

Dehydration

22.1 Introduction

Drying is one of the most ancient methods of food preservation known to mankind. Preservation of meat, fish and food plants by drying in the sun or in the naturally dry air of the deserts and mountains has been practiced since prehistoric times and is still a vital operation in the life of many rural communities.

Drying or dehydration is, by definition, the removal of water by *evaporation*, from a solid or liquid food, with the purpose of obtaining a solid product sufficiently low in water content. (Note: Osmotic dehydration, where removal of water takes place by virtue of a difference in osmotic pressure and not by evaporation, is briefly treated at the end of this chapter.)

The main technological objectives of food dehydration are:

- Preservation as a result of depression of water activity
- Reduction in weight and volume
- Transformation of a food to a form more convenient to store, package, transport and use, e.g. transformation of liquids such as milk or coffee extract, to a dry powder that can be reconstituted to the original form by addition of water (instant products)
- Imparting to a food product a particular desirable feature such as a different flavor, crispiness, chewiness etc., i.e. creating a new food (e.g. transformation of grapes to raisins).

Despite the importance of drying as an industrial operation and recent progress in drying research, the physical principles of the complex phenomena that occur in the course of dehydration and rehydration are not entirely understood. Modeling of drying is discouragingly difficult in the case of food materials. A completely satisfactory model of drying kinetics, applicable to foods, is not available at the moment. Yet, in the following treatment of drying, extensive use will be made of

models. It should be remembered that these theoretical models are only approximations. Their use as an exclusive tool for process development or design of equipment is not recommended. 'Food drying engineering' still largely relies on experience and experimentation.

The most important engineering and technological issues in food dehydration are the following:

- *The kinetics of drying*: with some notable exceptions such as spray drying, drying is a relatively slow process. Knowledge of the factors that affect the rate of drying is essential for the optimal design and operation of drying systems
- *Product quality*: removal of water is not the only consequence of most drying operations. Other important quality-related changes in taste, flavor, appearance, texture and nutritive value may occur in the course of drying. The extent of such changes depends on the process conditions
- *Energy consumption*: most common drying processes use extensive quantities of energy at relatively low efficiency. Energy wise, drying is a wasteful water removal process, compared to other water removal operations such as evaporation or membrane separation.

The mechanism of water removal by drying involves two simultaneous processes, namely, transfer of heat for the evaporation of water to the food and transport of the water vapors formed away from the food. Drying is, therefore, an operation based on *simultaneous heat and mass transfer*. As explained in the next sections, the rate limiting mechanism may be superficial evaporation or internal transport of water, depending on the conditions.

Depending on the mode of transfer, industrial drying processes can be grouped in two categories: convective drying and conductive (boiling) drying.

- *Convective drying*: hot and dry gas (usually air) is used both to supply the heat necessary for evaporation and to remove the water vapor from the surface of the food. Both heat and mass exchanges between the gas and the particle are essentially convective transfers, although conduction and radiation may also be involved to some extent. This widespread mode of drying is also known as *air drying*.
- *Conductive (boiling) drying*: the moist food is brought into contact with a hot surface (or, in a particular application, with superheated steam). The water in the food is 'boiled-off'. In essence, boiling drying is tantamount to 'evaporation to dryness'. Vacuum drying, drum drying and drying in superheated steam are cases of this mode of drying.

Freeze-drying (lyophilization) is another method of water removal based on the sublimation of water from a frozen material under high vacuum. In view of the peculiar mechanisms involved, freeze-drying will be discussed in a separate chapter.

22.2 Thermodynamics of Moist Air (Psychrometry)

22.2.1 Basic principles

In most food dehydration processes, air is the drying medium. It is therefore appropriate to review basic concepts pertaining to the thermodynamic behavior of air–water vapor mixtures (moist air), before discussing drying kinetics and drying processes.

Although dry air is, in itself, a mixture of gases (nitrogen, oxygen, carbon dioxide etc.), we shall consider moist air as consisting of two components only: dry air and water vapor. Gibbs' phase rule (Josiah Willard Gibbs, American mathematician-physicist, 1839–1903) establishes the number of possible 'degrees of freedom' in a system in equilibrium, as follows:

$$F = C - P + 2 \quad (22.1)$$

where:

F = number of degrees of freedom (number of possible variables)

C = number of components

P = number of phases.

For mono-phase (homogeneous) moist air $C = 2$ and $P = 1$. It follows that the number of independent variables is 3, e.g. temperature, pressure and moisture content. Psychrometry is most commonly applied to air at atmospheric pressure. In this case, the pressure is not a 'variable'. It follows that the state of homogeneous moist air at atmospheric pressure can be unequivocally defined by two variables: temperature and moisture content (humidity). Consequently, it is customary to represent the state of moist air graphically on a system of two axes: temperature as the abscissa and humidity as the ordinate. The resulting graph is known as the 'Psychrometric chart' (see Figure A.2).

The state of two-phase moist air (e.g. moist air containing mist) at atmospheric pressure can be unequivocally defined by one variable only. For processes at variable pressure one more degree of freedom has to be added.

22.2.2 Humidity

The moisture content of air (*humidity, absolute humidity*) is expressed as the mass ratio of water vapor to dry air (kg water vapor per kg dry air) and is dimensionless. At atmospheric pressure, the air–moisture mixture can be treated as an ideal mixture of gases and Dalton's law (Section 13.2) can be applied:

$$p = p_w + p_a \quad (22.2)$$

where:

P = total pressure (usually atmospheric)

p_w and p_a = partial pressures of water vapor and dry air, respectively.

Assuming perfect gas behavior, the mass of water vapor per unit volume of mixture is:

$$m_w = \frac{18p_w}{RT} \quad (22.3)$$

Similarly, the mass of dry air (average molecular weight = 29) per unit volume of mixture is:

$$m_a = \frac{29p_a}{RT} \quad (22.4)$$

It follows that the absolute humidity of the air, H , is:

$$H = \frac{m_w}{m_a} = \frac{18.0153p_w}{28.966p_a} = \frac{0.6219p_w}{P - p_w} \cong 0.62 \frac{p_w}{P} \quad (22.5)$$

Note: The approximation $P - p_w \approx P$, is permissible only at ambient temperatures and in relatively dry air where p_w is much smaller than the total pressure P .

22.2.3 Saturation, Relative Humidity (RH)

Moisture-saturated air is air in equilibrium with pure liquid water at a given temperature. Consequently, the partial vapor pressure of water vapor in moisture-saturated air is equal to the vapor pressure of liquid water at the same temperature:

$$(p_w)_{sat.} = p_0 \quad (22.6)$$

The saturation humidity H_s is the maximum quantity of water vapor that air can contain at a given temperature, without phase separation.

The relative humidity (ϕ or RH) is the ratio (as percentage) of the partial pressure of water vapor in air, to the vapor pressure of liquid water at the same temperature.

$$RH = \frac{p_w}{p_0} \times 100 \quad (22.7)$$

'Percent saturation' (S), often confounded with relative humidity, is:

$$S = \frac{H}{H_s} \times 100 \quad (22.8)$$

The saturation and constant percent saturation lines are indicated on the psychrometric chart.

22.2.4 Adiabatic saturation, wet-bulb temperature

If a mass of air is brought into contact with water under adiabatic conditions (no heat transfer with the exterior), the humidity of the air increases until saturation is reached.

Since there is no external source of heat, water is evaporated using heat from the air itself. Consequently, the air is cooled at the same time that it is humidified. The process described is called ‘adiabatic saturation’ and the temperature reached at saturation is called the ‘temperature of adiabatic saturation’. It can be demonstrated that:

1. The temperature of adiabatic saturation is a sole function of the initial conditions (temperature and humidity) of the air. It follows that the temperature of adiabatic saturation is a thermodynamic property of moist air.
2. The process of adiabatic saturation follows a straight line on the psychrometric chart.

If the bulb of a mercury thermometer is enveloped in wet tissue and air is flown over the bulb, the air (and the thermometer) will be eventually cooled to the adiabatic saturation temperature. For this reason, the *wet bulb temperature* is often confounded with the adiabatic saturation temperature of the air. One of the ways to describe unequivocally the state of moist air is to specify its dry bulb (normal) and wet-bulb temperatures. It should be remembered, however, that the ‘wet-bulb temperature’ is an empirical value (not a property) depending on the kinetics of the measurement, while the adiabatic saturation temperature is a thermodynamic property.

22.2.5 Dew point

‘Dew point’ is yet another property of moist air. If a mass of moist air is cooled *at constant humidity*, saturation is reached and liquid water is formed in the form of dew or mist. The temperature at which this occurs is called the ‘dew point’ of the air. Dew point is one of the phenomena on which certain hygrometers (instruments for measuring humidity in air) are based. The air is contacted with a mirror that is gradually cooled. The temperature at which the mirror becomes foggy is the dew point of the air. Based on the relation between the water activity of a food sample and the relative humidity of the air in equilibrium contact with the food, the ‘foggy mirror’ phenomenon is one of the methods for measuring water activity in the laboratory.

EXAMPLE 22.1

A mass of air at 30°C has a percent saturation of 60%. Find the absolute humidity, wet-bulb temperature and dew point of that air.

Solution:

All the properties are obtainable from the psychrometric chart in Appendices (Figure A.2)

The absolute humidity is $H = 0.016 \text{ kg/kg}$

The adiabatic saturation (wet-bulb) temperature is 23.6°C

The dew point is 21.2°C.

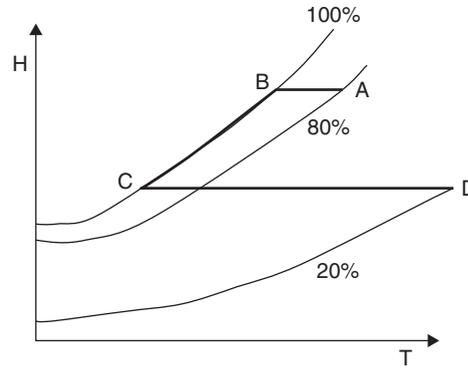


Figure 22.1 Dehumidification of air by cooling

EXAMPLE 22.2

The air in the room used for packaging instant coffee must have a percent saturation of 20% or less, in order to prevent moisture sorption and caking of the product. The dehumidified air is ‘produced’ by cooling air at 30°C and 80% percent saturation beyond saturation, separating the condensed water and heating the air back to 30°C.

Show the process on the psychrometric chart and find the temperature to which the air has to be cooled.

Solution:

The process is shown on the general form of the psychrometric chart (Figure 22.1). Line AB represents cooling at constant humidity until saturation is reached. The section BC represents condensation at 100% saturation. The line CD represents heating of the dehumidified air. Point C shows the temperature to which air has to be cooled to perform the desired dehumidification.

The result is 5°C.

22.3 Convective Drying (Air Drying)

A typical example of this category is tray drying. The moist food is placed on a tray. Hot and dry air is passed over the food (cross-flow drying) or through it (through-flow drying). Heat flows from the hot air to the cooler food by virtue of the temperature gradient and induces evaporation. Water vapor passes from the moist food to the dry air by virtue of the vapor pressure gradient.

22.3.1 The drying curve

As explained above, the rate of drying is of particular engineering and economic importance, mainly because it determines the production capacity of the dryer. Drying rate Φ is defined either as *the mass of water removed per unit time per unit mass of*

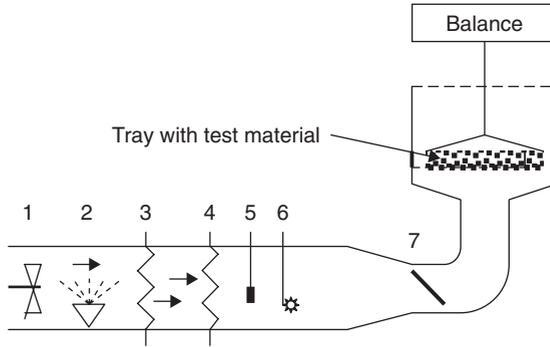


Figure 22.2 Schema of the setup for the determination of drying curves. 1: Fan; 2: humidification; 3: cooling (drying); 4: heating; 5: thermometer; 6: hygrometer; 7: damper

dry matter or as *the mass of water removed per unit time per unit area (water flux)* denoted by N . The factors that affect drying rate may be divided to the following two groups:

- ‘Internal’ conditions: variables of the material subjected to drying (shape and size, structure, e.g. porosity, moisture content, water vapor pressure as a function of composition and temperature, i.e. water vapor sorption isotherms at different temperatures)
- ‘External’ conditions: temperature, humidity and velocity of the air.

Drying rate data are usually represented in the form of *drying curves*. A drying curve is the plot of the drying rate Φ versus the remaining water content X . Water content X is expressed as kg of water per kg of dry matter.

$$\Phi = -\frac{dW}{Mdt} = -\frac{dX}{dt} \quad \text{with} \quad X = \frac{W}{M} \quad (22.9)$$

W = mass of water in the food

M = mass of dry matter in the food.

To build a drying curve experimentally, the food sample is placed in a drying chamber, on a tray attached to a balance (Figure 22.2). Air at known and constant temperature, humidity and velocity is passed in either cross-flow or through-flow direction and the weight of the tray is determined periodically and recorded (usually, air flow has to be stopped for a short time during weighing). If water is the only volatile substance in the sample, then the weight loss is taken to be equal to the loss of water mass $-\Delta W$. The mass of dry matter in the sample, M , is assumed to be constant.

