of water concentration profile, C(z, t), can be found as infinite series that converge rapidly for large values of a characteristic dimensionless time (Carslaw & Jaeger, 1959). Water concentration profile is then averaged over the whole sample thickness, thus obtaining an expression of the variable <C(t)> that can be also experimentally measured by monitoring the decrease in food weight with respect to time. The experimental value of <C(t)> is equated to the analytical expression of space average concentration profile so to obtain a single non-linear equation whose solution gives a value of effective diffusion coefficient of water, $D_{eff,i}$, at a definite time, t. This procedure has to be applied to all the experimental values of <C(t)>, measured when drying rate is controlled by internal mass transfer, thus providing a set of different $D_{eff,i}$. The actual estimation of effective diffusion coefficient is, finally, achieved calculating the average among all the $D_{eff,i}$ values.

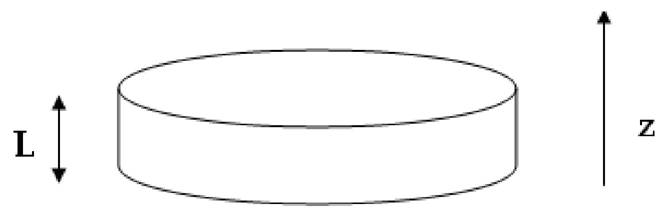

Fig. 1: Shape of the food submitted to drying

In the present case, Eq. 1 has been referred to a slice (Fig. 1), under the following general assumptions:

- The process rate is controlled by internal mass transfer and external resistance is negligible;
- The system is isothermal;
- The slice diameter is much larger than its thickness, so that water transfer prevailingly occurs in z direction;
- D_{eff} is a function of temperature only;
- Moisture distribution is initially uniform;
- The food slice is placed on a metal plate through which no water transport is allowed to occur.

The following boundary conditions apply:

No water transport through the lower surface (z = 0)

$$D_{eff} \left. \frac{\partial C}{\partial z} \right|_{z=0} = 0 \qquad 2$$

An equilibrium relationship between water concentration in the air and water concentration holds on the upper food surface at z = L

$$C_{z=L} = \rho_m y_{eq} \qquad 3$$

where ρ_m is the air molar density; y_{eq} (-) is the molar fraction of water, as a vapor, in air and is evaluated in terms of the vapor pressure of water, the activity of water on the food surface and the pressure in the drying chamber (Curcio, Aversa, Calabrò , Iorio, 2008). As it will be described in the following, drying has been performed by a halogen lamp that allows attaining very drastic drying conditions so that the activity of water on food surface tends rapidly to zero and $C_{z=L} = 0$.

On the basis of the above discussion, the water concentration profile has been determined and expressed according to the following general form:

$$C(t,z) = \sum_{n=0}^{\infty} K_{2n} \cos\left[\frac{\pi(1+2n)z}{2L(t)}\right] \exp\left[-\frac{\pi^2(1+2n)^2 D_{eff}\, t}{4L^2(t)}\right] \qquad 4$$

where K_2 is a generic constant, and L(t) is the sample thickness, accounting for its variation due to shrinkage effect. It should be remarked that a set of experimental information is necessary to estimate diffusion coefficient of water. First of all, a proper set of drying curves is needed to precisely specify when internal resistance is the limiting step of drying process so as to validate one of the hypotheses of the present model. The drying curves are also necessary to evaluate the average water concentration, <C(t)>, in the sample. Moreover, a measurement of the actual food thickness has to be performed to estimate the time evolution of L(t).

Materials and Methods

A Halogen Moisture Analyzer, HB43 Mettler Toledo, has been used to obtain the drying curves. The HB43 instrument consists of a small drying chamber placed on a precision balance that allows monitoring sample weight loss during drying process (Fig. 2). The heat necessary for water evaporation is provided by a halogen lamp; the drying temperature is set before drying takes place and maintained throughout all the process.

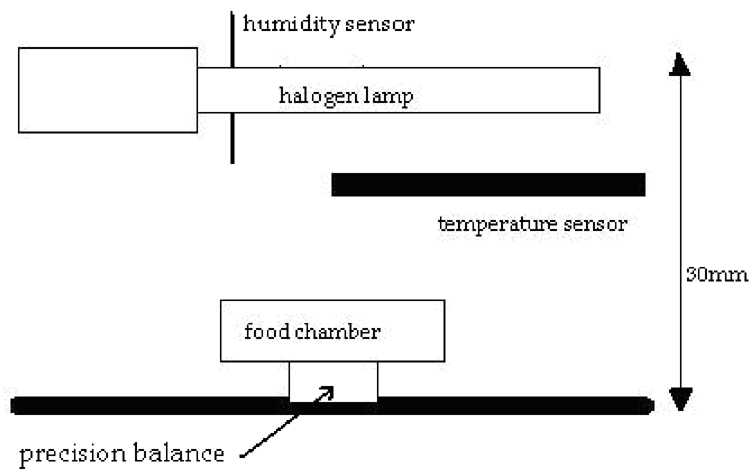

Fig. 2: Schematic of Halogen Moisture Analyzer.

Eggplants, purchased in local market, have been preliminarily sliced so to obtain a regular disk having an initial diameter of 8±0.01 cm. Three different thicknesses, 6±0.1, 8±0.1, 10±0.1 mm, have been tested to analyze the effect of this variable on process behavior. The above dimensions were measured by a vernier caliper providing a precision of 0.02 mm. The initial moisture content varied between 0.92 and 0.94 Kg water / Kg wet food.

Each slice has been dried in the Moisture Analyzer at fixed operating conditions, continuously monitored by a digital thermo-hygrometer whose sensors have been placed within the drying chamber (Fig. 2). For each value of eggplant thickness, three different temperatures, i.e. 70, 80 and

90 °C, have been chosen to perform the experiments. Each experiment has been repeated twice in order to ascertain the reproducibility of the experimental drying curves; the average value and the correspondent standard deviation have been calculated and henceforth used. During drying, the food sample main dimensions have been also determined: it has been periodically removed from the drying chamber and its thickness measured, at three different locations, by a vernier caliper. Moreover, the slice surface has been replicated on a paper sheet of know dimensions: the eggplant sample surface was then calculated by weighting the paper sheet and the eggplant image reported on it**.** All the above operations have been performed in less than 30 seconds so to avoid a significant food re-humidification. At regular time intervals, it was therefore possible to measure the sample weight, its thickness, to estimate its surface and, consequently, the volume. The ratio between the weight and the volume has been used to evaluate bulk density at each time. All the experimental points have been used to express volume versus water content so as to model eggplant shrinkage by a fitting procedure.

Before using the experimental data to evaluate water diffusion coefficient, it is necessary to verify if the assumptions on which the proposed mathematical model is based are correct. Inside the Moisture Analyzer, the food sample is placed on the precision balance plate through which no mass transfer can occur. Water transport is, instead, allowed on the upper surface; it should be remarked that this surface is exposed to the halogen lamp and gets dry quite soon, thus determining a water activity value rapidly tending to zero.

Moreover, it is necessary to adopt a procedure that allows identifying the actual time when internal resistance to mass transfer starts being the controlling step for the drying rate. Both the constant rate period and the second falling rate period of each drying curve, have been excluded by means of a graphical method shown, in a typical case, in Fig 3.

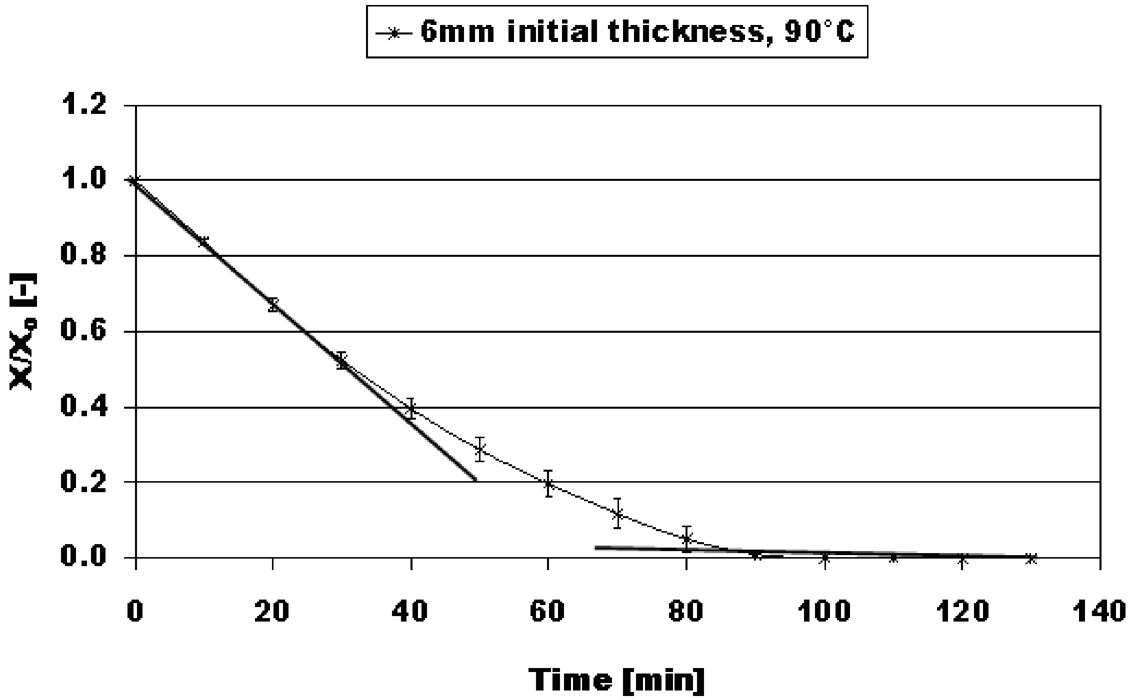

Fig 3: Graphical method for the identification of the experimental points belonging to the falling rate period (initial thickness: 6mm, temperature 90°C).

The constant rate period is characterized by a set of experimental points laying on a straight line. A similar behaviour is observed for the points belonging to the second falling rate period that have been excluded since, when evaporation front takes place, dry and wet matter coexist in the food and the assumption of isotropic material becomes hardly applicable. Only the experimental points belonging to the intermediate region (first falling rate period) have been henceforth used to estimate water diffusion coefficient.

Results and Discussion

Figs. 4-6 show the drying curves obtained at 70, 80 and 90 °C for an initial thickness equal to 6, 8 and 10 mm, respectively. X and X_0 are the actual and initial water contents on a dry basis (Kg water/Kg dry matter), respectively. All the experimental data exhibit low standard deviations; moreover, the drying curves show rather regular trends that allow identifying the falling rate period and, to estimate the effective diffusion coefficient of water in eggplant, according to the above described procedure.

