Simple Mathematical Model To Predict the Drying Rates of Potatoes

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A simple mathematical model to describe the drying kinetics of potato cubes with hot air has been developed. The influence of both air temperature and air flow rate on the drying of 1-cm potato cubes has been studied. The critical value above which air flow rate has no effect on the drying rates was approximately of 8000 kg/h m^2. Variation of the effective diffusivity with temperature was of the Arrhenius type. The agreement between calculated and experimental average moisture content was satisfactory.

INTRODUCTION

A possible method of storing potatoes and avoiding further deterioration is by dehydration treatment of the product. Decreasing the moisture content guards against degradative reactions, from both the physicochemical and microbiological points of view.

Food drying involves simultaneous mass- and heat-transfer phenomena. These transfer operations of food processing can be analyzed on the basis of physical and engineering principles. Exact application of these principles to food systems is usually difficult because of the complex and heterogeneous structure of foods and the physical, chemical, and biological changes that may take place during processing (Saravacos, 1986). Moisture transfer in food processing involves both diffusion of moisture in solid food and gas phase. Diffusion in solid food during drying is a complex process, which may involve molecular diffusion, capillary flow, Knudsen flow, hydrodynamic flow, or surface diffusion. Likewise, the physical structure of food plays a very important role in the diffusion of moisture. It is assumed that during the falling rate period of drying moisture is transferred mainly by molecular diffusion (Saravacos, 1986). The analysis of this mass-transfer phenomenon must be carried out under the assumption that the effective diffusivity will involve all of the parameters which can influence the process rate: air-drying temperature, moisture content, structure of the solid, and, in some cases, air-drying rate. The microscopic mass balance is usually employed in the modeling of this process. The differential equation requires the corresponding initial and boundary conditions for each particular transfer process. Numerical techniques are used for solving these types of problems. Studies mainly based on finite differences (Balaban and Pigott, 1988; Sereno and Medeiros, 1989; Mulet et al., 1989a,b) and finite element methods (Sastry et al., 1985; Hong et al., 1986; Lomauro and Bakshi, 1985; Vanegas and Marinos-Kouris, 1991) have been found in the literature. The former are known to be difficult techniques to apply, and their complexity involves a great amount of computer time.

Karathanos et al. (1990) performed a comparative study on two estimating methods of effective moisture diffusivity in starch model systems for different sample geometries (spheres and slabs), based on the solution of the Fick's law equation for unsteady-state diffusional processes. In the drying of food materials, effective diffusivity may vary considerably with regard to moisture, at short time periods. Nevertheless, effective diffusivity remains constant at dimensionless moisture values lower than 0.6 (Perry and Green, 1984), which in turn give rise to linear curves. In some cases, the drying curve can consist of two or more different linear sections from which several effective diffusivity values can be estimated (Cobbinah et al., 1987). The solutions obtained by using both a simple mathematical method of slopes of the nonlinear drying curve and a computer simulation technique were compared. Both methods gave similar results on high-amylopectin starch gels of low porosity, where liquid diffusion might predominate during the drying process. Potatoes have an approximate amylopectin content of about 77% with regard to the total starch content, which suggests the use of a simplified model representative of the drying rates.

In industrial operations, mass transfer in the gas phase may be very important and should be taken into consideration in the calculations of unsteady-state diffusion. Assuming the existence of an effective diffusivity and taking the concept of both external and internal resistance established by King (1968) into account, mass transfer can be considered the result of two resistances in series: the resistance corresponding to the mass transfer from the inside of the solid to the interface (internal resistance), where the diffusional coefficient varies only as a function of the air-drying temperature (Mazza and LeMaguer, 1980; Mulet et al., 1983), and the resistance to mass transfer from the interface to the gas (external resistance), which is dependent on both the air-drying rate and temperature.

The purpose of this work is to establish a mathematical model that is easy to apply, is representative of the rates of drying potatoes with hot air, and would not require computers of great capacity.

MATHEMATICAL MODEL

To establish a simplified model, it is necessary to make several simplifying assumptions. First, it will be assumed that the process that takes place is of an isotherm type, since in dehydration of food products heat transfer occurs very quickly (Tosi et al., 1987). It is assumed that no contraction or deformation of the solid particle occurs during the process (Alvarez and Legues, 1986; Tolaba et al., 1989). It will also be assumed that water displacement throughout the solid follows a diffusional mechanism. A constant effective diffusivity will be considered throughout the solid, representative of the global transport process of moisture, both in the inside and in the gas phase.