A hypothetical W versus time plot is shown in Figure 22.3. From the data, the drying rate is calculated and plotted against X , to obtain the drying curve. A drying curve corresponding to the data of Figure 22.3 is shown in Figure 22.4. (Note: Since the moisture content X decreases during the experiment, the course of the process on

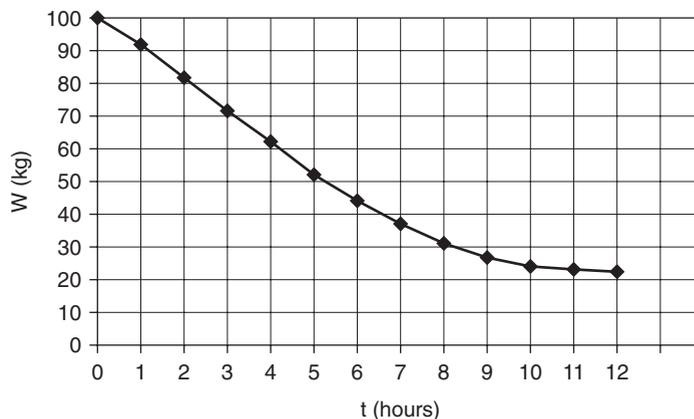


Figure 22.3 Theoretical W versus t curve

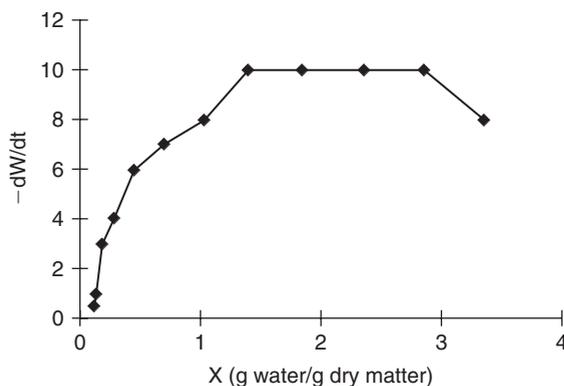


Figure 22.4 Drying curve corresponding to data in [Figure 22.3](#)

the graph is from right to left direction.) Practically, the most sophisticated part of the experimental determination of a drying curve is the production of an air stream having precisely constant temperature, humidity and velocity.

The drying curve of [Figure 22.4](#) is usually modeled so as to show three regions or phases:

1. Region I – Phase of *rising rate*: the rate of drying increases as water is removed. Physically, this behavior is attributed to the ‘conditioning’ of the sample, e.g. warming-up, opening the pores etc. This phase is usually short and not always observed in drying experiments. It is often omitted in the calculation of drying time.
2. Region II – Phase of *constant rate*: drying rate remains nearly constant as water is removed. Truly constant-rate drying may be observed when slowly drying wet sand or paper ([Krischer, 1993](#)) but seldom when drying real foods ([Bimbenet et al., 2002](#)). The physical background of the constant drying rate regime will be explained later.

3. Region III – Phase of *falling rate*: below a certain moisture content, called the ‘critical moisture content, X_C ’, drying rate drops sharply as water is removed. The probable mechanisms responsible for the falling rate will be discussed later.

22.3.2 The constant rate phase

Theory predicts that the rate of convective evaporation from a water-saturated surface is constant if the external conditions (temperature, humidity and velocity of the air) are constant.

Consider a mass of wet material subjected to convective drying by air (Figure 22.5). Heat is transferred convectively from the warm air to the cooler wet surface through the boundary layer at the interface and induces some evaporation. The water vapor formed is transported by convection from the moisture-saturated surface to the drier air.

The following transport equation can be written:

For heat transfer:

$$q = hA (T_a - T_s) \quad (22.10)$$

For mass transfer:

$$-\frac{dW}{dt} = k_g A (p_s - p_a) = k'_g A (H_s - H_a) \quad (22.11)$$

where:

$-dW/dt$ = rate of water transfer, kg.s^{-1}

h = coefficient of convective heat transfer, $\text{w.m}^{-2}.\text{K}^{-1}$

k_g and k'_g = coefficients of convective mass transfer, $\text{kg.m}^{-2}.\text{s}^{-1}.\text{Pa}^{-1}$ and $\text{kg.m}^{-2}.\text{s}^{-1}$ respectively

A = area of active transfer, m^2

p_a and p_s = partial pressure of water vapor in the air and on the wet surface, respectively, Pa.

H_a and H_s = humidity in the air and on the wet surface, respectively (dimensionless).

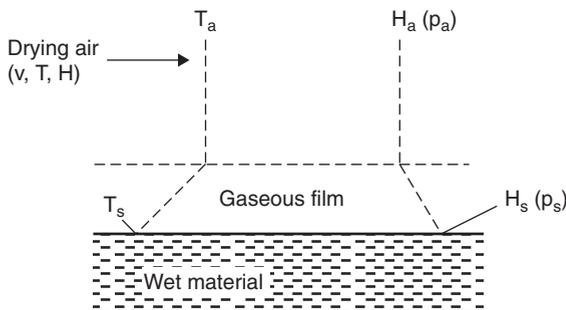


Figure 22.5 Heat and mass transfer in convective drying

Assume that, at steady state, the only effect of heat transfer is evaporation of water (no sensible thermal effects). Then:

$$N = -\frac{dW}{A dt} = k'_g(H_s - H_a) = \frac{h(T_a - T_s)}{\lambda} \quad (22.12)$$

As long as the wet surface is saturated with water, T_s is the ‘wet bulb temperature’ of the drying air and it is constant (see Thermodynamic Properties of Moist Air, Section 22.2). The humidity H_s is the adiabatic saturation humidity of the air and it is also constant. It follows that the drying rate *per unit area* $N = [-dW/Adt]$ should also be constant as long as the following conditions are satisfied:

1. The wet surface is water-saturated (i.e. behaves like the surface of pure water)
2. The air temperature, humidity and velocity are kept constant
3. Heat is transferred to the wet surface only by convection from the drying air.

To the extent that the conditions above are fully satisfied, the temperature of the food in the constant rate period does not exceed the adiabatic saturation temperature of the air, even if the actual temperature (dry bulb temperature T_a) of the air is quite high.

As mentioned, the theoretical analysis above predicts a period of constant drying rate *per unit area*. In the absence of shrinkage, the area is constant and so is the area per unit mass of dry matter. In this case, the drying rate *per unit mass of dry matter* Φ , as defined in 22.3.1 would also be constant.

Constant rate drying prevails as long as any amount of water lost by evaporation is replaced by water transported from the interior to the surface by diffusion. In other words, the constant rate drying period can be defined as the phase of drying during which evaporation from the surface and not internal diffusion is the rate-controlling mechanism (Chu et al., 1953).

In practical food dehydration, these conditions are seldom met and, therefore, truly constant rate drying is, at most, an approximation that can be applied only to foods with a high initial moisture content and an inert solid matrix. Real situations deviate from the ideal conditions required for constant rate drying in the following points:

- In real drying, heat is not transferred to the drying surface solely by convection. Conduction and radiation transfer also take place, to some extent. As a result, the temperature at the drying surface is higher than the wet bulb temperature of the air (see Section 22.3.7 below)
- Even in the case of foods with fairly high water content, the surface of the food does not behave like pure water, because of the water vapor pressure depression due to the soluble components, among other reasons
- The heat exchanged is not used only for evaporation. Sensible heat effects also occur.

Notwithstanding these deviations, the existence of a constant rate period is assumed in the modeling of convective drying.

EXAMPLE 22.3

'Filo' dough sheets, 2 mm thick, are air-dried from both sides, from an initial moisture content of $X_0 = 1$ kg water/kg dry matter to a final moisture content of $X = 0.25$ kg water/kg dry matter. The temperature of drying air is 50°C and its percent saturation is 20%. In order to avoid mechanical stress, deformation and cracks, the moisture gradient in the dough must be maintained below 200 m^{-1} . It is known that under the conditions of the process, the convective heat transfer coefficient is given by the correlation:

$$h = 10(v)^{0.8}$$

where:

h = convective heat transfer coefficient, $\text{W}/\text{m}^2\text{K}$

v = air velocity, m/s .

It is also known that the critical moisture content under the conditions of the process is less than 0.25 kg water/kg dry matter.

The density of the dough is $\rho = 1030\text{ kg}/\text{m}^3$, the diffusivity of water in the dough is $D = 3 \times 10^{-9}\text{ m}^2/\text{s}$ and the latent heat of evaporation of water is $\lambda = 2300\text{ kJ}/\text{kg}$.

What is the maximum permissible air velocity?

Solution:

During drying, moisture gradient is at its maximum at the surface. Hence the drying conditions must comply with the following condition:

$$-\left(\frac{\partial X}{\partial z}\right)_{z=0} \leq 200\text{ m}^{-1}$$

The rate of moisture transport to the surface is, by Fick's law:

$$\frac{dW}{Adt} = -D\rho\left(\frac{\partial X}{\partial z}\right)_{z=0}$$

Since there can be no accumulation on the interface, the rate of moisture transport to the surface must be equal to the rate of evaporation from the surface.

The moisture content of the dough is higher than the critical moisture. Hence, drying takes place at constant rate. The rate of evaporation from the surface is then given by Eq. (22.12):

$$N = -\frac{dW}{Adt} = \frac{h(T_a - T_s)}{\lambda}$$

It follows that:

$$\frac{h(T_a - T_s)}{\lambda} = D\rho\left(\frac{\partial X}{\partial z}\right)_{z=0}$$

From the psychrometric chart, we find the adiabatic saturation temperature of the air $T_s = 28.8^\circ\text{C}$. Substitution in the equation leads to:

$$\frac{h_{\max}(50 - 28.8)}{2300000} = 3 \times 10^{-9} \times 1030 \times 200 \Rightarrow h_{\max} = 67 \text{ Wm}^{-2}\text{K}^{-1}$$

But $h = 10 (v)^{0.8}$. Hence: $67 = 10 (v_{\max})^{0.8}$.

$$v_{\max} = 10.78 \text{ m/s}$$

22.3.3 The falling rate phase

The rate of water transfer from the interior of the particle to its surface decreases continuously as the product becomes drier. When the supply of water to the surface drops below the rate of evaporation, the moisture content of the surface begins to decrease rapidly and approaches quickly the equilibrium moisture content corresponding to the relative humidity of the air on the sorption isotherm of the material. From that moment, internal transport and not evaporation becomes the rate limiting factor and the falling rate period begins. The average moisture content of the food when this occurs is called the ‘critical moisture content X_c ’. Obviously, a sharp transition from constant to falling rate can only be observed if a real constant rate period exists, i.e. when drying inert, non-shrinking materials, but not in foods. In the case of foods, the negative slope of the drying curve becomes gradually steeper but there is no well-defined point of transition. Nevertheless, the concept of critical moisture is retained for the purpose of modeling.

Since the surface of the food is no longer water saturated, the temperature of the food rises during the falling rate period and approaches asymptotically the dry bulb temperature of the air. Theoretically, drying stops (i.e. the drying rate becomes 0) when the moisture content everywhere in the food has been reduced to the equilibrium moisture content X_e .

Many different theories concerning the mechanism of internal water transfer in drying have been proposed (Bruin and Luyben, 1980; Barbosa-Cánovas and Vega-Mercado, 1996). These include liquid water diffusion by virtue of concentration gradients, water vapor diffusion under the effect of vapor pressure differences, capillary transport, evaporation–condensation etc. Probably, a number of different mechanisms are simultaneously responsible for the internal movement of water molecules during drying.

The shape of the drying curve in the falling rate stage depends on the mechanism of internal water transfer. If molecular (Fickian) diffusion is assumed to be the principal mechanism, linear relationship should exist between the logarithm of the residual ‘removable’ water content and drying time (see Section 3.5.2):

$$\frac{d[\ln(X - X_e)]}{dt} = \text{Constant} \quad (22.13)$$

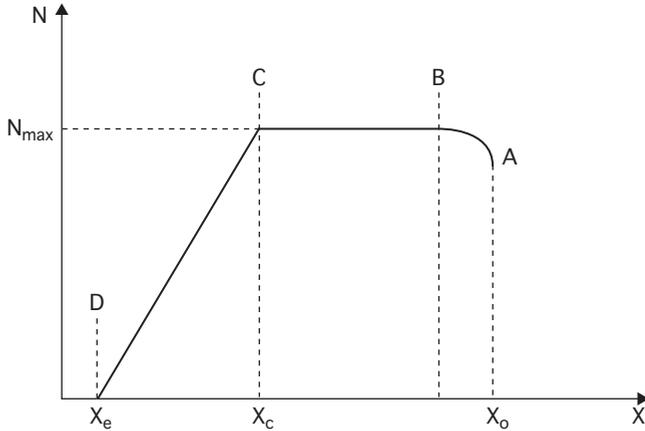


Figure 22.6 Theoretical drying curve model

Table 22.1 Water diffusion coefficient of skim milk

Temperature (°C)	Diffusion coefficient D ($m^2 \cdot s^{-1}$)		
	Moisture content (kg water/kg dry matter)		
	0.1	0.3	0.7
10	3×10^{-13}	1.6×10^{-11}	1.2×10^{-10}
30	1.2×10^{-12}	5×10^{-11}	2.2×10^{-10}
50	3.3×10^{-12}	1×10^{-10}	3.3×10^{-10}
70	1.3×10^{-11}	2.2×10^{-10}	7.4×10^{-10}

Calculated from a figure in [Bruin and Luyben, 1980](#)

Taking derivatives with respect to X on both sides and rearranging we obtain:

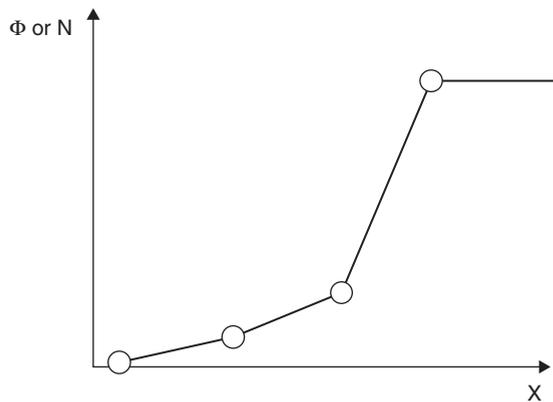
$$\frac{dX}{dt} = K(X - X_e) \quad (22.14)$$

Equation (22.14) indicates that the falling rate portion of the idealized drying curve should be a straight line between X_c and X_e (Figure 22.6). The theoretical curve describing the falling rate regime with the assumption of capillary transport would also be a straight line.

