NUMERICAL SIMULATION OF THE CONVECTIVE DRYING PROCESS OF APPLE SLICES

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Abstract: *The quality of dried products mainly depends on temperature and moisture content changes during the drying process, as they govern the quality determining reactions that take place during dehydration. The temperature and moisture fields developing in the drying material during drying can be determined by solving the governing differential equations. They describe the heat and mass transport in the material and on its surface. Therefor a lumped parameter model of the drying material was established, using discretization and calculation methods known from aerospace engineering. The paper describes these methods as well as the simulation results for the drying process. It provides information on the dependency of heat and mass transport coefficients on local temperature and moisture content. Finally, the results are compared to experimental results obtained in a previous study.*

Key words: *convective drying, heat and mass transfer, mathematical modeling, lumped parameters*

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1. Introduction

Most biological products are dried using hot air forced convection - a very energy intensive process that is responsible for up to 25% of national industrial energy consumption in the developed countries (Mujumdar, 1999). Due to economic and environmental reasons it is aimed to minimize energy demand.

Furthermore there is an increasing demand for high quality dried products that retain their natural appearance (Fernandes *et al.*, 2011; Krokida *et al.*, 1998; Kiranoudis and Markatos, 2000; Nijhuis *et al.*, 1998). As the quality of the drying process is a function of drying time, product quality and energy consumption, for process optimization all three aspects have to be accounted for (Sturm, 2010). To reach this aim, drying equipment should be specifically designed for every drying material.

An instrument for design can be computational fluid dynamics and numerical simulations of coupled heat and mass transfer within the product. They facilitate the prediction of the most convenient drying conditions if the dependency of the product's drying behaviour on process conditions is known.

Convective drying of fruits and vegetables is a complex process involving simultaneous coupled heat and mass transfer within the drying material and at the boundary to the drying air. Extensive research has been directed towards mathematical prediction of the drying process of biological materials. Barati and Esfahani (2009) stated that over 200 drying models for various foods had been offered in literature until 2009. The status of food process modeling and models for safety, quality, and competitiveness of the food processing sector was the topic of a workshop held in 2006 and summarized and edited by Datta in 2008 (Banga *et al.*, 2008; Black and Davidson, 2008; Datta, 2008; Datta and Halder, 2008; Ivester, 2008; Jousse, 2008; Marks, 2008; Sablani, 2008; van Boekel, 2008). Puri and Anantheswaran (1993) as well as Wang and Sun (2003) give a comprehensive overview of the use of the Finite-Element Method in food processing and developments in numerical modeling of heating and cooling processes in the food industry. There are also numerous more recent papers on simulation and modeling in the food industry, for fruits and vegetables (Toğrul, 2005; Lagunez-Rivera *et al.*, 2007; de Bonis and Ruocco, 2008; Janjai *et al.*, 2008b; Janjai *et al.*, 2008a; Janjai *et al.*, 2010; Oztop and Akpınar, 2008; Kaya *et al.*, 2009), meat (Diéguez *et al.*, 2010), bread (Purlis and Salvadori, 2009; Purlis, 2011) and coffee (Nilnont *et al.*, 2011) to mention a few.

However, the coupled heat and mass transfer is so complicated, that no universally applicable model has been formulated to describe the drying process for all biological materials. Most of the approaches presented in literature provide simplifying assumptions. Shrinkage is often neglected and/or thermo-physical properties are assumed to be constant (Barati and Esfahani, 2009; de Bonis and Ruocco, 2008; Oztop and Akpınar, 2008; Kaya *et al.*, 2009; Jun *et al.*, 1998; Queiroz and Nebra, 2001; Bravo *et al.*, 2009; Barati and Esfahani, 2011). In addition most published work in literature about mathematical modelling of the drying process of foodstuffs does not differentiate between mass transfer in liquid and gaseous state

(Janjai *et al.*, 2008a; Janjai *et al.*, 2010; Janjai *et al.*, 2008b; Oztop and Akpınar, 2008; Jun *et al.*, 1998; Kaya *et al.*, 2009; Białobrzewski, 2006).

