CHAPTER 16

Drying

16.1. INTRODUCTION

The drying of materials is often the final operation in a manufacturing process, carried out immediately prior to packaging or dispatch. Drying refers to the final removal of water, or another solute, and the operation often follows evaporation, filtration, or crystallisation. In some cases, drying is an essential part of the manufacturing process, as for instance in paper making or in the seasoning of timber, although, in the majority of processing industries, drying is carried out for one or more of the following reasons:

(a) To reduce the cost of transport.
(b) To make a material more suitable for handling as, for example, with soap powders, dyestuffs and fertilisers.
(c) To provide definite properties, such as, for example, maintaining the free-flowing nature of salt.
(d) To remove moisture which may otherwise lead to corrosion. One example is the drying of gaseous fuels or benzene prior to chlorination.

With a crystalline product, it is essential that the crystals are not damaged during the drying process, and, in the case of pharmaceutical products, care must be taken to avoid contamination. Shrinkage, as with paper, cracking, as with wood, or loss of flavour, as with fruit, must also be prevented. With the exception of the partial drying of a material by squeezing in a press or the removal of water by adsorption, almost all drying processes involve the removal of water by vaporisation, which requires the addition of heat. In assessing the efficiency of a drying process, the effective utilisation of the heat supplied is the major consideration.

16.2. GENERAL PRINCIPLES

The moisture content of a material is usually expressed in terms of its water content as a percentage of the mass of the dry material, though moisture content is sometimes expressed on a wet basis, as in Example 16.3. If a material is exposed to air at a given temperature and humidity, the material will either lose or gain water until an equilibrium condition is established. This equilibrium moisture content varies widely with the moisture content and the temperature of the air, as shown in Figure 16.1. A non-porous insoluble solid, such as sand or china clay, has an equilibrium moisture content approaching zero.
for all humidities and temperatures, although many organic materials, such as wood, textiles, and leather, show wide variations of equilibrium moisture content. Moisture may be present in two forms:

*Bound moisture.* This is water retained so that it exerts a vapour pressure less than that of free water at the same temperature. Such water may be retained in small capillaries, adsorbed on surfaces, or as a solution in cell walls.

*Free moisture.* This is water which is in excess of the equilibrium moisture content.

The water removed by vaporisation is generally carried away by air or hot gases, and the ability of these gases to pick up the water is determined by their temperature and humidity. In designing dryers using air, the properties of the air–water system are essential, and these are detailed in Volume 1, Chapter 13, where the development of the humidity chart is described. For the air–water system, the following definitions are of importance:

**Humidity** $\mathcal{H}$, mass of water per unit mass of dry air.

Since:

$$\frac{\text{moles of water vapour}}{\text{moles of dry air}} = \frac{P_w}{(P - P_w)}$$

then:

$$\mathcal{H} = \frac{18P_w}{29(P - P_w)}$$

where $P_w$ is the partial pressure of water vapour and $P$ is the total pressure.

**Humidity of saturated air** $\mathcal{H}_0$. This is the humidity of air when it is saturated with water vapour. The air then is in equilibrium with water at the given temperature and pressure.

**Percentage humidity**

$$= \frac{\text{Humidity of air}}{\text{Humidity of saturated air}} \times 100 = \frac{\mathcal{H}}{\mathcal{H}_0} \times 100$$
Percentage relative humidity, $\mathcal{R}$

$$\mathcal{R} = \frac{\text{Partial pressure of water vapour in air}}{\text{Vapour pressure of water at the same temperature}} \times 100$$

The distinction between percentage humidity and percentage relative humidity is of significance though, the difference in the values of the two quantities does not usually exceed 7 to 8 per cent. Reference may be made here to Volume 1, Section 13.2.1.

**Humid volume** is the volume of unit mass of dry air and its associated vapour. Then, under ideal conditions, at atmospheric pressure:

$$\text{humid volume} = \frac{22.4}{29} \left( \frac{T}{273}\right) + \frac{22.4 \mathcal{R}}{18} \left( \frac{T}{273}\right) \text{m}^3/\text{kg}$$

where $T$ is in degrees K.

or:

$$\frac{359}{29} \left( \frac{T}{492}\right) + \frac{359 \mathcal{R}}{18} \left( \frac{T}{492}\right) \text{ft}^3/\text{lb}$$

where $T$ is in degrees Rankine.

**Saturated volume** is the volume of unit mass of dry air, together with the water vapour required to saturate it.

**Humid heat** is the heat required to raise unit mass of dry air and associated vapour through 1 degree K at constant pressure or $1.00 + 1.88 \mathcal{R}$ kJ/kg K.

**Dew point** is the temperature at which condensation will first occur when air is cooled.

**Wet bulb temperature**. If a stream of air is passed rapidly over a water surface, vaporisation occurs, provided the temperature of the water is above the dew point of the air. The temperature of the water falls and heat flows from the air to the water. If the surface is sufficiently small for the condition of the air to change inappreciably and if the velocity is in excess of about 5 m/s, the water reaches the wet bulb temperature $\theta_w$ at equilibrium.

The rate of heat transfer from gas to liquid is given by:

$$Q = hA(\theta - \theta_w) \quad (16.1)$$

The mass rate of vaporisation is given by:

$$G_v = \frac{h_DAM_w}{RT} (P_{w0} - P_w)$$

$$= \frac{h_DAM_w}{RT} [(P - P_w)_{\text{mean}}(\mathcal{H}_w - \mathcal{H})]$$

$$= h_D\rho_A(\mathcal{H}_w - \mathcal{H}) \quad (16.2)$$

The rate of heat transfer required to effect vaporisation at this rate is given by:

$$G_v = h_D\rho_A(\mathcal{H}_w - \mathcal{H})\lambda \quad (16.3)$$
At equilibrium, the rates of heat transfer given by equations 16.1 and 16.3 must be equal, and hence:

\[ \dot{H} = \dot{H}_w = -\frac{h}{h_D \rho_A \lambda} (\theta - \theta_w) \]  

(16.4)

In this way, it is seen that the wet bulb temperature \( \theta_w \) depends only on the temperature and humidity of the drying air.

In these equations:

- \( h \) is the heat transfer coefficient,
- \( h_D \) is the mass transfer coefficient,
- \( A \) is the surface area,
- \( \theta \) is the temperature of the air stream,
- \( \theta_w \) is the wet bulb temperature,
- \( P_{w0} \) is the vapour pressure of water at temperature \( \theta_w \),
- \( M_A \) is the molecular weight of air,
- \( M_w \) is the molecular weight of water,
- \( R \) is the universal gas constant,
- \( T \) is the absolute temperature,
- \( \dot{H} \) is the humidity of the gas stream,
- \( \dot{H}_w \) is the humidity of saturated air at temperature \( \theta_w \),
- \( \rho_A \) is the density of air at its mean partial pressure, and
- \( \lambda \) is the latent heat of vaporisation of unit mass of water.

Equation 16.4 is identical with equation 13.8 in Volume 1, and reference may be made to that chapter for a more detailed discussion.

### 16.3. RATE OF DRYING

#### 16.3.1. Drying periods

In drying, it is necessary to remove free moisture from the surface and also moisture from the interior of the material. If the change in moisture content for a material is determined as a function of time, a smooth curve is obtained from which the rate of drying at any given moisture content may be evaluated. The form of the drying rate curve varies with the structure and type of material, and two typical curves are shown in Figure 16.2. In curve 1, there are two well-defined zones: AB, where the rate of drying is constant and BC, where there is a steady fall in the rate of drying as the moisture content is reduced. The moisture content at the end of the constant rate period is represented by point B, and this is known as the critical moisture content. Curve 2 shows three stages, DE, EF and FC. The stage DE represents a constant rate period, and EF and FC are falling rate periods. In this case, the Section EF is a straight line, however, and only the portion FC is curved. Section EF is known as the first falling rate period and the final stage, shown as FC, as the second falling rate period. The drying of soap gives rise to a curve of type 1, and sand to a curve of type 2. A number of workers, including Sherwood\(^{(1)}\) and Newitt and co-workers\(^{(2-7)}\), have contributed various theories on the rate of drying at these various stages.
During the constant rate period, it is assumed that drying takes place from a saturated surface of the material by diffusion of the water vapour through a stationary air film into the air stream. GILLILAND\(^{(8)}\) has shown that the rates of drying of a variety of materials in this stage are substantially the same as shown in Table 16.1.

**Table 16.1. Evaporation rates for various materials under constant conditions\(^{(8)}\)**

<table>
<thead>
<tr>
<th>Material</th>
<th>Rate of evaporation (kg/m(^2) h)</th>
<th>Rate of evaporation (kg/m(^2) s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Water</td>
<td>2.7</td>
<td>0.00075</td>
</tr>
<tr>
<td>Whiting pigment</td>
<td>2.1</td>
<td>0.00058</td>
</tr>
<tr>
<td>Brass filings</td>
<td>2.4</td>
<td>0.00067</td>
</tr>
<tr>
<td>Brass turnings</td>
<td>2.4</td>
<td>0.00067</td>
</tr>
<tr>
<td>Sand (fine)</td>
<td>2.0(-)2.4</td>
<td>0.00055(-)0.00067</td>
</tr>
<tr>
<td>Clays</td>
<td>2.3(-)2.7</td>
<td>0.00064(-)0.00075</td>
</tr>
</tbody>
</table>

In order to calculate the rate of drying under these conditions, the relationships obtained in Volume 1 for diffusion of a vapour from a liquid surface into a gas may be used. The simplest equation of this type is:

\[
W = k_G A (P_s - P_w) \quad (16.5)
\]

where \(k_G\) is the mass transfer coefficient.

Since the rate of transfer depends on the velocity \(u\) of the air stream, raised to a power of about 0.8, then the mass rate of evaporation is:

\[
W = k_G A (P_s - P_w) u^{0.8} \quad (16.6)
\]

where: \(A\) is the surface area,

\(P_s\) is the vapour pressure of the water, and

\(P_w\) is the partial pressure of water vapour in the air stream.
The type of equation, used in Volume 1 for the rate of vaporisation into an air stream, simply states that the rate of transfer is equal to the transfer coefficient multiplied by the driving force. It may be noted, however, that \((P_s - P_w)\) is not only a driving force, but it is also related to the capacity of the air stream to absorb moisture.

These equations suggest that the rate of drying is independent of the geometrical shape of the surface. Work by Powell and Griffiths\(^9\) has shown, however, that the ratio of the length to the width of the surface is of some importance, and that the evaporation rate is given more accurately as:

\[
W = 5.53 \times 10^{-9} L^{0.77} B(P_s - P_w)(1 + 61u^{0.85}) \text{ kg/s} \quad (16.7)
\]

\[
W = 3.72 \times 10^{-9} L^{0.23} B^{0.8}(P_s - P_w)(1 + 61u^{0.85}) \text{ kg/s} \quad (16.8)
\]

where: \(P_s\), the saturation pressure at the temperature of the surface (N/m\(^2\)); \(P_w\), the vapour pressure in the air stream (N/m\(^2\)), and \(L\) and \(B\) are the length and width of the surface, respectively (m).

For most design purposes, it may be assumed that the rate of drying is proportional to the transfer coefficient multiplied by \((P_s - P_w)\). Chakravorty\(^10\) has shown that, if the temperature of the surface is greater than that of the air stream, then \(P_w\) may easily reach a value corresponding to saturation of the air. Under these conditions, the capacity of the air to take up moisture is zero, while the force causing evaporation is \((P_s - P_w)\). As a result, a mist will form and water may be redeposited on the surface. In all drying equipment, care must therefore be taken to ensure that the air or gas used does not become saturated with moisture at any stage.

The rate of drying in the constant rate period is given by:

\[
W = \frac{\text{d} m}{\text{d} t} = \frac{h A \Delta T}{\lambda} = k_G A (P_s - P_w) \quad (16.9)
\]

where: \(W\) is the rate of loss of water; \(h\) is the heat transfer coefficient from air to the wet surface; \(\Delta T\) is the temperature difference between the air and the surface; \(\lambda\) is the latent heat of vaporisation per unit mass; \(k_G\) is the mass transfer coefficient for diffusion from the wet surface through the gas film; \(A\) is the area of interface for heat and mass transfer, and \((P_s - P_w)\) is the difference between the vapour pressure of water at the surface and the partial pressure in the air.
It is more convenient to express the mass transfer coefficient in terms of a humidity difference, so that \( k_G A (P_s - P_w) \approx k A (\frac{\Delta h}{\Delta T}) \). The rate of drying is thus determined by the values of \( h \), \( \Delta T \) and \( A \), and is not influenced by the conditions inside the solid. \( h \) depends on the air velocity and the direction of flow of the air, and it has been found that \( h = C G' / A \), where \( G' \) is the mass rate of flow of air in kg/s m\(^2\). For air flowing parallel to plane surfaces, Shepherd et al.\(^{(11)}\) have given the value of \( C \) as 14.5 where the heat transfer coefficient is expressed in W/m\(^2\) K.

If the gas temperature is high, then a considerable proportion of the heat will pass to the solid by radiation, and the heat transfer coefficient will increase. This may result in the temperature of the solid rising above the wet bulb temperature.

**First falling-rate period**

The points B and E in Figure 16.2 represent conditions where the surface is no longer capable of supplying sufficient free moisture to saturate the air in contact with it. Under these conditions, the rate of drying depends very much on the mechanism by which the moisture from inside the material is transferred to the surface. In general, the curves in Figure 16.2 will apply, although for a type 1 solid, a simplified expression for the rate of drying in this period may be obtained.

**Second falling-rate period**

At the conclusion of the first falling rate period it may be assumed that the surface is dry and that the plane of separation has moved into the solid. In this case, evaporation takes place from within the solid and the vapour reaches the surface by molecular diffusion through the material. The forces controlling the vapour diffusion determine the final rate of drying, and these are largely independent of the conditions outside the material.

**16.3.2. Time for drying**

If a material is dried by passing hot air over a surface which is initially wet, the rate of drying curve in its simplest form is represented by BCE, shown in Figure 16.3.
where: $w$ is the total moisture,
$w_e$ is the equilibrium moisture content (point E),
$w - w_e$ is the free moisture content, and
$w_c$ is the critical moisture content (point C).

**Constant-rate period**

During the period of drying from the initial moisture content $w_1$ to the critical moisture content $w_c$, the rate of drying is constant, and the time of drying $t_c$ is given by:

$$t_c = \frac{w_1 - w_c}{R_c A} \quad (16.10)$$

where: $R_c$ is the rate of drying per unit area in the constant rate period, and $A$ is the area of exposed surface.

**Falling-rate period**

During this period the rate of drying is, approximately, directly proportional to the free moisture content $(w - w_e)$, or:

$$-\left(\frac{1}{A}\right) \frac{dw}{dt} = m(w - w_e) = mf \quad \text{(say)} \quad (16.11)$$

Thus:

$$-\frac{1}{mA} \int_{w_e}^{w} \frac{dw}{(w - w_e)} = \int_{0}^{t_f} dr$$

or:

$$\frac{1}{mA} \ln \left[ \frac{w_c - w_e}{w - w_e} \right] = t_f$$

and:

$$t_f = \frac{1}{mA} \ln \left( \frac{f_c}{f} \right) \quad (16.12)$$

**Total time of drying**

The total time $t$ of drying from $w_1$ to $w$ is given by $t = (t_c + t_f)$.

The rate of drying $R_c$ over the constant rate period is equal to the initial rate of drying in the falling rate period, so that $R_c = mf_c$.