In reality, even if the mechanism of molecular diffusion is taken for granted, the application of Eq. (22.13) to food drying is problematic. The diffusivity of water in the food is strongly temperature and composition dependent (Table 22.1). The concentration dependence of diffusivity is a logical consequence of the sorption isotherm (dependence of water activity on concentration). Other phenomena, such as migration of solutes, shrinkage and thermal effects also contribute to the deviation from the simplified Fickian model.

Table 22.2 Effective diffusivity in some foods, calculated from drying rate

Material	Temp (°C)	D_{eff} ($\text{m}^2\cdot\text{s}^{-1}$)	Reference
Avocado	50	2.3×10^{-9}	Daudin, 1983
Sugar beet	50	5×10^{-10}	Daudin, 1983
Manioc	50	3×10^{-10}	Daudin, 1983
Apple	50	1.1×10^{-9}	Daudin, 1983
Potato	50	2×10^{-10}	Bruin and Luyben, 1980
Apple, Granny Smith	76	1.15×10^{-10}	Bruin and Luyben, 1980
Pepperoni sausage, 13.3% fat	12	5.7×10^{-11}	

**Figure 22.7** Drying curve approximated as a series of linear falling rate periods

An average diffusivity value can be calculated from experimental drying curves for the entire curve or parts of it. The values obtained are reported as *effective diffusivity*. Some effective diffusivity values, calculated by different authors from experimental drying curves, are given in [Table 22.2](#).

In some models, the concave falling rate portion of the drying curve is approximated by two or more linear segments, representing so-called second, third etc. falling rate periods ([Figure 22.7](#) and [22.6](#)).

22.3.4 Calculation of drying time

What is the time required for drying a food from an initial moisture content X_1 to a final moisture content X_2 , under constant external drying conditions? [Equation \(22.9\)](#) can be written as follows:

$$-dw = -MdX = \Phi Mdt \quad (22.15)$$

Integration gives the drying time:

$$t = -\int_{X_1}^{X_2} \frac{dX}{\Phi} \quad (22.16)$$

The integral can be calculated if $\Phi = f(X)$ is known as a drying curve or as an algebraic expression. In the absence of such information, a model of drying kinetics must be assumed. As an example, it will be assumed that the drying curve consists of a constant rate portion from X_1 to X_c , at a known constant rate Φ_0 , followed by a linear falling rate portion from a known (or assumed) X_c to X_e .

If both X_1 and X_2 are larger than X_c , the entire drying period takes place at constant rate. Then:

$$t = \frac{X_1 - X_2}{\Phi_0} \quad (22.17)$$

If both X_1 and X_2 are smaller than X_c , the entire drying period takes place during the falling rate phase. Then:

$$t = \frac{X_c - X_e}{\Phi_0} \int_{X_2}^{X_1} \frac{dX}{X - X_e} = \frac{X_c - X_e}{\Phi_0} \cdot \ln \left(\frac{X_1 - X_e}{X_2 - X_e} \right) \quad (22.18)$$

If X_1 is larger than X_c and X_2 is smaller than X_c , the drying period consists of a constant rate phase and a falling rate phase. Then:

$$t = \frac{X_1 - X_c}{\Phi_0} + \frac{X_c - X_e}{\Phi_0} \cdot \ln \left(\frac{X_c - X_e}{X_2 - X_e} \right) \quad (22.19)$$

EXAMPLE 22.4

Estimate the time necessary to dry dates from 75% to 20% moisture content (on wet basis) under constant external conditions. Under the conditions of the process, falling rate regime is known to prevail during drying. Linear relation between drying rate and residual moisture is assumed. The initial drying rate (when the moisture content of the dates is 75%) is 0.5 kg water removed per kg dry matter per hour. The moisture content of the dates at equilibrium with the drying air is 8% (wet basis).

Solution:

We convert the 'wet basis' moisture content data to 'dry basis' values:

$$X_1 = \frac{75}{100 - 75} = 3 \quad X_2 = \frac{20}{100 - 20} = 0.25 \quad X_E = \frac{8}{100 - 8} = 0.087$$

$$t = \frac{X_c - X_e}{\Phi_0} \cdot \ln \left(\frac{X_1 - X_e}{X_2 - X_e} \right) = \frac{3 - 0.087}{0.5} \ln \left(\frac{3 - 0.087}{0.25 - 0.087} \right) = 5.89 \text{ h.}$$

The drying time is 5.89 hours.

EXAMPLE 22.5

Calculate the time necessary to dry a food material from 80% to 20% moisture content. The food is dried on trays, from one side. Loading rate is 10 kg per m².

Drying air data: Temperature DB = 70°C WB = 30°C, Velocity, $v = 10 \text{ m/s}$
Density $\rho = 1 \text{ kg/m}^3$.

Critical moisture of the food = 45%.

Equilibrium moisture of the food = 0

All moisture data are w/w, wet basis. Assume that there is no shrinkage.

Assume

$$h = 20 \cdot G^{0.8}$$

where $G = v\rho$

Solution:

Drying is at constant rate from 80% to 45% moisture and at falling rate from 45% to 20% moisture. Standard drying curve with sharp change at critical moisture and linear falling rate will be assumed.

For the constant rate period, drying rate N is:

$$N = -\frac{dW}{Adt} = \frac{h(T_a - T_s)}{\lambda}$$

$$h = (v\rho)^{0.8} = (10 \times 1)^{0.8} = 6.31 \text{ Wm}^{-2}\text{K}^{-1}$$

From water vapor tables, at 30°C: $\lambda = 2430 \text{ kJ/kg}$.

$$N = \frac{6.31(70 - 30)}{2430 \times 10^3} = 0.103 \times 10^{-3} \text{ kg.m}^{-2}.\text{s}^{-1}$$

The loading rate is 10 kg/m² wet material at 80% moisture i.e. 2 kg/m² dry material. Converting N to Φ :

$$\Phi = N \times 2 = 0.206 \times 10^{-3} \text{ kg/kg.s}$$

The moisture content data are converted to dry basis:

$$X_1 = \frac{80}{100 - 80} = 4 \qquad X_2 = \frac{20}{100 - 20} = 0.25$$

$$X_c = \frac{45}{100 - 45} = 0.82 \qquad X_e = 0$$

For drying time, Eq. (22.19) is applied:

$$t = \frac{X_1 - X_c}{\Phi_0} + \frac{X_c - X_e}{\Phi_0} \cdot \ln \left(\frac{X_c - X_e}{X_2 - X_e} \right)$$

$$t = \frac{4 - 0.82}{0.206 \times 10^{-3}} + \frac{0.82}{0.206 \times 10^{-3}} \ln \left(\frac{0.82}{0.25} \right) = 20.17 \times 10^3 \text{ s} = 5.6 \text{ h.}$$

The total drying time is 5.6 hours.

22.3.5 Effect of external conditions on the drying rate

1. Air velocity: increasing the air velocity accelerates both heat and mass transfer at the interface and therefore increases the drying rate *as long as the rate controlling mechanism is evaporation at the surface*. The following empirical correlation gives approximate values of the convective heat transfer coefficient as a function of air flow:

$$h = 20.G^{0.8} \qquad (22.20)$$

where:

h = coefficient of convective heat transfer $\text{w.m}^{-2}.\text{K}^{-1}$

$G = v.\rho$ = superficial mass flow rate of the air, $\text{kg.m}^{-2}.\text{s}^{-1}$ (to obtain G , multiply air velocity by the density of the air at the prevailing temperature and pressure).

Air velocity has no direct effect on the internal water transport and therefore should not affect drying rate at the falling rate period. For the entire drying period, the following approximate correlation may be used:

$$h = 20.G^{0.5} \qquad (22.21)$$

2. Air temperature: according to Eq. (22.12), drying rate in the constant rate phase is proportional to the wet-bulb temperature depression $T - T_s$. In the falling rate period, the effect of air temperature is indirect, mainly through its influence on the diffusivity of water.

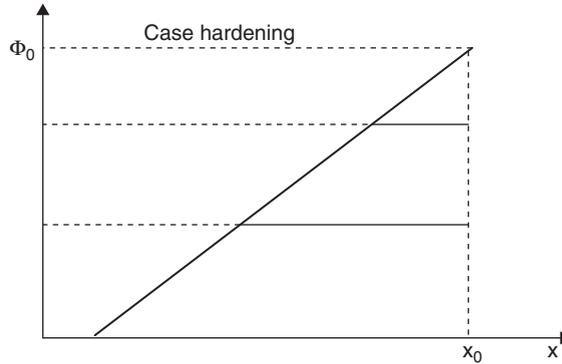


Figure 22.8 Conditions for case hardening

- Air humidity: in the constant rate period, drying rate is proportional to the difference $H_s - H$ between the adiabatic saturation humidity of the air and its actual humidity. Theoretically, the humidity of the air has no effect on the falling rate phase, except for its obvious relation with X_e through the sorption isotherm of the food.

In summary, all the external conditions affect the shape of the drying curve. Theoretically, every external factor that increases the initial drying rate (higher air temperature, lower air humidity, higher air velocity) is susceptible of increasing the critical moisture content and thus anticipating the onset of falling rate, as shown schematically in Figure 22.8. Too rapid initial drying is one of the reasons responsible for the phenomenon known as *case hardening*, whereby a dry, hard crust impermeable to moisture forms on the surface while the water content at the center is still quite high. Case hardening is generally undesirable in most food dehydration processes, but may be desirable in other cases such as bread baking and frying.

22.3.6 Relationship between film coefficients in convective drying

As stated before, the convective heat and mass transfer coefficients in air drying are interrelated.

The relationship between the mass transfer coefficients k_g and k'_g is given by the following approximate formula:

$$k'_g/k_g = 1600 \text{ Pa} \quad (22.22)$$

The heat and mass transfer coefficients are interrelated as follows:

$$h/k'_g = C_H \text{ J} \cdot \text{kg}^{-1} \quad (22.23)$$

where C_H is the specific heat of the humid air.

Following is another practical correlation between coefficients (Bimbenet et al., 2002):

$$h/k_g \approx 65 \lambda \text{ J.Pa.K}^{-1}.\text{kg}^{-1} \quad (22.24)$$

where λ stands for the latent heat of evaporation of water at the wet bulb temperature of the air.

22.3.7 Effect of radiation heating

In practice, a certain proportion of the heat is transferred to or from the drying surface by radiation, in addition to convection from the air. In some dryers, radiating bodies are installed to accelerate drying and, in this case, the proportion of heat transferred by radiation may be considerable.

Combined convection–radiation heat transfer was treated in Chapter 3. A ‘pseudo’ combined heat transfer coefficient h_r was defined. Thus, for the calculation of the constant drying rate, Eq. (22.12) is rewritten, adding the portion of the heat transferred by radiation:

$$-\frac{dW}{Adt} = K'_g(H_s - H_a) = \frac{h(T_a - T_s)}{\lambda} + \frac{h_r(T_r - T_s)}{\lambda} \quad (22.25)$$

where T_r is the temperature (in °C) of the radiating hot surface.

However, h_r depends on the temperatures of the surfaces emitting and receiving radiation. Note that the surface temperature T_s is no longer the wet-bulb temperature but higher. Thus, the solution of Eq. (22.25) requires trial and error. A surface temperature is assumed, the corresponding H_s is found on the psychrometric chart, h_r is calculated and the process is repeated until the values of H_s , T_s and h_r satisfy Eq. (22.25). For exchange between two parallel plates behaving like black bodies, h_r is given by Eq. (3.61), which becomes in this case:

$$h_r = \sigma \frac{T_r^4 - T_s^4}{T_r - T_s} \quad (22.26)$$

22.3.8 Characteristic drying curves

As explained in Section 22.3.5, the drying curve is affected by the external drying conditions (air temperature, moisture and velocity). If a given material is dried under different external conditions, the result is a family of drying curves. In practice, it is impossible and often undesirable to maintain constant external conditions. The conditions change by virtue of the process itself. As the food is dried, the air is humidified and cooled. The nearly constant air conditions required for the experimental determination of drying curves are obtained by maintaining a very high air/food mass ratio, so that the changes in air properties as a result of drying itself are negligible. How then can drying curves be used for the analysis of drying processes (calculation of drying time, estimation of the temperature and moisture distribution within the mass etc.)

where the air/food mass ratio is more realistic, such as drying a thick bed of food by air flowing through it or drying food on a long tray by air flowing longitudinally over its surface? The accepted method of calculation is to consider the bed (or the tray) as consisting of a number of finite elements (thin layers or short lengths), to calculate the extent of drying in each element as drying under constant conditions, to determine the changed conditions of the air emerging from each element and calculate drying in the next element using the changed conditions of the air. That procedure requires access to a large number of experimental drying curves.

This problem led a number of researchers to define a single characteristic curve, capable of describing the drying profile of a given product under different external conditions (Daudin, 1983). The characteristic curve is constructed by applying certain transformations on the variables of the drying curve, X and Φ . In one of those empirical models (Fornell, 1979), the following transformations are made:

$$\begin{aligned}\Phi &\Rightarrow [\Phi] = \frac{\Phi}{(T_a - T_s) \cdot v^{0.5}} \\ X &\Rightarrow [X] = \frac{X}{X_0}\end{aligned}\tag{22.27}$$

If the results of drying experiments, obtained under different external conditions are plotted as $[\Phi]$ versus $[X]$ instead of Φ versus X , all the points are supposed to fall on the same single curve. Successful grouping of the drying curves to a single characteristic curve has been reported for pieces of carrot, apple, potato and sugar beet but the method does not seem to be universally applicable (Daudin, 1983).