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Assuming an isotropic behavior of the solid with regard to the water diffusivity, the process can be defined as a function of the Fick's law combined with the microscopic balance of mass transfer. The differential equation obtained for a cubic symmetry is

$$\frac{\partial W_p}{\partial t} = -D_{eff}\left(\frac{\partial^2 W_p}{\partial x^2} + \frac{\partial^2 W_p}{\partial y^2} + \frac{\partial^2 W_p}{\partial z^2}\right) \quad (1)$$

To solve this differential equation (eq 1), the following should be assumed: the initial moisture content is uniform throughout the solid, and the air is at equilibrium with the surface of the solid for the time considered. The geometry of the solid remains constant during the whole process, i.e., no shrinkage or deformation is detected.

The boundary conditions considered in this study are those related to both the thermodynamic equilibrium and the symmetry of the solid.

The differential equation can be solved by separation of variables and expression of the moisture content in a dimensionless way:

$$\Psi(x,y,z,t) = \frac{W_p - W_e}{W_c - W_e} \quad (2)$$

As an initial condition it is assumed that the critical moisture content (W_c) corresponds to the initial moisture content of the solid. It is an approximate value which does not modify the values obtained for the effective diffusivity according to the proposed model. A further evaluation of the proposed model will lead to the determination of a moisture critical value nearer to the real value.

The solution for the x axis when the external resistance is negligible is

$$X(x,t) = 2\sum_{\nu=0}^{\infty} \frac{(-1)^{\nu}}{(2\nu+1)\frac{\pi}{2}}\left(\exp\left(-\left((2\nu+1)\frac{\pi}{2}\right)^2\frac{D_{eff}}{L^2}t\right)\cos\left((2\nu+1)\frac{\pi}{2}\frac{x}{L}\right)\right) \quad (3)$$

The solutions for both y and z axes are similar to that of x. The three-dimensional solution is obtained as the product of the three one-dimensional expressions:

$$\Psi(x,y,z,t) = X(x,t)Y(y,t)Z(z,t) \quad (4)$$

Calculation of the Drying Time. Considering the first term of each of the one-dimensional expressions as an approximate solution to the model and subsequently integrating, it is possible to obtain a new expression for the dimensionless average moisture (Ψ) of the solid.

$$\ln\Psi = -\frac{3\pi^2(D_{eff}/L^2)}{4}t \quad (5)$$

From eq 5 it is possible to calculate the time required in a drying process to obtain a certain value of the average moisture content of the solid, as a function of the effective diffusivity. This expression is valid when a monolayer loading of 1 cm is used.

Calculation of the Mass Flux. Calculation of the mass flux (N_A) can be carried out in different ways depending on which resistance governs the process, being either external (gas phase) or internal (solid phase). The external resistance to mass transfer is clearly related to air flow rate by means of the mass-transfer coefficient, k_G, between air and solid.

If the resistance which controls the mass transfer is the external, the mass flux can be calculated as

$$N_A = k_G(C^* - C_{AG}) \quad (6)$$

The mass-transfer coefficient can be obtained from the numerous empirical correlations found in the literature, some of which have been compiled by Skelland (1974). In general, those correlations are based on the heat- and mass-transfer analogy related to a particle shape. The most adequate correlation for the experimental conditions of the present work was that of Williams:

$$Sh = 0.43Re^{0.56}Sc^{1/3} \qquad 500 < Re < 5000 \quad (7)$$

The characteristic dimension used for the calculation of Sh and Re numbers is defined as the total surface area of the body divided by the perimeter of the maximum projected area perpendicular to the air flow.

An expression (eq 8) is obtained from eq 7 in which can be noticed the dependence of the mass-transfer coefficient on the average velocity.

$$k_G \propto v^{0.56} \quad (8)$$

When the governing resistance is the internal resistance, Fick's law can be applied in the calculation of mass flux by assuming a diffusional transport mechanism. Calculation of the mean concentration gradient of the solid can be performed by derivation of the average dimensionless moisture of the solid with regard to one of the three directions (in $x = L$) and subsequently integrating in relation to the other two directions.