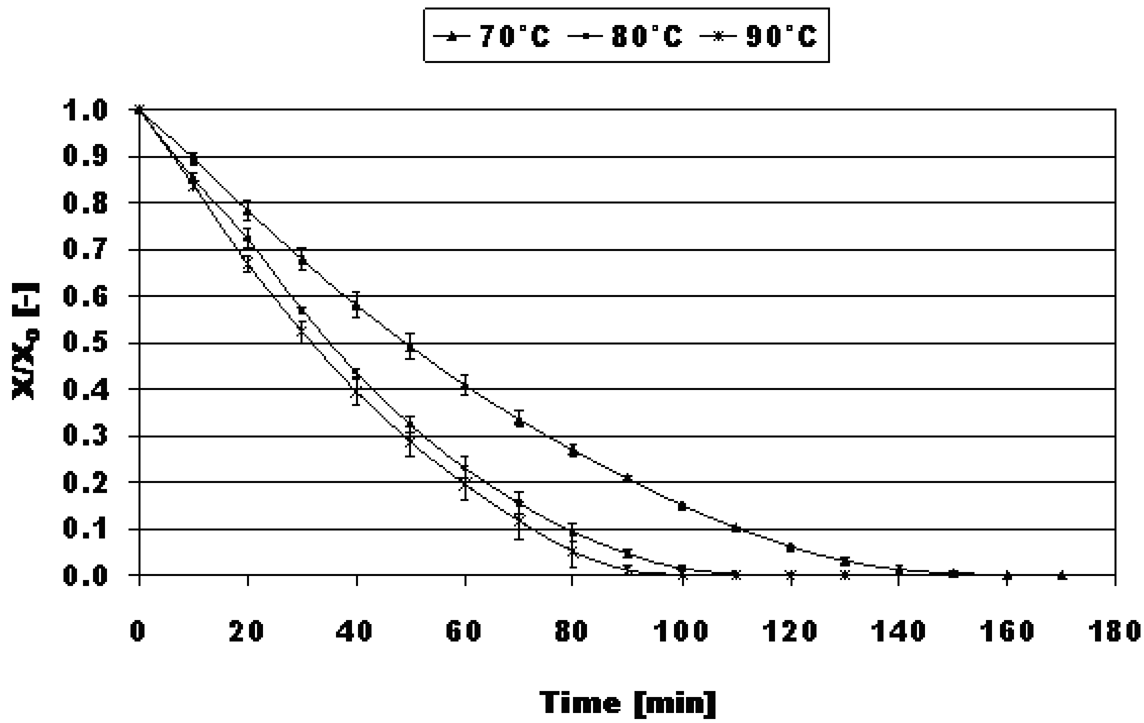

Fig. 4: Eggplant drying curves as a function of Temperature (initial thickness: 6 mm).

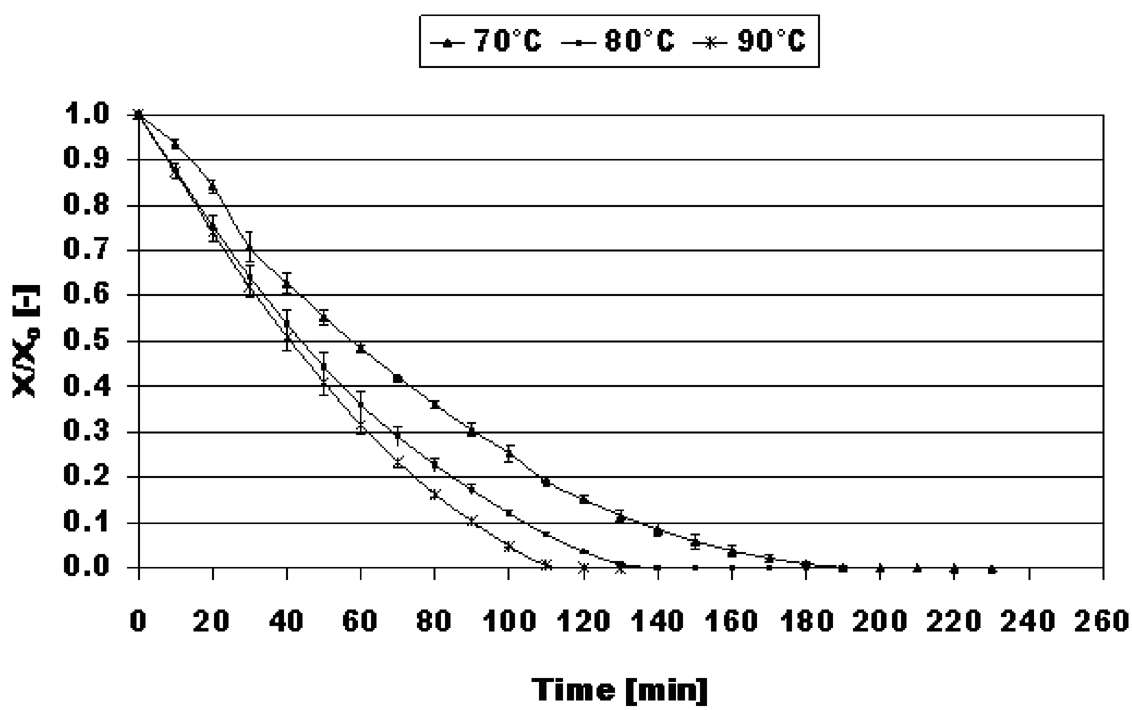

Fig. 5: Eggplant drying curves as a function of Temperature (initial thickness: 8 mm).

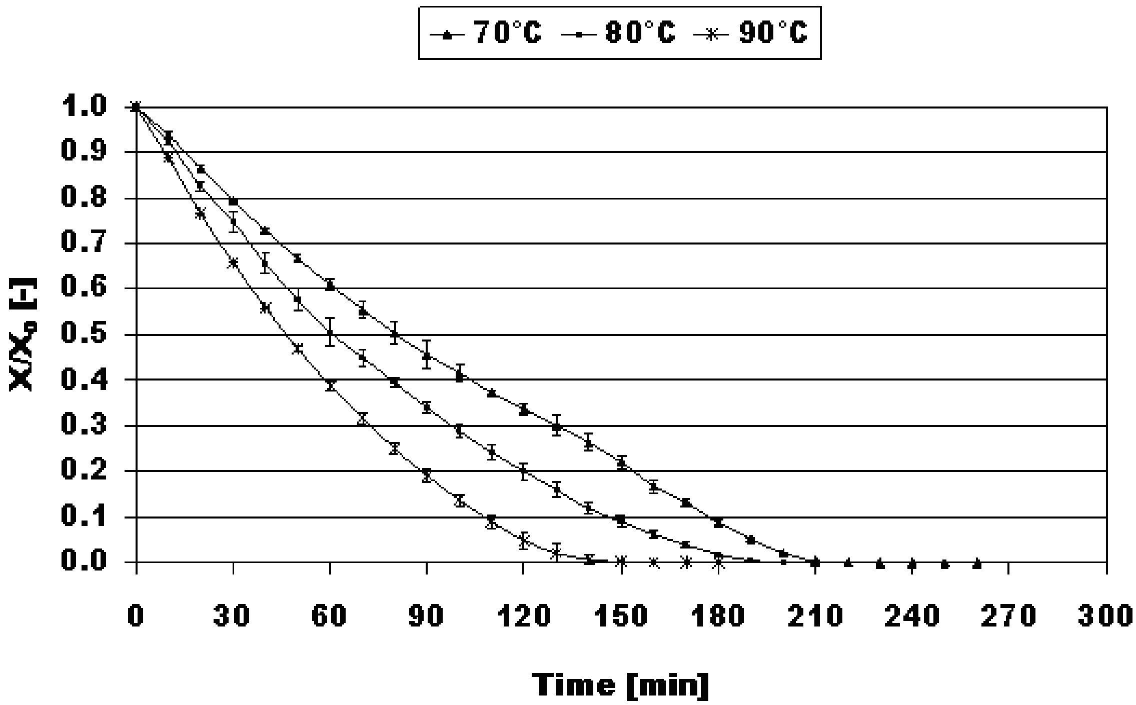

Fig 6: Eggplant drying curves as a function of Temperature (initial thickness: 10 mm).

Tab. 1 summarizes the average values of D_{eff}, as estimated in all the tested conditions. In the case of an initial thickness equal to 8 and 10 mm a monotonous increase in D_{eff} with temperature is observed, thus confirming that temperature improves both molecular and capillary diffusion (Bird, Stewart & Lightfoot, 1960).

Initial thickness [mm]	Temperature[°C]	Diffusion coefficient [m^2/s]
6	70	1.67E-10
	80	1.55E-10
	90	2.92E-10
8	70	1.36E-10
	80	4.15E-10
	90	5.65E-10
10	70	1.13E-10
	80	1.78E-10
	90	2.92E-10

Tab. 1: Effective diffusion coefficients of water in eggplants at different temperature and for different values of initial thickness.

The results of D_{eff} obtained for the same temperature and a different initial thickness show that a precise relationship between D_{eff} and L cannot be identified. This could be ascribed to two opposing effects. Some authors, indeed, found that in a thicker sample the formation of channels and pores, actually representing a preferential path for water transport, is facilitated; this leads to an overall increase of water diffusion coefficient (Saravacos et al. 2001). Conversely, it should be remarked that, generally, lower resistance to mass transfer are observed when water transport occurs along a shorter path.

As far as shrinkage characterization is concerned, the experimental results obtained in a typical case are shown in Fig. 7, where the ratio between the actual and the initial volume (V/V_0) is plotted (for both the experiments performed at the same temperature) versus X/X_0.