22.4 Drying Under Varying External Conditions

The process conditions in a dryer vary with location and time. Frequently, the conditions are changed intentionally for process optimization. For example, vegetables with high initial moisture content (e.g. onions) are usually dried by a multistage process. In the first stage, the temperature and velocity of the air are relatively high. At the final stage, during which falling rate regime prevails, air at moderate temperature and low velocity is used. In this section, however, only the changes resulting from drying itself, without external intervention will be discussed with the help of some examples. In all the examples, only drying under constant rate regime (water-saturated surfaces) will be considered. This is because at moisture content below the critical moisture, drying follows the course dictated by internal transport, with almost no effect of the external conditions on drying kinetics.

22.4.1 Batch drying on trays

Consider a tray (Figure 22.9) on which a layer of wet food is being dried with a stream of air flowing over the surface of the food (cross-flow drying). In passing over the food, the air is humidified and cooled. It follows that portions of the food are exposed

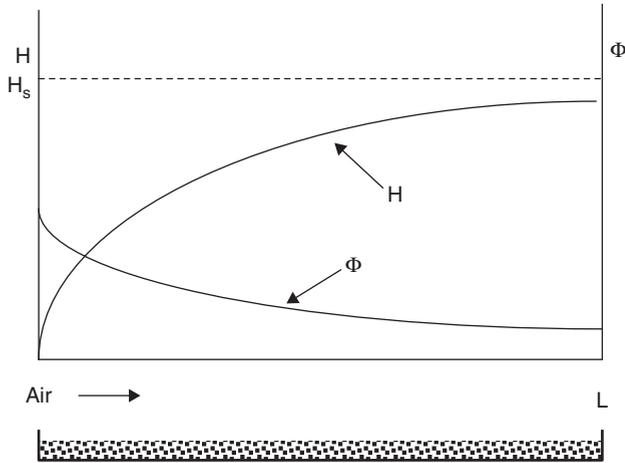


Figure 22.9 Drying rate and air humidity distribution in tray drying

to different air conditions, depending on their position on the tray. Assume that the rate limiting mechanism, everywhere on the tray, is evaporation, i.e. nowhere on the tray is the moisture content of the food below the critical moisture. It will be also assumed that the system is adiabatic.

Let dH be the increase in the humidity of the air as it passes over a segment of the tray of length dL . A material balance can be written, equating the moisture gained by the air with the water lost by the food on that tray segment:

$$G.dH = k'_g(a.dL)(H_s - H) = \left(-\frac{dW}{Adt} \right) a.dL \quad (22.28)$$

where:

G = mass flow rate of the air (dry basis), kg.s^{-1}

H = local humidity of the air (dimensionless)

H_s = adiabatic saturation humidity of the air (dimensionless)

a = width of the tray, m.

Separation of the variables and integration gives:

$$\int_{H_0}^H \frac{dH}{H_s - H} = \int \frac{k'_g a.dL}{G} \Rightarrow \ln \frac{H_s - H_0}{H_s - H} = \frac{k'_g a.L}{G} \quad (22.29)$$

Equation (22.29) indicates that $\ln(H_s - H)$ is linear with L (Figure 22.9) but, since drying is assumed to take place from water-saturated surfaces everywhere on the tray, the rate of drying is proportional to $(H_s - H)$. Hence:

$$N_L = N_0 \exp \left(-\frac{k'_g a.L}{G} \right) \quad (22.30)$$

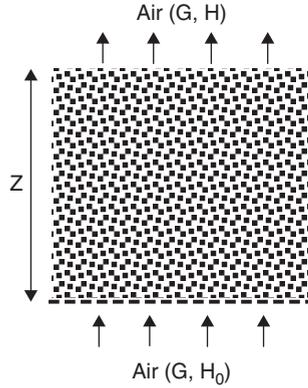


Figure 22.10 Through-flow drying

where:

$N_L = (-dW/Adt)_L$ = local drying rate per unit area at location L on the tray.

$N_0 = (-dW/Adt)_0$ = local drying rate per unit area at location 0 on the tray.

It follows that the drying rate, although constant with time, varies with location. At any time, the food closest to the entrance of the air will also be the driest and vice versa. The residual moisture content of the food will be a function of location and time.

22.4.2 Through-flow batch drying in a fixed bed

Consider a bed of humid food particles, dried with air flowing through the bed (Figure 22.10).

Assume, as in the example before, that the moisture content of every particle in the bed is still above the critical moisture and that the bed is adiabatic. Let dH be the increase in the humidity of the air as it passes through a bed thickness of dz . Let dA be the total area of the particles confined in the bed thickness dz . If the particles are assumed to be non-shrinking spheres of equal diameter d_p , one can write:

$$dA = \frac{S(1 - \epsilon)dz}{\pi d_p^3/6} \times \pi d^2 = \frac{6S(1 - \epsilon)dz}{d_p} \tag{22.31}$$

where S is the cross-sectional area of the bed.

A material balance on water is written as before:

$$G dH = k'_g \cdot dA \cdot (H_s - H) = \left[\frac{6S(1 - \epsilon)k'_g}{d_p} \right] (H_s - H) dz \tag{22.32}$$

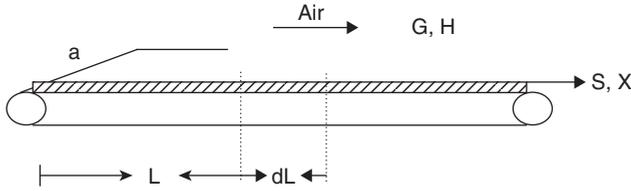


Figure 22.11 Continuous drying on a belt

Denoting as K the group of invariables in the bracket, integration gives:

$$\int_{H_0}^{H_s} \frac{dH}{H_s - H} = \frac{K}{G} \cdot \int_0^Z dz \Rightarrow \ln \left(\frac{H_s - H_0}{H_s - H} \right) = \frac{KZ}{G} \quad (22.33)$$

Again, linear relationship is found between $\ln(H_s - H)$ and the distance Z .

22.4.3 Continuous air drying on a belt or in a tunnel

Consider a belt conveyor (Figure 22.11) carrying food at the rate of $F \text{ kg}\cdot\text{s}^{-1}$ (dry basis). Air, at a flow rate of G , flows over the surface of the food.

Material balance on water, for a belt segment dL gives:

$$-F dX = N \cdot a \cdot dL \quad (22.34)$$

N is the local drying rate. Assuming again drying from a water-saturated surface, one can write N as a function of the position L , as formulated in Eq. (22.30). Combining Eq. (22.30) and Eq. (22.34) and integrating between $L = 0$ and $L = L$, we get:

$$X_0 - X_L = \pm \frac{G}{F} \cdot \frac{N_0 - N_L}{k'_g} \quad (22.35)$$

The plus and minus signs apply to co-current and countercurrent operation, respectively.

22.5 Conductive (boiling) Drying

22.5.1 Basic principles

‘Drying by boiling’ or ‘contact drying’ is a process whereby heat is transferred to the wet material from a heated surface in contact with it. The principal mechanism of heat transfer in boiling drying is conduction, with the exception of superheated steam drying where heat is transported by convection. Boiling drying is applied most commonly to foods in liquid or slurry form. Unlike air drying, the temperature of the material during most of the drying period is at or above the boiling point of the

liquid at the prevailing pressure, hence the name of ‘drying by boiling’. As stated in the introduction to this chapter, boiling drying is similar to evaporation. The main difference between the two operations is in that ‘boiling drying’ is carried on to a much lower final moisture content. Like in evaporation, the rate-controlling factor is heat transfer and mass transfer has a much weaker influence on the process.

In order to transfer heat to the boiling liquid at a sufficiently high rate, the heated surface must be at a temperature much higher than the boiling point at the process pressure. Consequently, the temperature of the food in contact with the surface may reach very high temperatures, particularly towards the end of drying. One of the two following measures, or both, are taken in order to limit product overheating:

1. The material (liquid, slurry, paste) is applied on the heated surface as a very thin layer so that drying time is fairly short
2. Drying takes place under reduced pressure (vacuum) so as to lower the boiling point of the material.

Regardless of these measures, thermal damage is generally more serious in the case of contact drying. On the positive side, contact drying is more economical in energy expenditure because heat is transferred from the primary source of energy (steam, electricity) directly to the food, without air as intermediate transfer medium.

22.5.2 Kinetics

The kinetics of contact drying, like that of air drying, shows a pattern consisting of three stages (Figure 22.12):

1. Stage 1: this is the stage of heating the feed up to the boiling point. Only a negligible proportion of the water is evaporated during this stage.
2. Stage 2: the temperature of the material is maintained at the boiling point which rises slightly as the product becomes more concentrated. The viscosity

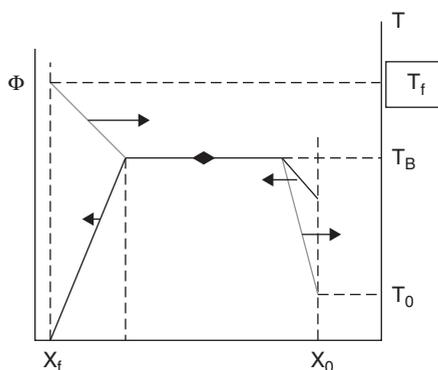


Figure 22.12 Stages in contact drying

of the material increases. Most of the water is evaporated during this stage. The rate of drying is governed by the rate of heat transfer, as shown in Eq. (22.36):

$$\frac{-dW}{dt} = \frac{-M.dX}{dt} = \frac{A.U.(T_h - T_B)}{\lambda} \quad (22.36)$$

where:

A = area of the contact surface, m².

U = coefficient of contact heat exchange, w.m⁻².K⁻¹

T_h = temperature of the heated surface, °C

T_B = boiling point of the material, °C

λ = latent heat of evaporation of water, J.kg⁻¹

Assuming constant boiling temperature and constant heat transfer coefficient, integration of Eq. (22.36) yields:

$$t = \left(\frac{\lambda}{U.\Delta T} \right) \left(\frac{M}{A} \right) (X_0 - X) \quad (22.37)$$

The time t required to reduce the water content of the food from X₀ to X during Stage 2 is therefore proportional to the temperature difference ΔT between the heated surface and the boiling liquid and to the 'loading factor' M/A, i.e. the mass of the feed (dry basis) per unit area of heated surface – hence the thickness of the food layer.

3. Stage 3: this is the stage of falling rate of drying. Because of the high viscosity of the concentrated liquid, the overall coefficient of heat transfer U is reduced and the rate of heat dissipation through the drying layer is slowed down. The temperature of the product now rises and tends asymptotically to that of the contact surface. Some porosity may be created in the drying material, causing further decrease in the rate of heat transfer but improving somewhat mass transfer. The simultaneous drop in ΔT and U results in rapid decay of drying rate. At this stage, most of the residual water in the product is adsorbed on the dry matter, therefore removal of water now takes place by desorption, until the desired final dryness is reached.

22.5.3 Systems and applications

1. Drum drying: in drum drying, the heated surface is the envelope of a rotating horizontal metal cylinder. The cylinder is heated by steam condensing inside, at a pressure in the range of 200 to 500 kPa bringing the temperature of the cylinder wall to 120–155°C. The wet material is applied on the drum surface as a relatively thin layer by a variety of different methods to be described later. The dried product is removed from the drum with the help of a blade (Figure 22.13). In vacuum drum drying, applied to materials highly sensitive to heat, the drum

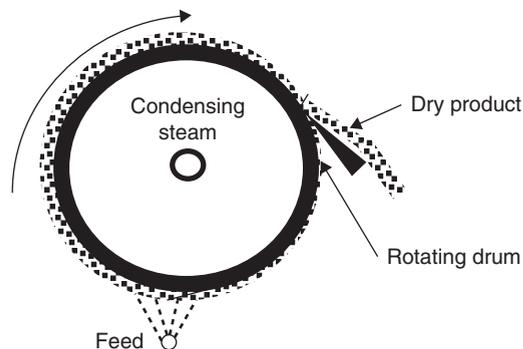


Figure 22.13 Principle of drum drying

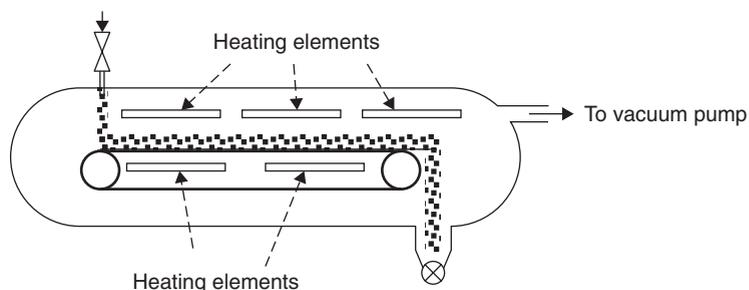


Figure 22.14 Vacuum belt dryer

and its accessories are enclosed in a vacuum chamber. Drum drying is extensively used in the production of instant mashed potatoes, pre-cooked cereals, soup mixtures and low-grade milk powder.

2. Belt drying: in this application, the heated surface is a metal belt conveyor heated by contact or radiation by hot elements installed on both sides. In one case, known as 'foam-mat drying', the belt is used for drying concentrated juices. Because of the high viscosity of the feed material, mass transfer is important right from the start of the process. In order to improve mass transfer, the concentrates are first foamed and placed on the belt as a porous mat. Some of the tomato powders available in commerce are made by foam-mat drying of tomato paste. Belt drying can also be carried-out under vacuum (Figure 22.14).
3. Vacuum tray drying: industrial vacuum tray drying is essentially a scaled-up version of the laboratory vacuum dryer. The material to be dried is spread on heat-conducting trays. The trays are placed on heated shelves enclosed in a vacuum chamber. This is a batch process, seldom used for industrial scale dehydration of foods.
4. Drying with superheated steam: in this method of dehydration, the heating medium is superheated steam. The material to be dried is brought in contact with water vapors at a temperature considerably higher than the saturation temperature at the prevailing pressure. Heat is transferred to the wet material by convection.