The present work deals with the simulation of apple drying, as apples are the most consumed fruit in several western countries (Aprikian *et al.*, 2001) and their consumption is on the rise in many other countries, too. An example for this is China, with a per capita consumption of 19 kg in 2004 and an upward trend (Blanke, 2005). A multitude of publications on simulation of apple drying can be found (Chiang and Petersen, 1987; Vergara *et al.*, 1997; Jun *et al.*, 1998; Krokida *et al.*, 2000; Lewicki and ŁUKASZUK, 2000; Moreira *et al.*, 2000; Białobrzewski, 2006; Bravo *et al.*, 2009; Doymaz, 2009; Wang *et al.*, 2007; Verma *et al.*, 1985; Lin *et al.*, 2009; Putranto *et al.*, 2011) although only Lin *et al.* (2009) proposed an approach similar to the model presented in this paper, but for the simulation of a novel simultaneous infrared dry blanching and dehydration (SIRDBD) process.

In the work presented, a lumped parameter model of the drying material was established, using discretization and calculation methods known from aerospace engineering. The aim of this study was to develop a model that takes into account temperature and moisture dependent material and transport properties, shrinkage and mass transfer occurring in liquid and gaseous states respectively. This work is focused on apple drying. Nevertheless the model presented can be adapted for every biological material with known drying curves and sorption isotherms.

2. Mathematic model

In order to solve the coupled partial differential equations describing simultaneous heat and mass transport within the drying material, apple slices are discretized in nodes, representing the finite elements of the problem area. The apple slice geometry as shown in figure 1 therefore was segmented into 31 nodes, of which for clarity reasons only 9 are depicted. Figure 2 shows a schematic sectional view through the drying material.

The nodes are treated as isothermal, and also as of uniform moisture content. By this means, the governing partial differential equations can be transformed into a system of non-linear equations, which is solved by using an implicit forward backward differencing method (Crank-Nicholson method).

Fig. 1. Segmentation of an apple slice into nodes

Fig. 2. Schematic sectional view through the drying material

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In this work, the following assumptions are made:

- one-dimensional heat and mass transfer within the drying material
- transport of water vapour within the drying food is not negligible
- heat, liquid and vapour fluxes can be described with identical mathematical approaches
- shrinkage is not negligible
- material properties are assumed to be functions of local moisture content and temperature
- sample consists of nodes that are treated as isothermal and as of uniform moisture

With respect to heat transfer, three different nodes can be distinguished:

- i. Diffusion nodes D have a thermal capacity C that is calculated from the specific heat c and the mass of the node consisting of the mass of the dry matter M_{dm} and the moisture in liquid and gaseous state M_w :

$$C = c \cdot (M_{dm} + M_w) \quad (1)$$

- ii. Arithmetic nodes D^0 differ from diffusion nodes, as they do not have a thermal capacity.
- iii. Boundary nodes B are characterized by having a constant temperature

Each node can interchange heat, water and vapour with the adjacent nodes, except the arithmetic nodes at the edges, because no liquid water but vapour passes to the drying air. An internal source of heat H_i can be allocated to every node allowing for taking into account loss of heat caused by phase transition.

2.1 Heat transfer

The nodes are coupled through conductors describing heat conduction as well as heat transfer through convection. Their conductance values are calculated in different ways:

$$G_{\text{cond}} = \lambda \cdot \frac{A}{h} \quad (2)$$

$$G_{\text{conv}} = \alpha \cdot A \quad (3)$$

In these equations, h is the distance between coupled nodes and A the heat exchange area. As heat transfer is assumed to occur only perpendicularly to the direction of airstream conductive heat flow between nodes $\dot{H}_j$ and convective heat transfer at the lower and upper surface $\dot{H}_s$ can be described as follows:

$$\dot{H}_j = \lambda_j \cdot \frac{A}{h} \cdot (\theta_j - \theta_{j+1}) = G_j \cdot (\theta_j - \theta_{j+1}) \quad (4)$$

$$\dot{H}_{sl} = k_H \cdot \alpha \cdot A \cdot (\theta_a - \theta_{sl}) = G_{sl} \cdot (\theta_a - \theta_{sl}) \quad (5)$$

$$\dot{H}_{su} = \alpha \cdot A \cdot (\theta_{su} - \theta_a) = G_{su} \cdot (\theta_{su} - \theta_a) \quad (6)$$

In Equation (4) λ_j means a mean thermal conductivity calculated with the average temperature and moisture content of the involved nodes. The convective boundary conditions at the upper surface are different from those on the lower surface. The apple slices lay on a grid reducing heat and mass transfer at the lower surface. This is accounted for by the parameter k_H , which has to be < 1 . The drying surface is represented by an arithmetic node and the drying air by a boundary node.