Thus:

$$t_c = \frac{(w_1 - w_c)}{mA f_c} \quad (16.13)$$

and the total drying time,

$$t = \frac{(w_1 - w_c)}{mA f_c} + \frac{1}{mA} \ln \left( \frac{f_c}{f} \right)$$

$$= \frac{1}{mA} \left[ \frac{(f_1 - f_c)}{f_c} + \ln \left( \frac{f_c}{f} \right) \right] \quad (16.14)$$
Example 16.1

A wet solid is dried from 25 to 10 per cent moisture under constant drying conditions in 15 ks (4.17 h). If the critical and the equilibrium moisture contents are 15 and 5 per cent respectively, how long will it take to dry the solid from 30 to 8 per cent moisture under the same conditions?

Solution

For the first drying operation:

\[ w_1 = 0.25 \text{ kg/kg}, \quad w = 0.10 \text{ kg/kg}, \quad w_c = 0.15 \text{ kg/kg and } w_e = 0.05 \text{ kg/kg} \]

Thus:

\[ f_1 = (w_1 - w_e) = (0.25 - 0.05) = 0.20 \text{ kg/kg} \]
\[ f_c = (w_c - w_e) = (0.15 - 0.05) = 0.10 \text{ kg/kg} \]
\[ f = (w - w_e) = (0.10 - 0.05) = 0.05 \text{ kg/kg} \]

From equation 16.14, the total drying time is:

\[ t = \left( \frac{1}{mA} \right) \left[ \left( \frac{f_1 - f_c}{f_c} \right) + \ln(f_c/f) \right] \]

or:

\[ 15 = \left( \frac{1}{mA} \right) \left[ (0.20 - 0.10)/0.10 + \ln(0.10/0.05) \right] \]

and:

\[ mA = 0.0667(1.0 + 0.693) = 0.113 \text{ kg/s} \]

For the second drying operation:

\[ w_1 = 0.30 \text{ kg/kg}, \quad w = 0.08 \text{ kg/kg}, \quad w_c = 0.15 \text{ kg/kg and } w_e = 0.05 \text{ kg/kg} \]

Thus:

\[ f_1 = (w_1 - w_e) = (0.30 - 0.05) = 0.25 \text{ kg/kg} \]
\[ f_c = (w_c - w_e) = (0.15 - 0.05) = 0.10 \text{ kg/kg} \]
\[ f = (w - w_e) = (0.08 - 0.05) = 0.03 \text{ kg/kg} \]

The total drying time is then:

\[ t = \left( \frac{1}{0.113} \right) \left[ (0.25 - 0.10)/0.10 + \ln(0.10/0.03) \right] \]

\[ = 8.856(1.5 + 1.204) \]

\[ = 23.9 \text{ ks (6.65 h)} \]

Example 16.2

Strips of material 10 mm thick are dried under constant drying conditions from 28 to 13 per cent moisture in 25 ks (7 h). If the equilibrium moisture content is 7 per cent, what is the time taken to dry 60 mm planks from 22 to 10 per cent moisture under the same conditions assuming no loss from the edges? All moistures are given on a wet basis.

The relation between \( E \), the ratio of the average free moisture content at time \( t \) to the initial free moisture content, and the parameter \( J \) is given by:

\[
\begin{array}{cccccccc}
E & 1 & 0.64 & 0.49 & 0.38 & 0.295 & 0.22 & 0.14 \\
J & 0 & 0.1 & 0.2 & 0.3 & 0.5 & 0.6 & 0.7 \\
\end{array}
\]
It may be noted that $J = \frac{kt}{L^2}$, where $k$ is a constant, $t$ the time in ks and $2L$ the thickness of the sheet of material in millimetres.

**Solution**

*For the 10 mm strips*

Initial free moisture content $= (0.28 - 0.07) = 0.21$ kg/kg.

Final free moisture content $= (0.13 - 0.07) = 0.06$ kg/kg.

Thus: when $t = 25$ ks, $E = (0.06/0.21) = 0.286$

and from Figure 16.4, a plot of the given data,

$$J = 0.52$$

Thus: $0.52 = (k \times 25)/((10/2)^2)$

and:

$$k = 0.52$$

![Figure 16.4: Drying data for Example 6.2](image)

*For the 60 mm planks*

Initial free moisture content $= (0.22 - 0.07) = 0.15$ kg/kg.

Final free moisture content $= (0.10 - 0.07) = 0.03$ kg/kg.

$E = (0.03/0.15) = 0.20$

From Figure 16.4: $J = 0.63$

and hence:

$$k = \frac{J L^2}{2}$$

$$= \frac{0.63(60/2)^2}{0.52} = 1090 \text{ ks (12.6 days)}$$
Example 16.3

A granular material containing 40 per cent moisture is fed to a countercurrent rotary dryer at a temperature of 295 K and is withdrawn at 305 K, containing 5 per cent moisture. The air supplied, which contains 0.006 kg water vapour/kg dry air, enters at 385 K and leaves at 310 K. The dryer handles 0.125 kg/s wet stock.

Assuming that radiation losses amount to 20 kJ/kg dry air used, determine the mass flowrate of dry air supplied to the dryer and the humidity of the exit air.

The latent heat of water vapour at 295 K = 2449 kJ/kg, specific heat capacity of dried material = 0.88 kJ/kg K, the specific heat capacity of dry air = 1.00 kJ/kg K, and the specific heat capacity of water vapour = 2.01 kJ/kg K.

Solution

This example involves a heat balance over the system. 273 K will be chosen as the datum temperature, and it will be assumed that the flow rate of dry air = G kg/s.

*Heat in:*

(a) *Air*

G kg/s dry air enter with 0.006G kg/s water vapour and hence the heat content of this stream

\[ = [(1.00G) + (0.006G \times 2.01)](385 - 273) = 113.35G \text{ kW} \]

(b) *Wet solid*

0.125 kg/s enter containing 0.40 kg water/kg wet solid, assuming the moisture is expressed on a wet basis.

Thus:

mass flowrate of water = (0.125 \times 0.40) = 0.050 kg/s

and:

mass flowrate of dry solid = (0.125 - 0.050) = 0.075 kg/s

Hence:

the heat content of this stream = [(0.050 \times 4.18) + (0.075 \times 0.88)](295 - 273) = 6.05 kW

*Heat out:*

(a) *Air*

Heat in exit air = [(1.00 G) + (0.006 G \times 2.01)](310 - 273) = 37.45G kW.

Mass flow rate of dry solids = 0.075 kg/s containing 0.05 kg water/kg wet solids.

Hence:

water in the dried solids leaving = (0.05 \times 0.075)/(1 + 0.05) = 0.0036 kg/s

and:

the water evaporated into gas steam = (0.050 - 0.0036) = 0.0464 kg/s.

Assuming evaporation takes place at 295 K, then:

heat in the water vapour = 0.0464[2.01(310 - 295) + 2449 + 4.18(295 - 273)]

\[ = 119.3 \text{ kW} \]
and:

the total heat in this stream $= (119.30 + 37.45G)$ kW.

(b) Dried solids

The dried solids contain 0.0036 kg/s water and hence heat content of this stream is:

$$\left[(0.075 \times 0.88) + (0.0036 \times 4.18)/(305 - 273)\right] = 2.59 \text{ kW}$$

(c) Losses

These amount to 20 kJ/kg dry air or 20$m$ kW.

Heat balance

$$(113.35 \ G + 6.05) = (119.30 + 37.45 \ G + 2.59 + 20 \ G)$$

and:

$$G = 2.07 \text{ kg/s}$$

Water in the outlet air stream $= (0.006 \times 2.07) + 0.0464 = 0.0588 \text{ kg/s}$

and:

the humidity $\gamma = (0.0588/2.07) = 0.0284 \text{ kg/kg dry air}$

16.4. THE MECHANISM OF MOISTURE MOVEMENT DURING DRYING

16.4.1. Diffusion theory of drying

In the general form of the curve for the rate of drying of a solid shown in Figure 16.2, there are two and sometimes three distinct sections. During the constant-rate period, moisture vapourises into the air stream and the controlling factor is the transfer coefficient for diffusion across the gas film. It is important to understand how the moisture moves to the drying surface during the falling-rate period, and two models have been used to describe the physical nature of this process, the diffusion theory and the capillary theory. In the diffusion theory, the rate of movement of water to the air interface is governed by rate equations similar to those for heat transfer, whilst in the capillary theory the forces controlling the movement of water are capillary in origin, arising from the minute pore spaces between the individual particles.

Falling rate period, diffusion control

In the falling-rate period, the surface is no longer completely wetted and the rate of drying steadily falls. In the previous analysis, it has been assumed that the rate of drying per unit effective wetted area is a linear function of the water content, so that the rate of drying is given by:

$$\left(\frac{1}{A}\right) \frac{dw}{dr} = -m(w - w_r) \quad \text{ (equation 16.11)}$$

In many cases, however, the rate of drying is governed by the rate of internal movement of the moisture to the surface. It was initially assumed that this movement was a process
of diffusion and would follow the same laws as heat transfer. This approach has been examined by a number of workers, and in particular by Sherwood\textsuperscript{12} and Newman\textsuperscript{13}.

Considering a slab with the edges coated to prevent evaporation, which is dried by evaporation from two opposite faces, the \( Y \)-direction being taken perpendicular to the drying face, the central plane being taken as \( y = 0 \), and the slab thickness \( 2l \), then on drying, the moisture movement by diffusion will be in the \( Y \)-direction, and hence from Volume 1, equation 10.66:

\[
\frac{\partial C_w}{\partial t} = D_L \frac{\partial^2 C_w}{\partial y^2}
\]

where \( C_w \) is the concentration of water at any point and any time in the solid, and \( D_L \) is the coefficient of diffusion for the liquid.

If \( w \) is the liquid content of the solid, integrated over the whole depth, \( w_1 \) the initial content, and \( w_e \) the equilibrium content, then:

\[
\frac{(w - w_e)}{(w_1 - w_e)} = \frac{\text{Free liquid content at any time}}{\text{Initial free liquid content}}
\]

Sherwood\textsuperscript{12} and Newman\textsuperscript{13} have presented the following solution assuming an initially uniform water distribution and zero water-concentration at the surface once drying has started:

\[
\frac{(w - w_e)}{(w_1 - w_e)} = \frac{8}{\pi^2} e^{-\frac{D_L(\pi/2)^2}{l^2}} + \frac{1}{9} e^{-\frac{9D_L(\pi/2)^2}{l^2}} + \frac{1}{25} e^{-\frac{25D_L(\pi/2)^2}{l^2}} + \cdots \quad (16.15)
\]

This equation assumes an initially uniform distribution of moisture, and that the drying is from both surfaces. When drying is from one surface only, then \( l \) is the total thickness. If the time of drying is long, then only the first term of the equation need be used and thus, differentiating equation 16.15 gives:

\[
\frac{dw}{dt} = -\frac{2D_L}{l^2} e^{-\frac{D_L(\pi/2)^2}{l^2}} (w_1 - w_e) \quad (16.16)
\]

In the drying of materials such as wood or clay, the moisture concentration at the end of the constant rate period is not uniform, and is more nearly parabolic. Sherwood has presented an analysis for this case, and has given experimental values for the drying of brick clay.

In this case, it is assumed that the rate of movement of water is proportional to a concentration gradient, and capillary and gravitational forces are neglected. Water may, however, flow from regions of low concentration to those of high concentration if the pore sizes are suitable, and for this and other reasons, Ceglske and Hougen\textsuperscript{14,15} have proposed a capillary theory which is now considered.

### 16.4.2. Capillary theory of drying

**Principles of the theory**

The capillary theory of drying has been proposed in order to explain the movement of moisture in the bed during surface drying. The basic importance of the pore space between
granular particles was first pointed out by Slichter\(^{(16)}\) in connection with the movement of moisture in soils, and this work has been modified and considerably expanded by Haines\(^{(17)}\). The principles are now outlined and applied to the problem of drying. Considering a systematic packing of uniform spherical particles, these may be arranged in six different regular ways, ranging from the most open to the closest packing. In the former, the spheres are arranged as if at the corners of a cube with each sphere touching six others. In the latter arrangement, each sphere rests in the hollow of three spheres in adjacent layers, and touches twelve other spheres. These configurations are shown in Figure 16.5. The densities of packing of the other four arrangements will lie between those illustrated.

![Figure 16.5. Packing of spherical particles. (a) Cubic arrangement, one sphere touching six others. (b) Rhombohedral packing, one sphere touching twelve others, with layers arranged in rhombic formation.](image)

In each case, a regular group of spheres surrounds a space which is called a pore, and the bed is made up of a series of these elemental groupings. The pores are connected together by passages of various sizes, the smallest portions of which are known as waists. The size of a pore is defined as the diameter of the largest sphere which can be fitted into it, and the size of a waist as the diameter of the inscribed circle. The sizes of the pores and waists will differ for each form of packing, as shown in Table 16.2.

<table>
<thead>
<tr>
<th>Packing arrangement</th>
<th>Pore space (per cent total volume)</th>
<th>Radius of pore</th>
<th>Radius of waist</th>
<th>Value of (x) in equation 16.20 for:</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cubical</td>
<td>47.64</td>
<td>0.700(r)</td>
<td>0.414(r)</td>
<td>Limiting suction potential of pores = 2.86, Entry suction potential of waists = 4.82</td>
</tr>
<tr>
<td>Rhombohedral</td>
<td>25.95</td>
<td>0.288(r)</td>
<td>0.155(r)</td>
<td>Limiting suction potential of pores = 6.90, Entry suction potential of waists = 12.90</td>
</tr>
</tbody>
</table>

The continuous variation in the diameter of each passage is the essential difference between a granular packing and a series of capillary tubes. If a clean capillary of uniform
diameter $2r'$ is placed in a liquid, the level will rise in the capillary to a height $h_s$ given by:

$$h_s = \left( \frac{2\sigma}{r' \rho g} \right) \cos \alpha$$

(16.17)

where: $\rho$ is the density of the liquid,
$\sigma$ is the surface tension, and
$\alpha$ is the angle of contact.

A negative pressure, known as a *suction potential*, will exist in the liquid in the capillary. Immediately below the meniscus, the suction potential will be equivalent to the height of the liquid column $h_s$ and, if water is used, this will have the value:

$$h_s = \frac{2\sigma}{r' \rho g}$$

(16.18)

If equilibrium conditions exist, the suction potential $h_1$ at any other level in the liquid, a distance $z_1$ below the meniscus, will be given by:

$$h_s = h_1 + z_1$$

(16.19)

Similarly, if a uniform capillary is filled to a height greater than $h_s$, as given by equation 16.17, and its lower end is immersed, the liquid column will recede to this height.

The non-uniform passages in a porous material will also display the same characteristics as a uniform capillary, with the important difference that the rise of water in the passages will be limited by the pore size, whilst the depletion of saturated passages will be controlled by the size of the waists. The height of rise is controlled by the pore size, since this approximates to the largest section of a varying capillary, whilst the depletion of water is controlled by the narrow waists which are capable of a higher suction potential than the pores.

The theoretical suction potential of a pore or waist containing water is given by:

$$h_i = \frac{x \sigma}{r \rho g}$$

(16.20)

where: $x$ is a factor depending on the type of packing, shown in Table 16.2, and $r$ is the radius of the spheres.

For an idealised bed of uniform rhombohedrally packed spheres of radius $r$, for example, the waists are of radius $0.155r$, from Table 16.2, and the maximum theoretical suction potential of which such a waist is capable is:

$$\frac{2\sigma}{0.155r \rho g} = \frac{12.9\sigma}{r \rho g}$$

from which $x = 12.9$.

The maximum suction potential that can be developed by a waist is known as the *entry suction potential*. This is the controlling potential required to open a saturated
pore protected by a meniscus in an adjoining waist and some values for $x$ are given in Table 16.2.