Water evaporates and the vapors released mix with the heating medium. Since both consist of water vapor, there is no diffusional resistance to mass transfer and the vapors from the product simply flow into the medium by virtue of a pressure difference. The temperature of the product rises up to the boiling point at the prevailing pressure and is maintained at that level as long as the water content is high. Thereafter, the temperature of the product rises above the boiling point and tends towards the temperature of the superheated steam. In a system operating at atmospheric pressure, this may result in final product temperatures in the range of 120–150°C. Although saturated steam drying can also be performed at reduced pressure (vacuum) to prevent product overheating, high pressure operation is preferred because of the greatly increased heat transfer coefficient (Svensson, 1980). Consequently, superheated steam drying is mainly used for the dehydration of materials that are not particularly prone to thermal damage, such as wood, paper and cellulose pulp (Svensson, 1980; Mujumdar, 1992). The main industrial application in the food area to date is in the dehydration of spent beet pulp in the production of sugar (Jensen et al., 1987).

The main advantage of steam drying is its excellent energy economy. Unlike air drying, all the heat supplied to the system, which is part of the excess enthalpy of superheat, is utilized for bringing the product to the boiling temperature and for evaporation. The gas leaving the dryer is live steam that can be used for heating duties elsewhere in the plant. Thus, in the case of the sugar industry, the ‘spent’ steam leaving the pulp dryer can be used for heating the evaporators, with or without recompression (see Chapter 19). The energy cost in steam drying has been reported to be 50% of that of air drying, provided that the ‘spent’ steam is utilized (Svensson, 1980). On the other hand, the capital cost of steam drying is considerably higher than that of air drying. Steam drying may have additional advantages in specific applications in the food industry, for example where sterilization of the product is an objective. In summary, although the present utilization of superheated steam drying in food processing is limited, the potential advantages of the method in specific applications cannot be ignored.

22.6 Dryers in the Food Processing Industry

The numerous types of dryers in use in the food industry may be classified in relation to different criteria:

- By method of operation: batch, continuous
- By the mechanism of heat transfer: convection (air), convection (steam), conduction (contact), radiation (infrared, microwave, sun)
- By the physical state of the feed material: solid, liquid, paste
- By the movement of the material during drying: static, moving, fluidized
- By pressure of operation: atmospheric, vacuum, high pressure.

Table 22.3 lists the main types of dryers used by the food industry.

Table 22.3 Principal types of dryers in the food industry

Dryer type	Operation	State of feed	Movement of bulk	Product examples
Cabinet	B	S	0	Fruit, veg, meat, fish
Tunnel	C	S	0	Fruit, veg
Belt	C	S, P	0	Fruit, veg, tomato
Belt-trough	C	S	M	Veg
Rotary	C	S	M	Animal feed, waste
Bin	B	S	0	Veg
Grain dryers	B, C	S	0, M	Grain
Spray	C	L, P	M	Milk, coffee, tea
Fluid bed	B, C	S	F	Veg, grain, yeast
Pneumatic	C	S	M	Flour
Drum	C	L, P	0	Mashed potato, soup
Screw conveyor	C	S, P	M	Grain, waste
Mixer	B	S	M	Particles, powders
Solar	B, C	All	All	All
Sun drying	B	S	0	Fruit, veg, fish

B = Batch; C = continuous; S = solid, L = liquid, P = paste; 0 = static, M = moving, F = fluidized

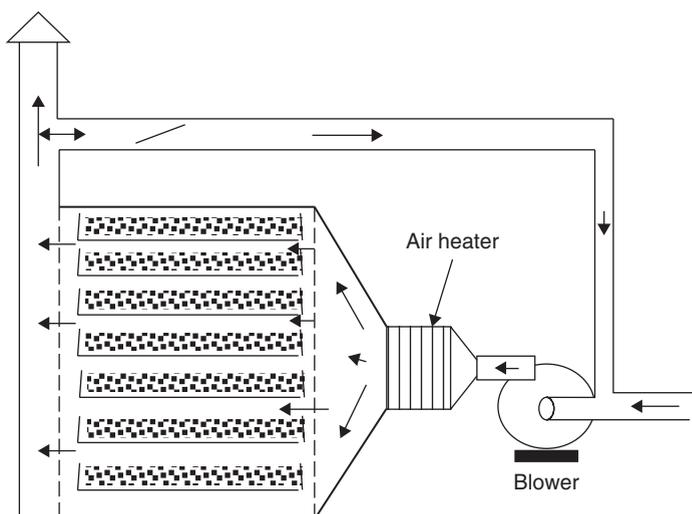


Figure 22.15 Cabinet dryer

22.6.1 Cabinet dryers

Cabinet dryers are used for batch drying of solid foods at small to moderate scale (say, 2000 to 20000kg per day). They are inexpensive and simple to construct.

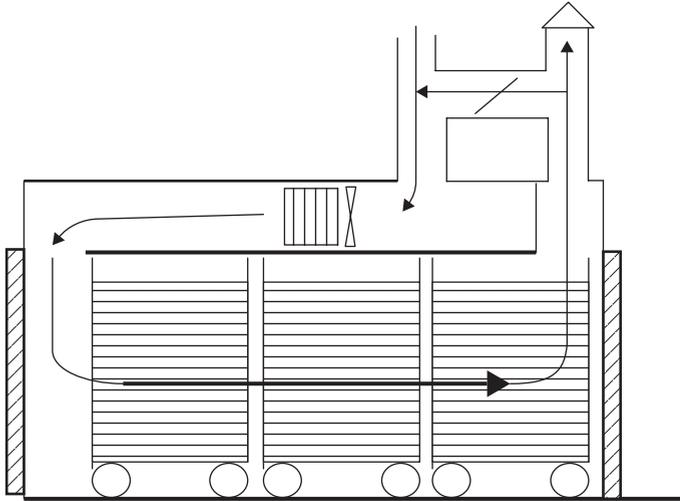


Figure 22.16 Short tunnel dryer

Cabinet dryers consist of a closed compartment in which are placed trays containing the food to be dried (Figure 22.15).

The trays rest on shelves with adequate spacing between them. Heated dry air circulates between the shelves. Very often, tray bottoms are slatted or perforated, in order to provide some air flow also through the trays. As illustrated in the example in Section 22.4.1, the drying rate, hence the moisture content of the material, depends on its position on the tray. The material located closest to the entrance of dry air has the lowest moisture content. In order to secure more uniform drying, the direction of air flow may be reversed or the trays may be rotated periodically. The cabinet is usually equipped with movable baffles, adjusted so as to have uniform distribution of the drying air throughout the cabinet. Cabinet driers are frequently found in rural installations where they are used for drying fruits (grapes, dates, apples), vegetables (onion, cabbage) and herbs (parsley, basil, mint, dill). Air inlet temperatures are usually in the range of 60–80°C. Air velocity is a few $\text{m}\cdot\text{s}^{-1}$ and must be adjusted according to the size, shape and density of the food particles so as to avoid entrainment of dry particles with the wind. Depending on the product and the conditions, the duration of a batch is typically 2 to 10 hours. Most cabinet dryers feature means for adjustable recirculation of the air. The rate of recirculation is increased as drying progresses, when the air exiting the cabinet is warmer and less humid. Recirculation results in considerable saving in energy cost.

22.6.2 Tunnel dryers

Tunnel dryers consist of long tunnels through which trucks carrying stacks of trays travel with or against a stream of drying air (Figure 22.16). The material to be dried is evenly spread on the trays. Typical tray loading for wet vegetables is in the order of 10–30 kg per m^2 . As one truck with wet material is introduced into the tunnel at one end, another truck, carrying dehydrated product, exits at the other end. Depending on

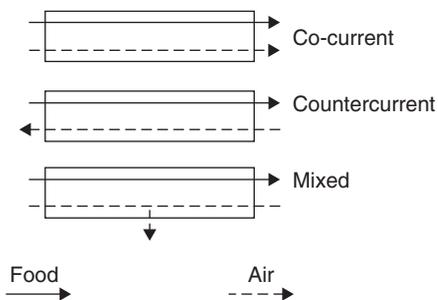


Figure 22.17 Flow patterns in tunnel dryers

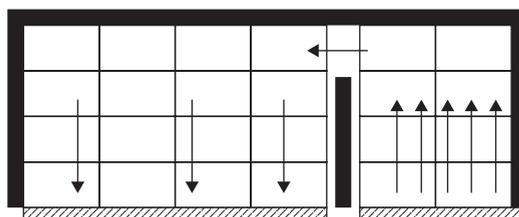


Figure 22.18 Two-stage tunnel dryer

the size of the trucks and the tunnel, the trucks are moved manually or mechanically, e.g. with the help of chains.

With respect to the relative direction of movement of the air and the trucks, tunnel dryers operate in co-current, countercurrent or mixed current fashion (Figure 22.17). In the case of the co-current tunnel, air with the highest temperature and lowest humidity meets the food with the highest humidity and lowest temperature. This provides the highest ‘driving force’ for drying and therefore the most rapid rate of water transfer at the entrance to the tunnel. If the feed material is sufficiently humid, its temperature remains low despite the contact with hot air. The ‘driving force’, however, diminishes as the food travels towards the exit. The air at the exit end of the tunnel is the most humid and the coolest. Consequently, the final residual moisture content of the product may not be as low as desired. The contrary occurs in the case of countercurrent tunnels. The starting rate of drying is lower but it is possible to dehydrate the product to the lower final moisture content. The mixed flow, central exhaust tunnel functions as two tunnels in series. Its first portion is co-current and provides the desired high initial drying rate. Its last portion is countercurrent and gives the desired finishing effect. Unlike cabinet drying, tunnel drying provides the possibility of exposing the product to a changing profile of external conditions. In addition to air temperature and humidity, it is possible to vary air velocity.

In one model used for drying fruit, the tunnel is designed as two units in series, with a smallest cross-section, resulting in higher air velocity in the first unit (Figure 22.18).

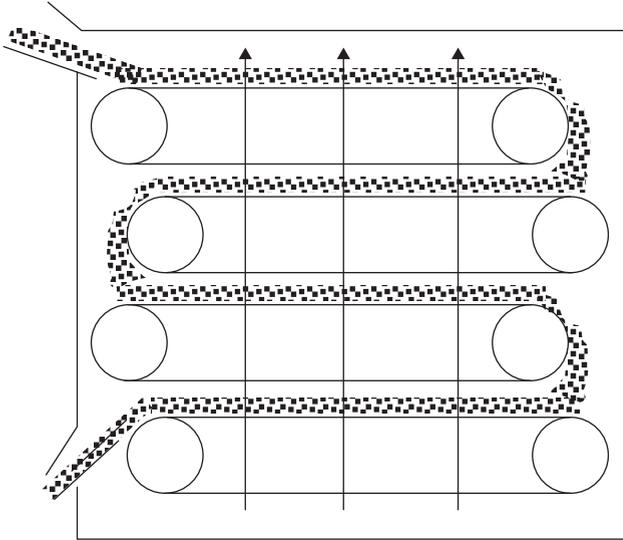


Figure 22.19 Multistage belt dryer

22.6.3 Belt dryers

Belt dryers are among the most versatile continuous dryers for solid foods. They are extensively used, mainly for large-scale dehydration of vegetables. Essentially, a belt dryer functions like a tunnel dryer with the difference that trays and trucks have been replaced by belt conveyors. These dryers can be operated in cross-flow or through-flow mode or a combination of both. In dryers for through-flow operation, the belt is made of metal mesh to allow air circulation through the bed. Belt dryers can be single-stage or multistage. Multistage dryers consist of a number of belts in series.

The advantages of belt dryers over tunnel dryers are:

- Continuous feeding and discharging of the belts are much easier and less labor-intensive than loading and unloading trays
- In tray dryers (tunnel or cabinet), the bed is static during the entire process. In the absence of mixing, the moisture content of the food on a tray is uneven. The top surface of the bed may be quite dry while the material at half-depth is still wet. In multistage belt dryers, the bed is refilled and mixed at each transfer from one belt to another (Figure 22.19).
- Most food materials shrink upon drying. In tray dryers, shrinking results in diminishing rate of utilization of the tray surface as drying progresses. In multistage belt dryers, it is possible to maintain adequate bed thickness by moving each stage at a slower speed than the previous one (Figure 22.20). Thus, higher loading per m^2 can be achieved in belt dryers.

22.6.4 Belt-trough dryers

This is a special type of dryer, designed for the initial drying of vegetables in small pieces. The unit consists of a wide mesh belt, freely hanging over two cylindrical

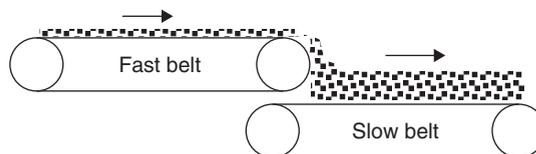


Figure 22.20 Effect of varying belt speed on bed thickness

rollers so as to form a trough. The wet material is fed onto the belt at one end of the trough and forms a fairly thick bed. The belt is driven slowly by rotating the supporting rollers. The movement of the bed causes gentle and continuous mixing of the bed, while a vertical blast of hot air circulates through the material. The blast is sufficiently strong to expand the bed slightly without fluidizing it. Evaporation is rapid and most of the water in the wet material is removed, typically in less than one hour. The trough assembly is tilted towards one end, causing the material to move slowly down the slope towards the discharge. The rate of discharge, hence the holdup of the unit is regulated by an adjustable weir placed at the discharge end.