The overall resistance will be the sum of the external and internal resistances. With regard to this fact three different zones depending on the air-drying flow rate value can be found. In the first zone the external resistance controls the process, in the second (a transitional zone) both resistances are important, and in the third zone the process is controlled by the internal resistance. Given a certain temperature value and for a certain type of material, size, and geometric shape, there is an air flow rate value called critical rate, above which the drying rate does not increase even if the air flow rate value is increased (Mulet et al., 1987).

MATERIALS AND METHODS

Claustar potato from Majorca was the raw material used in all experiments. Before drying, the potatoes were peeled, cut into 1-cm-edged cubes, and blanched for 3 min by steam. A laboratory-scale hot air drier described previously by Mulet et al. (1987) was used for dehydration of the samples.

A 0.5-hp fan was used, and the air flow rate was measured with a rotameter which allowed measurements up to 40 m^3/h, with an accuracy of 0.5 m^3/h. Air flow was perpendicular to the drying bed. The heating system consisted of an electrical heater (1000 W) placed inside the air pipe, and the heating control was achieved by a proportional-integral controller (PI). A load density corresponding to a monolayer was used.

The average room air characteristics were 26 °C and 35% humidity. Water losses were measured by weighing the basket and its content. The moisture content of the dried product was obtained according to the AOAC (1984) method.

Drying experiments were carried out in two sets. In the first, to obtain the air critical flow rate value, experiments were performed at constant air temperature (50 °C) by changing the air flow rate between 800 and 25 000 kg/m^2 h. In the second, experiments were carried out to determine the influence of temperature on the drying rates. The latter experiments were performed on the samples at different temperatures (30, 40, 50, 60, 70, 80, and 90 °C) by using higher air flow rate values than the critical value.

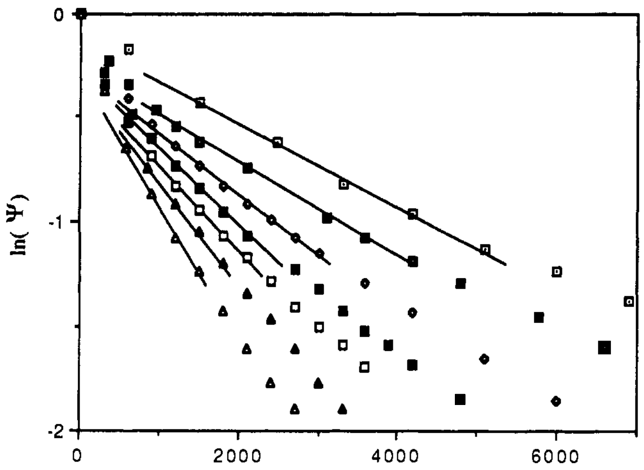

Figure 1. Influence of air-drying temperature on the drying kinetics ($G > 8000$ kg/m² h). (◫) $T = 30$ °C; (■) $T = 40$ °C; (◇) $T = 50$ °C; (■) $T = 60$ °C; (□) $T = 70$ °C; (▲) $T = 80$ °C; (△) $T = 90$ °C.

Water activity measurements were carried out on dehydrated samples with different moisture content, according to the method of polyols (Steele, 1987).

RESULTS AND DISCUSSION

Water activity data were fitted to the Henderson model (Iglesias and Chirife, 1982), recommended by numerous authors (Yanniotis et al., 1989; Schuchmann et al., 1990; Vidal and Bornhardt, 1991). Using the Marquardt method (Kuester and Mize, 1983) the following equation was obtained:

$$a_w = 1 - \exp(-6.11(W)^{0.74}) \quad (9)$$

To evaluate the dimensionless moisture, the equilibrium moisture content should be stated. Using eq 9, the equilibrium moisture content of the solid with the drying air can be calculated.

In the dehydration kinetics of food products, induction periods and constant drying rate are usually very short (Alvarez and Legues, 1986); therefore, the most representative period of the process, which is the period addressed in the present work, is that of the falling drying rate.

The results of the moisture content obtained at different times for each of the drying experiments have been expressed as dimensionless moisture. According to the model proposed for the calculation of the drying time (eq 5), when representing the Naperian logarithm of the dimensionless moisture vs time, it is possible to obtain the D_{eff}/L^2 parameter value. When the experimental points corresponding to the first falling drying rate period are fit to a straight line, the former value can be obtained from the corresponding slope.