When a bed is composed of granular material with particles of mixed sizes, the suction potential cannot be calculated and it must be measured by methods such as those given by Haines(17) and Oliver and Newitt(18).

**Drying of a granular material according to the capillary theory**

If a bed of uniform spheres, initially saturated, is to be surface dried in a current of air of constant temperature, velocity and humidity, then the rate of drying is given by:

$$\frac{dw}{dt} = k_G A(P_w^0 - P_w)$$

where $P_w^0$ is the saturation partial pressure of water vapour at the wet bulb temperature of the air, and $P_w$ is the partial pressure of the water vapour in the air stream. This rate of drying will remain constant so long as the inner surface of the “stationary” air film remains saturated.

As evaporation proceeds, the water surface recedes into the waists between the top layer of particles, and an increasing suction potential is developed in the liquid. When the menisci on the cubical waists, that is the largest, have receded to the narrowest section, the suction potential $h_s$ at the surface is equal to $4.82\sigma/r\rho g$, from Table 16.2. Further evaporation will result in $h_s$ increasing so that the menisci on the surface cubical waists will collapse, and the larger pores below will open. As $h_s$ steadily increases, the entry suction of progressively finer surface waists is reached, so that the menisci collapse into the adjacent pores which are thereby opened.

In considering the conditions below the surface, the suction potential $h_1$ at a distance $z_1$ from the surface is given by:

$$h_1 = h_s + z_1$$

The flow of water through waists surrounding an open pore is governed by the size of the waist as follows:

(a) If the size of the waist is such that its entry suction potential exceeds the suction potential at that level within the bed, it will remain full by the establishment of a meniscus therein, in equilibrium with the effective suction potential to which it is subjected. This waist will then protect adjoining full pores which cannot be opened until one of the waists to which it is connected collapses.

(b) If the size of the waist is such that its entry suction potential is less than the suction potential existing at that level, it will in turn collapse and open the adjoining pore. In addition, this successive collapse of pores and waists will progressively continue so long as the pores so opened expose waists having entry suction potentials of less than the suction potentials existing at that depth.

As drying proceeds, two processes take place simultaneously:

(a) The collapse of progressively finer surface waists, and the resulting opening of pores and waists connected to them, which they previously protected, and
(b) The collapse of further full waists within the bed adjoining opened pores, and the consequent opening of adjacent pores.

Even though the effective suction potential at a waist or pore within the bed may be in excess of its entry or limiting suction potential, this will not necessarily collapse or open. Such a waist can only collapse if it adjoins an opened pore, and the pore in question can only open upon the collapse of an adjoining waist.

**Effect of particle size.** Reducing the particle size in the bed will reduce the size of the pores and the waists, and will increase the entry suction potential of the waists. This increase means that the percentage variation in suction potentials with depth is reduced, and the moisture distribution is more uniform with small particles.

As the pore sizes are reduced, the frictional forces opposing the movement of water through these pores and waists may become significant, so that equation 16.19 is more accurately represented by:

\[ h_s = h_1 + z_1 + h_f \]  

(16.22)

where \( h_f \), the frictional head opposing the flow over a depth \( z_1 \) from the surface, will depend on the particle size. It has been found\(^{(1)}\) that, with coarse particles when only low suction potentials are found, the gravity effect is important though \( h_f \) is small, whilst with fine particles \( h_f \) becomes large.

(a) For particles of 0.1–1 mm radius, the values of \( h_1 \) are independent of the rate of drying, and vary appreciably with depth. Frictional forces are, therefore, negligible whilst capillary and gravitational forces are in equilibrium throughout the bed and are the controlling forces. Under such circumstances the percentage moisture loss at the critical point at which the constant rate period ends is independent of the drying rate, and varies with the depth of bed.

(b) For particles of 0.001–0.01 mm radius, the values of \( h_1 \) vary only slightly with rate of drying and depth, indicating that both gravitational and frictional forces are negligible whilst capillary forces are controlling. The critical point here will be independent of drying rate and depth of bed.

(c) For particles of less than 0.001 mm (1 \( \mu \)m) radius, gravitational forces are negligible, whilst frictional forces are of increasing importance and capillary and frictional forces may then be controlling. In such circumstances, the percentage moisture loss at the critical point diminishes with increased rate of drying and depth of bed. With beds of very fine particles an additional factor comes into play. The very high suction potentials which are developed cause a sufficient reduction of the pressure for vaporisation of water to take place inside the bed. This internal vaporisation results in a breaking up of the continuous liquid phase and a consequent interruption in the free flow of liquid by capillary action. Hence, the rate of drying is still further reduced.

Some of the experimental data of Newitt et al.\(^{(2)}\) are illustrated in Figure 16.6.

**Freeze drying**

Special considerations apply to the movement of moisture in freeze drying. Since the water is frozen, liquid flow under capillary action is impossible, and movement must be
Figure 16.6. Rates of drying of various materials as a function of percentage saturation. A – 60 µm glass spheres, bed 51 mm deep. B – 23.5 µm silica flour, bed 51 mm deep. C – 7.5 µm silica flour, bed 51 mm deep. D – 2.5 µm silica flour, bed 65 mm deep. Subscripts: 1. Low drying rate 2. High drying rate by vapour diffusion, analogous to the “second falling rate period” of the normal case. In addition, at very low pressures the mean free path of the water molecules may be comparable with the pore size of the material. In these circumstances the flow is said to be of the ‘Knudsen’ type, referred to in Volume 1, Section 10.1.

16.5. DRYING EQUIPMENT

16.5.1. Classification and selection of dryers

Because of the very wide range of dryer designs available, classification is a virtually impossible task. PARKER\(^{(18)}\) takes into account, however, the means by which material is transferred through the dryer as a basis of his classification, with a view to presenting a guide to the selection of dryers. Probably the most thorough classification of dryer types has been made by KROLL\(^{(19)}\) who has presented a decimalised system based on the following factors:

(a) Temperature and pressure in the dryer,
(b) The method of heating,
(c) The means by which moist material is transported through the dryer,
(d) Any mechanical aids aimed at improving drying,
(e) The method by which the air is circulated,
(f) The way in which the moist material is supported,
(g) The heating medium, and
(h) The nature of the wet feed and the way it is introduced into the dryer.
In selecting a dryer for a particular application, as Sloan\textsuperscript{(20)} has pointed out, two steps are of primary importance:

(a) A listing of the dryers which are capable of handling the material to be dried,
(b) Eliminating the more costly alternatives on the basis of annual costs, capital charges + operating costs. A summary of dryer types, together with cost data, has been presented by Backhurst and Harker\textsuperscript{(21)} and the whole question of dryer selection is discussed further in Volume 6.

Once a group of possible dryers has been selected, the choice may be narrowed by deciding whether batch or continuous operation is to be employed and, in addition to restraints imposed by the nature of the material, whether heating by contact with a solid surface or directly by convection and radiation is preferred.

In general, continuous operation has the important advantage of ease of integration into the rest of the process coupled with a lower unit cost of drying. As the rate of throughput of material becomes smaller, however, the capital cost becomes the major component in the total running costs and the relative cheapness of batch plant becomes more attractive. This is illustrated in Figure 16.7 which is taken from the work of KEEY\textsuperscript{(22)}.

In general, throughputs of 5000 kg/day (0.06 kg/s) and over are best handled in batches whilst throughputs of 50,000 kg/day (0.06 kg/s) and over are better handled continuously. The ease of construction of a small batch dryer compared with the sophistication of the

![Figure 16.7: Variation of unit costs of drying with production rate](22)
A continuous dryer should also be taken into account. In addition, a batch dryer is much more versatile and it can often be used for different materials. The humidity may be controlled during the drying operation, and this is especially important in cases where the humidity has to be maintained at different levels for varying periods of time.

Direct heating, in which the material is heated primarily by convection from hot gases has several advantages. Firstly, directly heated dryers are, in general, less costly, mainly because of the absence of tubes or jackets within which the heating medium must be contained. Secondly, it is possible to control the temperature of the gas within very fine limits, and indeed it is relatively simple to ensure that the material is not heated beyond a specified temperature. This is especially important with heat-sensitive materials. Against this, the overall thermal efficiency of directly heated dryers is generally low due to the loss of energy in the exhaust gas and, where an expensive solvent is evaporated from the solid, the operation is often difficult and costly. Losses also occur in the case of fluffy and powdery materials, and further problems are encountered where either the product or the solvent reacts with oxygen in the air.

A major cost in the operation of a dryer is in heating the air or gas. Frequently, the hot gases are produced by combustion of a fuel gas or atomised liquid, and considerable economy may be effected by using a combined heat and power system in which the hot gases are first passed through a turbine connected to an electrical generator.

Many of these disadvantages may be overcome by modifications to the design, although these increase the cost, and often an indirectly heated dryer may prove to be more economical. This is especially the case when thermal efficiency, solvent recovery or maximum cleanliness is of paramount importance and, with indirectly heated dryers, there is always the danger of overheating the product, since the heat is transferred through the material itself.

The maximum temperature at which the drying material may be held is controlled by the thermal sensitivity of the product and this varies inversely with the time of retention. Where lengthy drying times are employed, as for example in a batch shelf dryer, it is necessary to operate under vacuum in order to maintain evaporative temperatures at acceptable levels. In most continuous dryers, the retention time is very low, however, and operation at atmospheric pressure is usually satisfactory. As noted previously, dryer selection is considered in some detail in Volume 6.

16.5.2. Tray or shelf dryers

Tray or shelf dryers are commonly used for granular materials and for individual articles. The material is placed on a series of trays which may be heated from below by steam coils and drying is carried out by the circulation of air over the material. In some cases, the air is heated and then passed once through the oven, although, in the majority of dryers, some recirculation of air takes place, and the air is reheated before it is passed over each shelf. As air is passed over the wet material, both its temperature and its humidity change. This process of air humidification is discussed in Volume 1, Chapter 13.

If air of humidity $\mathcal{H}_1$ is passed over heating coils so that its temperature rises to $\theta_1$, this operation may be represented by the line AB on the humidity chart shown in Figure 16.8. This air then passes over the wet material and leaves at, say 90 per cent relative humidity,
with its temperature falling to some value $\theta_2$. This change in the condition of the air is shown by the line BC, and the humidity has risen to $\mathcal{H}_2$. The wet-bulb temperature of the air will not change appreciably and therefore BC will coincide with an adiabatic cooling line. Each kg of air removes $(\mathcal{H}_2 - \mathcal{H}_1)$ kg of water, and the air required to remove a given amount of water from the material may easily be found. If the air at $\theta_2$ is now passed over a second series of heating coils and is heated to the original temperature $\theta_1$, the heating operation is shown by the line CD. This reheated air can then be passed over wet material on a second tray in the dryer, and pick up moisture until its relative humidity rises again to 90 per cent at a temperature $\theta_3$. This is at point E. In this way each kilogram of air has picked up water amounting to $(\mathcal{H}_3 - \mathcal{H}_1)$ kg of water. Reheating in this way may be effected a number of times, as shown in Figure 16.8, so that the moisture removed per kilogram of air can be considerably increased compared with that for a single pass. Thus, for three passes of air over the material, the total moisture removed is $(\mathcal{H}_4 - \mathcal{H}_1)$ kg/kg air.

If the air of humidity $\mathcal{H}_1$ had been heated initially to a temperature $\theta_5$, the same amount of moisture would have been removed by a single passage over the material, assuming that the air again leaves at a relative humidity of 90 per cent.

This reheating technique has two main advantages. Firstly, very much less air is required, because each kilogram of air picks up far more water than in a single stage system, and secondly, in order to pick up as much water in a single stage, it would be necessary to heat the air to a very much higher temperature. This reduction in the amount of air, needed simplifies the heating system, and reduces the tendency of the air to carry away any small particles.

A modern tray dryer consists of a well-insulated cabinet with integral fans and trays which are stacked on racks, or loaded on to trucks which are pushed into the dryer. Tray areas are 0.3–1 m$^2$ with a depth of material of 10–100 mm, depending on the particle size of the product. Air velocities of 1–10 m/s are used and, in order to conserve heat,
85–95 per cent of the air is recirculated. Even at these high values, the steam consumption may be 2.5–3.0 kg/kg moisture removed. The capacity of tray dryers depends on many factors including the nature of the material, the loading and external conditions, although for dyestuffs an evaporative capacity of 0.03–0.3 kg/m²·ks (0.1–1 kg/m²·h) has been quoted with air at 300–360 K(22).

**Example 16.4**

A 100 kg batch of granular solids containing 30 per cent moisture is to be dried in a tray drier to 15.5 per cent of moisture by passing a current of air at 350 K tangentially across its surface at a velocity of 1.8 m/s. If the constant rate of drying under these conditions is 0.0007 kg/s·m² and the critical moisture content is 15 per cent, calculate the approximate drying time. Assume the drying surface to be 0.03 m²/kg dry mass.

**Solution**

In 100 kg feed, mass of water = \((100 \times 30/100) = 30\) kg

and:

mass of dry solids = \((100 - 30) = 70\) kg

For \(b\) kg water in the dried solids: \(100b/(b + 70) = 15.5\)

and the water in the product, \(b = 12.8\) kg

Thus:

initial moisture content, \(u_1 = (30/70) = 0.429\) kg/kg dry solids

final moisture content, \(u_2 = (12.8/70) = 0.183\) kg/kg dry solids

and water to be removed = \((30 - 12.8) = 17.2\) kg

The surface area available for drying = \((0.03 \times 70) = 2.1\) m² and hence the rate of drying during the constant period = \((0.0007 \times 2.1) = 0.00147\) kg/s.

As the final moisture content is above the critical value, all the drying is at this constant rate and the time of drying is:

\[ t = (17.2/0.00147) = 11,700\) s or \(11.7\) ks \((3.25\) h\)

**16.5.3. Tunnel Dryers**

In tunnel dryers, a series of trays or trolleys is moved slowly through a long tunnel, which may or may not be heated, and drying takes place in a current of warm air. Tunnel dryers are used for drying paraffin wax, gelatine, soap, pottery ware, and wherever the throughput is so large that individual cabinet dryers would involve too much handling. Alternatively, material is placed on a belt conveyor passing through the tunnel, an arrangement which is well suited to vacuum operation. In typical tunnel arrangements, shown in Figure 16.9,
the construction is of block or sheet metal and the size varies over a wide range, with lengths sometimes exceeding 30 m.

### 16.5.4. Rotary Dryers

For the continuous drying of materials on a large scale, 0.3 kg/s (1 tonne/h) or greater, a rotary dryer, which consists of a relatively long cylindrical shell mounted on rollers and driven at a low speed, up to 0.4 Hz is suitable. The shell is supported at a small angle to the horizontal so that material fed in at the higher end will travel through the dryer under gravity, and hot gases or air used as the drying medium are fed in either at the upper end of the dryer to give co-current flow or at the discharge end of the machine to give countercurrent flow. One of two methods of heating is used:

(a) Direct heating, where the hot gases or air pass through the material in the dryer.

(b) Indirect heating, where the material is in an inner shell, heated externally by hot gases. Alternatively, steam may be fed to a series of tubes inside the shell of the dryer.

The shell of a rotary dryer is usually constructed by welding rolled plate, thick enough for the transmission of the torque required to cause rotation, and to support its own weight and the weight of material in the dryer. The shell is usually supported on large tyres which run on wide rollers, as shown in Figure 16.10, and although mild steel is the usual material of construction, alloy steels are used, and if necessary the shell may be coated with a plastics material to avoid contamination of the product.