2.2 Mass transfer

Analogue to heat conduction, liquid moisture fluxes $\dot{M}_w$ are assumed to be proportional to a constant of proportionality, namely the liquid diffusivity κ_j , that's determination is explained in equation (16). The driving force is the difference in moisture content between adjacent nodes. With respect to particle density ρ_{pj} liquid moisture streams going from node j to $j+1$ are calculated:

$$\dot{M}_{wj} = \kappa_j \cdot \rho_{pj} \cdot \frac{A}{h} \cdot (X_j - X_{j+1}) \quad (7)$$

At the drying surface water stream (liquid state) is zero.

Water vapor transport is described by Fick's law of diffusion. This is a valid assumption, as water vapor pressure within hygroscopic materials is low compared to ambient pressure. Therefore, the unidirectional diffusion can be approximated to the stationary equimolar case with a water vapor pressure difference ($P_{vj} - P_{vj+1}$) as the driving force:

$$\dot{M}_{vj} = \mu_j \cdot \delta_{va} \cdot \frac{A}{h} \cdot \left(\frac{P_{vj} - P_{vj+1}}{R_v \cdot T_j} \right) \quad (8)$$

In this approach, R_v is the individual gas constant for water vapor, T_j is the mean absolute temperature, and δ_{va} is the mean diffusion coefficient for water vapor in air (diffusion is not hindered) of involved nodes. The internal resistances abating vapor diffusion are summarized in the resistant coefficient μ_j (Krischer and Kast, 1978). The convective boundary conditions for heat transfer and vapor streams leaving the drying material at the lower and upper surface are given by:

$$\dot{M}_{vsl} = k_M \cdot \beta \cdot \frac{P^\circ}{P_{vsl} - P_{va}} \cdot \ln \left(\frac{P^\circ - P_{va}}{P^\circ - P_{vsl}} \right) \cdot \frac{A}{R_v \cdot T_{sl}} \cdot (P_{va} - P_{vsl}) \quad (9)$$

$$\dot{M}_{vsu} = \beta \cdot \frac{P^\circ}{P_{vsu} - P_{va}} \cdot \ln \left(\frac{P^\circ - P_{va}}{P^\circ - P_{vsu}} \right) \cdot \frac{A}{R_v \cdot T_{su}} \cdot (P_{vsu} - P_{va}) \quad (10)$$

The parameter k_M is introduced considering lower mass transfer at the lower surface. As no mass transfer but heat exchange can take part where the apple lies on the grid, k_M has to be < 1 , and also smaller than k_H . T_{sl} and T_{su} are mean values between surface and air temperature, P_{vsl} and P_{vsu} are water vapor partial pressures at the lower and upper surface and P_{va} is the water vapor pressure of the drying air. The water vapor partial pressures within the drying material are determined by means of moisture sorption isotherms. This is valid if two conditions are complied:

- i. water activity within the drying material is identical to relative humidity
- ii. relationship for sorption isotherm can be applied for parts of the drying material

In this case water vapor pressure partial pressure in node j is calculated from water activity a_{wj} and saturation pressure:

$$P_{vj} = a_{wj} \cdot P_v'' \quad (11)$$

2.3 Heat transfer through phase transition

During convective drying, water vapor diffuses into regions of higher temperatures. Therefore recondensation can be excluded. Due to heat of vaporization of water h_{evap} and heat of sorption h_{sorp} the difference of vapor streams entering and leaving the node ($\Delta \dot{M}_{vj} = \dot{M}_{vj-1} - \dot{M}_{vj}$) leads to a loss of heat HI_j within a node:

$$HI_j = \Delta \dot{M}_{vj} \cdot (h_{\text{evap}} + h_{\text{sorp}}) = \Delta \dot{M}_{vj} \cdot h_{\text{tot}} \quad (12)$$