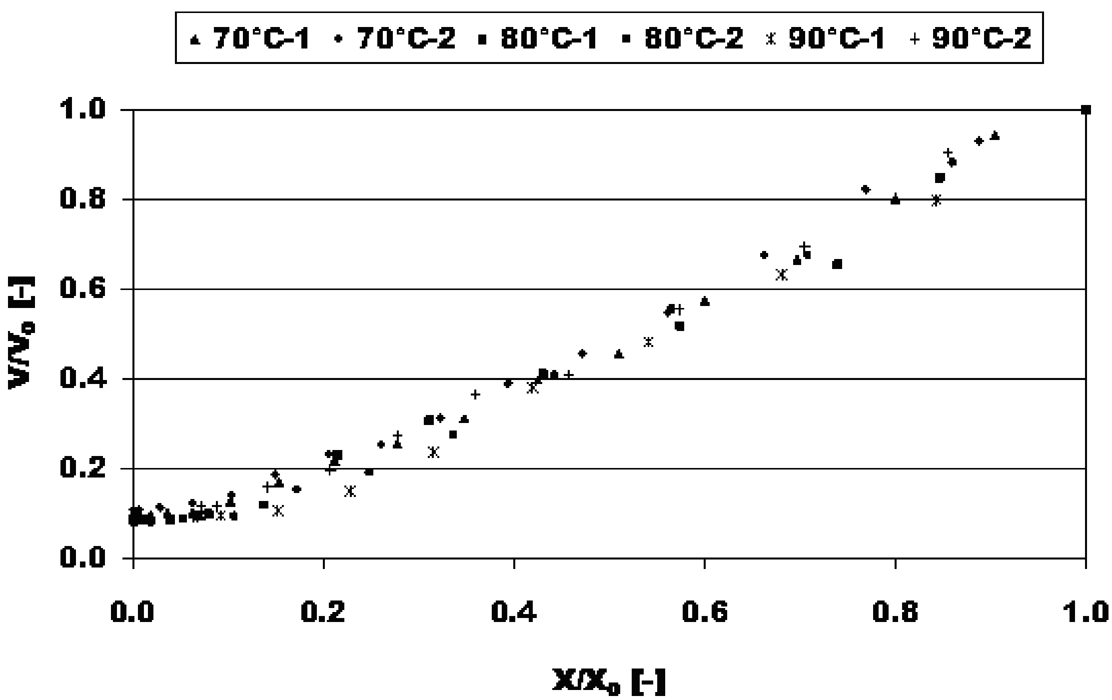

Fig. 7: Volume variation vs. eggplant moisture as a function of drying temperature (initial thickness: 6 mm; 1 and 2 indicate the first and the second experiment performed at the same temperature).

A considerable decrease of food volume is observed during both the constant and the falling rate period (X/X_0 between 1 and 0.1); on the contrary, in the final part of the process (X/X_0 lower than 0.1) the increased rigidity of the solid matrix does not allow noticing remarkable variations in food sample external dimensions as it was also reported by Mayor et al. (2004). Moreover, although different drying temperatures have been tested, no remarkable difference is observed so that it can be assumed that operating temperature does not significantly affect eggplant shrinkage, as it was also discussed by Zogzas, Maroulis and Marinos-Kouris (1994). As reported in the literature, shrinkage is modeled by fitting the experimental data of volume variation versus food moisture content. Linear models are usually suitable to describe materials and process conditions leading to a linear decrease of volume in the whole range of humidity; this phenomenon can be, generally, ascribed to the development of negligible or uniform porosity during drying process. Linearity, however, does not fit the actual behaviour when pores formation increases sharply, so especially during the final stage of drying; in these cases, the time evolution of V/V_0 versus X/X_0 is better described by more complicated models, for instance exponential or quadratic as suggested by Mayor et al. (2004) and by Zogzas et al. (1994). The experimental data of V/V_0 versus X/X_0, collected for the three food thicknesses tested in the present study, have been fitted by two models: a linear one, and a slight modification of a linear model, namely a power law with a third parameter, c, as the exponent of X/X_0. Fig. 8 shows, in the case of an eggplant slice with an initial thickness of 8 mm, a comparison between both the tested models.

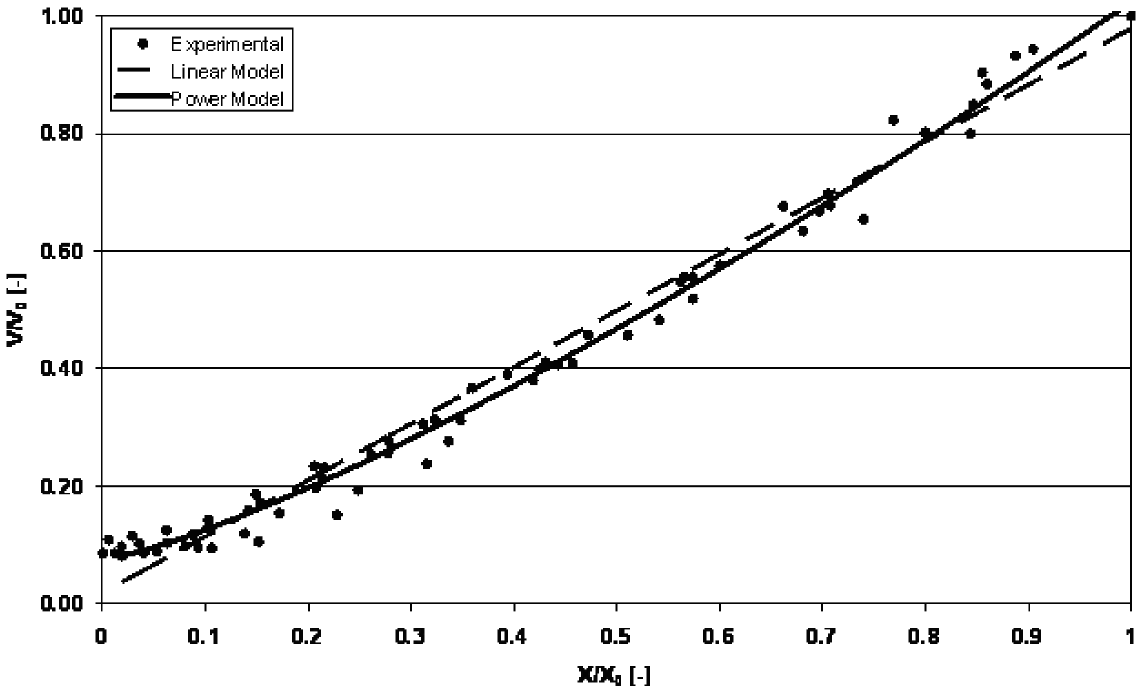

Fig. 8: Fitting of volume variation by both the proposed models (initial thickness 6 mm, temperature ranging between 70 and 90 °C).

As it is also reported in Tab. 2, the power model gives better performance in terms of statistical significance and should be preferred to describe shrinkage over the whole range of sample moisture content, especially when volume decrease is no longer proportional to water content decrease. Linear models can be actually adopted in a wide range of X/X_0 but do actually fail, as expected, when sample gets dry and X/X_0 is lower than about 0.1.

Initial thickness	Model	r^2	Parameter	Value	Standard error
6mm	$V/V_0 = a + b\left(X/X_0\right)^c$	0.998	a	7.50 10^{-2}	3.65 10^{-3}
			b	0.949	4.80 10^{-3}
			c	1.272	1.66 10^{-2}
	$V/V_0 = a + b\left(X/X_0\right)$	0.990	a	1.866 10^{-2}	6.12 10^{-3}
			b	0.960	1.17 10^{-2}
8mm	$V/V_0 = a + b\left(X/X_0\right)^c$	0.996	a	0.109	4.48 10^{-3}
			b	0.925	6.18 10^{-3}
			c	1.230	2.07 10^{-2}
	$V/V_0 = a + b\left(X/X_0\right)$	0.989	a	6.46 10^{-2}	5.22 10^{-3}
			b	0.929	1.03 10^{-2}
10mm	$V/V_0 = a + b\left(X/X_0\right)^c$	0.995	a	5.97 10^{-2}	5.74 10^{-3}
			b	0.973	6.83 10^{-3}
			c	1.402	2.59 10^{-2}
	$V/V_0 = a + b\left(X/X_0\right)$	0.975	a	1.001 10^{-3}	8.81 10^{-3}
			b	0.948	1.63 10^{-2}

Tab. 2: Parameter estimation for the two fitting models at three different initial thickness.

Fig. 9 shows the dependence of bulk density on water content in some typical cases. A rather significant variation of bulk density is observed as drying proceeds: at the beginning, it ranges between 400 and 500 Kg/m^3, in agreement with the values found by Ali, Ramaswamy and Awuah (2002) and by Jha and Matsuoka (2002). Subsequently, a slight increase is observed up to a value close to 700 Kg/m^3.

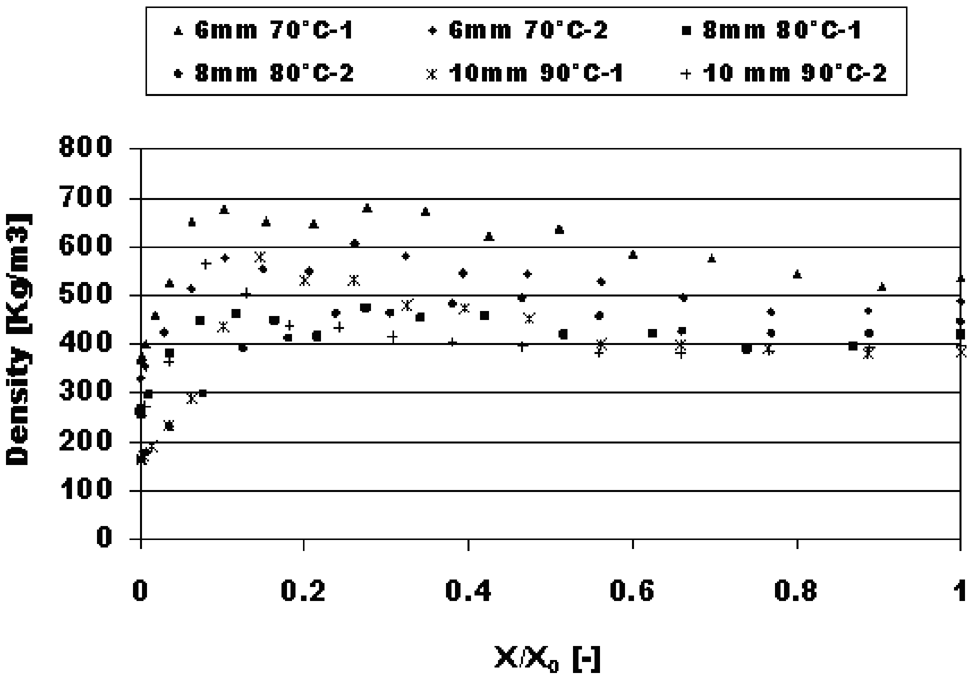

Fig. 9: Bulk Density variation during drying of eggplant slabs (1 and 2 indicate the first and the second experiment performed at the same temperature).

When eggplant gets dry (X/X_0 lower than 0.1) a steep decrease in bulk density occurs; this phenomenon could be probably ascribed to the formation of pores within the solid matrix, as it was found by May and Perré (2002) and Wang and Brennan (1995). For the sake of brevity, some other experimental data have been omitted since their general trend is similar to those already presented.