Figures 1 and 2 represent ln Ψ vs time for the experiments conducted at different temperatures (Figure 1) and at different air flow rates (Figure 2). As can be observed, increasing the air-drying temperature caused an important increase in the drying rate. Nevertheless, the air flow rate has no effect on the drying process of the product above a certain air flow value (critical air flow rate), which corresponds approximately to a value between 7000 and 10 000 kg/m² h. Figures 1 and 2 show the existence of more than one drying falling period. In the present work only the first diffusional period will be taken into account, which corresponds approximately to $0.7 \geq \Psi \geq 0.3$.

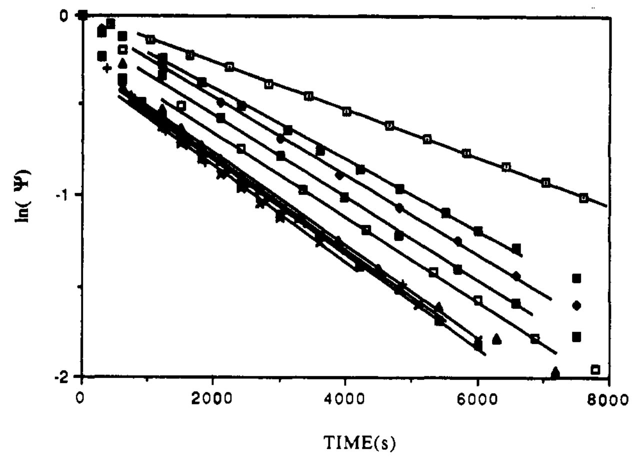

Figure 2. Influence of air flow rate on the drying kinetics ($T = 50$ °C). (◫) $G = 882$ kg/m² h; (■) $G = 1764$ kg/m² h; (◆) $G = 2747$ kg/m² h; (■) $G = 3298$ kg/m² h; (□) $G = 5524$ kg/m² h; (△) $G = 8305$ kg/m² h; (+) $G = 9998$ kg/m² h; (■) $G = 15\,916$ kg/m² h; (×) $G = 24\,924$ kg/m² h.

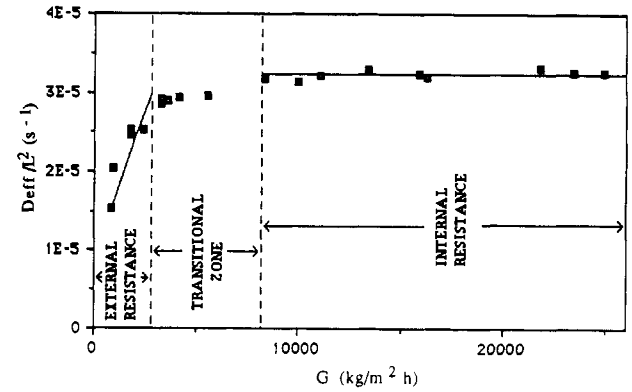

Figure 3. Variation of D_{eff}/L^2 with the air flow rate ($T = 50$ °C).

Table I. Influence of Air Flow Rate on the D_{eff}/L^2 Value ($T = 50$ °C)

G, kg/m² h	D_{eff}/L^2 (10^5), s⁻¹	G, kg/m² h	D_{eff}/L^2 (10^5), s⁻¹
882	2.03	8305	3.17
1764	2.52	9998	3.14
2747	2.79	13369	3.26
3298	2.85	15916	3.22
5524	2.95	24924	3.24

Influence of Air Flow Rate. In Figure 3 and Table I are shown the values obtained for D_{eff}/L^2 vs the air flow rate for the experiments performed at constant temperature. Three different zones can be considered. In the first, when the air flow rate is low, D_{eff}/L^2 is clearly influenced by it, and the external resistance then controls the process. In the same conditions, it is possible to obtain a relationship between D_{eff}/L^2 and the air flow rate similar to the one proposed by Mulet et al. (1987). At a constant temperature of 50 °C the following relationship was obtained:

$$D_{eff}/L^2 = \exp(-14.849 + 0.56 \ln G) \quad (10)$$

There is an intermediate zone in which both resistances contribute significantly. The critical value of the air mass flow is then about 8000 kg/m² h. In the dehydration of 1-cm-edged carrots, Mulet et al. (1987) obtained a critical value of approximately 6000 kg/m² h and Mitchell and Potts (1958), 4200 kg/m² h.