3. Numerical techniques

The lumped parameter model provides algorithms to solve the heat transfer problem numerically. The thermal system is brought into a stable state with respect to the energy balance for every iteration loop. The new temperatures of diffusion nodes $\theta_{j,\text{new}}$ (and arithmetic nodes, considering $C = 0$) are calculated using the following equation:

$$\sum_j^n G_j \cdot \left(\frac{\theta_{j,\text{new}} - \theta_{j,\text{old}}}{2} - \frac{\theta_{j+1,\text{new}} - \theta_{j+1,\text{old}}}{2} \right) + HI_j = C_j \cdot \frac{\theta_{j,\text{new}} - \theta_{j,\text{old}}}{2} \quad (13)$$

The software allows to implementing individual FORTRAN code. It is executed either before, or after every iteration step. This is the key to integration of mass transfer into the model. It also allows determining material and transport properties with temperatures and moisture content calculated in the last iteration loop. While new nodal temperatures are determined by the thermal software, new moisture masses are calculated by producing a mass balance for every node:

$$M_{\text{wj}} = M_{\text{wj}}^{\text{old}} + (\dot{M}_{\text{wj}} + \dot{M}_{\text{vj}}) \cdot dt - (\dot{M}_{\text{wj}+1} + \dot{M}_{\text{vj}+1}) \cdot dt \quad (14)$$

4. Parameters used in the model

In this work relationships found in literature as well as own empirically determined ones were applied (Table 1).

4.1 Thermal conductivity

There is a wide variation of reported experimental data of thermal conductivity of foodstuffs. Saravacos and Maroulis (Saravacos and Maroulis, 2001) evaluated information given in literature statistically, and suggested an empirical mathematical model for calculating thermal conductivity in foods as a function of moisture content X and absolute temperature of the drying material T :

$$\lambda(X, T) = \frac{1}{1+X} \cdot 0.287 \cdot e^{\left(-1390 \cdot \left(\frac{1}{T} - \frac{1}{333}\right)\right)} + \frac{X}{1+X} \cdot 0.589 \cdot e^{\left(-285 \cdot \left(\frac{1}{T} - \frac{1}{333}\right)\right)} \quad (15)$$

4.2 Heat and Mass Transfer coefficients

A common approach in practical engineering is applied, expressing the heat transfer coefficients without dimension via Nusselt-number (as a function of Reynolds- and Prandtl-number) and the mass transfer coefficient via Sherwood-number (as a function of Reynolds- and Schmidt-number). Depending on the flow condition, they are calculated by different empirical equations.

4.3 Mass transport coefficients

As the approaches in literature do not differentiate between mass transfer in liquid and gaseous state, an empirical equation for liquid diffusivity was established. It takes into account its temperature as well as moisture dependency and comprises three parameters:

$$\kappa(\theta, X) = a_1 \cdot \theta \cdot \left(1 - \exp \left[- \left(\frac{X}{a_2} \right)^{a_3} \right] \right) \quad (16)$$

The resistance coefficient is assumed to be moisture dependent and the diffusion coefficient for water vapour in air reflects temperature dependency. For the resistance coefficient a similar approach to liquid diffusivity was developed, whereas an empirical equation for the diffusion coefficient was given by Schirmer (1938).

$$\mu(X) = b_1 \cdot \exp \left[- \left(\frac{X}{b_2} \right)^{b_3} \right] \quad (17)$$

$$\delta_{va}(\theta) = \frac{2.26}{P^\circ} \cdot \left(\frac{\theta + 273.15}{273.15} \right)^{1.81} \quad (18)$$

variable	equation
Particle density (Krokida <i>et al.</i> , 2000)	$\rho_p = \frac{1 + X}{\frac{1}{1690} + \frac{X}{999}} \quad \left[\frac{\text{kg}}{\text{m}^3} \right]$
Saturation pressure (Tetens, 1939)	$P_v'' = 133.289474 \cdot 10^{\left(\frac{7.5 \cdot \theta}{\theta + 237.3} + 0.6609 \right)} \quad [\text{Pa}]$
Specific heat (Hugenschmidt, 2010)	$c = 1750 + 2500 \cdot \frac{X}{1 + X} \quad \left[\frac{\text{J}}{\text{kg} \cdot \text{K}} \right]$
Water activity (Hugenschmidt, 2010)	$a_w = 1 - \exp \left[- 0.0002926 \cdot (\theta + 288.4) \cdot \left(\frac{\ln \left(\frac{X - 0.08502}{0.001347} \right)}{0.4861} \right) \right] \quad [-]$
Heat of sorption (Hugenschmidt, 2010)	$h_{\text{sorp}} = 1.2 \cdot 10^6 \cdot e^{\left(- \frac{X}{10} \right)} \quad \left[\frac{\text{J}}{\text{kg}} \right]$