Conclusions

A novel experimental technique has been proposed to evaluate some the most important parameters involved in food drying process, i.e. effective diffusion coefficient of water, shrinkage and bulk density. Generally, effective diffusion coefficient increased as operating temperature was raised, but a precise relationship between D_{eff} and initial sample thickness could not be identified. During the first period of drying, food bulk density did not change appreciably with sample moisture content, whereas a linear relationship between food sample volume and its moisture content has been recognized. During the final period, instead, a remarkable decrease in bulk density and a strong deviation form linear proportionality between volume and water content have been observed.

The huge variability of vegetables properties imposes to make use of reliable, but also simple, versatile and cheap experimental techniques aimed at the estimation of foods main properties. The proposed method has several advantages over those reported in the literature since it is very cheap, it does not require high energy consumptions and, in principle, can be applicable to different types of foods. Moreover, it provides trustworthy estimations as it is confirmed by comparing the present results with those reported in other papers and performed on eggplants, when available, or on other vegetables under similar operating conditions.

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Experimental evaluation of quality parameters during drying of carrots

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Abstract

The aim of this work is to evaluate the effect of operating conditions on the quality of dried foods, monitoring, with respect to time, shrinkage, re-hydration capacity and colour changes. Carrot samples, having different shapes and dimensions, have been dried in a convective oven by air whose temperature and velocity have been changed in a range of physical significance. A series of experimental drying curves have been obtained and eventually fitted by an exponential model. Food volume variations, instead, have been expressed in terms of food moisture content by a linear relationship. Both the models showed a remarkable agreement with the experimental data. It has been also observed that more drastic drying conditions do indeed improve the drying rate, thus increasing shrinkage effects and decreasing food re-hydration capacity. A lower correlation between drying conditions and food colour has been noticed.

Keywords: drying, carrot, shrinkage, colour, rehydration .

Introduction

Thermal processing affects food quality, so that a substantial degradation in some essential attributes, such as colour, flavour, texture, nutrients and rehydration capacity is generally observed

As drying goes by, the initial mechanical equilibrium, which determines the food size and its shape, may fail due to the water transport from the solid matrix. A field of contracting stresses develops, thus promoting the attainment of a new equilibrium situation that leads to a change in food shape and dimensions. This change is known as Shrinkage (Mayor & Sereno, 2004, Ratti, 1994). Optimization of drying process is strictly related to the identification of a particular set of operating conditions that, in the cheapest way, allow attaining a very high quality dried food, i.e. a product exhibiting a fast re-hydration capacity, a small shrinkage, and an attractive colour (Maskan, 2001-b).

Dried foods are susceptible to colour deterioration. Colour is considered a key quality attribute due to its relation with flavour and aroma. Food discoloration and browning are the results of various reactions, including pigment degradation, especially carotenoids and chlorophyll, and browning reactions such as Maillard condensation of both hexoses and amino components and oxidation of ascorbic acid (Krokida et al., 2001, Krokida et al. ,1998). All the above unwanted reactions depend on the operating conditions chosen to carry out drying process; in order to minimize colour deterioration both a suitable design of manufacturing and treatment equipments and a proper choice of process conditions are needed (Maskan, 2001- b).

Re-hydration process, usually used to restore raw material properties, takes place when dried food comes in contact with water or water vapour. Re-hydration of dried plant tissues consists of three simultaneous processes: imbibitions of water into dried material, swelling and leaching of soluble. Generally, dried foods are not so prone to uptake water; in the literature, several studies showed that the rate and the degree of rehydration are strictly correlated to the duration and the severity of drying. The degree of rehydration is dependent on the extent of cellular and structural disruption and can be considered as a measure of the damage caused to the material by drying process. A minimization of shrinkage and, therefore, the presence of well defined intercellular voids may increase re-hydration rate (Krokida, 1999).

Shrinkage, rehydration and colour changes have been used by several authors to evaluate the damages of processing on food matter. Sanjùan et al. (2004) analyzed the effect of storage temperature on the quality of dehydrated broccoli florets. Maskan (2001-a) studied drying, shrinkage and rehydration characteristics of kiwi fruits during drying by hot air and by microwaves. Colour was utilized as a quality parameter of orange juice (Tiwari et al., 2008) of dried Chempedak (Chong et al., 2008), of dried Bay Leaves (Gunhan et al., 2004), of dried coconut (Niamnuy & Devahastin, 2005) and during drying of potato, carrot and banana samples (Chua et al., 2004). Prachayawarakorn, et al. (2008) considered shrinkage and colour as quality parameters to analyze drying of banana samples. Re-hydration and colour were used as quality parameters (Sanjùan et al.2004) for dried peppers by Vega-Gálvez et al. (2008). Shrinkage, re-hydration capacity and colour were all utilized during drying of cooked rice (Luangmalawata et al.2008), while shrinkage and re-hydratability were chosen for potato drying by Khraisheh et al. (2004) as quality parameters.

The aim of this work is to evaluate the effect of operating conditions on the quality of dried carrots, monitoring, with respect to time, shrinkage, re-hydration capacity and colour changes.

Experimental

Carrot samples of different shapes and dimensions have been dried by air in a convective oven. The carrots, bought in a local market, have been cut in cylindrical and slab shapes as shown in the following Fig. 1.

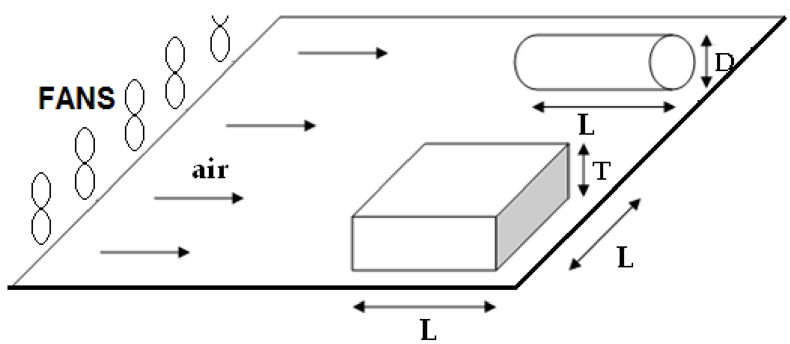

Fig. 1: Carrot samples shapes and drying airflow.

Both cylinder length and slab side (L) had an initial value of 30mm. Three different values of initial cylinder diameter (D) and slab thickness (T), i.e.: 5, 10 and 15mm, have been chosen to perform the present experimental analysis thus leading to a total number of six kinds of food samples.

Each sample has dried in lab-scale convective oven (Memmert Universal Oven model UFP 400) monitoring, with respect to time, the following food parameters:

- weight, by a precision balance (Mettler AE 160), having an accuracy of $\pm 10^{-4}$g;
- color, using a Minolta colour reader (CR-300, Japan). The colour readings were expressed by ICI coordinates (L^*,a^*,b^*) system. L^*, a^* and b^* indicates whiteness/darkness,

redness/greenness, blueness/yellowness values, respectively. A standard white colour has bees used as a reference. Three replicate readings have been carried out.

- dimensions, using a vernier caliper having an accuracy of ± 0.02 cm.

The lab-scale oven allowed monitoring, by a Dostmann electronic Precision Measuring Instrument P 655, air temperature and its humidity (by a rh 071073 probe) and air velocity (by a H 113828 probe). The convective flow of drying air has been obtained by a line of fans placed along the edge of oven internal tray. Two values of the air velocity, *u*, have been chosen, i.e. 2.8 and 2.2 m/s. Air Absolute Humidity, AH, was kept constant throughout all the experiments and equal to 10.32 g water/m^3 dry air; air temperature, T_a, was chosen equal to 50°C, 70°C, 85°C that corresponded to a relative humidity of 12.4%, 5.21% and 2.93%, respectively. Food samples have been placed on a wide-mesh perforated tray; as shown in Fig. 1, air flowed parallel to cylinder axis or parallel to one of the two slab sides having a length equal to L.

Food weight has been periodically measured, over a time horizon of 5 hours, also to perform re-hydration tests, carried out by immersion of dried samples in water at 20°C. Before each measurement, the carrot samples were taken out and blotted with paper towel to eliminate the exceeding water on their surface. Each test has been repeated twice to ascertain its reproducibility.

The main dimensions of the samples have been experimental evaluated and on the basis of the above measurements, food volume has been calculated.

Results and Discussion

Drying Curves

In the following, the effect of air velocity and air temperature on carrots convective drying is presented. Figs. 2 and 3 show the time evolution of normalized food moisture content X/X_0 in the case of a slab-shaped and of a cylindrical sample, respectively. The reported standard deviations, and the experimental error never exceeding the 15%, indicate a good reproducibility of experimental tests.

The so-called Newton Model is widely adopted to describe drying process evolution by means of an exponential equation according to a pseudo first order reaction kinetics, (Ertekin & Yaldiz 2004, Senadeera et al. 2003):

$$MR = (X - Xe)/(Xo - Xe) = \exp(-kt) \qquad 1$$

Where X (Kg $_{water}$/Kg $_{dry\ solid}$) is the moisture content on a dry basis, X_0 is the corresponding initial value, X_e is the moisture content in equilibrium with drying air and t is the drying time. The parameter k is the so-called *drying constant* and represents a measure of drying process rate.

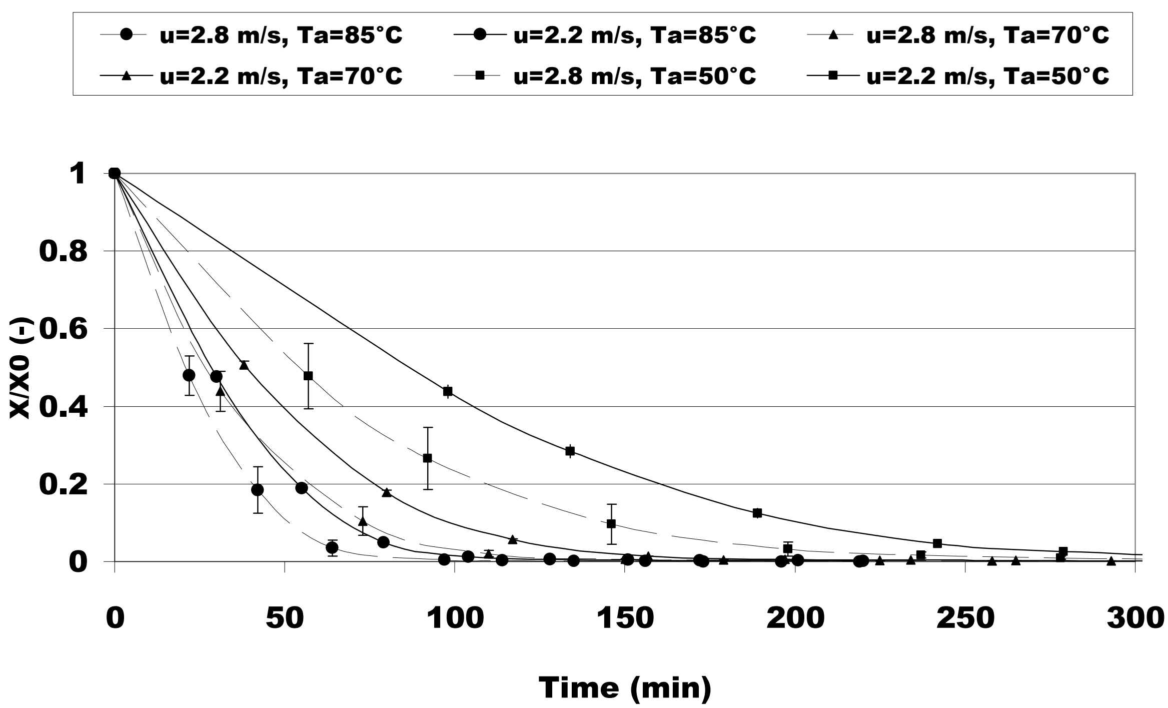

Fig. 2: Drying curves obtained for a slab-shaped carrot sample under different conditions (AH = 10.32 g $_{water}$/m^3 $_{dry\ air}$, thickness = 5 mm).