With countercurrent operation, since the gases are often exhausted by a fan, there is a slight vacuum in the dryer, and dust-laden gases are in this way prevented from escaping.
This arrangement is suitable for sand, salt, ammonium nitrate and other inorganic salts, and is particularly convenient when the product is discharged at a high temperature. In this case, gas or oil firing is used and, where air is used as a drying medium, this may be filtered before heating, in order to minimise contamination of the product. As the gases leaving the dryer generally carry away very fine material, some form of cyclone or scrubber is usually fitted. Since the hot gases come into immediate contact with the dried material, the moisture content may be reduced to a minimum, though the charge may become excessively heated. Further, since the rate of heat transfer is a minimum at the feed end, a great deal of space is taken up with heating the material.

With co-current flow, the rate of passage of the material through the dryer tends to be greater since the gas is travelling in the same direction. Contact between the wet material and the inlet gases gives rise to rapid surface drying, and this is an advantage if the material tends to stick to the walls. This rapid surface drying is also helpful with materials containing water of crystallisation. The dried product leaves at a lower temperature than with countercurrent systems, and this may also be an advantage. The rapid lowering of the gas temperature as a result of immediate contact with the wet material also enables heat sensitive materials to be handled rather more satisfactorily.

Since the drying action arises mainly from direct contact with hot gases, some form of lifter is essential to distribute the material in the gas stream. This may take the form of flights, as shown in Figure 16.11, or of louvres. In the former case, the flights lift the material and then shower it across the gas stream, whilst in the latter the gas stream enters the shell along the louvres. In Figure 16.12 it may be seen that, in the rotary louvre dryer, the hot air enters through the louvres, and carries away the moisture at the end of the dryer. This is not strictly a co-current flow unit, but rather a through circulation unit, since the material continually meets fresh streams of air. The rotation of the shell, at about 0.05 Hz (3 rpm), maintains the material in agitation and conveys it through the dryer. Rotary dryers are 0.75 – 3.5 m in diameter and up to 9 m in length{23}.

The thermal efficiency of rotary dryers is a function of the temperature levels, and ranges from 30 per cent in the handling of crystalline foodstuffs to 60 – 80 per cent in the
Figure 16.11. Helical and angled lifting flights in a rotary dryer
case of inert materials. Evaporative capacities of 0.0015–0.0080 kg/m³ s may be achieved and these are increased by up to 50 per cent in louvre dryers.

In one form of indirectly heated dryer, shown in Figure 16.13, hot gases pass through the innermost cylinder, and then return through the annular space between the outer cylinders. This form of dryer can be arranged to give direct contact with the wet material during the return passage of the gases. Flights on the outer surface of the inner cylinder, and the inner surface of the outer cylinder, assist in moving the material along the dryer. This form of unit gives a better heat recovery than the single flow direct dryer, though it is more expensive. In a simpler arrangement, a single shell is mounted inside a brickwork chamber, through which the hot gases are introduced.

The steam-tube dryer, shown in Figure 16.14, incorporates a series of steam tubes, fitted along the shell in concentric circles and rotating with the shell. These tubes may be fitted with fins to increase the heat transfer surface although material may then stick to the tubes. The solids pass along the inclined shell, and leave through suitable ports at the other end. A small current of air is passed through the dryer to carry away the moisture, and the air leaves almost saturated. In this arrangement, the wet material comes in contact with very humid air, and surface drying is therefore minimised. This type of unit has a high thermal efficiency, and can be made from corrosion resisting materials without difficulty.
Design considerations

Many of the design problems associated with rotary dryers have been discussed by Friedman and Marshall\(^{(24)}\), by Prutton et al.\(^{(25)}\) and by Miller et al.\(^{(26)}\).

Heat from the air stream passes to the solid material during its fall through the air stream, and also from the hot walls of the shell, although the first mechanism is much the more important. The heat transfer equation may be written as:

\[
Q = U_a V \Delta T
\]  

(16.23)

where:
- \(Q\) is the rate of heat transfer,
- \(U\) is the overall heat transfer coefficient,
- \(V\) is the volume of the dryer,
- \(a\) is the area of contact between the particles and the gas per unit volume of dryer, and
- \(\Delta T\) is the mean temperature difference between the gas and material.

The combined group \(Ua\) has been shown\(^{(25,26)}\) to be influenced by the feed rate of solids, the air rate and the properties of the material, and a useful approximation is given by:

\[
Ua = \bar{\kappa} G^m / D
\]  

(16.24)

where \(\bar{\kappa}\) is a dimensional coefficient. Typical values for a 300 mm diameter dryer revolving at 0.08–0.58 Hz (5–35 rpm) show that \(n = 0.67\) for specific gas rates in the range 0.37–1.86 kg/m\(^2\), as given by Saeman\(^{(27)}\). The coefficient \(\bar{\kappa}\) is a function of the number of flights and, using SI units, this is given approximately by:

\[
\bar{\kappa} = 20(n_f - 1)
\]  

(16.25)
Equation 16.25 was derived for a 200 mm diameter dryer with between 6 and 16 flights \(^{(28)}\).

Combining equations 16.24 and 16.25 gives:

\[
U_a = 20(a_f - 1) G^0.67 f / D \tag{16.26}
\]

and hence for a 1 m diameter dryer with 8 flights, \(U_a\) would be about 140 W/m\(^3\) K for a gas rate of 1 kg/m\(^2\) s.

\(\text{SAEMAN}^{(27)}\) has investigated the countercurrent drying of sand in a dryer of 0.3 m diameter and 2.0 m long with 8 flights rotating at 0.17 Hz (10 rpm) and has found that the volumetric heat transfer coefficient may be correlated in terms of the hold-up of solids, as shown in Figure 16.15, and is independent of the gas rate in the range 0.25–20 kg/m\(^2\) s.

![Figure 16.15: Correlation of volumetric heat-transfer coefficients with hold-up\(^{(27)}\)](image)

Other typical values of heat transfer coefficients achieved in drying operations are:

<table>
<thead>
<tr>
<th>Application</th>
<th>Parameter</th>
<th>Value</th>
<th>Reference</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heated surface (fairly dry solid)</td>
<td>(U)</td>
<td>5–12 W/m(^2) K</td>
<td>28</td>
</tr>
<tr>
<td>Unagitated dryers</td>
<td>(U)</td>
<td>5–29 W/m(^2) K</td>
<td>29</td>
</tr>
<tr>
<td>Moderate agitation</td>
<td>(U)</td>
<td>29–85 W/m(^2) K</td>
<td>29</td>
</tr>
<tr>
<td>High agitation</td>
<td>(U)</td>
<td>85–140 W/m(^2) K</td>
<td>29</td>
</tr>
<tr>
<td>Cake-encrusted heating surface</td>
<td>(U)</td>
<td>5–29 W/m(^2) K</td>
<td>29</td>
</tr>
<tr>
<td>Light powdery materials</td>
<td>(U\Delta T)</td>
<td>950 W/m(^2)</td>
<td>24</td>
</tr>
<tr>
<td>Coarse granular materials</td>
<td>(U\Delta T)</td>
<td>6300 W/m(^2)</td>
<td>24</td>
</tr>
</tbody>
</table>

The hold-up in a rotary dryer varies with the feed rate, the number of flights, the shell diameter and the air rate. For zero air flow, \(\text{FRIEDMAN and MARSHALL}^{(24)}\) give the hold-up as:

\[
X = \frac{25.7 F'}{SN^{0.9} D} \text{ per cent of dryer volume} \tag{16.27}
\]
where: $D$ is the diameter of the drum (m),
$F'$ is the feed rate ($m^3/s \cdot m^2$),
$S$ is the slope of the dryer (m/m length),
$N$ is the rate of rotation (Hz), and
$X$ is the holdup, expressed as a percentage of the drum volume.

As the air flowrate is increased, $X$ changes and an empirical relation for the hold-up $X_a$ with air flow is:

$$X_a = X \pm KG'$$

(16.28)

although the values of $K$ are poorly defined; $X_a$ usually has a value of about 3 per cent, when working with a slope of about 0.1. In equation 16.28, the positive sign refers to countercurrent flow and the negative sign to co-current flow.

A more general approach has been made by Sharple et al.\(^{30}\) who have solved differential moisture and heat-balance equations coupled with expressions for the forward transport of solids, allowing for solids being cascaded out of lifting baffles. Typical results, shown in Figure 16.16, were obtained using a 2.6 m diameter dryer, 16 m long with a 1° slope (60 mm/m). Commercial equipment is available with diameters up to 3 m and lengths up to 30 m, and hence the correlations outlined in this section must be used with caution beyond the range used in the experimental investigations.

![Figure 16.16](image)

**Figure 16.16.** Conditions in a 2.6 m diameter × 16 m long rotary dryer with 1° slope, Co-current drying of 11.3 kg/s of fertiliser granules with 9.1 kg/s of air\(^{30}\)

In general, the allowable mass velocity of the gas in a direct-contact rotary dryer depends on the dust content of the solids and is 0.55–7.0 kg/m\(^2\) s with coarse particles. Inlet gas temperatures are typically 390–450 K for steam-heated air and 800–1100 K for flue gas, and the peripheral speed of the shell is 0.3–0.4 m/s.
Example 16.5

A flow of 0.35 kg/s of a solid is to be dried from 15 per cent to 0.5 per cent moisture on a dry basis. The mean heat capacity of the solids is 2.2 kJ/kg deg K and it is proposed that a co-current adiabatic dryer should be used with the solids entering at 300 K and, because of the heat sensitive nature of the solids, leaving at 325 K. Hot air is available at 400 K with a humidity of 0.01 kg/kg dry air and the maximum allowable mass velocity of the air is 0.95 kg/m^2 s. What diameter and length should be specified for the proposed dryer?

Solution

With an inlet air temperature and humidity of 400 K and 0.01 kg/kg dry air respectively from Figure 13.4 in Volume 1, the inlet wet-bulb temperature \( T_1 \) is given by:

\[ T_1 = \ln\left(\frac{400 - 312}{(T_0 - 312)}\right) \]

The solids outlet temperature will be taken as the maximum allowable, 325 K.

From the steam tables in the Appendix, the latent heat of vaporisation of water at 312 K is 2410 kJ/kg. Again from steam tables, the specific heat capacity of water vapour is 4.187 kJ/kg K and that of the solids will be taken as 2.18 kJ/kg K.

For a mass flow of solids of 0.35 kg/s and inlet and outlet moisture contents of 0.15 and 0.005 kg/kg dry solids respectively, the mass of water evaporated is:

\[ 0.35(0.15 - 0.005) = 0.0508 \text{ kg/s} \]

For unit mass of solids, the heat duty includes:

- heat to the solids
- heat to raise the moisture to the dew point
- heat of vaporisation
- heat to raise remaining moisture to the solids outlet temperature
- heat to raise evaporated moisture to the air outlet temperature

A total of:

\[ (54.5 + 7.5 + 349.5 + 0.3 + 5.3) = 417.1 \text{ kJ/kg} \]

or:

\[ (417.1 \times 0.35) = 146 \text{ kW} \]

From Figure 13.4 in Volume 1, the humid heat of the entering air is 1.03 kJ/kg K and making a heat balance:

\[ G_1 (1 + 0.01) = Q/C_p (T_1 - T_2) \]

where:

- \( G_1 \) (kg/s) is the mass flowrate of inlet air,
- \( \rho_1 \) (kg/kg) is the humidity of inlet air,
- \( Q \) (kW) is the heat duty,
- \( C_p \) (kJ/kg K) is the humid heat of inlet air

and:

- \( T_1 \) and \( T_2 \) (K) are the inlet and outlet air temperatures respectively.

In this case:

\[ G_1(1 + 0.01) = 146/(1.03(400 - 331.5)) = 2.07 \text{ kg/s} \]

and:

mass flowrate of dry air, \( G_0 = (2.07/1.01) = 2.05 \text{ kg/s} \)
The humidity of the outlet air is then:

\[ H_2 = 0.01 + \frac{(0.0508/2.05)}{0.347} = 0.0347 \text{ kg/kg} \]

At a dry bulb temperature of 331.5 K, with a humidity of 0.0347 kg/kg, the wet-bulb temperature of the outlet air, from Figure 13.4 in Volume 1, is 312 K, the same as the inlet, which is the case for adiabatic drying.

The dryer diameter is then found from the allowable mass velocity of the air and the entering air flow and for a mass velocity of 0.95 kg/m\(^2\)s, the cross sectional area of the dryer is:

\[ \left( \frac{2.07}{0.95} \right)^2 = 2.18 \text{ m}^2 \]

equivalent to a diameter of \((4 \times 2.18)/\pi)^{0.5} = 1.67 \text{ m} \]

With a constant drying temperature of 312 K,

at the inlet: \( \Delta T_1 = (400 - 312) = 88 \text{ deg K} \)

and at the outlet: \( \Delta T_2 = (331.5 - 312) = 19.5 \text{ deg K} \)

and the logarithmic mean, \( \Delta T_{lm} = (88 - 19.5)/\ln(88/19.5) = 45.5 \text{ deg K} \).

The length of the dryer, \( L \), is then: \( L = \frac{Q}{(0.0625\pi D G' \Delta T_{lm})} \)

where \( D \) (m) is the diameter

and \( G' \) (kg/m\(^2\)s) is the air mass velocity.

In this case: \( L = \frac{146/[0.0625\pi \times 1.67(0.95)^{0.67} \times 45.5)] = 10.1 \text{ m} \)

This gives a length/diameter ratio of \((10.1/1.67) = 6\), which is a reasonable value for rotary dryers.

**16.5.5. Drum dryers**

If a solution or slurry is run on to a slowly rotating steam-heated drum, evaporation takes place and solids may be obtained in a dry form. This is the basic principle used in all drum dryers, some forms of which are illustrated in Figure 16.17. The feed to a single drum dryer may be of the dip, pan, or splash type. The dip-feed system, the earliest design, is still used where liquid can be picked up from a shallow pan. The agitator prevents settling of particles, and the spreader is sometimes used to produce a uniform coating on the drum. The knife, which is employed for removing the dried material, functions in a similar manner to the doctor blade on a rotary filter. If the material is dried to give a free-flowing powder, this comes away from the drum quite easily. The splash-feed type is used for materials, such as calcium arsenate, lead arsenate, and iron oxide, where a light fluffy product is desired. The revolving cylinder throws the liquor against the drum, and a uniform coating is formed with materials which do not stick to the hot surface of the drum.