22.6.5 Rotary dryers

Rotary dryers are mainly used in the chemical and mineral industry. In the area of food their most common applications are for dehydrating waste materials (citrus peels, vegetable trimmings) and animal feedstuffs (alfalfa). Rotary dryers consist of a metal cylinder with internal flights or louvers (Figure 22.21). The cylinder is slightly inclined. The material is fed at the high end and discharged at the low end. Hot air is blown in a co-current or countercurrent direction. As the cylinder rotates, the material climbs in the direction of rotation. When it reaches a position where its angle of repose has been exceeded, the material falls back to the bottom of the cylinder. Most of the drying takes place while the material falls through the air blast. Using very hot air or combustion gases, rotary dryers can also function as roasters for nuts, sesame seeds and cocoa beans. A detailed method for the design of rotary dryers, based on a heat exchange approach has been described by Nonhebel (1971).

22.6.6 Bin dryers

Tray, belt and belt-trough dryers are quite efficient in removing most of the water of high-moisture materials in the initial stage of drying. In the last stages of the falling rate period, however, removal of the residual moisture takes a long time and external turbulence and mixing cannot accelerate the process. In the case of vegetables, using the more expensive types of dryers for reducing the moisture content below 15–20% would be uneconomical. Bin dryers provide the ideal solution for ‘finishing’ the process to the desired final moisture content of 3–6%.

As their name implies, bin dryers consist of simple containers with a perforated or mesh bottom (Figure 22.22). The partially dehydrated material is placed in the bin as a deep bed. A slow current of air at moderate temperature is passed through the bed

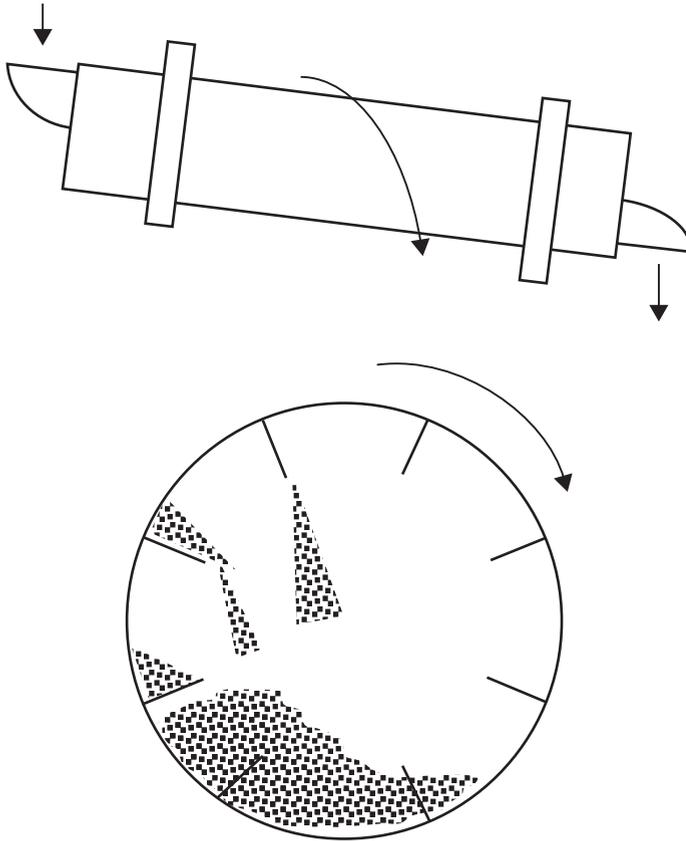


Figure 22.21 (a), (b) Generalized schema of the rotary dryer

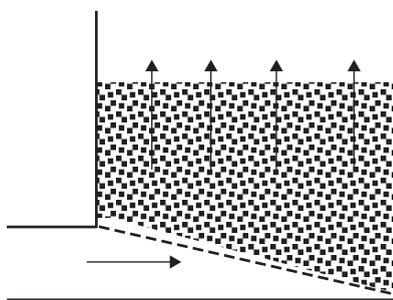


Figure 22.22 Bin dryer

until the desired final dryness is reached. Another function of the bin dryer is moisture equilibration. As explained before, at the exit from the more ‘rapid’ dryers, the moisture content distribution between and within particles is not uniform. The long residence in the bins allows moisture equilibration in the product as a result of internal transport.

22.6.7 Grain dryers

Cereal grains and oilseeds are often harvested with a moisture content that does not warrant safe storage for a long period. In this case, it is necessary to dry the grains before storage. In terms of tonnage, grain drying is probably the largest drying operation in food industry and agriculture. Grain drying is also called 'conditioning'. The initial moisture content depends on the climatic conditions during the harvest season. Typically, grain may contain 25% moisture when harvested and must be dried to 12–15% moisture before storage. Because of the relatively low moisture content throughout, drying takes place entirely under the falling rate regime and takes a long time, e.g. several days. Air temperatures used in grain drying are, typically, 50 to 70°C, the lower temperatures being applied to grain for seed and the upper temperatures to grain for processing.

Grain dryers may be batch or continuous. Batch dryers function as very large bin dryers, except that special precautions must be taken for the proper distribution through the bed which can be many meters high. The wet grain enters through the top. Often, the upper section of the bin or tower is separated from the rest. There, the wet grain is pre-dried and then dumped to the section below. In order to avoid heat damage, dried grain is usually cooled by passing unheated air through the bed. Frequently, the batch dryer also serves as a long-term storage silo. In continuous grain dryers, hot air is passed through continuously flowing or agitated grain. After drying, the grain is continuously cooled with air before being discharged. Drying rates are high. Precise control is required in order to avoid over-drying and over-heating. A period of 'tempering' is provided for moisture equilibration.

22.6.8 Spray dryers

Spray dryers are used for drying liquid solutions and suspensions, with the objective of producing light, porous powders. Spray drying is the accepted method for the production of milk and whey powders, coffee creamers, cheese powder, dehydrated yeast extract, instant coffee and tea, isolated soybean protein, enzymes, maltodextrin, egg powder and many other products in powder form. Spray drying is also one of the methods used for micro-encapsulation.

The liquid is dispersed (atomized) as a spray of fine droplets into very hot air inside a large chamber. Because of their small size and the high temperature of the air, the droplets are dried in a matter of seconds and transformed into particles of solid powder. At the exit from the chamber, the solid particles are separated from the humid air. A spray dryer system consists of the following elements (Figure 22.23):

- An air heater
- A device for forming the spray (atomizer)
- A pump for feeding the liquid to the atomizer
- A drying chamber
- Solid–gas separators (cyclones)
- Fans for moving the air through the system

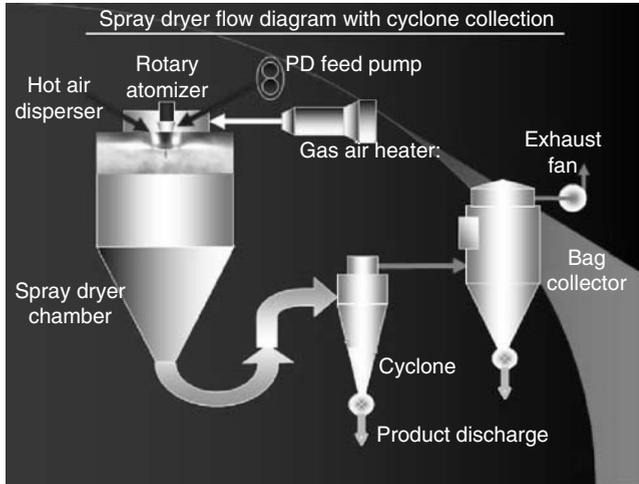


Figure 22.23 Spray dryer. (Courtesy of GEA-Niro)



Figure 22.24 Rotary atomizer. (Courtesy of GEA-Niro)

- Control and measurement instruments.
 1. *Air heater*: inlet air temperatures in spray drying are in the order of 200–250°C and sometimes higher. Steam cannot be used for heating the air, because of the high pressure that would be required to deliver heat at such high temperatures. Electrical heating is practiced in small installations. Combustion gases are the preferred source of heat at industrial scale. In direct heating, the combustion gases at 400–500°C are mixed directly with fresh air to produce a gas mixture at the desired temperature. In indirect heating, air is heated by combustion gases in heat exchangers, without direct contact between the two streams. Direct heating is simple, less expensive, provides excellent energy utilization but the direct contact between the food and combustion gases may be problematic.
 2. *Formation of the spray*: there are five methods for dispersing the feed as a spray: centrifugal (turbine) atomizers, pressure nozzles, bi-fluid nozzles, hot air (Leafflash) dispersers and ultrasonic atomizers.

Centrifugal (rotary) atomizers (Figure 22.24) consist of a wheel, similar to a closed impeller in a centrifugal pump.



Figure 22.25 Pressure nozzle. (Courtesy of GEA-Niro)

The atomizer, which can be motor-driven or moved by compressed air, spins at high speed, corresponding to tip velocities, in the order of $10^2 \text{ m} \cdot \text{s}^{-1}$. The liquid is fed to the center of the wheel, flows through the channels to the periphery and is ejected as a film that disintegrates to droplets. Centrifugal atomizers form sprays with fairly narrow droplet size distribution. Uniformity of drop size is important for even drying and for producing a powder with uniform particle size distribution. The mean Sauter diameter (see Section 6.2.1) of the droplets depends on the wheel diameter D , speed of rotation N , mass flow rate G , liquid viscosity μ , density ρ and surface tension σ , according to the following approximate empirical equation (Masters, 1991):

$$\bar{d}_{SV} = k(N)^{-0.8}(G)^{0.2}(D)^{-0.6}(\mu)^{0.2}(\rho)^{0.5}(\sigma)^{0.2} \quad (22.38)$$

Equation (22.38) indicates that viscosity, surface tension and feed flow rate have little influence on the drop size. Uniformity of drop size is the main advantage of centrifugal atomizers. Their disadvantages are high capital and maintenance cost and high energy consumption.

Pressure nozzles (Figure 22.25): The liquid is fed to a narrow nozzle at high pressure. At the exit from the nozzle the liquid jet disintegrates to form the spray. Nozzles that deliver a conical spray are selected. The droplets are relatively large and size distribution is wide.

The following approximate relation expresses the dependence of mean droplet size on pressure drop ΔP and other variables (Masters, 1991). Note the strong dependence on viscosity.

$$\bar{d}_{VS} = k \left(\frac{G}{\rho} \right)^{0.25} \mu (\Delta P)^{-0.5} \quad (22.39)$$



Figure 22.26 Bi-fluid nozzle. (Courtesy of GEA-Niro)

Bi-fluid (kinematic) nozzles: the liquid and pressurized air are supplied separately to the nozzle (Figure 22.26). The two fluids meet at the exit from the nozzle. The liquid is disintegrated by the high velocity air jet. The droplets are finer and more uniform than those in sprays delivered by pressure nozzles.

'Leaflash' atomization: 'Leaflash' is a novel spray-drying system with a special atomization mechanism. In the atomizer head of this system, very hot air (typically at 300–400°C; Bhandari et al., 1992) at high velocity tears a film of fluid into droplets. The air that induces atomization is also the air used for drying. Because of the intimate contact with very hot air, the droplets are dried even more rapidly than in classical spray drying. Consequently, the residence time of the product in the dryer is particularly short.

Ultrasonic atomization: if a surface, such as a rod or a disc, wetted with liquid, is set to vibrations at ultrasound frequency and appropriate amplitude, the liquid film is atomized as very fine droplets. This is the operating principle of the ultrasonic nozzle (Figure 22.27), used for spray-drying and micro-encapsulation (Bittner and Kissel, 1999; Yeo and Park, 2004).

3. *Feed pump:* the starting materials for spray drying are often highly viscous solutions or suspensions. Frequently, the feed has to be pumped to a considerable height, from ground floor to the top of the spray dryer. If the spraying devices are pressure nozzles, the liquid must be delivered to the nozzle at MPa range pressure. For all these reasons, positive displacement pumps are used as feed pumps. Furthermore, the controlled variable in spray drying is the feed flow rate (see below). Therefore, feeding pumps must be equipped with automatically controlled variable speed drive.
4. *Drying chamber:* the drying chamber consists, most commonly, of a vertical cylindrical section with a conical bottom. Industrial spray dryers are very large in diameter and in height and often occupy an entire building. The large volume of the drying chamber is required in order to provide the residence time

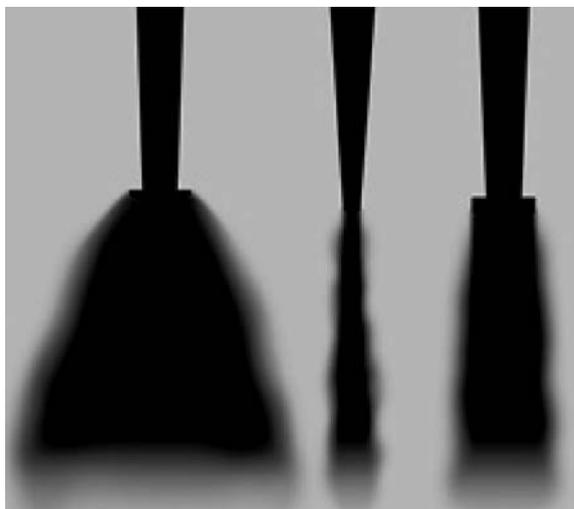


Figure 22.27 Different types of sprays obtained by ultrasonic atomization. (Courtesy of SonoTek Corporation)

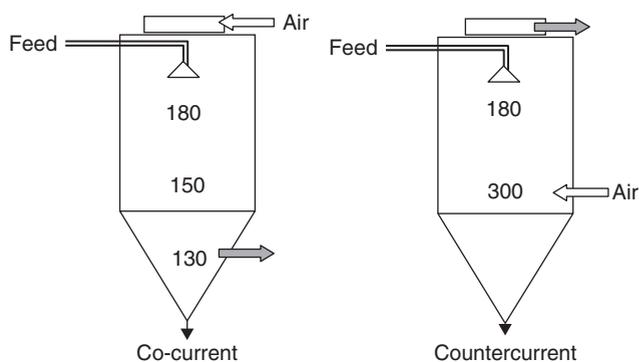


Figure 22.28 Typical temperature distribution in co-current and countercurrent spray dryers (inlet air temperature in both cases is 350°C). (Adapted from Masters, 1991)

necessary for complete drying. In spray dryers with centrifugal atomizers, the droplets are ejected in a radial direction. In this case, a large diameter is required in order to prevent the droplets from reaching the chamber walls while they are still wet and sticky. Dryers with pressure or bi-fluid nozzles are tall but less wide, because of the narrower spray angle. The conical bottom serves to collect the major part of the dried product but adds to the height of the dryer. A spray dryer of a different model has a nearly flat bottom and uses a rotating suction duct (air broom) for collecting the product.