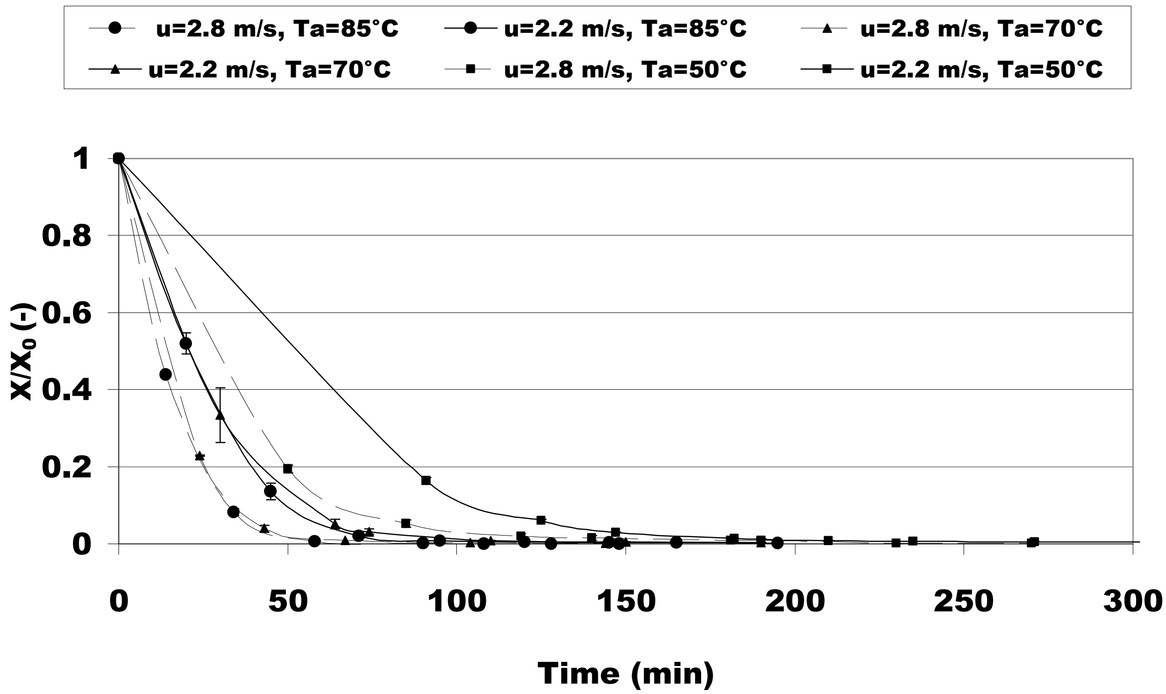

Fig. 3: Drying curves obtained for a cylindrical carrot under different conditions (AH = 10.32 g $_{water}$/m^3 $_{dry\ air}$, diameter = 5 mm)

Assuming that, as reported by Ertekin & Yaldiz (2004), X_e tends to zero, the experimental data collected in all the tested conditions have been fitted by Newton model so as to evaluate the effect of operating conditions on k parameter and, therefore, on drying rate.

The results of fitting procedure are synthetically shown in Tabs 1 and 2 for the case of slab-shaped and cylindrical samples, respectively.

	T = 5mm			T = 10mm			T = 15mm		
Air conditions (u, Ta)	r^2	k[1/min]	k standard error	r^2	k[1/min]	k standard error	r^2	k[1/min]	K standard error
2.8m/s, 85°C	0.992	0.034	1.15E-03	0.991	0.018	4.50E-04	0.998	0.013	1.26E-04
2.8m/s, 70°C	0.991	0.023	9.27E-04	0.997	0.012	1.84E-04	0.997	0.008	1.05E-04
2.8m/s, 50°C	0.996	0.013	2.93E-04	0.994	0.008	1.46E-04	0.996	0.006	6.98E-05
2.2m/s, 85°C	0.990	0.027	9.87E-04	0.997	0.012	1.36E-04	0.999	0.009	4.16E-05
2.2m/s, 70°C	0.992	0.018	6.25E-04	0.995	0.010	1.95E-04	0.999	0.007	5.79E-05
2.2m/s, 50°C	0.991	0.010	2.73E-04	0.984	0.006	1.62E-04	0.999	0.004	1.87E-05

Tab. 1: Results of drying curves fitting procedure using Newton model (slab samples).

The high value of r^2 coefficient indicates a very good agreement between Newton Model and the experimental drying curves. For all the experiments an increase in drying constant value is observed, as both air temperature and its velocity increase and food characteristic dimension decreases.

In the case of carrot slabs having a thickness of 5mm, an increase in drying temperature from 50˚C to 85˚C allows a drying time reduction of 88% and of 84% when air velocity was equal to 2.8 m/s and 2.2 m/s, respectively.

At a temperature of 85 °C, an increase of air velocity from 2.2 to 2.8 m/s determined a 29% decrease in drying time; whereas at 50 °C a 6.5% decrease was observed for the same variation of air velocity. The results shown in Tabs 1 and 2 evidence that drying rate increases as food dimension (thickness or diameter) decreases. This could be ascribed to three main factors occurring at lower diameter (thickness): the shorter path through which water is transported within the solid matter; the increased exposed surface per unit of volume; the decrease of both internal and external mass transfer resistances (Bird, 1960).

	D = 5mm			D = 10mm			D = 15mm		
Air conditions (u, Ta)	r^2	k[1/min]	k standard error	r^2	k[1/min]	k standard error	r^2	k[1/min]	K standard error
2.8m/s, 85°C	0.969	0.044	3.93E-03	0.994	0.029	7.84E-04	0.996	0.018	2.88E-04
2.8m/s, 70°C	0.955	0.042	4.52E-03	0.995	0.020	5.41E-04	0.998	0.011	1.25E-04
2.8m/s, 50°C	0.994	0.026	9.66E-04	0.997	0.014	2.49E-04	0.989	0.007	1.48E-04
2.2m/s, 85°C	0.992	0.033	1.26E-03	0.992	0.022	6.37E-04	0.996	0.011	1.35E-04
2.2m/s, 70°C	0.993	0.035	1.43E-03	0.995	0.014	3.37E-04	0.999	0.009	5.65E-05
2.2m/s, 50°C	1.000	0.021	2.06E-04	0.992	0.009	2.12E-04	1.000	0.005	8.33E-06

Tab. 2 : Results of drying curves fitting procedure using Newton model (cylindrical samples).

In a typical case, i.e. drying performed by air at 85 °C flowing at 2.8 m/s, a decrease in food characteristic of 66% determines a decrease in drying time of 74% and of 65% for a slab-shaped and a cylindrical sample, respectively.

Re-hydration Curves

Food re-hydration capacity can provide useful knowledge of damage occurred during drying. The results, presented as moisture content on a dry basis, X, versus hydration time show that food dimensions strongly affect re-hydration process since a thinner sample reaches a stationary value earlier than a thicker one (Fig. 4).

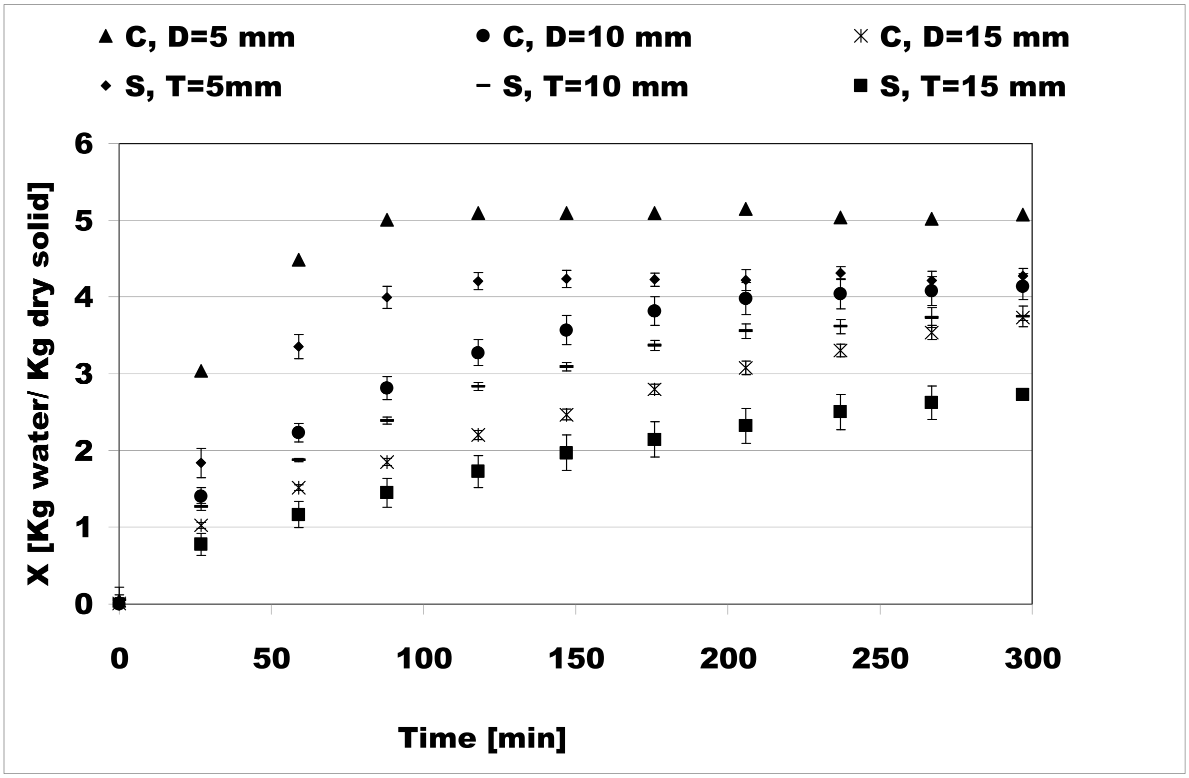

Fig. 4: Re-hydration curves (S: slab-shaped sample; C: cylindrical sample) obtained after the same drying conditions (u= 2.8 m/s, Ta= 85°C).