Double drum dryers may be used in much the same way, and Figure 16.17 shows dip-feed and top-feed designs. Top-feed gives a larger capacity as a thicker coating is obtained. It is important to arrange for a uniform feed to a top-feed machine, and this may be effected by using a perforated pipe for solutions and a travelling trough for suspensions.
Drums are usually made from cast iron, although chromium-plated steel or alloy steel is often used where contamination of the product must be avoided, such as with pharmaceuticals or food products. Arrangements must be made for accurate adjustment of the separation of the drums, and the driving gears should be totally enclosed. A range of speeds is usually obtained by selecting the gears, rather than by fitting a variable speed drive. Removal of the steam condensate is important, and an internal syphon is often fitted to keep the drum free of condensate. In some cases, it is better for the drums to be rotated upwards at the point of closest proximity, and the knives are then fitted at the bottom. By this means, the dry material is kept away from the vapour evolved. Some indication of the sizes of this type of dryer is given in Table 16.3, and it may be noted that the surface of each drum is limited to about 35 m$^2$. This, coupled with developments in the design of spray dryers, renders the latter more economically attractive, especially where large throughputs are to be handled. For steam-heated drum dryers, normal evaporative capacities are proportional to the active drum area and are 0.003–0.02 kg/m$^2$ s, although higher rates are claimed for grooved drum dryers.
Table 16.3: Sizes of double drum dryers

<table>
<thead>
<tr>
<th>Drum dimensions (m)</th>
<th>Length (m)</th>
<th>Width (m)</th>
<th>Height (m)</th>
<th>Mass (kg)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.61 × 0.61</td>
<td>3.5</td>
<td>2.05</td>
<td>2.3</td>
<td>3.850</td>
</tr>
<tr>
<td>0.61 × 0.91</td>
<td>3.8</td>
<td>2.05</td>
<td>2.4</td>
<td>4.170</td>
</tr>
<tr>
<td>0.81 × 1.32</td>
<td>5.0</td>
<td>2.5</td>
<td>2.8</td>
<td>7.620</td>
</tr>
<tr>
<td>0.81 × 1.83</td>
<td>5.5</td>
<td>2.5</td>
<td>2.8</td>
<td>8.350</td>
</tr>
<tr>
<td>0.81 × 2.28</td>
<td>5.9</td>
<td>2.5</td>
<td>2.8</td>
<td>8.890</td>
</tr>
<tr>
<td>0.81 × 2.54</td>
<td>6.25</td>
<td>2.5</td>
<td>2.8</td>
<td>9.300</td>
</tr>
<tr>
<td>0.81 × 3.05</td>
<td>6.7</td>
<td>2.5</td>
<td>2.8</td>
<td>10.120</td>
</tr>
<tr>
<td>1.07 × 2.28</td>
<td>6.4</td>
<td>3.0</td>
<td>3.0</td>
<td>14.740</td>
</tr>
<tr>
<td>1.07 × 2.54</td>
<td>6.7</td>
<td>3.0</td>
<td>3.0</td>
<td>15.420</td>
</tr>
<tr>
<td>1.07 × 3.05</td>
<td>7.2</td>
<td>3.0</td>
<td>3.0</td>
<td>16.780</td>
</tr>
<tr>
<td>1.52 × 3.66</td>
<td>7.9</td>
<td>4.1</td>
<td>4.25</td>
<td>27.220</td>
</tr>
</tbody>
</table>

The solid is usually in contact with the hot metal for 6–15 s, short enough to prevent significant decomposition of heat sensitive materials, and heat transfer coefficients are 1–2 kW/m² K.

When the temperature of the drying material must be kept as low as possible, vacuum drying is used, and one form of vacuum dryer is shown in Figure 16.18. The dried material is collected in two screw conveyors and carried usually to two receivers so that one can be filled while the other is emptied.

16.5.6. Spray dryers

Water may be evaporated from a solution or a suspension of solid particles by spraying the mixture into a vessel through which a current of hot gases is passed. In this way, a
large interfacial area is produced and consequently a high rate of evaporation is obtained. Drop temperatures remain below the wet bulb temperature of the drying gas until drying is almost complete, and the process thus affords a convenient means of drying substances which may deteriorate if their temperatures rise too high, such as milk, coffee, and plasma. Furthermore, because of the fine state of subdivision of the liquid, the dried material is obtained in a finely divided state.

In spray drying, it is necessary to atomise and distribute, under controlled conditions, a wide variety of liquids, the properties of which range from those of solutions, emulsions, and dispersions, to slurries and even gels. Most of the atomisers commonly employed are designed for simple liquids, that is mobile Newtonian liquids. When atomisers are employed for slurries, pastes, and liquids having anomalous properties, there is a great deterioration in performance and, in many cases, atomisers may be rapidly eroded and damaged so as to become useless. There is therefore much to be gained by considering various types and designs of atomiser so that a suitable selection can be made for the given duty.

The performance of a spray dryer or reaction system is critically dependent on the drop size produced by the atomiser and the manner in which the gaseous medium mixes with the drops. In this context an atomiser is defined as a device which causes liquid to be disintegrated into drops lying within a specified size range, and which controls their spatial distribution.

**Atomisers**

Atomisers are classified in Table 16.4(31) according to the three basic forms of energy commonly employed — pressure energy, centrifugal kinetic energy, and gaseous energy. Where greater control is required over disintegration or spatial dispersion, combinations of atomiser types may be employed, and, for example, swirl-spray nozzles or spinning discs may be incorporated in a blast atomiser, their primary functions being to produce thin liquid sheets which are then eventually atomised by low, medium or high velocity gas streams.

The fundamental principle of disintegrating a liquid consists in increasing its surface area until it becomes unstable and disintegrates. The theoretical energy requirement is the increase in surface energy plus the energy required to overcome viscous forces, although in practice this is only a small fraction of the energy required. The process by which drops are produced from a liquid stream depends upon the nature of the flow in the atomiser, that is whether it is laminar or turbulent, the way in which energy is imparted to the liquid, the physical properties of the liquid, and the properties of the ambient atmosphere. The basic mechanism is, however, unaffected by these variables and consists essentially of the breaking down of unstable threads of liquid into rows of drops and conforms to the classical mechanism postulated by Lord Rayleigh(32). This theory states that a free column of liquid is unstable if its length is greater than its circumference, and that, for a non-viscous liquid, the wavelength of that disturbance which will grow most rapidly in amplitude is 4.5 times the diameter. This corresponds to the formation of droplets of diameter approximately 1.89 times that of the jet $d_j$. Weber(33) has shown that the
Table 16.4. Classification of atomisers

<table>
<thead>
<tr>
<th>Type of Atomiser</th>
<th>Pressure Used in Centrifugal Energy Used in Rotary Atomisers</th>
<th>Gaseous Energy Used in Twin-fluid or Blast Atomisers</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fan-spray nozzles</td>
<td>Spinning cups</td>
<td>External mixing (fluid pressures independent)</td>
</tr>
<tr>
<td>0.25–1.0 MN/m²</td>
<td>6–30 m/s</td>
<td>Internal mixing (fluid pressures interdependent)</td>
</tr>
<tr>
<td>Impact nozzles</td>
<td>Spinning discs</td>
<td>Low velocity</td>
</tr>
<tr>
<td>Impinging jet nozzles</td>
<td>Flat discs</td>
<td>Gas velocity</td>
</tr>
<tr>
<td>0.25–1.0 MN/m²</td>
<td>Saucer-shaped discs</td>
<td>30–120 m/s</td>
</tr>
<tr>
<td>Impact plate nozzles</td>
<td>Radial-vaned discs</td>
<td>Gas–liquid ratio</td>
</tr>
<tr>
<td>Up to 3.0 MN/m²</td>
<td>Multiple discs</td>
<td>2–25 kg/kg</td>
</tr>
<tr>
<td>Deflector nozzles</td>
<td>30–180 m/s</td>
<td></td>
</tr>
<tr>
<td>Swirl-spray nozzles</td>
<td>Hollow cone, full cone</td>
<td></td>
</tr>
<tr>
<td>0.4–70 MN/m²</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Divergent pintle nozzles</td>
<td></td>
<td></td>
</tr>
<tr>
<td>0.25–7.0 MN/m²</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fixed or vibrating pintle</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Optimum wavelength for the disruption of jets of viscous liquid is:

$$\lambda_{opt} = \sqrt{\frac{2\pi d_j}{1 + \frac{3\mu}{\sqrt{\rho \sigma d_j}}}}^{0.5} \quad (16.29)$$

A uniform thread will break down into a series of drops of uniform diameter, each separated by one or more satellite drops. Because of the heterogeneous character of the atomisation process, however, non-uniform threads are produced and this results in a wide range of drop sizes. An example of a part of a laminar sheet collapsing into a network of threads and drops is shown in Figure 16.19. Only when the formation and disintegration of threads are controlled can a homogeneous spray cloud be produced. One method by which this can be achieved is by using a rotary cup atomiser operating within a critical range of liquid flow rates and rotor speeds, although, as shown later, this range falls outside that normally employed in practice.

Although certain features are unique to particular atomiser types, many of the detailed mechanisms of disintegration are common to most forms of atomiser. The most effective way of utilising energy imparted to a liquid is to arrange that the liquid mass has a large specific surface before it commences to break down into drops. Thus the primary function of an atomiser is to transpose bulk liquid into thin liquid sheets. Three modes of disintegration of such spray sheets have been established, namely rim, perforated sheet, and wave. Because of surface tension, the free edge of any sheet contracts into a thick rim, and rim disintegration occurs as it breaks up by instabilities similar to those

$\mu$ - viscosity of the liquid
$\rho$ - density of the liquid
$\sigma$ - surface tension of the liquid
Figure 16.19. Modes of disintegration of liquid films

of free jets. Figure 16.19a illustrates a fan spray sheet and shows that, as the liquid in each edge moves along the curved boundary, the latter becomes disturbed and disintegrates. When this occurs, the resulting drops sustain the direction of flow of the edge at the point at which the drops are formed, and remain attached to the receding surface by thin threads which rapidly disintegrate into streams of drops.

In perforated-sheet disintegration shown in Figure 16.19b, small holes suddenly appear in the sheet as it advances into the atmosphere. They rapidly grow in size until the thickening rims of adjacent holes coalesce to form threads of varying diameter. The threads finally break down into drops.

Disintegration can also occur through the superimposition of a wave motion on the sheet, as shown in Figure 16.19c. Sheets of liquid corresponding to half or full wavelengths of liquid are torn off and tend to draw up under the action of surface tension, although these may suffer disintegration by air action or liquid turbulence before a regular network of threads can be formed.

In a pressure atomiser, liquid is forced under pressure through an orifice, and the form of the resulting liquid sheet can be controlled by varying the direction of flow towards the orifice. By this method, flat and conical spray sheets can be produced. From the Bernoulli equation, given in Volume 1, Chapter 6, the mass rate of flow through a nozzle may be derived as:

$$ G = C_D \rho A_N \sqrt{2(-\Delta P)} \rho $$  \hspace{1cm} (16.30)

For a given nozzle and fluid, and an approximately constant coefficient of discharge, $C_D$, then:

$$ G = \text{constant} \sqrt{(-\Delta P)} $$  \hspace{1cm} (16.31)

The capacity of a nozzle is conveniently described by the flow number, $FN$ a dimensional constant defined by:

$$ FN = \frac{\text{Volumetric flow (gal/h)}}{\sqrt{\text{Pressure (lb/in}^2\text{)}}} = 2.08 \times 10^6 \frac{\text{Volume flow (m}^3/\text{s})}{\sqrt{\text{Pressure (kN/m}^2\text{)}}} $$  \hspace{1cm} (16.32)
In the fan-spray drop nozzle shown in Figure 16.20a, two streams of liquid are made to impinge behind an orifice by specially designed approach passages, and a sheet is formed in a plane perpendicular to the plane of the streams. The orifice runs full and, since the functional portion is sharp-edged, high discharge coefficients are obtained which are substantially constant over wide ranges of Reynolds number.

The influence of conditions on the droplet size where the spray sheet disintegrates through aerodynamic wave motion may be represented by the following expression proposed by Dombrowski and Munday\(^{(34)}\) for ambient densities around normal atmospheric conditions:

\[
d_m = \left( \frac{0.000156}{C_p} \right) \left( \frac{F_N \sigma \rho}{\sin \phi (-\Delta P)} \right)^{1/3} \rho_A^{-1/6}\]

(16.33)

where \(F_N\) is given in equation 16.32, \(\phi\) is the half-angle of the sheet, \(-\Delta P\) is in kN/m\(^2\) and all other quantities are expressed in SI units.

The principle of operation of the impinging jet nozzle resembles that of the fan spray nozzle with the exception that two or more independent jets are caused to impinge in the atmosphere. In impact atomisers, one jet is caused to strike against a solid surface, and for two jets impinging at 180°\(^{(34)}\), using SI units:

\[
d_m = 1.73 \left( \frac{d_{0.75}}{d_{0.5}} \right)^{0.5} \left( \frac{\alpha}{\rho} \right)^{0.25}\]

(16.34)
With this atomiser, the drop size is effectively independent of viscosity, and the size spectrum is narrower than with other types of pressure nozzle.

When liquid is caused to flow through a narrow divergent annular orifice or around a pintle against a divergent surface on the end of the pintle, a conical sheet of liquid is produced where the liquid is flowing in radial lines. Such a sheet generally disintegrates by an aerodynamic wave motion. The angle of the cone and the root thickness of the sheet may be controlled by the divergence of the spreading surface and the width of the annulus. For small outputs this method is not so favourable because of difficulties in making an accurate annulus. A conical sheet is also produced when the liquid is caused to emerge from an orifice with a tangential or swirling velocity resulting from its path through one or more tangential or helical passages before the orifice. Figure 16.20 shows a typical nozzle used for a spray dryer. In such swirl-spray nozzles the rotational velocity is sufficiently high to cause the formation of an air core throughout the nozzle, resulting in low discharge coefficients for this type of atomiser.

Several empirical relations have been proposed to express drop size in terms of the operating variables. One suitable for small atomisers with $85^\circ$ spray cone angles, at atmospheric pressure is

$$d_m = 0.0134 \left( \frac{FN^{0.209} (\mu/\rho)^{0.215}}{(-\Delta P)^{0.348}} \right)$$  \hspace{1cm} (16.35)

where $-\Delta P$ is in kN/m$^2$, FN is given in equation 16.32, and all other quantities are in SI units.

Pressure nozzles are somewhat inflexible since large ranges of flowrate require excessive variations in differential pressure. For example, for an atomiser operating satisfactorily at 275 kN/m$^2$, a pressure differential of 17.25 MN/m$^2$ is required to increase the flowrate to ten times its initial value. These limitations, inherent in all pressure-type nozzles, have been overcome in swirl spray nozzles by the development of spill, duplex, multi-orifice, and variable port atomisers, in which ratios of maximum to minimum outputs in excess of 50 can be easily achieved\textsuperscript{[34].}

In a rotary atomiser, liquid is fed on to a rotating surface and spread out by centrifugal force. Under normal operating conditions the liquid extends from the periphery in the form of a thin sheet which breaks down some distance away from the periphery, either freely by aerodynamic action or by the action of an additional gas blast. Since the accelerating force can be controlled independently, this type of atomiser is extremely versatile and it can handle successfully a wide range of feed rates with liquids having a wide range of properties. The rotating component may be a simple flat disc though slippage may then occur, and consequently it is more usual to use bowls, vaned discs, and slotted wheels, as shown in Figure 16.21. Diameters are 25–460 mm and small discs rotate at up to 1000 Hz while the larger discs rotate at up to 200 Hz with capacities of 1.5 kg/s. Where a coaxial gas blast is used to effect atomisation, lower speeds of the order of 50 Hz may be used. At very low flowrates, such as 30 mg/s, the liquid spreads out towards the cup lip where it forms a ring. As liquid continues to flow into the ring, its inertia increases, overcomes the restraining surface tension and viscous forces and is centrifuged off as discrete drops of uniform size, which initially remain attached to the rim by a fine attenuating thread. When the drop is finally detached, the thread breaks down into a chain of small satellite drops.
Since the satellite drops constitute only a small proportion of the total liquid flowrate, a cup operating under these conditions effectively produces a monodisperse spray. Under these conditions the drop size from sharp-edged discs has been given by Walton and Prewett as:

$$d_m = \frac{0.52}{N} \left( \frac{\sigma}{D \rho} \right)$$

(16.36)

When the liquid flowrate is increased, the retaining threads grow in thickness and form long jets. As they extend into the atmosphere, these jets are stretched and finally break down into strings of drops. Under more practical ranges of flowrate, the jets are unable to remove all the liquid, the ring is forced away and a thin sheet extends around the cup lip and eventually breaks up into a polydisperse spray, as shown in Figure 16.22.