5. *Air flow*: air is driven through the heater and the dryer by low pressure, high capacity blowers. Air movement may be co-current or countercurrent to the direction of the product. The typical temperature profile in the dryer for the two types of flow pattern is shown in Figure 22.28. A certain portion of the dry

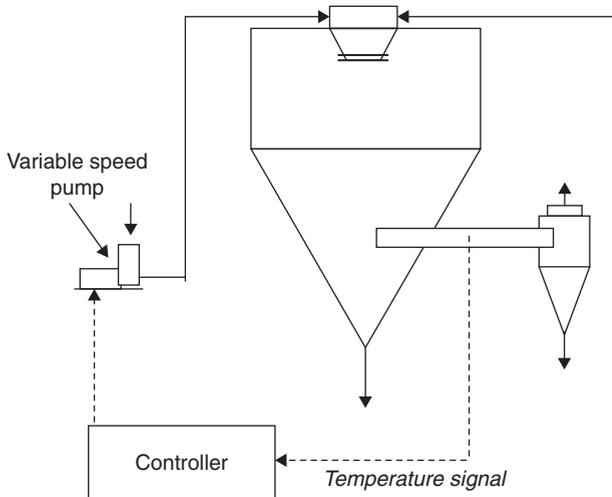


Figure 22.29 Control of a spray dryer

particles and particularly the fines are carried away with the exhaust air and recovered in cyclones. For environmental reasons, it is often necessary to pass the exhaust air through a scrubber before discharge to the atmosphere.

6. *Control:* because of the rapidity of the process, it is difficult to control exactly the residual moisture content of the product and to adjust operating conditions accordingly. The approximate method of control most commonly applied assumes that the temperature of exhaust air is related to the residual moisture content of the product. For foods, exhaust temperatures in the range of 90 to 110°C are specified. If the exhaust temperature is too low, it is assumed that the product is too moist. In this case, the controller reduces the feed rate by regulating the speed of the feed pump. Contrarily, high exhaust temperature indicates poor utilization of the drying power of the air, with possible thermal damage to the product. The controller increases the feed rate accordingly (Figure 22.29). An additional independent control loop regulates the inlet temperature of the air.

Due to the difficulty of precise control of the final moisture content, it is often necessary to include a second stage of drying, usually in a fluidized bed dryer (Figure 22.30). This second stage may also serve for agglomeration.

The relationship between product quality and process conditions in spray drying is discussed in Section 22.7.

22.6.9 Fluidized bed dryer

The properties of fluidized beds and the conditions for fluidization have been discussed in Section 2.5.3. In fluidized bed dryers, hot and dry air is used both for fluidization and for drying. Fluidized bed drying can be applied to particulate, non-sticky

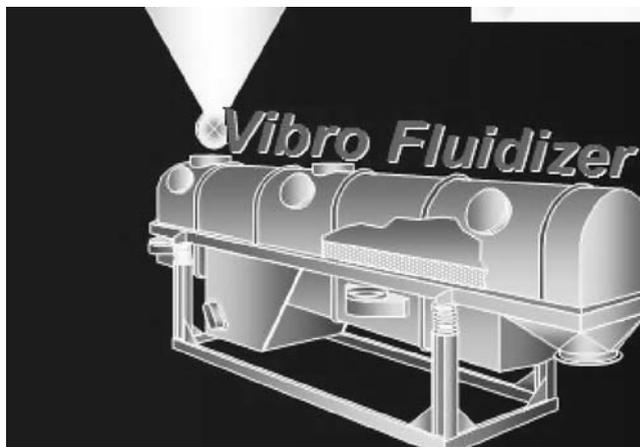


Figure 22.30 Vibrating belt fluidized bed after-dryer. (Courtesy of GEA-Niro)

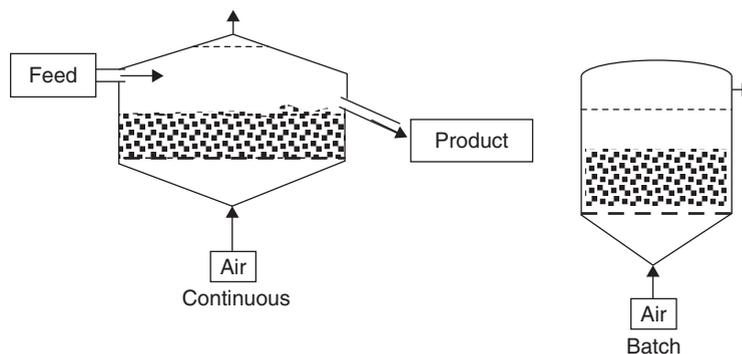


Figure 22.31 Schema of fluidized bed drying

foods with particle size within the range of 0.05 to 10 mm depending on the density. Fluidized bed drying can be in batches or continuous (Figure 22.31). Due to the efficient heat and mass transfer, the product is dried rapidly. Sticking and product accumulation in continuous dryers is largely prevented by vibrating the fluid bed. Fluidized bed dryers are also used for powder agglomeration and coating.

22.6.10 Pneumatic dryer

The basic principles of pneumatic transport of particulate solids were discussed in Section 2.5.4. In pneumatic dryers, the particles are dried while being moved in a stream of hot, dry air. This method is used for removing 'free' moisture in the constant drying rate period. High drying rates can be achieved (hence the name of 'flash dryer'), but residence time is too short for complete drying. For this reason, pneumatic drying is often used as a method of 'pre-drying', followed by another type of drying. It is also possible to recycle part of the product in order to achieve the desired final moisture content. The use of pneumatic drying in the food industry is limited to the drying of flours, starch, gluten powder, casein powder etc.



Figure 22.32 Drum dryer. (Courtesy of the Dept. of Biotechnology and Food Engineering, Technion, I.I.T.)

22.6.11 Drum dryers

The basic principle of operation of the drum dryer was treated in [Section 22.5.3](#). The different types of drum dryers vary in the method used for applying the wet material on the surface of the drum ([Figure 22.32](#)).

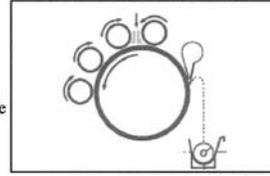
Drum dryers are classified into two types, namely, single drum and double-drum dryers. Double-drum dryers consist of two drums rotating in opposite directions, with a narrow, adjustable gap between the two. The so-called ‘twin’ drum dryer consists, in fact, of two independent co-rotating single drums sharing some of the accessory devices.

The simplest method of application is dip feeding. The drum is partially immersed in the feed fluid contained in a tray. A film of fluid adheres to the immersed segment of the drum. Fresh material is continuously supplied to the tray. Because the mass of material in the tray is heated by the drum, this method is less suitable for heat sensitive products. In the double-drum dryer with nip feed, the feed fluid is introduced into the pool formed between the two drums. The thickness of the adhering film is controlled by adjusting the gap. This type of feeding is used with less viscous materials, such as milk and other dairy products. Applicator rolls are used for spreading viscous liquids, purees and pastes on the drum surface. Single drum dryers with applicator rolls are extensively used in the production of instant mashed potato. Multiple applications, resulting in thicker layers, are made possible by mounting a number of applicator rolls on the drum periphery ([Figure 22.33](#)). Application rolls are also useful for pressing the film upon the drum, to restore good contact and to reduce porosity.

SINGLE DRUM DRYER WITH APPLICATOR ROLLS

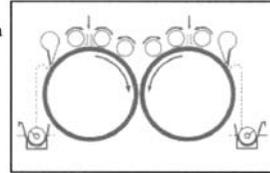
The wet product is applied to the drum by means of applicator rolls. According to the number of the applicator rolls used the layer formed on the drying drum is thicker or thinner. This arrangement is suitable for the processing of pulpy or pasty products.

Typical applications are: Cereal based breakfast foods, Babyfood products, Pre-gelatinized starches, Fruit pulp and pastes, Potato flakes.

**DOUBLE-DRUM DRYER WITH FEED ROLLS**

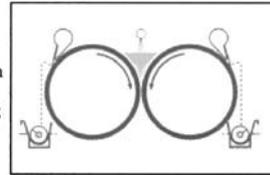
This is a special 'Hybrid' system used to impart certain physical properties, particularly density, to specific products

Principle applications are: Cereal based breakfast foods, Baby food products, Fruit pulp and pastes.

**TWIN CYLINDER DRYER WITH NIP FEED**

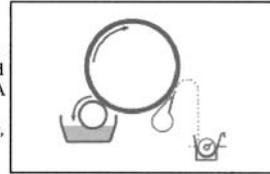
The material to be dried is pumped, either directly or through spray nozzles, into the nip formed between two drying drums. This is the oldest and simplest form of drum dryer. The thickness of the product film may be varied by adjustment of the gap between the drying drums or cylinders.

Typical applications are: Drying of Yeast, Milk products, Detergents, Dyestuff manufacture.

**SINGLE DRUM DRYER WITH APPLICATOR ROLL**

This is a more specialized type of drying, for example in the Chemical Industry. The applicator roll is located underneath the drum dryer and dips into the product. A liquid film is then transferred to the drying drum.

Typical applications are: Animal based glue, Gelatine, Pesticides.

**DIP FEED DRUM DRYER**

One of the most basic forms of the drum dryer, where a film of the product to be dried is picked up on the surface of the dryer drum as it rotates through a feed tray mounted below. The feed tray may be cooled or fitted with a recirculation system to prevent overheating or settling out of product from suspension.

Typical applications are: Drying of cereals, Spent Yeast.

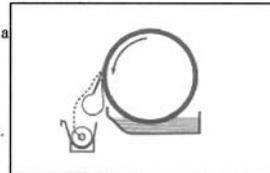


Figure 22.33 Drum drying feeding methods. (Courtesy of Simon Dryers)

Removal of the water vapors from the vicinity of the drums is essential in order to prevent moisture adsorption by the dry product. To this end, drum dryer installations are usually equipped with venting hoods of adequate size.

22.6.12 Screw conveyor and mixer dryers

Screw conveyors with jacketed troughs and hollow augers are used for drying slurries and wet particulate solids (Waje et al., 2006). The trough and screw auger are steam heated. Heat transfer is mainly by conduction. These dryers, also known as *hollow flight* dryers, are frequently used for drying waste materials and biomass (sludge) from wastewater treatment processes.

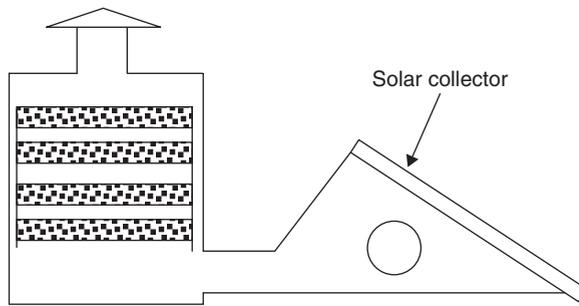


Figure 22.34 A simple solar dryer

Dryers built on the principle of ribbon, double-cone or V tumbler mixers (Section 7.5.3) are used for small-scale batch drying of solids. The mixers are jacketed for heating with steam or hot water. Some models operate under vacuum.

22.6.13 Sun drying, solar drying

Sun drying (Bansal and Garg, 1987; Ekechukwu and Norton, 1994) refers to the dehydration of foods by direct exposure to radiation from the sun. Important quantities of fruits, vegetables, grains and fish are dried by this method. Sun dried tomatoes are a specialty product of increasing popularity. A large proportion of the raisins and practically all of the dried apricots and figs produced in the world are sun dried. Sun drying of fish at village level is common in tropical regions. The term 'solar dryer' is reserved for a large variety of convective dryers whereby the products are not exposed to sun but dried indirectly by air heated by solar energy. An example of solar dryer is shown in Figure 22.34. Obviously, both types of drying are feasible only if sufficient and reliable insolation can be assumed.

22.7 Issues in Food Drying Technology

22.7.1 Pre-drying treatments

Most vegetables are blanched before drying. The objectives of blanching, exposed in the chapter on freezing, are also applicable to drying. In addition, blanching often has a beneficial effect on the drying rate. In order to preserve their characteristic pungency, onions are usually not blanched.

Fruits with tough skins, such as prunes, figs and grapes are dipped in hot water or a hot solution of potassium carbonate. This treatment (checking) induces small cracks, removes the waxy layer covering the skin and accelerates drying. Dipping in a solution of sodium bisulfite is often practiced in the case of grapes (bleached raisins), apricots and some vegetables. Some producers dip carrot pieces in a hot solution of starch. Upon drying, the starch forms a protective coating that retards oxidation of the carotenoid pigments (Zhao and Chang, 1995). Dipping in solutions

of ascorbic acid, citric acid or a mixture of both is effective in preventing browning discoloration in vegetables.