Fig. 5 shows the influence of drying conditions on cylindrical samples re-hydration curves; actually, the effect of air temperature and of its velocity chosen to perform the preceding drying process is not immediately evident, since the curves are very close. A more informative comprehension of process behavior can be instead obtained comparing the so-called *water regain*, defined as the percentage ratio between the water regained at the end of re-hydration (5 hours) and the water lost during drying. In order to perform a significant comparison, only the samples having the same value of initial and final moisture contents will be henceforth used (Tab. 3). On a wet basis, the considered values of initial and final moisture contents ranged between 0.87 and 0.9 and between 0.002 and 0.007, respectively.

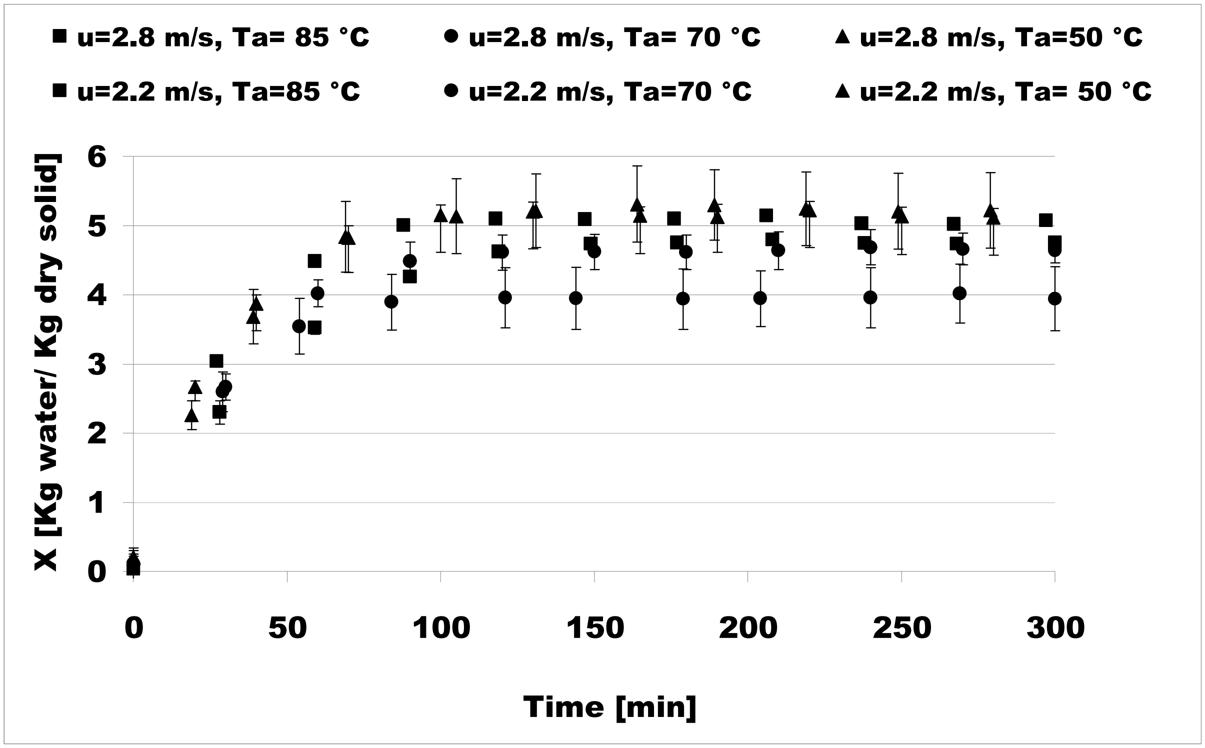

Fig. 5: Re-hydration curves for a cylindrical carrot previously dried under different conditions (D = 5mm).

Air conditions during drying (u, Ta)	**Cylinders D = 5mm**	**Cylinders D = 10mm**	**Slabs T = 5mm**
2.8m/s, 85°C	54%	45%	47%
2.8m/s, 70°C	53%	50%	64%
2.8m/s, 50°C	77%	53%	70%
2.2m/s, 85°C	-	-	-
2.2m/s, 70°C	-	-	58%
2.2m/s, 50°C	-	-	71%

Tab. 3: Water regain of different samples after re-hydration.

As expected, the samples having the same initial dimension generally exhibit an increasing water regain as the operating conditions chosen to perform the preceding drying process were milder. The effect of drying temperature is however stronger than that of air velocity. Tab. 3 confirms one of the results already shown in Fig. 4: a thinner sample is characterized by a larger water regain than a thicker one.

Shrinkage

Fig. 6 shows the experimental results of food total volume versus its moisture content on a wet basis as obtained in a typical case, i.e. for a slab-shaped carrot sample having an initial thickness of 15 mm. A significant variation of total volume, V, is observed even though the effect of air temperature and of its velocity is rather insignificant since the points obtained at the same value of U ($Kg_{water}/ Kg_{wet\ food}$) tend to overlap. Similar results have been obtained also for the other samples taken in consideration. To verify the actual effect of temperature and of air velocity from a quantitative point of view, the experimental data are generally fitted by a linear model that, as reported by Mayor & Sereno (2004) for many vegetables, was found to provide the best agreement between normalized volume (V/V_0) and normalized moisture content on a dry basis (X/X_0). The subscript 0 is referred to the initial values of both volume and moisture content. Fig. 7 shows, starting from the experimental data presented in Fig. 6, a plot of V/V_0 vs. X/X_0 that, also in the present case, may suggest the existence of a linear relationship (Eq. 2) between volume and moisture content, both normalized with respect to their corresponding initial values:

$$\frac{V}{V_0} = k_1 \frac{X}{X_0} + k_2 \qquad 2$$

The parameter k_1 can be regarded as an index of drying rate, k_2 is the normalized volume corresponding to a completely dried material (X≈0).

Eq. 2 has been, therefore, tested to fit all the experimental data collected as functions of the operating conditions and of sample dimensions.

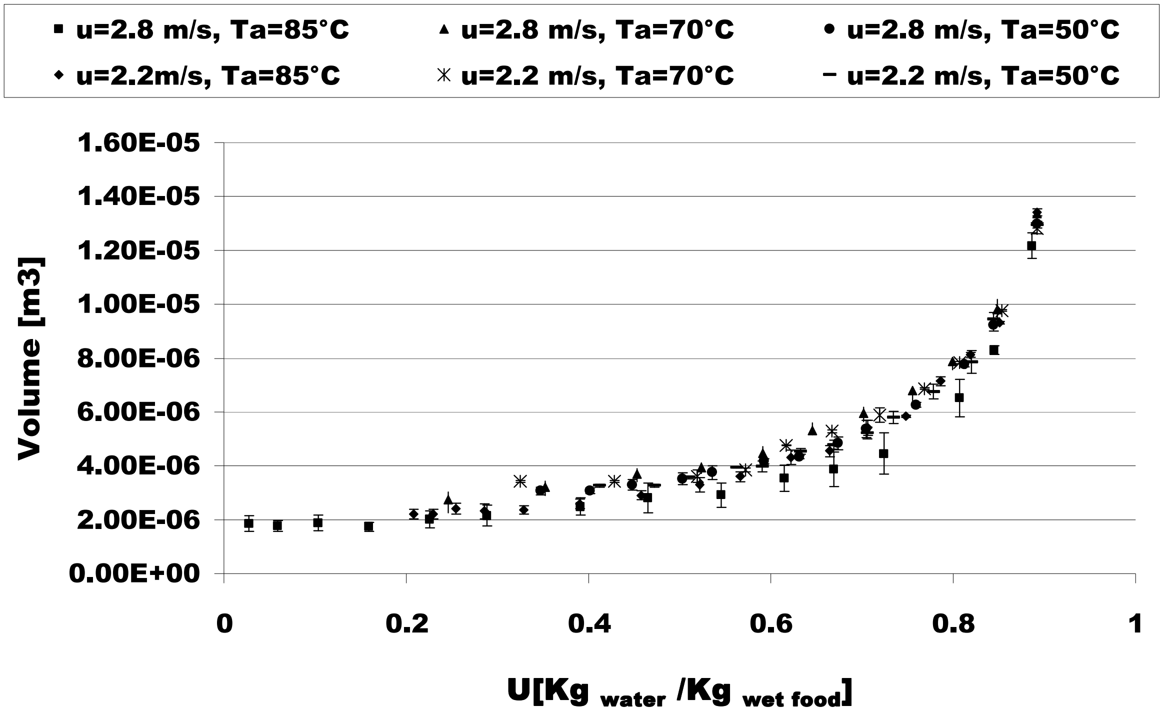

Fig. 6: Volume variation as a function of food moisture content on a wet basis at different drying conditions (slab-shaped carrot having an initial thickness of 15 mm)