Tab. 1. Variables used in the model

4.4 Shrinkage

Assuming that the decrease of drying material volume is equal to the volume of water removed, a linear approach describing shrinkage can be applied. Experimental results (Sturm, 2010) showed that the average ratios of shrinkage were 0.34 for thickness (h_x/h_0) and 0.84 for diameter (d_x/d_0), leading to different equations:

$$h(X) = h_0 \cdot \left(0.66 \cdot \frac{X}{X_0} + 0.34 \right) \quad (19)$$

$$d(X) = d_0 \cdot \left(0.16 \cdot \frac{X}{X_0} + 0.84 \right) \quad (20)$$

5. Results and discussion

In order to determine the empirical constants used in the equations for transport coefficients, the drying curves simulated were adjusted to experimental data provided by Sturm (2010) who examined the drying behavior of apples, variety Jonagold in a wide parameter range. An experiment with drying parameters in the middle of the parameter space was chosen to fit calculated data. Once the empirical parameters were determined, the model was verified by comparing predicted drying curves with experimental ones using different process parameters. The parameters found are listed in the Notations. It was also examined if the influence of temperature, velocity and relative humidity of drying air are all reproduced correctly by the model.

5.1 Transport properties

As shown in Figure 3, liquid diffusivity has got a maximum at high moisture content. Inside the apple slices there is a continuous liquid film preventing vapor diffusion completely.

Fig. 3. Mass transport coefficients as function of moisture content for different temperatures

As drying proceeds moisture content decreases leading to a microstructural change that is responsible for increasing transport resistance. At very low moisture content mass transport in liquid state is negligible as equalizing between the cells stops. In this range of moisture content vapor diffusion becomes the crucial transport mechanism. Pores become empty, allowing vapor to move through the emerging hollows. Both transport phenomena are enforced by temperature.

5.2 Drying curves

Figures 4 and 5 represent the drying behavior as a function of temperature θ_a and velocity w_a respectively, the parameters which have the greatest influence on the drying process. Those parameters have been varied according to the available experimental data. Dew point temperature or humidity of air respectively has little effect on drying characteristics in the range investigated (Sturm, 2010). Figure 4 shows the predicted and experimentally determined drying curves for different air temperatures (35°C, 60°C and 85°C), whereas velocity and dew point temperature were held constant ($w_a = 3.4$ m/s, $\theta_{dp} = 17.5^\circ\text{C}$). As reported by Sturm (2010) air temperature has the most significant influence on drying time. Drying time can be shortened considerably by increasing air temperature due to the higher temperature difference between the drying surface, and air leading to a higher amount of heat being transferred to the drying material. Increasing product temperature also leads to higher water vapour pressures within the product, and consequently convective mass transfer increases.

Fig. 4. Predicted and experimental determined drying curves for different air temperatures θ_a

Fig. 5. Predicted and experimental determined drying curves for different air velocities w_a

The effect of air temperature on reduction of drying time is higher at low temperatures than at high temperatures, if changing temperature for the same amount. Additionally, product quality can deteriorate significantly if exceeding product specific temperature limits. The model shows high accuracy over the whole range of moisture content and temperature. Figure 5 depicts predicted and experimental drying curves at different air velocities (2.0 m/s, 3.4 m/s and 4.8 m/s) but constant temperature ($\theta_a = 60^\circ\text{C}$) and dew point temperature ($\theta_{dp} = 17.5^\circ\text{C}$). Qualitatively the influence of air velocity on drying behaviour is similar to that of temperature. Increasing air velocity reduces drying time significantly. This effect decreases with higher velocities. Air velocity influences transfer coefficients at the drying surface. For calculating Nusselt- and Sherwood-number (Nu, Sh) Reynolds-number (Re) is used being proportional to velocity. Again predicted data shows a high degree of accuracy. Therefore it can be assumed, that the model reproduces the physical effects correctly.