The time–temperature profile of the product during drying is usually insufficient for the destruction of microorganisms. Therefore, thermal processing before drying is required in critical cases such as milk and dairy products.

The most common treatment given to liquid foods before drying is pre-concentration. As explained in [Section 22.8](#), removal of water by drying is more costly than by evaporation. Milk and coffee extract are pre-concentrated before spray drying. The upper limit of concentration before drying is dictated by the viscosity of the concentrate that can still be successfully spray dried.

22.7.2 Effect of drying conditions on quality

Heat and mass transport phenomena may have profound effects on the quality of dehydrated foods. Texture, appearance, color, flavor, taste and nutritional value are all subject to change as a result of the drying process.

Food structure is changed by drying ([Wang and Brennan, 1995](#); [Aguillera et al., 2003](#)). The most general and the most evident effect, also probably the most difficult to avoid, is shrinkage. In foods consisting of soft, pliable, hydrophilic gels not containing rigid fillers, the decrease in volume corresponds roughly to the volume of the removed water and shrinkage is isotropic. The shape of the particle is therefore not affected. This case is, however, not very frequent in solid foods. Usually the solid matrix has some resilience and therefore shrinkage is less than the volume of removed water. Mechanical stresses are induced and deformation occurs ([Earle and Ceaglske, 1949](#); [Eichler et al., 1998](#); [Bar et al., 2002](#)). The dried product is porous ([Krokida and Maroulis, 1997](#)).

During drying, solutes are transported along with water. As a result, the distribution of components in the dried product may be different from that of the starting material. The concentration of solutes such as sugars and proteins may be higher at the surface of the dehydrated product ([Fäldt, 1995](#)).

The most common thermal effects are non-enzymatic browning, denaturation of proteins and thermal destruction of heat sensitive vitamins and pigments. Protein denaturation is the main cause of decrease in the dispersibility/solubility of milk powder obtained by drying at high temperature.

With respect to the effect of drying on product quality, spray drying occupies a peculiar position ([Patel and Chen, 2005](#)). Because of the extremely fast drying rate, short residence time of the food in the dryer and the relatively low temperature during most of the drying period, thermal damage in spray drying is remarkably low. Even when suspensions of live cells (starter cultures, yeast) are spray dried, the survival rate of the cells in the process is remarkably high ([Bonazzi et al., 1996](#)).

The retention of volatile aromas in spray drying is also much better than what could be expected from the volatility of the aroma relative to water ([Senoussi et al., 1995](#); [Hecht and King, 2000](#)). The retention of organic trace components in drying in general and in spray drying in particular has been extensively investigated ([Coulter and Reineccius, 1969](#); [Rulkens, 1973](#); [Thijssen, 1975](#); [Etzet and King, 1984](#); [Bangs and](#)

Reineccius, 1990). Tijssen proposed the 'Selective Diffusion Theory' to explain the retention of volatiles in spray drying. According to this theory, as the concentration of carbohydrates increases during drying, the diffusivity of the organic trace components decreases faster than the diffusivity of water. One of the practical consequences of his theory is that the retention of volatile aroma increases as the solids content of the feed solution is increased.

22.7.3 Post-drying treatments

As a result of thermal damage during drying, the dried product may contain defective particles. These are removed by visual or automatic inspection and sorting. Automatic color sorters are used in large-scale production of dehydrated vegetables.

Some dehydrated products in powder form (e.g. milk powder, infant formulae, soluble coffee) are agglomerated. Agglomeration as a unit operation will be discussed later in this section.

Considerable quantities of dehydrated herbs and spices are irradiated with ionizing radiations in order to meet microbiological standards. Some dried products, such as soluble coffee, are highly hygroscopic and must be protected against adsorption of moisture immediately after drying. For the intermediate storage and packaging of such products, air conditioned rooms kept at low relative humidity must be provided. Proper packaging and adequate storage conditions are, of course, the most important factors in post-drying handling of all dehydrated products.

22.7.4 Rehydration characteristics

Dehydrated products may be classified into those that are used as such (e.g. most dried fruits) and those that must be 'reconstituted' by rehydration (e.g. milk powder, mashed potato flakes, most dried vegetables). Ideally, dehydrated foods should be able to regain their original moisture content, volume, shape and quality when rehydrated. This, however, seldom occurs. A number of indices are used to define the rehydration characteristics of dehydrated foods (Lewicki, 1998). 'Rehydration ratio' is the ratio of the mass of rehydrated and drained food to the mass of the original material. It is usually expressed as a percentage. 'Rehydration rate' or conversely 'rehydration time' refer to the kinetics of rehydration. The 'rehydratability' of dehydrated foods depends largely on the conditions of drying (Saravacos, 1967; Saguy and Marabi, 2003; Bilbao-Sáinz et al., 2005). High porosity seems to be among the most important requisites for complete and fast rehydration (Marabi and Saguy, 2004). The rate limiting factor is adsorption and internal transport of the water. Consequently, agitation has only a slight effect on rehydration kinetics, except if the rehydration medium is highly viscous (Marabi et al., 2004).

Rehydration of dried foods involves a sequence of events: wetting of the surface, penetration of the water into the pores, adsorption on the surface of the matrix, diffusion into the solid matrix and equilibration. It should be kept in mind that, simultaneously with the penetration of water, there may be leaching of soluble components out of the food particle. In the case of spray dried 'instant' powder, the rehydration characteristics

of the product are evaluated in terms of the following properties ([Barbosa-Cánovas and Vega-Mercado, 1996](#)):

- **Wettability:** this is the ability of the powder to absorb water on its surface. This property can be evaluated quantitatively in terms of surface tension and wetting angle, or empirically by measuring the rate of penetration of water into a column of powder
- **Sinkability:** for rapid reconstitution, a quantity of powder placed on the surface of water should sink down into the water as quickly as possible
- **Dispersibility:** this property represents the ability of the powder to disperse quickly and evenly into the water, without forming lumps
- **Solubility:** this property is a function of the chemical composition but may be affected by drying (e.g. denaturation of proteins).

22.7.5 Agglomeration

Agglomeration ([Schubert, 1981](#); [Ormos, 1994](#)) is a ‘size augmentation’ process applied to powders. The most commonly applied process (‘rewet agglomeration’) consists of wetting the powder with water or with an aqueous solution, followed by re-drying under agitation. Wetting creates ‘liquid bridges’ between powder particles. These bridges contain solutes, either dissolved from the powder or added previously to the wetting liquid. Upon re-drying the liquid bridges are converted to ‘solid bridges’, binding powder particles together. Agitation during drying is essential for controlling the ultimate size of the agglomerates.

The objectives of agglomeration are:

- To produce a rigid porosity and thus prevent lump formation and improve rehydration properties of ‘instant’ powders ([Schubert, 1993](#))
- To prevent segregation of components in composite powders
- To reduce the proportion of ‘fines’
- To control the bulk density of the powder
- To improve flow characteristics.

The methods of agglomeration differ from each other, mainly in the manner that agitation is applied. Thus, agglomeration can be performed in a rotating pan, in a tumbler mixer or in fluidized bed. An example of a fluidized bed method for powder agglomeration is shown in [Figure 22.35](#).

22.8 Energy Consumption in Drying

Drying processes usually consume considerable quantities of energy. A number of different sources of energy are used in drying processes: fuel, steam, hot water,

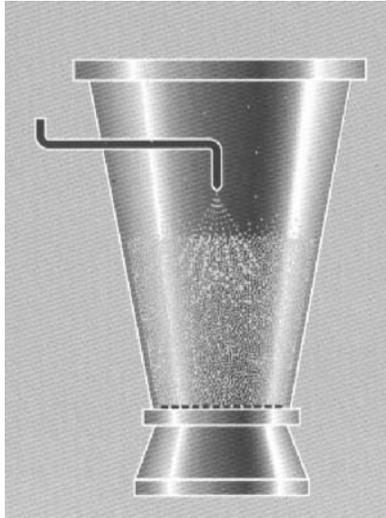


Figure 22.35 Top-spray fluidized bed agglomeration. (Courtesy of Niro)

electric current, solar heat etc. The main parameters used to characterize the energy consumption of drying processes are:

1. The specific energy consumption: this is the energy consumed per unit mass of product. This parameter has economic significance
2. The efficiency of energy usage: this is the proportion of the energy used for the evaporation of water only. This parameter has similarities with the 'steam economy' of evaporators. It is related to the engineering aspects of the process.

The energy usage efficiency of conduction (boiling) drying is similar to that of single-effect evaporation without vapor recompression. Recompression of the water vapor released in drying and its recycling as heating steam is theoretically possible in some cases, but has not been applied commercially. The energy efficiency of drying with superheated steam is particularly high but the use of this method for drying foods has limited possibilities, for reasons explained before.

The energy usage efficiency of convective (air) drying is significantly lower, mainly for two reasons:

1. While in conductive drying, heat is supplied directly to the material being dried, in convective drying heat is supplied to air. The heated air, in its turn, transfers heat to the material being dried. Air serves here as an intermediate heat transport agent. Any time an intermediate agent is used to transfer energy the efficiency of the transfer is reduced
2. In order to maintain the necessary driving force for drying, the air is not allowed to reach saturation. 'Spent air' leaving the dryer is still hot and only partially humidified, i.e. it has unutilized drying capacity. Spent air is usually released to the environment and represents considerable loss of energy.

One of the methods to save energy in drying liquid foods is to remove most of the water by pre-concentrating the feed through an energy-efficient method (e.g. multi-effect evaporation) before drying. Such combined processes are extensively used in drying milk and other dairy products, instant coffee etc.

EXAMPLE 22.6

It is desired to compute the specific energy consumption and the energy consumption efficiency of a spray dryer for milk, according to the following data:

- Feed: pre-concentrated milk, 35% w/w dry matter, 30°C, 3600 kg/h
- Product: milk powder, 2.5% residual moisture, 90°C
- Drying air: ambient air at 22°C, 40% RH is heated indirectly by fuel burners to 180°C
- Spent air: leaves the dryer at 90°C.

The specific heat of the feed and the product: 3.1 and 1.2 kJ/kg.K respectively
Heat losses are negligible.

Solution:

In: 3600 kg/h = 1 kg/s milk at 35% solids. This yields 0.36 kg/s milk powder at 2.5% moisture. $1 - 0.36 = 0.64$ kg of water evaporated.

The air is heated at constant moisture (H_1) from 22°C to 180°C, then adiabatically humidified and cooled to 95°C. At this point its moisture content is H_2 .

H_1 and H_2 are read from the psychrometric chart:

$$H_1 = 0.007 \text{ kg/kg}$$

$$H_2 = 0.046 \text{ kg/kg}$$

Note: Because the temperature of 180°C is outside the range of the psychrometric chart (see Figure A.2), H_2 is found by extrapolation. Alternatively, the chart at:

(www.EngineeringToolBox.com/psycrometricchart) that covers high temperatures may be consulted.

The mass flow rate of air is then:

$$G_a = \frac{0.64}{0.046 - 0.007} = 16.4 \text{ kg/s}$$

At an average C_p of 1 kJ/kg.K, the heat transfer rate to the air is:

$$q_{in} = 16.4 \times 1 \times (180 - 22) = 2591 \text{ kW}$$

The energy spend for evaporation (at an average latent heat of 2270 kJ/kg) is:

$$q_{eva} = 0.64 \times 2270 = 1453 \text{ kJ/kg}$$

The specific energy expenditure (per kg of product is):

$$E = \frac{2591}{0.36} = 7197 \text{ kJ/kg}$$

The energy utilization efficiency is:

$$\eta_E = \frac{1453}{2591} \times 100 = 56\%$$

7197 kJ of energy are consumed for each kg of milk powder leaving the dryer. Even this high figure is an under-estimation because heat losses and inefficiencies in the indirect heating of the air have not been considered. Only 56% of this energy is utilized for the removal of water. Most of the rest is ejected with the spent air.

22.9 Osmotic Dehydration

Osmotic dehydration is the removal of water by immersing the food in a solution of salt or sugars of high osmotic pressure. Water is transferred from the food to the solution by virtue of the difference in osmotic pressure.

Essentially, the process of partial removal of water by osmosis is an operation that has been known and practiced for centuries. Salting fish and candying fruit are examples of long established food processing techniques, where removal of water takes place together with solute penetration. Nevertheless, the process, now named ‘osmotic dehydration’ or ‘dewatering-impregnation soaking in concentrated solutions’ (Raoult-Wack et al., 1991, 1992), has recently attracted considerable research interest and a voluminous literature on the subject has accumulated (Torreggiani, 1995). The industrial application of the process, outside the classical processes mentioned above, has been extremely limited.

Physically, the process of osmotic dehydration is simple. The prepared (peeled, sliced or cut etc.) material is immersed in the ‘osmotic solution’, a relatively concentrated solution of sugar, or salt or both. Water and some of the natural solutes of the food pass to the osmotic solution while, at the same time, a certain amount of the ‘osmotic solute’ penetrates the food. The selection of the osmotic solution composition and the process conditions aims at maximizing the water removal and minimizing all other transports. The osmotic solution is recycled, after concentration by evaporation. Energy consumption per kg of water removed is, of course, much lower than in ordinary drying, particularly if the osmotic solution is concentrated by multieffect evaporation. Continuous operation has been tried at pilot-plant level (Barbosa-Cánovas and Vega-Mercado, 1996).

The rate of water removal is fast in the beginning but slows down considerably as the osmotic pressure difference becomes gradually smaller. Consequently, water removal by osmosis cannot be carried too far. Osmotically dehydrated products still

contain too much moisture to be shelf stable. It is suggested that these materials can actually serve as intermediate products, intended for further processing such as drying, freezing or thermal processing (Raoult-Wack et al., 1989).

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