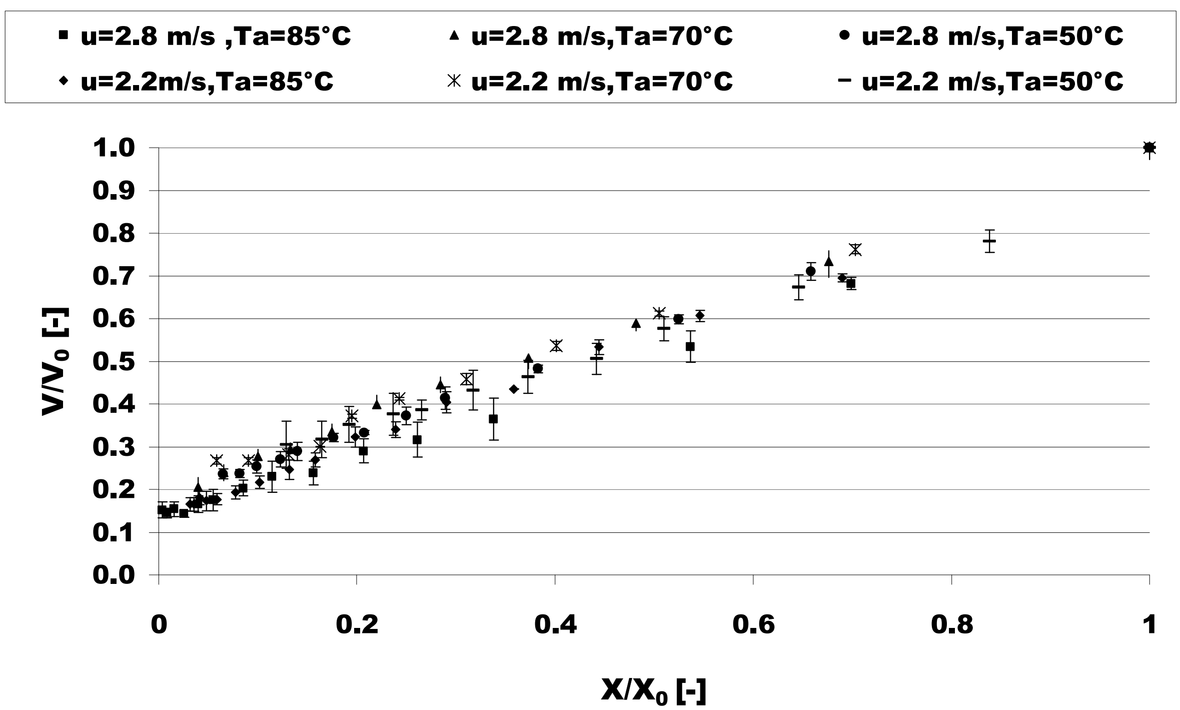

Fig. 7: Normalized volume versus normalized moisture content at different drying conditions for a slab sample (T=15mm)

The results of fitting procedure are summarized in Tabs. 4 and 5. Both the high value of correlation coefficient, r^2, and the relatively low standard error (as compared with the corresponding value of model parameters), confirm that the proposed linear model exhibits a remarkable agreement between V/V_0 and X/X_0, in all the tested conditions. The thinnest cylindrical samples (data not shown) have not been used for the shrinkage characterization due to the difficulty to correctly measure the geometrical characteristics of the samples that, during drying experiments, did not maintain their initial cylindrical shape and presented a significant curvature. Except in some cases, an increase of k_1 is observed if, for the same shape and the same initial dimension, both air temperature and air velocity raise. As far as food quality is concerned, milder drying conditions may reduce the negative effects related to food shrinkage since the rate of volume variation (k_1) is lower.

Cylinders	**10mm**					**15mm**				
	r^2	k_1	k_1 standard error	k_2	k_2 standard error	r^2	k_1	k_1 standard error	k_2	k_2 standard error
2.8m/s, 85°C	0.974	0.908	0.037	0.066	0.015	0.960	0.848	0.041	0.055	0.018
2.8m/s, 70°C	0.913	0.866	0.085	0.255	0.040	0.987	0.742	0.019	0.260	0.008
2.8m/s, 50°C	0.942	0.785	0.052	0.142	0.021	0.985	0.724	0.019	0.230	0.008
2.2m/s, 85°C	0.986	0.844	0.032	0.090	0.016	0.976	0.753	0.023	0.130	0.009
2.2m/s, 70°C	0.999	0.848	0.010	0.165	0.005	0.994	0.864	0.015	0.165	0.006
2.2m/s, 50°C	0.980	0.746	0.028	0.295	0.012	0.984	0.835	0.025	0.181	0.012

Tab. 4: Results of shrinkage curves fitting procedure using linear model (cylindrical samples).

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Slabs	**5mm**					**10mm**					**15mm**				
	r^2	k_1	$k_1$standard error	k_2	k_2 standard error	r^2	k_1	k_1 standard error	k_2	k_2 standard error	r^2	k_1	k_1 standard error	k_2	k_2 standard error
2.8m/s, 85°C	0.984	0.845	0.027	0.120	0.010	0.994	0.843	0.012	0.120	0.004	0.994	0.829	0.013	0.124	0.005
2.8m/s, 70°C	0.990	0.778	0.035	0.241	0.020	0.997	0.779	0.011	0.247	0.004	0.997	0.813	0.010	0.194	0.004
2.8m/s, 50°C	0.999	0.733	0.010	0.259	0.005	0.997	0.835	0.010	0.195	0.004	1.000	0.821	0.003	0.172	0.001
2.2m/s, 85°C	0.992	0.767	0.022	0.205	0.010	0.997	0.842	0.009	0.178	0.003	0.998	0.852	0.007	0.137	0.002
2.2m/s, 70°C	1.000	0.747	0.002	0.251	0.001	0.994	0.762	0.013	0.206	0.006	0.998	0.800	0.009	0.203	0.004
2.2m/s, 50°C	1.000	0.723	0.005	0.273	0.003	0.996	0.826	0.011	0.200	0.004	0.997	0.788	0.008	0.233	0.004

Tab. 5: Results of shrinkage curves fitting procedure using linear model (slab samples).

Color parameters

The preservation of color during drying is usually regarded as an index of low food degradation. (Krokida, et al. 2001, Sumnu, et al. 2005). Different colorimeters and different measurements units do exist. The characterization of food color is generally achieved by a set of three numbers: the most common being $L^{*,}$ the lightness, a^*, the redness-greenness and b^*, the yellowness-blueness. Figs. 8-10 present some of the experimental results obtained, as the time evolutions of L^*, a^* and, b^* during carrots drying process performed in different conditions. Actually, the experimental data do not exhibit a regular trend. The values are very close, at least at the beginning of the process, to those obtained by Krokida et al.(1998) in the case of carrots drying at 70°C, i.e. $L^* = 45$, $a^* = 15$, $b^* = 22$.

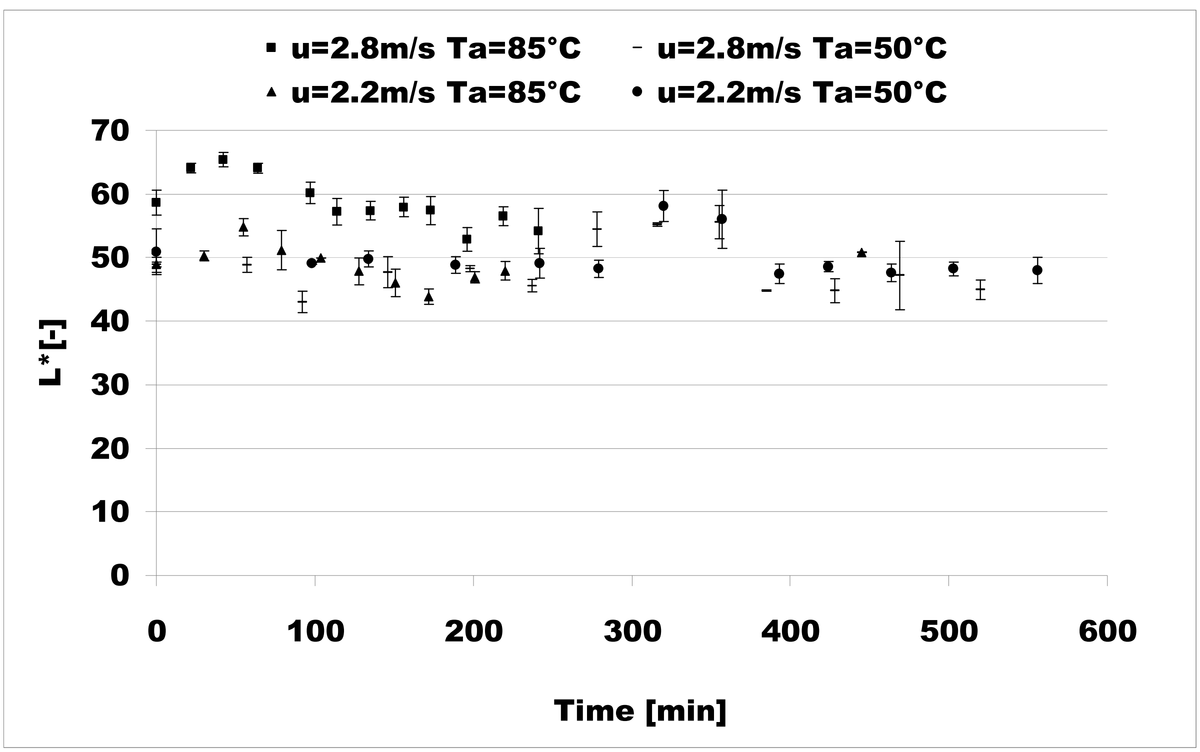

Fig. 8 : "L*" variations for carrot slabs (T= 5mm) during drying at different drying conditions.

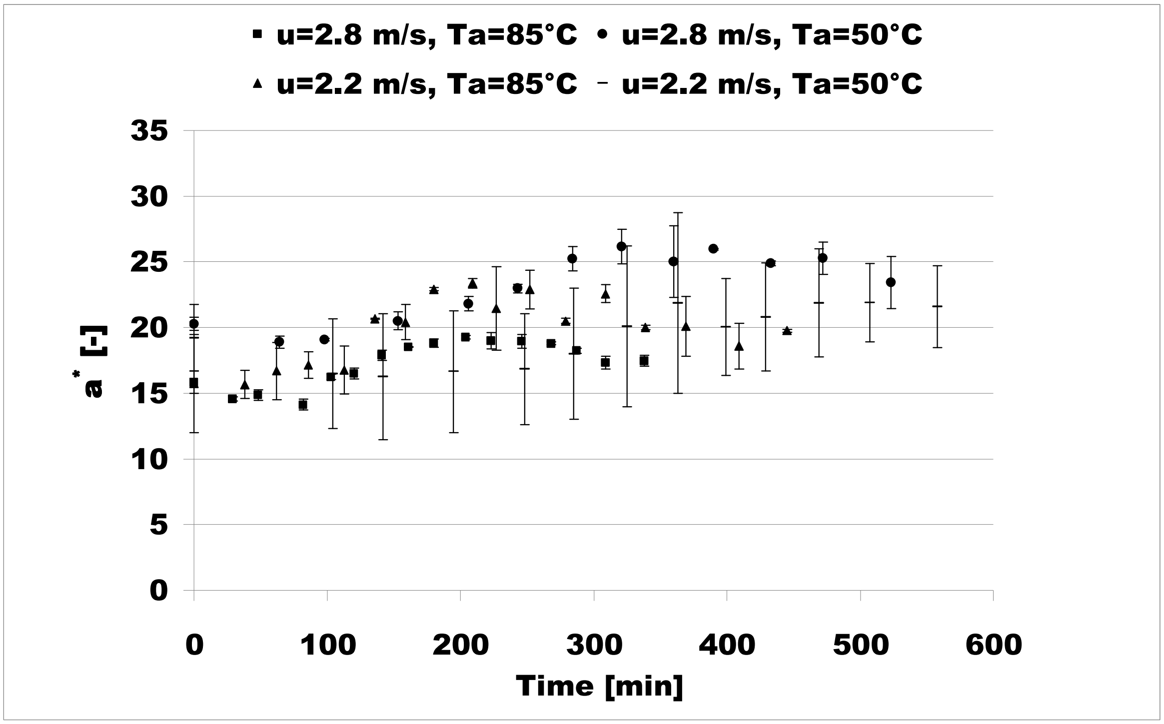

Fig. 9 : "a*" variations for carrot slabs (T=15mm) during drying at different drying conditions.