A far greater supply of energy for disintegration of the liquid jet can be provided by using a high-speed gas stream which impinges on a liquid jet or film. By this means a greater surface area is formed and drops of average size less than 20 µm can be produced. This is appreciably smaller than is possible by the methods previously described, although the energy requirements are much greater. The range of flowrates which can be used is wide, because the supply of liquid and the energy for atomisation can be controlled independently. Gas velocities ranging from 50 m/s to sonic velocity are common, and sometimes the gas is given a vortex motion.

Break-up of the jet occurs as follows. Ligaments of liquid are torn off, which collapse to form drops. These may be subsequently blown out into films, which in turn further collapse to give a fine spray. Generally, this spray has a small cone angle and is capable of penetrating far greater distances than the pressure nozzle. Small atomisers of this type have been used in spray-drying units of low capacities.
Where the gas is impacted on to a liquid jet, the mean droplet size is given approximately by:

\[
d_s = \frac{0.585}{u_c} \sqrt{\left( \frac{\sigma}{\rho} \right) + 0.0017 \left( \frac{\mu}{\sqrt{\sigma \rho}} \right)^{0.45} \left( \frac{1000}{j} \right)^{1.5}} \tag{16.37}
\]
where: $d_0$ is the surface mean diameter (m),

- $j$ is the volumetric ratio of gas to liquid rates at the pressure of the surrounding atmosphere,

- $u_r$ is the velocity of the gas relative to the liquid (m/s),

- $\mu$ is the liquid viscosity (Ns/m$^2$),

- $\rho$ is the liquid density (kg/m$^3$), and

- $\sigma$ is the surface tension (N/m).

**As Fraser et al. point out, more efficient atomisation is achieved if the liquid is spread out into a film before impact.**

**Drying of drops**

The amount of drying that a drop undergoes depends upon the rate of evaporation and the contact time, the latter depending upon the velocity of fall and the length of path through the dryer. The terminal velocity and the transfer rate depend upon the flow conditions around the drop. Because of the nature of the flow pattern, the latter also varies with angular position around the drop, although no practical design method has incorporated such detail and the drop is always treated as if it evaporates uniformly from all its surface.

There are two main periods of evaporation. When a drop is ejected from an atomiser its initial velocity relative to the surrounding gas is generally high and very high rates of transfer are achieved. The drop is rapidly decelerated to its terminal velocity, however, and the larger proportion of mass transfer takes place during the free-fall period. Little error is therefore incurred in basing the total evaporation time on this period.

An expression for the evaporation time for a pure liquid drop falling freely in air has been presented by Marshall$^{38,39}$. For drop diameters less than 100 $\mu$m this may be simplified to give:

$$t = \frac{\rho \lambda (d_0^3 - d_t^3)}{8k_f \Delta T} \quad \text{(16.38)}$$

Sprays generally contain a wide range of drop sizes, and a stepwise procedure can be used$^{39}$ to determine the size spectrum as evaporation proceeds.

When single drops containing solids in suspension or solutions are suspended in hot gas streams, it is found that evaporation initially proceeds in accordance with equation 16.38 for pure liquid, although when solids deposition commences, a crust or solid film is rapidly formed which increases the resistance to transfer. Although this suggests the existence of a falling rate period similar to that found in tray-drying, the available published data indicate that it has little effect on the total drying time. As a result of crust formation, the dried particles may be in the form of hollow spheres.

**Industrial spray dryers**

Spray dryers are used in a variety of applications where a fairly high grade product is to be made in granular form. In the drying chamber the gas and liquid streams are brought into contact, and the efficiency of mixing depends upon the flow patterns induced in the chamber. Rotating disc atomisers are most commonly used. Countercurrent dryers give
the highest thermal efficiencies although product temperatures are higher in these units. This limits their use to materials which are not affected by overheating. Co-current dryers suffer from relatively low efficiencies, although they have the advantage of low product temperatures unless back-mixing occurs. In the case of materials which are extremely sensitive to heat, great care has to be taken in the design of the chamber to avoid overheating. Combustion gases are frequently used directly although, in some cases, such as the preparation of food products, indirectly heated air is used. Maximum temperatures are then normally limited to lower values than those with direct heating. Typical flow arrangements in spray drying are shown in Figure 16.23.

Figure 16.23. Flow arrangements in spray-drying (21)
The drying time and size of the particles are directly related to the droplet size, and therefore the initial formation of the spray is of great importance. The factors which govern the choice of atomisers for any specific drying application are principally dependent upon the characteristics of the liquid feed and upon the required drying characteristics of the drying chamber. A general guide is given in Table 16.5.

Table 16.5. Choice of atomiser

<table>
<thead>
<tr>
<th>Atomiser</th>
<th>Pressure</th>
<th>Rotary</th>
<th>Twin-fluid</th>
<th>Spinning-cup plus gas blast</th>
</tr>
</thead>
<tbody>
<tr>
<td>Drying chamber</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Co-current</td>
<td>*</td>
<td>*</td>
<td>*</td>
<td>*</td>
</tr>
<tr>
<td>Countercurrent</td>
<td>*</td>
<td>*</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Feed</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Low viscosity solutions</td>
<td>*</td>
<td></td>
<td>*</td>
<td></td>
</tr>
<tr>
<td>High viscosity solutions</td>
<td></td>
<td>*</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Slurries</td>
<td>*</td>
<td>*</td>
<td>*</td>
<td></td>
</tr>
<tr>
<td>Pastes</td>
<td>*</td>
<td>*</td>
<td>*</td>
<td></td>
</tr>
<tr>
<td>Flexibility</td>
<td>Flowrate $\alpha \sqrt{(-\Delta P)}$</td>
<td>Flowrate independent of cup speed</td>
<td>Liquid flow independent of gas energy</td>
<td>Liquid flow independent of cup speed and gas energy</td>
</tr>
</tbody>
</table>

Pressure nozzles are most suited to low viscosity liquids and, where possible, viscous liquids should be preheated to ensure the minimum viscosity at the nozzle. Because of their simplicity, pressure nozzles are also employed to atomise viscous liquids with a kinematic viscosity up to 0.01 m²/s, depending upon the nozzle capacity. Under these conditions, injection pressures of up to 50 MN/m² may be required to produce the required particle size. With slurries, the resulting high liquid velocities may cause severe erosion of the orifice and thus necessitate frequent replacement.

Spinning discs are very suitable for slurries and pastes, while high viscosity liquids tend to produce a stringy product. Care must also be taken in design to minimise incrustation around the lip and subsequent out-of-balance as drying takes place.

The simple gas atomiser is inherently fairly flexible although it has not yet found widespread application. This is a result of its tendency to produce a dusty product containing a large proportion of very small particles.

Often, little difficulty is experienced in removing the majority of the dried product, though in most cases the smaller particles that may be carried over in the exit gases must be reclaimed. Cyclones are the simplest form of separator though bag filters or even electrostatic precipitators may be required. With heat-sensitive materials, and in cases where sterility is of prime importance, more elaborate methods are required. For example, cooling streams of air may be used to aid the extraction of product while maintaining the required low temperature. Mechanical aids are often incorporated to prevent particles adhering to the chamber walls, and, in one design, the cooling air also operates a revolving device which sweeps the walls.

In some cases all the product is conveyed from the dryer by the exhaust gases and collected outside the drying chamber. This method is liable to cause breakage of the
particles though it is particularly suited for heat-sensitive materials which may deteriorate if left in contact with hot surfaces inside the dryer.

Spray drying has generally been regarded as a relatively expensive process, especially when indirect heating is used. The data given in Table 16.6 taken from Gross and Duffield using 1990 costs, illustrate the cost penalties associated with indirect heating or with low inlet temperatures in direct heating.

**Table 16.6. Operating costs for a spray dryer evaporating 0.28 kg/s (1 tonne/h) of water**

<table>
<thead>
<tr>
<th>Air inlet temperature (K)</th>
<th>Steam flow (kg/s)</th>
<th>Oil flow (kg/s)</th>
<th>Power consumption (kW)</th>
<th>Cost (£/100 h)</th>
<th>Cost (£/Mg)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Air heated:</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Indirectly</td>
<td>453</td>
<td>0.708</td>
<td>71</td>
<td>2550</td>
<td>8150</td>
</tr>
<tr>
<td>Direct combustion</td>
<td>523</td>
<td>0.033</td>
<td>55</td>
<td>1800</td>
<td>5820</td>
</tr>
<tr>
<td>Direct combustion</td>
<td>603</td>
<td>0.031</td>
<td>47</td>
<td>1650</td>
<td>5300</td>
</tr>
<tr>
<td></td>
<td>353</td>
<td></td>
<td></td>
<td>2934</td>
<td></td>
</tr>
<tr>
<td></td>
<td>353</td>
<td>0.033</td>
<td>55</td>
<td>2096</td>
<td></td>
</tr>
<tr>
<td></td>
<td>353</td>
<td>0.031</td>
<td>47</td>
<td>1904</td>
<td></td>
</tr>
</tbody>
</table>

Quinn has drawn attention to the advantages with larger modern units using higher air inlet temperatures, 675 K for organic products and 925 K for inorganic products.

In spray dryers, using either a nozzle or rotating disc as the atomiser, volumetric evaporative capacities are 0.0003–0.0014 kg/m³ s for cross-and co-current flows, with drying temperatures of 420–470 K. For handling large volumes of solutions, spray dryers are unsurpassed, and it is only at feed rates below 0.1 kg/s, that a drum dryer becomes more economic. Indeed the economy of spray drying improves with capacity until, at evaporative capacities of greater than 0.6 kg/s, the unit running cost is largely independent of scale.

In the jet spray dryer, cold feed is introduced into preheated primary air which is blown through a nozzle at velocities up to 400 m/s. Very fine droplets are obtained with residence times of around 0.01 s, and an air temperature of 620 K. This equipment has been used for evaporating milk without adverse effect on flavour and, although operating costs are likely to be high, the system is well suited to the handling of heat-sensitive materials.

### 16.5.7. Pneumatic dryers

In pneumatic dryers, the material to be dried is kept in a state of fine division, so that the surface per unit volume is high and high rates of heat transfer are obtained. Solids are introduced into the dryer by some form of mechanical feeder, such as a rotating star wheel, or by an extrusion machine arranged with a high-speed guillotine to give short lengths of material, such as 5–10 mm. Hot gases from a furnace, or more frequently from an oil burner, are passed into the bottom of the dryer, and these pick up the particles and carry them up the column. The stream of particles leaves the dryer through a cyclone.
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separator and the hot gases pass out of the system. In some instances, final collection of the fine particles is by way of a series of bag filters. The time of contact of particles with the gases is small, typically 5 s, even with a lengthy duct—and the particle temperature does not approach the temperature of the hot gas stream. In some cases the material is recycled, especially where bound moisture is involved. Evaporative capacities, which are high, are greatly affected by the solids—air ratio. Typical thermal and power requirements are given by Quinn as 4.5 MJ/kg moisture evaporated and 0.2 MJ/kg, respectively.

A typical installation with the associated equipment is shown in Figure 16.24. Wet feed is delivered on to a bed of previously dried material in a double-paddle mixer to produce a friable mixture. This then passes to a cage mill where it comes into contact with hot gases from the furnace. Surface and some inherent moisture is immediately flash evaporated. The cage mill breaks up any agglomerates to ensure uniform drying of each individual particle. Gases and product are drawn up the drying column, where inherent moisture continues to be evaporated, before passing into the cyclone collector. Separated solids discharge through a rotary air lock into a dry divider which is set to recycle an
adequate percentage of solids for conditioning the new wet feed in the double paddle mixer. Gases from the cyclone are vented to atmosphere through a suitable dust collector or wet scrubber. The system operates under suction and dust is therefore reduced to a minimum. A direct oil or gas fired furnace is generally employed and the heat input is controlled according to the vent stack gas temperature. Indirect heating may be used where contamination of the product is undesirable.

Materials handled include food products, chalk, coal, organic chemicals, clays, spent coffee grounds, sewage sludge and chicken manure. Where exhaust gases have unpleasant odours, after-burners can be supplied to raise the temperature and burn off the organic and particulate content causing the problem.

Convex dryers are continuously operating pneumatic dryers with an inherent classifying action in the drying chamber which gives residence times for the individual particles differing according to particle size and moisture content. Such units offer the processing advantages of short-time, co-current dryers and are used primarily for drying reasonably to highly free-flowing moist products that can be conveyed pneumatically and do not tend to stick together. By virtue of the pronounced classifying action, such dryers are also well suited to the drying of thermally sensitive moist materials with widely differing particle sizes where the large particles have to be completely dried without any overheating of the small ones. Basically, this form of pneumatic dryer consists of a truncated cyclone with a bottom outlet that acts as a combined classifier and dryer.

16.5.8. Fluidised bed dryers

The principles of fluidisation, discussed in Chapter 6, are applied in this type of dryer, shown typically in Figure 16.25. Heated air, or hot gas from a burner, is passed by way of a plenum chamber and a diffuser plate, fitted with suitable nozzles to prevent any back-flow of solids, into the fluidised bed of material, from which it passes to a dust separator. Wet material is fed continuously into the bed through a rotary valve, and this mixes immediately with the dry charge. Dry material overflows through a downcomer to an integral after-cooler. An alternative design of this type of dryer is one in which a thin bed is used.

Quinn\(^{43}\) points out that, whilst it would seem impossible to obtain very low product moisture levels when the incoming feed is very wet and at the same time ensure that the feed point is well away from the discharge point, this is not borne out by operating experience. Mixing in the bed is so rapid that it may be regarded as homogeneous, and baffles or physical separation between feed and discharge points are largely ineffective. The very high mass-transfer rates achieved make it possible to maintain the whole bed in a dry condition. Some rectangular fluidised-bed dryers have separately fluidised compartments through which the solids move in sequence. The residence time is very similar for all particles and the units, known as plug-flow dryers, often employ cold air to effect product cooling in the last stage.

Many large-scale uses include the drying of fertilisers, plastics materials, foundry sand, and inorganic salts, and Agarwal, Davis, and King\(^{44}\) describe a plant consisting of two units, each drying 10.5 kg/s of fine coal. Small fluidised-bed dryers also find use in, for example, the drying of tablet granulations in the pharmaceutical industry\(^{45}\).
Dryers with grid areas of up to 14 m$^2$ have been built and evaporative capacities vary from 0.02 kg/s m$^2$ grid area for the low-temperature drying of food grains to 0.3 kg/s m$^2$ for the drying of pulverised coal by direct contact with flue gases. Specific air rates are usually 0.5–2.0 kg/s m$^2$ grid area and the total energy demand is 2.5–7.5 MJ/kg moisture evaporated. The exit gas is nearly always saturated with vapour for all allowable fluidisation velocities.

The choice between spray, pneumatic and fluidised dryers depends very much on the properties of the particles and some guidance in this respect is given in Table 16.7.

<table>
<thead>
<tr>
<th>Particle property</th>
<th>Spray dryer</th>
<th>Pneumatic dryer</th>
<th>Fluidised-bed dryer</th>
</tr>
</thead>
<tbody>
<tr>
<td>Initial moisture greater than 80 per cent</td>
<td>Yes</td>
<td>No</td>
<td>No</td>
</tr>
<tr>
<td>Too dry to pump</td>
<td>No</td>
<td>Yes</td>
<td>Yes</td>
</tr>
<tr>
<td>Wet enough to pump but moisture less than 80 per cent</td>
<td>Yes</td>
<td>Yes</td>
<td>No</td>
</tr>
<tr>
<td>Solids in dissolved state</td>
<td>Yes</td>
<td>No</td>
<td>No</td>
</tr>
<tr>
<td>Partially dry but sticky particles</td>
<td>Yes</td>
<td>No</td>
<td>No</td>
</tr>
<tr>
<td>Fragile particles</td>
<td>Yes</td>
<td>No</td>
<td>Possible</td>
</tr>
<tr>
<td>Very small particles</td>
<td>Yes</td>
<td>Yes</td>
<td>No</td>
</tr>
<tr>
<td>Residence time (s)</td>
<td>3–10</td>
<td>1–10, often</td>
<td>Widely variable,</td>
</tr>
<tr>
<td>Heat sensitive material</td>
<td>Yes</td>
<td>Yes</td>
<td>No</td>
</tr>
<tr>
<td>Relative drying speed</td>
<td>Third</td>
<td>First</td>
<td>Second</td>
</tr>
</tbody>
</table>
Design considerations

A simple, concise method for the preliminary sizing of a fluidised bed dryer has been proposed by Clark\(^{47}\) and this is now considered.