5.3 Temperature

Figure 6 shows the predicted and experimental measured upper surface temperature of food as a function of relative moisture content X/X_0 . Air properties were constant and equal to 60°C , 3.4 m/s and 17.5°C for temperature, velocity and dew point temperature. At the beginning of the drying process surface temperature rapidly increases to wet bulb temperature ($\approx 30^\circ\text{C}$). After this warm-up stage temperature increase is significantly reduced due to the cooling effect caused by

evaporation at the surface. When moisture content falls below approximately 10% of the initial content, temperature rises again and approaches that of air temperature. According to Pavón-Melendez *et al.* (2002), predicted temperature differences between the drying surface and interior are marginal as Biot-number is < 1 during the whole drying process. The heat is transferred rapidly into the interior, as the apple slices are quite thin. Because of shrinkage, the initial thickness of only 3.8 mm is even reduced to nearly one third during the drying process.

Fig. 6. Predicted and experimental measured upper surface temperature θ_{su}

Figure 7 shows the predicted moisture content for different times as function of dimensionless height, which takes into account shrinkage.

Fig. 7. Predicted moisture distribution for different times of drying

The upper surface of the apple slices dries much faster than the lower one. Especially at the beginning of the drying process moisture is unequally distributed and after about 40 minutes the upper surface layer tends to equilibrium moisture content, whereas the lower one still contains 70% of the initial moisture content. In the final stages of drying moisture distribution gets more and more symmetric. After about 120 minutes, the lower surface layer reaches equilibrium moisture content and the drying process is increasingly controlled by internal resistances.

6. Conclusions

The present work introduces a lumped parameter model describing coupled heat and mass transfer phenomena. With respect to most work published in literature, the main innovation introduced by this study was to differentiate between mass transfer in liquid and gaseous state. Shrinkage and dependency of thermo physical properties on temperature and moisture content was also taken into account. The mathematical model enables the prediction of moisture content as a function of time within the examined range of process parameters with high accuracy. It was shown that the influences of external drying parameters are predicted correctly. At the same time, it enables the prediction of moisture content as well as temperature at different depths in the slab of apple slices at different drying times. Further investigation is necessary, as the present model neglects three dimensional transport phenomena, local heat and also mass transfer coefficients, and traces back shrinkage only to moisture content, but not to geometry. Due to ambiguous heat and mass transfer conditions at the lower surface of the apple slices, further experimental data excluding these uncertainties should be generated. Then a second verification of the model with this data can be executed. Later on, quality aspects might be simulated and integrated into the model as well, as they mostly depend on temperature and moisture content. Future research aims at the amplification of the model presented, for the simulation of instationary drying processes such as belt dryers.

7. Notation

A	exchange area	m^2
a_w	water activity in food	-
c	specific heat of food	$\text{J}/(\text{kg}\cdot\text{K})$
C	thermal capacity of food	J/K
d	diameter of apple slices	m
G	conductance value	W/K
h	height of a node	m
h_{evap}	heat of evaporation	J/kg
h_{sorp}	heat of sorption	J/kg

h_{tot}	total heat of evaporation	J/kg
M	mass	kg
$\dot{M}$	mass flow	kg/s
P_v	water vapour partial pressure	Pa
P°	system pressure	Pa
$\dot{H}$	heat flow	W
H_I	loss of heat (evaporation)	W
X	moisture content (dry basis)	$\text{kg}_w/\text{kg}_{\text{dm}}$

Greek Symbols

α	heat transfer coefficient	$\text{W}/(\text{m}^2 \cdot \text{K})$
β	mass transfer coefficient	m/s
δ_{va}	diffusion coefficient (vapour in air)	m^2/s
θ	food temperature	$^\circ\text{C}$
κ	liquid diffusivity	m^2/s
λ	thermal conductivity of food	$\text{W}/(\text{m} \cdot \text{K})$
μ	diffusion resistance coefficient	-
ρ_p	particle density of food	kg/m^3
ϕ	relative humidity of air	-

Subscripts

0	initial
a	air
dm	dry matter
l	lower side
s	surface
u	upper side
v	vapour
w	water (liquid state)

Empirical Constants

$k_H = 0.10$	heat transfer
$k_M = 0.35$	mass transfer
$a_1 = 6.5e-12$	liquid diffusivity (magnitude)
$a_2 = 2.0$	liquid diffusivity (location)
$a_3 = 1.1$	liquid diffusivity (slope)

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