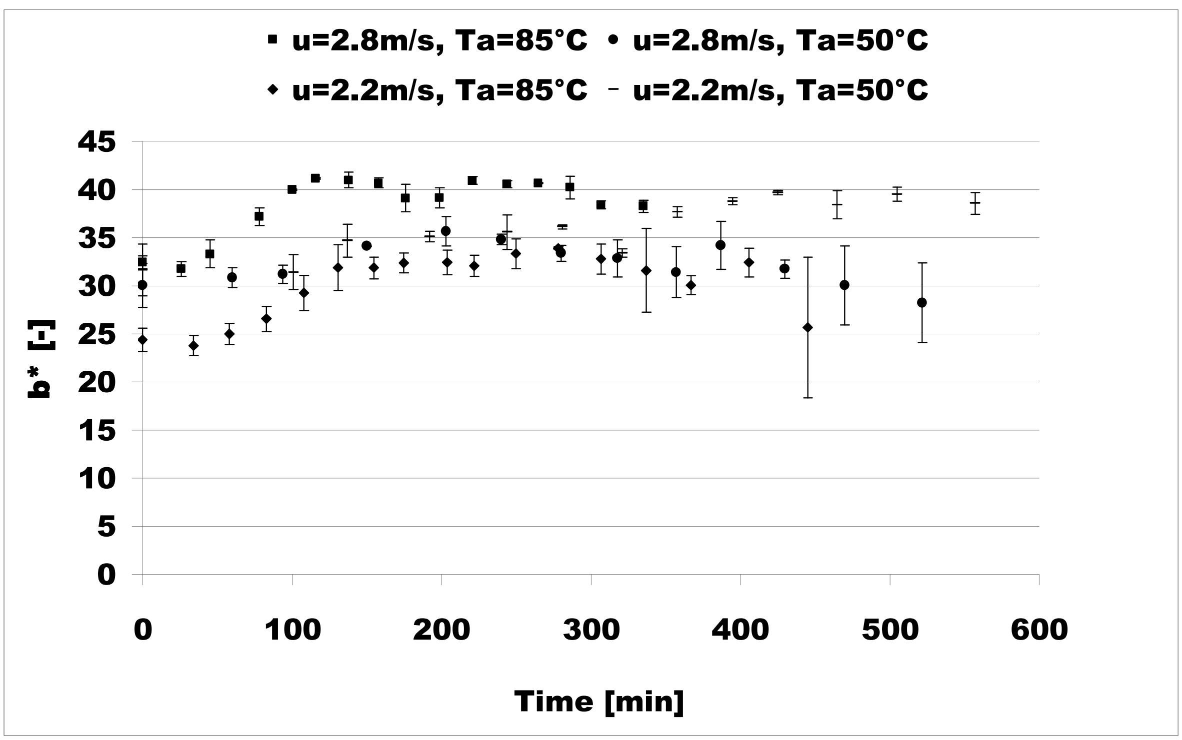

Fig. 10: "b*" variations for carrot slabs (T=10mm) during drying at different drying conditions.

Tab. 6 shows the color variations of the carrot samples as measured after 4 hours of drying. It should be remarked that slab samples, due to the larger exposed surface, exhibit a higher accuracy than the cylindrical ones as far as color measurement is concerned. For this reason, only the results related to slab samples are reported. Experimental tests did not show a monotonous trend of L^* variation, whereas the majority of the tests showed an increase of both a^* and b^* values, both related the presence of carotenes (G. Sumnu et al.2005), as drying process proceeds. In the range of operating conditions tested in this work, carotenes seem to be well preserved. However, a dependence of color parameters upon air temperature or velocity variation cannot be identified since a monotonous trend has not been observed.

	Slabs T=5mm			**Slabs T=10mm**			**Slabs T=15mm**		
Air conditions (u, Ta)	L^*	a^*	b^*	L^*	a^*	b^*	L^*	a^*	b^*
2.8m/s, 85°C	-8%	-28%	16%	-2%	8%	25%	2%	20%	32%
2.8m/s, 70°C	3%	36%	24%	5%	35%	24%	6%	46%	29%
2.8m/s, 50°C	-7%	30%	12%	3%	33%	16%	7%	13%	8%
2.2m/s, 85°C	-2%	-4%	12%	20%	10%	37%	-5%	45%	42%
2.2m/s, 70°C	0%	31%	28%	10%	48%	40%	6%	24%	12%
2.2m/s, 50°C	-4%	18%	9%	9%	26%	12%	17%	-12%	-5%

Tab. 6: Variations of color parameters of carrot slab samples after 4 hours of drying.

Conclusions

The effects of operating conditions on carrots drying process have been presented. Three important quality parameters, i.e. shrinkage, colour changes occurring as drying proceeds and re-hydration capacity after drying, have been monitored. It has been observed that the utilization of milder conditions allows achieving a better product quality, but longer drying time. An increase in air temperature and/ or in air velocity does increase the drying rate but, at the same time, increases the shrinkage effect, thus determining larger volume variations in food and decreases re-hydration capacity after drying. Both the above effects are, generally, an index of a high damaged food and have to be avoided if food quality has to be improved. As far as color variation is concerned, it has been observed that the experimental data do not exhibit a unique trend and, therefore, a useful correlation upon operating variables, thus advising the utilization of different and more accurate experimental techniques.

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Conclusions

The aim of the present work has been the modelling of one of the most common methods of food preservation: the drying process.

In many countries, dried vegetables have gained an important market share and are utilized as the main ingredients in many food products. With the increasing request for safer and high-quality foods, the process-modelling field is, at present, at its highest development phase.

The drying process modelling, performed in this work, aimed at the creation of a versatile and accurate predicting model, potentially utilizable by industries during the preliminary analysis and/or the optimization stage of the process.

A significant improvement with respect to the models available in the literature has been achieved since a number of simplifying assumptions, that affected the rigour and the potentialities of available models, have been dropped out. The simultaneous modelling of transport phenomena occurring in both the domains involved in the process (air and food) has been carried out. The complex system of partial differential equations has been solved by Comsol Multiphysics®, a powerful commercial package based on the finite elements method.

The model also accounts for food shrinkage and allows estimating the local values of heat and mass transfer coefficients at air-food interface, without resorting to any literature correlations.

The proposed models have been validated by a suitable experimental validation, which has confirmed a remarkable agreement between real values and theoretical predictions. The experimental part of the present work has concerned also the analysis and the evaluation of some properties, necessary to assess food quality during drying and to properly solve the balance equations on which the proposed modelling analysis was founded. Food properties like shrinkage, colour, re-hydratation capacity, bulk density and effective diffusion coefficient of water in vegetable have been experimentally evaluated. It has been found that drastic operating conditions do increase drying rate but, at the same time, determine a decrease of dried food quality.

Actually, it might be possible to minimize expensive pilot test-runs using the proposed models in order to have good indications on the trade-off conditions allowing short drying time and high food quality. The results obtained in this work can also be used, on an industrial scale, to choose a particular set of operating conditions that allows achieving improved food quality and higher customer satisfaction, at lower costs.

Conclusioni

Nel presente lavoro di tesi è stata affrontata la modellazione di uno dei metodi più diffusi per la conservazione degli alimenti: l'essiccazione.

I vegetali essiccati ricoprono una grossa fetta di mercato in molti paesi e sono utilizzati come ingredienti in molti cibi pronti e semipronti. Il settore della modellazione di processo nell'industria agroalimentare è attualmente nella sua fase di maggior sviluppo, a causa delle richieste sempre maggiori di alimenti sicuri e di qualità.

La modellazione del processo di essiccazione, realizzata nel presente lavoro di tesi, ha avuto lo scopo di creare un modello predittivo, versatile ed accurato, applicabile industrialmente in fase di analisi preliminare e/o in fase di ottimizzazione del processo.

Un notevole passo in avanti è stato realizzato rispetto allo stato dell'arte della modellazione in esame, in cui una serie di ipotesi semplificative comprometteva la rigorosità e le potenzialità dei modelli proposti. Per ridurre al minimo tale rischio è stata realizzata la modellazione contemporanea dei fenomeni di trasporto che avvengono in entrambi i domini coinvolti nel processo: aria ed alimento. Tale approccio ha aumentato notevolmente la complessità matematica affrontata attraverso l'utilizzo di un potente software commerciale basato sull'utilizzo del metodo degli elementi finiti, Comsol Multiphysics ®. Le caratteristiche del modello comprendono, tra l'altro, la modellazione dello *shrinkage* e la capacità di stimare localmente ed in ogni istante, durante il processo, i coefficienti di scambio all'interfaccia aria-alimento , senza avere bisogno di correlazioni di letteratura. La modellazione è stata affiancata da una opportuna validazione sperimentale che ne ha confermato le potenzialità. La parte sperimentale della tesi ha inoltre riguardato lo studio e la valutazione di alcune proprietà fondamentali sia per la caratterizzazione della qualità dell'alimento durante l'essiccazione sia per la risoluzione delle equazioni di bilancio che stanno alla base della modellazione proposta. Proprietà dell'alimento come lo *shrinkage*, il colore, la capacità di reidratazione, le densità di *bulk* ed il coefficiente di diffusione effettiva dell'acqua in matrici vegetali, sono stati oggetto di analisi sperimentale.

I risultati ottenuti hanno mostrato, oltre alla validità del modello, anche la sua utilità visto che l'utilizzo di condizioni operative particolarmente drastiche determina invero un aumento della velocità di essiccazione ma, allo stesso tempo, un peggioramento della qualità dell'alimento essiccato. In questo tipo di situazioni diventa particolarmente utile avere a disposizione un modello predittivo che consenta di fare una serie di simulazioni del processo (evitando, così, lunghi e dispendiosi test reali) e che trovi la condizione di ottimo tra le due esigenze (tempi brevi di processo ed qualità elevata dell'alimento).

I risultati ottenuti dal presente lavoro di tesi si candidano ad essere la base per le potenziali applicazioni sul processo reale, allo scopo di realizzare minori costi di produzione e maggiore gradimento, da parte dei consumatori, verso gli alimenti essiccati.

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