The minimum bed diameter is a function of the operating velocity, the particle characteristics and the humidity of the drying gas. The hot gas at the inlet rapidly loses heat and gains moisture as it passes through the bed which it eventually leaves at the bed temperature \(T_b\) and with a relative humidity \(R_h\) which is approximately equal to the relative humidity which would be in equilibrium with the dried product at the bed temperature. The operating velocity may be taken as twice the minimum fluidising velocity, obtained from the equations in Section 6.1.3, by laboratory tests, or more conveniently from Figure 16.26.

![Figure 16.26: Superficial operating velocity in fluidised bed dryers\(^{47}\)](image)

For drying media other than air at approximately atmospheric pressure, the velocity obtained from Figure 16.26 should be multiplied by \(0.00975 \rho^{-0.29} \mu^{-0.43}\) where \(\rho\) and \(\mu\) are the density (kg/m\(^3\)) and the viscosity (Ns/m\(^2\)) of the fluidising gas.

From a mass balance across the bed:

\[
(\%100)(P_w/P) = [W + (G/(1 + \mathcal{X})\mathcal{X})]/[W + (G/(1 + \mathcal{X})(0.625 + \mathcal{X})] \quad (16.39)
\]

where:
\(\%\) = exit gas relative humidity (per cent),
\(P_w\) = vapour pressure of water at the exit gas temperature (N/m\(^2\)),
\(P\) = total static pressure of gases leaving the bed (N/m\(^2\)),
\(W\) = evaporative capacity (kg/s),
\(G\) = inlet flow of air (kg/s), and
\(\mathcal{X}\) = humidity of inlet air (kg/kg dry air)
The term $G/(1 + \mathcal{F})$ is in effect the flow of dry air, and 0.625 is approximately the ratio of molecular masses in the case of water and air.

Values of $P_w$ may be obtained from Figure 16.27 and, for indirect heating, $\mathcal{F}$ is the humidity of the ambient air. For direct heating, $\mathcal{F}$ may be obtained from Figure 16.28 which assumes that the air at 293 K, with a humidity of 0.01 kg/kg, is heated by the combustion of methane.

Figure 16.27. Water vapour pressure at the bed exit\(^{(47)}\)

A heat balance (in W) across the bed gives:

$$C_m G (T_m - T_b) = \lambda_b W + C_f F (T_b - T_f)$$

(16.40)

where:

- $C_m$ = mean thermal capacity of the gas between $T_m$ and $T_b$ (J/kg deg K) (1000 J/kg deg K for air),
- $C_f = [(X_f c_x + c_s)/(X_f + 1)]$ average thermal capacity of the wet solid between $T_f$ and $T_b$ (J/kg deg K),
- $X_f$ = moisture content of wet feed (kg/kg dry solid),
- $C_x$ = heat capacity of liquid being evaporated (J/kg deg K),
- $C_s$ = heat capacity of dry solid (J/kg deg K),
- $\lambda_b$ = mean latent heat of liquid at $T_b$ (J/kg),
- $T_m, T_b, T_f$ = temperatures of inlet gas, bed and wet feed respectively (K), and
- $F$ = rate of wet solid feed (kg/s)
Values of $G$ and $T_b$ are obtained from equations 16.39 and 16.40 and the bed diameter is then obtained from:

$$D^2 = \frac{(G + 1.58 \text{ W}) T_b}{278 u_f} \text{ m}^2$$

(16.41)

noting that:

$$\text{Mean molecular weight of inlet air} / \text{molecular weight of water} = \frac{29}{18} \approx 1.58$$

and:

$$\pi \left( \frac{\text{molecular weight or air} \times \text{datum temperature}}{4 \times \text{kilomolecular volume}} \right) = \frac{\pi}{4} \left( \frac{29 \times 273}{22.4} \right) = 278 \text{ kg deg K/m}^3$$

**Example 16.6**

A granular material of density 5000 kg/m$^3$ is to be dried in a fluidised bed dryer using directly heated air at 811 K. The particle size is 0.5 mm and 1.26 kg/s of water is to be removed from 12.6 kg/s of solid feed at 293 K. What diameter of bed should be specified?

$x_b = 2325 \text{ kJ/kg}, \text{ ambient air is at } 293 \text{ K}, \rho = 0.01 \text{ kg/kg and } C_f = 1.67 \text{ kJ/kg deg K}$

**Solution**

From Figure 16.28; $\mathcal{F} = 0.036 \text{ kg/kg at 811 K}$
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and the right-hand side of equation 16.39 is:

\[ = \frac{(1.26 + 0.0347 G)}{(1.26 + 0.638 G)} \]

Taking \( \% \) as 90 per cent and \( P \) as 101.3 kN/m², then, for assumed values of \( T_b \) of 321, 333 and 344 K:

\[ P = 13, 20 \text{ and } 32 \text{ kN/m}^2, \text{ respectively} \]

and:

\[ G = 27.8, 12.9 \text{ and } 6.02 \text{ kg/s, respectively} \]

Using equation 16.40, for \( T_b = 321, 333 \) and 344 K,

\[ G = 7.16, 7.8 \text{ and } 7.54 \text{ kg/s respectively} \]

Plotting \( G \) against \( T_b \) for each equation on the same axis, then:

\[ G = 8.3 \text{ kg/s when } T_b = 340 \text{ K} \]

From Figure 16.27, \( u_f = 0.61 \text{ m/s} \). Hence, in equation 16.41:

\[ D^2 = \frac{340(8.3 + (1.58 \times 1.26))}{(278 \times 0.61)} \]

and:

\[ D = 4.60 \text{ m} \]

This is a very large diameter bed and it may be worthwhile increasing the fluidising velocity provided any increased elutriation is acceptable. If \( u_f \) is increased to 1.52 m, the diameter then becomes 2.88 m with a subsequent reduction in capital costs.

One development in the field of fluidised bed drying is what is known as sub-fluidised conditioning. A fluidised-bed dryer normally works with a maximum residence time of some 1200 s. If this is increased, then the spread of product residence times increases excessively due to the fact that axial and longitudinal mixing occur in the bed during the fluidisation process and, the more vigorous the fluidisation, the greater the spread of residence times. If fluidisation continues for too long, breakage and other product damage is likely to occur. A solution to this problem, developed by Ventilex, is sub-fluidised conditioning where forward movement of the product is created by a shaking mechanism in which the product is kept at the threshold of fluidisation. As a result, there is little or no axial or longitudinal mixing and the spread of residence time is kept to a minimum. In essence, this process provides a situation which approaches that of ideal plug flow, and many products can be conditioned by this combination of fluidised and sub-fluidised techniques. In addition, even less air is required for drying and thus small auxiliary equipment is required and energy requirements are less. Nominal bed thicknesses are 150–250 mm and long residence times, of up to 7 ks, are possible with a minimal time spread. In the drying of drugs, nuts, meats and rice, no degradation takes place and, in this respect, the process is comparable with traditional installations such as conveyor dryers, with the added benefits of fluidisation. The system is also applicable to the drying of minerals such as china clay and sand, and also to animal wastes and sludges. A typical installation together with a flow diagram is shown in Figure 16.29.
Figure 16.29: Unit for sub-fluidised drying
16.5.9. Turbo-shelf dryers

The handling of sticky materials can present difficulties, and one type of dryer which is useful for this type of material is the turbo-dryer. As shown in Figure 16.30, wet solid is fed in a thin layer to the top member of a series of annular shelves each made of a number of segmental plates with slots between them. These shelves rotate and, by means of suitably placed arms, the material is pushed through a slot on to a shelf below. After repeated movements, the solid leaves at the bottom of the dryer. The shelves are heated by a row of steam pipes, and in the centre there are three or more fans which suck the hot air over the material and remove it at the top.

The accelerated drying induced by the raking of the material gives evaporative capacities of 0.0002–0.0014 kg/s m\(^2\) shelf area which are comparable with those obtained by through circulation on perforated belts. Shelf areas are 0.7–200 m\(^2\) in a single unit and the dryer may easily be converted to closed-circuit operation, either to prevent emission of fumes or in order to recover valuable solvents. Typical air velocities are 0.6–2.5 m/s, and the lower trays are often used to cool the dry solids. A turbo-dryer combines cross-circulation drying, as in a tray dryer, with drying by showering the particles through the hot air as they tumble from one tray to another.
**16.5.10. Disc dryers**

A disc dryer provides a further way in which pasty and sticky materials may be handled, and it can also cope satisfactorily with materials which tend to form a hard crust or pass through a rheologically-difficult phase during the drying operation.

As shown in Figure 16.31, a single-agitator disc contact dryer, consists of a heated cylindrical housing (1) assembled from unit sections, and a heated hollow agitating rotor (2) which has a simultaneous rotating and oscillating movement produced by means of a rotating drive (3) and a reciprocating drive (4). The drive (3 and 4) and stuffing box (8) are located at the dry product end, and the stuffing box is protected by a reverse acting flight (12). The wet product is introduced, drawn in by the screw flight (13) and continuously conveyed through the dryer. The vapour passes through the vapour filter (15) to the condenser. This vapour filter system has a back-scavenging action, and is specially designed for removing dust from vapour in vacuum drying plants. The hollow agitator (2) is fitted, over its whole length, with heated flights, which are arranged equidistantly in pairs. Between every two axially neighbouring agitator flights, there are, projecting inwards from the housing fixed wiping pegs (10) or annular weir/kneading elements (11) which extend inward to the agitator core.
The self-cleaning of the heating surfaces is achieved by the combined rotating-reciprocating movement of the agitator. The stationary elements (10 and 11) clean the faces of the agitator flights, at the end of each oscillating movement, during one rotation of the agitator. By the forward and backward movement, the edges of the agitator flights (9) clean the inner surface of the housing and the fixed elements projecting inwards to the core of the agitator clean the agitator (2). In all, about 95 per cent of the heating surface is cleaned.

The rotating and reciprocating motions need not be synchronised, because the individual agitator flights (9) oscillate only between two adjacent fixed elements (10 and 11). This is so that the speed of rotation, the frequency of reciprocation and the forward and backward speed of reciprocation can be adjusted independently of one another over a wide range of settings. The housing sections (1) which are supported by frames (5) on rails, can be drawn forward when cleaning is necessary. The transport of pasty products through the dryer may be achieved by the differential forward and backward oscillatory motion, combined with the action of the bevelled edges of the agitator flights.

Contivac dryers, for example, have heating surfaces of 4–60 \( \text{m}^2 \) and volumes of 0.1–3 \( \text{m}^3 \). They may be operated under vacuum or up to 400 kN/m\(^2\) with heating fluids at 330 K–670 K. The evaporative capacity is 0.03–0.55 kg/s and the agitator speed ranges from 0.1 Hz (6 rpm) for rheologically difficult materials, to 1 Hz (60 rpm) for easier applications.

Operating on a similar principle is the Buss paddle dryer which effects batch drying of liquids, pasty and sandy materials. The paddle dryer, which is the main element of the drying unit. It consists of a horizontal, cylindrical housing which contains a paddle agitator in the form of a hollow shaft carrying agitator arms. The jacket, the hollow shaft and the agitator arms are steam-heated. The paddle agitator is driven by an electric motor and gear unit. A built-on torque bracket with a microswitch protects the paddle agitator against overloading. The product to be dried is passed through the charging nozzle into the paddle dryer, where it is distributed uniformly by the rotating paddle agitator. The drying proceeds under vacuum while intensive intermixing by the agitator causes continual renewal of the product particles in contact with the heated surfaces. This guarantees efficient heat transfer and uniform product quality. Vapours are purged of dust in passing through a vapour filter and are then fed to a condenser. Non-condensable gases are drawn off by the vacuum system. The vapour filter is equipped with a removable filter insert and, to prevent excessive pressure drop across the fabric from thick, and possibly moist dust build-up on the filter sacks, these are provided with a reverse jet arrangement, that is the individual filter sacks are cleaned in turn automatically during operation by short though powerful countercurrent blasts of steam blown through the filter sack. This serves to blow and shake off the dust layer and keep the filter sack dry. After the drying process is complete, the dried material may be cooled by applying cooling water to the dryer jacket and paddle agitator. The dryer is emptied by the arms of the paddle agitator, which are designed to move the material in the vessel towards the outlet when rotated backwards. The discharge outlet is specially constructed to prevent the formation of a plug of material. At 373 K, an evaporation rate of over 4 g/s \( \text{m}^2 \) can be achieved for a steam consumption of 1.5 kg/kg of evaporated water.
16.5.11. Centrifuge dryers

Where a product of a high quality and purity is required, this can, in many cases, be achieved only by washing the solids during a centrifugation process in order to flush out the suspension mother liquor or to dissolve salt crystals. The washed crystals are then usually dried in a separate physical unit operation involving different equipment and problems can be encountered due to the exposure to the atmosphere of the wet cake as it is transferred between the centrifuge and the dryer. These problems may be largely eliminated by using a combined centrifuge–dryer which it is claimed, has the following advantages:

(a) It is hermetically sealed from the environment and is therefore easy to inert.
(b) There is no human contact with the product at any stage of the operation and no possibility of foreign materials being introduced.
(c) The handling of the product is simple and gentle.
(d) Drying times can be reduced by using a jet-pulsed bed drying technique.
(e) The unit can be emptied completely with no residual cake in the basket or on the wall of the drum.

All these advantages are inherent in the FIMA, TZT centrifuge–dryer which is operated as follows.

The feed suspension is introduced into the centrifuge through a hollow drive shaft, and the liquid from the suspension is separated from the crystals and discharged from a sealed filter basket by passing it though a metal filter which retains the solids inside a sealed chamber. Any contaminants are then washed away from the filling pipe, the centrifuge basket and the separated solids by introducing wash liquid through the hollow drive shaft. This operation may be repeated as required. One advantage of the TZT system is that, with no internal pipes or structures that might accumulate unwashed solids, all the separated solids make contact with the wash water. The ring of separated solids is then removed from the wall of the centrifuge by introducing powerful gas blasts beneath the filter medium and the wet en-masse solids accumulate at the bottom of the filter drum. The drying process is achieved by rotating the drum at slow speeds and injecting heated drying gas as process into the closed centrifuge chamber. The moisture-laden gas leaves passing through the filter medium so that the solids are retained within the basket. Such a process is extremely gentle to the solids and dries at low product temperatures even when using higher gas temperatures as a result of the cooling effect of the surface evaporation of the moisture. Dry powder is discharged from the unit by opening the sealing centrifuge housing and fluidising the solids which then enter an integrated powder conveying system. In this system the gas used for drying may be recycled in a closed gas loop, or it may be discharged to atmosphere in an open system.

Units of this type have a capacity of 20–400 kg in terms of filling mass, filter areas of 0.37–2.4 m² and drum diameters of 400–1300 mm and lengths of 300–600 mm.