At high air flow rates ($G > 8000$ kg/m² h) the effect of the external resistance on the dehydration kinetics is negligible and the internal resistance to mass transfer then

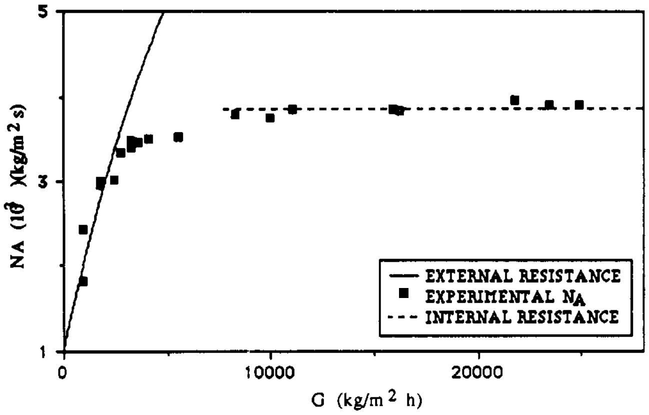

Figure 4. Variation of mass flux with the air flow rate (T = 50 °C; Ψ = 0.5).

Table II. Influence of Air-Drying Temperature on the D_{eff}/L^2 Value (G > 8000 kg/m^2 h)

temp, °C	D_{eff}/L^2 (10^5), s^{-1}	temp, °C	D_{eff}/L^2 (10^5), s^{-1}
30	2.20	70	4.93
40	2.68	80	6.13
50	3.29	90	7.46
60	4.00		

controls the process. Since D_{eff}/L^2 is constant throughout the process, at 50 °C this parameter has a value of 3.233 10^{-5} s^{-1}.

To calculate the mass flux, knowledge of a relationship between the volume of the solid and its moisture content is required. In the literature this relationship is usually of the linear type in fruits and vegetables (Mazza and LeMaguer, 1980; Madarro et al., 1981). The relationship obtained in this work was

$$V/V_0 = 0.1233 + 0.1522W \qquad r^2 = 0.998 \tag{11}$$

Following the proposed model, in Figure 4 are plotted both the experimental mass flux values, using Fick's law, and the theoretical values obtained according to the correlation of Williams (eq 7) vs the air flow rate. It is noted that at low air flow rates the experimental points fit well to the correlation of Williams. Above a certain flow value, air flow rate has no effect on the mass flux. The value corresponding to the air flow rate was found to be approximately 8000 kg/m^2 h.

Influence of Air Temperature. The study of the influence of air-drying temperature on the drying rates agrees with the fact that mass transfer within the solid follows a diffusional mechanism, when the concept of effective diffusivity is used. The D_{eff}/L^2 values obtained by applying the proposed model to the results of the experiments performed at different temperatures (Table II) were plotted according to the Arrhenius equation (Figure 5). Variation of the effective diffusivity with temperature was of the Arrhenius type. It is possible to calculate the activation energy of the process from the corresponding slope, which is approximately 1068 kJ/kg. In the literature it can be found that this parameter is present in a large range due to the influence of numerous factors (maturity rate, geometry, initial moisture content, size, etc.). Yusheng and Porsdal (1988) reported a list of the activation energies found by different authors for potatoes: 1784, 2910, and 1543 kJ/kg.

$$\ln (D_{eff}/L^2) = -3.35 - 2247.7/(T + 273.16) \qquad r^2 = 0.996 \tag{12}$$

This diffusivity equation, when inserted in the proposed model, should be able to reproduce the experimental results

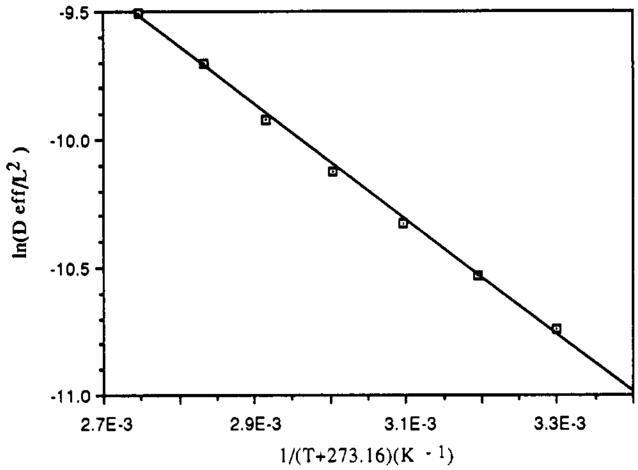

Figure 5. Influence of temperature on the D_{eff}/L^2 parameter (G > 8000 kg/m^2 h).

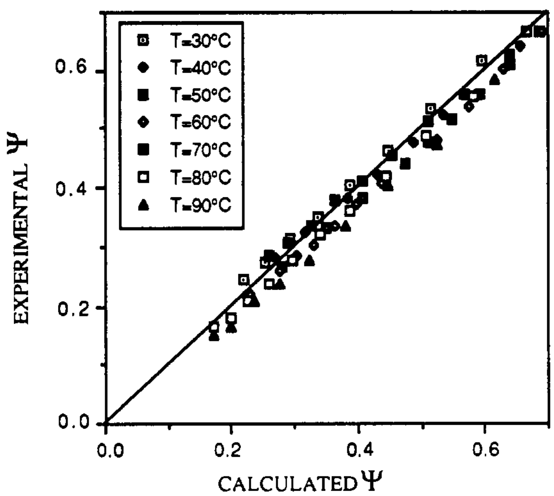

Figure 6. Comparison of the experimental dimensionless moisture with those obtained using the proposed model.

on the average moisture content at any temperature considered.

$$\ln \Psi = -\frac{3\pi^2(\exp(-3.35 - 2247.7/(T + 273.16)))}{4}t \tag{13}$$

This model gives an analytical description of the drying process of cube-shaped potatoes for the first falling drying rate period.

The maximum mean error when considering the established model for the first term of each one-dimensional expression of the dimensionless moisture (eq 4) at 5 min was approximately 11%.

The critical moisture contents considered for calculations of the experimental dimensionless moisture were those corresponding to the average value of the origin coordinates obtained in the plotting of ln Ψ vs time for the experiments conducted at different temperatures.

In Figure 6 are reported the calculated dimensionless (Ψ_{calc}) vs the experimental dimensionless moisture data (Ψ_{exp}). Good agreement between the former variables is observed.

In Table III are given the root mean square errors expressed as

$$\text{RMSD} = (\sum_{i=1}^{n}(\Psi_{exp} - \Psi_{calc})_i^2/n)^{1/2} \tag{14}$$

for the experiments conducted at different temperatures.

Table III. Comparison of the Experimental vs the Calculated Average Dimensionless Moisture Data and Calculated Root Mean Square Errors and Maximum Deviation

dehydration temp, °C	RMSD (10^2)	% deviation
30	2.552	5.10
40	3.438	6.88
50	2.664	5.33
60	2.399	4.80
70	1.451	2.90
80	2.541	5.08
90	3.668	7.33

The maximum deviation between the model and the experimental data expressed as percentage of the root mean square error with regard to the average dimensionless moisture (0.5) was 7.3%.

ABBREVIATIONS USED

a_w, water activity; C^*, interface water concentration (kg/m^3); C_{AG}, gas-phase water concentration (kg/m^3); D_{eff}, effective diffusivity (m^2/s); k_G, mass-transfer coefficient (m/s); L, half-thickness of the solid (m); N_A, mass flux (kg/m^2 s); G, air flow rate (kg/m^2 h); Re, Reynolds number; RMSD, root mean square deviation; Sh, Sherwood number; Sc, Schmidt number; t, time (s); T, temperature (°C); v, average velocity (m/s); V, volume (m^3); V_0, initial volume (m^3); W, average moisture content (kg of water/kg of dm); W_c, critical moisture content (kg of water/kg of dm); W_e, equilibrium moisture content (kg of water/kg of dm); W_p, local moisture content (kg of water/kg of dm); x, x-axis distance (m); X, dimensionless moisture on x axis; y, y-axis distance (m); z, z-axis distance (m); Ψ average dimensionless moisture.

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