Although this is, in essence a batch operation, close reproducibility between batches overcomes many of the problems associated with batch identification which can be a problem with more conventional drying equipment.
16.6. SPECIALISED DRYING METHODS

16.6.1. Solvent drying

Two processes are used here:

*Superheated-solvent drying* in which a material containing non-aqueous moisture is dried by contact with superheated vapours of its own associated liquid, and,

*Solvent dehydration* in which water-wet substances are exposed to an atmosphere of a saturated organic solvent vapour.

The first of these has advantages where a material containing a flammable liquid such as butanol is involved. Drying is effected with a gas with an air–moisture ratio of 90 kg/kg moisture in order to ensure that the composition is well below the lower explosive limit. The heat requirement is as great as those for superheated solvent drying at the same gas outlet temperature of 400 K\(^{\text{48}}\). Superheated solvent drying in a fluidised bed has been used for the drying of polypropylene pellets to eliminate the need for water-washing and for fractionating. A flowsheet of the installation is shown\(^{\text{49}}\) in Figure 16.32.

![Flowsheet for drying polypropylene by fluidising with a hot stream of solvent vapour](Image)

The most important applications of solvent dehydration lie within the field of kiln drying for seasoning timber where substantial reductions in drying times have been achieved\(^{\text{50}}\).

16.6.2. Superheated steam drying

The replacement of air by superheated steam to take up evaporating moisture is attractive in that it provides a high temperature heat source which also gives rise to a much higher
driving force for mass transfer since it does not become saturated at relatively low moisture contents as is the case with air. In the drying of foodstuffs, a further advantage is the fact that the steam is completely clean and there is much less oxidation damage. In the seasoning of timber, for example, drying times can be reduced quite significantly. Although the principles involved have been understood for some considerable time, as Basel and Gray \(^{(49)}\) points out, applications have been limited to due to corrosion problems and the lack of suitable equipment. A flowsheet of a batch dryer using superheated steam is shown in Figure 16.33. The dryer is initially filled with air circulated by a blower, together with evaporated moisture. Any excess moisture is vented to atmosphere so that the air is gradually replaced by steam. For an evaporation rate of 10 kg/m\(^3\) volume of the dryer chamber, the composition of the gas phase would reach about 90 per cent steam in about 600 s.

![Figure 16.33: Superheated-steam dryer](image)

As Lukoy \(^{(51)}\) reports, superheated steam drying may also be used to dry wet material by heating it in a sealed autoclave, and periodically releasing the steam which is generated. This pressure release causes flash evaporation of moisture throughout the whole extent of the material, thus avoiding drying stresses and severe moisture gradients. Such an operation has been applied to the drying of thermal insulating materials by Bosn et al. \(^{(52)}\) who have investigated the drying of paper. In this work, impinging jets of superheated steam were used at 293–740 K during the constant drying period, with jet Reynolds numbers, 100–12000. Above 450 K, steam drying was found to be much faster than air drying for the same mass velocity of gas. The specific blower power was found to be much lower than for air drying at temperatures of industrial importance. It was concluded that steam-impingement drying can lead to significant reductions in both capital investment and energy costs.

In tests on the drying of sand, Wenzel and White \(^{(53)}\) found that the use of steam rather than air did not alter the general characteristics of the drying process, and that the drying rate during the constant rate period was determined by the heat transfer rate. In these tests, the heat transferred by radiation from the steam and surrounding surfaces was 7.5–31
per cent of the total heat transferred, and coefficients of convective heat transfer were 13–100 W/m²K. It was concluded that higher drying rates and greater thermal efficiencies are possible when drying with superheated steam as opposed to air, and that the choice of steam drying must depend on a balance between the savings in operating costs and the higher capital investment attributable to the higher temperatures and pressures.

Schwartz and Brocker, who has made a theoretical study of the evaporation of water into mixtures of superheated steam and air, has calculated the inversion temperature above which the evaporation rate into pure superheated steam is higher than that into dry air under otherwise similar conditions. The data obtained are given in Figure 16.34 which shows quite clearly the enhanced evaporation rates at gas temperatures above about 475 K. This inversion temperature is given by the point of intersection of the curves for evaporation rate with dry air and superheated steam. The Nusselt and Sherwood equations for heat and mass transfer coefficients for the relevant geometrical configuration, given in Volume 1, were used to calculate the evaporation rates, and these were found to be in excellent agreement with experimental data in the literature. Taylor and Krishna points out that this approach may be used for a wide range of applications, and Vidaurre and Martinez show that the model may be extended to include specialised applications, such as the evaporation of multicomponent liquids.

16.6.3. Freeze drying

In this process, the material is first frozen and then dried by sublimation in a very high vacuum, 10–40 N/m², at a temperature of 240–260 K. During the sublimation of the ice, a dry surface layer is left, though this is not free to move because it has been frozen,
and a honeycomb structure is formed. With normal vacuum drying of biological solutions containing dissolved salts, a high local salt concentration is formed at the skin, although with freeze drying this does not occur because of the freezing of the solid. Thus, freeze drying is useful not only as a method of working at low temperatures, but also as a method of avoiding surface hardening.

As described by Chambers, heat has to be supplied to the material to provide the heat of sublimation. The rate of supply of heat should be such that the highest possible water vapour pressure is obtained at the ice surface without danger of melting the material. During this stage, well over 95 per cent of the water present should be removed, and in order to complete the drying, the material should be allowed to rise in temperature, to say ambient temperature. The great attraction of this technique is that it does not harm heat-sensitive materials, and it is suitable for the drying of penicillin and other biological materials. Initially, high costs restricted the application of the process though economic advances have reduced these considerably and foodstuffs are now freeze-dried on a large scale. Maguire has estimated a total cost of £0.06/kg of water evaporated, in a plant handling 0.5–0.75 kg/s (2–3 tonne/h) of meat.

A typical layout of a freeze-drying installation is shown in Figure 16.35. Heat is supplied either by conduction, or by radiation from hot platens which interleave with trays containing the product, and the sublimed moisture condenses on to a refrigerated coil at the far end of the drying chamber. The use of dielectric-heating has been investigated, though uneven loading of the trays can lead to scorching, and ionisation of the residual gases in the dryer results in browning of the food.

![Figure 16.35. Flowsheet of freeze-drying plant](image)

Continuous freeze-drying equipment has been developed, and chopped meat and vegetables may be dried in a rotating steam-jacketed tube enclosed in the vacuum chamber. A model of the freeze-drying process has been presented by Dyer and Sunderland, and further details are given in Section 15.8 on freeze crystallisation.

### 16.6.4. Flash drying

Conventional drying consists of mixing and heating the solids to achieve even drying, and simultaneously transporting the vapours away from the surface of the solids so as to maintain a high rate of mass transfer. Although these two stages can be achieved
simultaneously by contacting the solids with a hot gas stream, heat-sensitive materials, such as foods and pharmaceuticals, may suffer thermal degradation, and other solids may lose some of their water of hydration when subjected to high temperatures. This problem may often be overcome by using a flash drying technique to ensure that the solids are in contact with hot gas in a highly turbulent environment for only a very short time, perhaps only for a few seconds.

A typical flash-drying process consists of a modified pneumatic conveyor in which the wet solids are introduced into a pipe through which they are transported in a high velocity hot gas stream. In such a process, sticky sludges must first be mixed with dry solids so that the resultant mixture breaks up into smaller particles in the gas stream. Such a process can be used for a wide range of applications. In the drying of the gypsum by-product of flue gas desulphurisation, for example, operating at a minimum solids temperature allows the gypsum to retain its water of hydration, and in the conversion of sewerage sludge into dry fertiliser flash drying prevents the fertilizer from oxidising. Because waste flue gases can often be used as the drying medium, little or no external heat energy is required and, when the dried sludge is burned as a fuel, the heat generated provides much of the energy required for drying.

Even when back-mixed with dry solids, some materials are too sticky or paste-like to feed to a flash dryer, though these can often be handled in a spin dryer consisting of a high shear agitator inside a specially designed drying chamber. In this way, the wet paste may be subjected simultaneously to high shear to break up the solid mass, and to gas at a high velocity to dry the smaller particles. As the particles become small enough and dry enough, they are carried out of the chamber with the spent air and, as with conventional flash drying, the contact time is but a few seconds.

Neither flash nor spin-drying produces highly uniform particles and, even though the bulk of the particles remain at or near the wet bulb temperature, small particles are often over-dried and overheated. Particles of uniform size can be formed by spray drying, however, although it is then necessary that the feed should be in the form of a pumpable liquid which can be atomised so that the particles are evenly exposed to the hot air.

**16.6.5. Partial-recycle dryers**

The majority of drying operations depend on direct heating using a high flowrate of hot air and/or combustion products which is passed once through or over the wet material and then exhausted to the atmosphere. A variation of this arrangement is the closed-loop system where the entire air or gas stream is confined and recycled after condensing out the vapour. Such a system is justified only where the vapourised liquid or gas has to be recovered for economic or environmental reasons. In addition to the once-through and the closed-loop systems, there is a third class of dryers which incorporates the partial-recycle mode of operation. In this system, a substantial proportion, typically 40–60 per cent, of the outlet air is returned to the dryer in order to minimise the heating requirements and the amount of effluent treatment required. Conveyor, flash-pneumatic conveying-, fluidised-bed, rotary, spray and tray types of direct dryers can all be designed for closed-loop or partial-recycle operation and Cook has claimed that recycling part of the exhaust stream has the following advantages:
(a) smaller collectors are needed for removing dried product from the exhaust gases,
(b) less heat is lost in the exhausts which leave, typically, at 340–420 K,
(c) since the total volume of exhaust gases is reduced, the cost of preventing undesirable
gases or particles from entering the atmosphere is reduced;
(d) when using direct burners to heat the air, the level of oxygen content is minimised,
thus enabling sensitive products to be handled and reducing potential fire and/or
explosion hazards.

In conveyor and tray types of dryer, air is often recirculated inside the drying vessel in
an attempt to save energy or to maintain a relatively high moisture content in the drying
air. In other direct dryers such as flash, fluidised-bed, rotary and spray units, any recycle
of exhaust air must be returned to the dryer using external ducting, the cost of which is
offset by the net savings from the lower volumes of exhaust streams which have to be
handled.

As shown in Figure 16.36, in a partial-recycle drying process, the total airflow leaving
the dryer is first passed to a particulate or dust collector in order to remove entrained
product or fines, and is then split into two streams—a bleed stream which is vented to
atmosphere and a recycle stream which is returned to the system. The bleed stream is
large enough to carry all of the moisture that has entered the system, and the recycle
portion is returned to the inlet of either the heater or the drying vessel. In the condensing
operation, shown in Figure 16.36, the exhaust stream from the dust collector is passed
through a condenser, usually a wet scrubber, to remove most of the entrained moisture
before it is returned to the inlet of the heater or the dryer. The bleed stream may also
be passed through the condenser for gas-cleaning prior to discharge to atmosphere, thus
removing moisture and thereby reducing the load on the condenser. Although condensers
minimise the amount of bleed to the atmosphere and the moisture content of the drying
air, they also cool the recycle stream and this increases costs due to the need to reheat
the stream. The total volume of the bleed must be balanced by the heater output and any
leakage into the system, noting that moisture can enter the dryer from three sources—the feed, combustion air and the supply air—and can leave in the bleed in the product and as condensate, if any. It may be noted that, in Figure 16.36, the scrubber—condenser also serves as a collector for particulates. If an oxygen content of less than 21 per cent by volume or a temperature above 270–640 K is required, then a direct-fired heater may be used. If non-aqueous solvents are to be evaporated in the dryer, the effluent may require incineration or other treatment prior to discharge to the atmosphere, and dual blowers are often needed to maintain the required pressures throughout the system.

Cook(63) has presented data on the performance of a partial-recycle units based on a direct-fired burner using natural gas. These are compared with the once-through operation in Table 16.8:

<table>
<thead>
<tr>
<th>Example</th>
<th>Recycle mode</th>
<th>Burner outlet temperature (K)</th>
<th>Recycle (per cent)</th>
<th>Bleed (per cent)</th>
<th>Moisture (kg/kg air)</th>
<th>Heat Load (MW)</th>
<th>Air flow (m³/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>condensing</td>
<td>1090</td>
<td>46.0</td>
<td>54.0</td>
<td>0.206</td>
<td>2.74</td>
<td>8.14</td>
</tr>
<tr>
<td>2</td>
<td>condensing</td>
<td>810</td>
<td>19.7</td>
<td>80.3</td>
<td>0.173</td>
<td>2.83</td>
<td>8.22</td>
</tr>
<tr>
<td>3</td>
<td>condensing</td>
<td>1370</td>
<td>58.9</td>
<td>41.1</td>
<td>0.222</td>
<td>2.68</td>
<td>8.11</td>
</tr>
<tr>
<td>4</td>
<td>bleed only</td>
<td>1160</td>
<td>46.0</td>
<td>54.0</td>
<td>0.325</td>
<td>2.52</td>
<td>7.92</td>
</tr>
<tr>
<td>5</td>
<td>bleed only</td>
<td>858</td>
<td>25.4</td>
<td>74.6</td>
<td>0.206</td>
<td>2.68</td>
<td>8.14</td>
</tr>
<tr>
<td>6</td>
<td>once-through</td>
<td>700</td>
<td>0.0</td>
<td>100.0</td>
<td>0.148</td>
<td>2.90</td>
<td>8.28</td>
</tr>
</tbody>
</table>

(Dryer inlet = 700 K, dryer outlet = 365 K)

In these systems, the total collection efficiencies of the dry product are 85 per cent for the drying vessel, 90 per cent for the cyclone collector and 98 per cent for the scrubber-condenser. The net efficiency of the system may be as high as 99.97 per cent if the scrubber effluent is considered as product. All the runs are based on 1.25 kg/s product and 0.75 kg/s evaporation at an elevation of 300 m above sea level. The total air flow is measured at the outlet before the stream is split into the recycle and bleed portions and, for such flows, the design of suitable fans is outlined by Jorgensen(64). The calculations outlined here may be confirmed by the use of psychometric charts, and this procedure has been considered in some detail by Cook and Dumbourn(65).

In recent years, air-recycling in drying systems has become more widely adopted as a means of reducing total dryer discharges to the atmosphere. It would seem that this trend is likely to continue because of environmental concerns and the ever-increasing costs of treating emissions.

16.7. THE DRYING OF GASES

The drying of gases is carried out on a very large scale. The most important applications of the process are:

(a) in order to reduce the tendency of the gas to cause corrosion,
(b) preparation of dry gas for use in a chemical reaction, and
(c) reduction of the humidity of air in air-conditioning plants.
The problems involved in the drying or dehumidification of gases are referred to in Volume 1, Chapter 13, and the most important methods available are now summarised.

**Cooling.** A gas stream may be dehumidified by bringing it into contact with a cold liquid or a cold solid surface. If the temperature of the surface is lower than the dew point of the gas, condensation will take place, and the temperature of the surface will tend to rise by virtue of the liberation of latent heat. It is therefore necessary to remove heat constantly from the surface. Because a far larger interfacial surface can be produced with a liquid, it is usual to spray a liquid into the gas and then to cool it again before it is recycled. In many cases, countercurrent flow of the gas and liquid is obtained by introducing the liquid at the top of a column and allowing the gas to pass upwards.

**Compression.** The humidity of a gas may be reduced by compressing it, cooling it down to near its original temperature, and then draining off the water which has condensed. During compression, the partial pressure of the water vapour increases and condensation occurs as soon as the saturation value is exceeded.

**Liquid absorbents.** If the partial pressure of the water in the gas is greater than the equilibrium partial pressure at the surface of a liquid, condensation will take place as a result of contact between the gas and liquid. Thus, water vapour is frequently removed from a gas by bringing it into contact with concentrated sulphuric acid, phosphoric acid, or glycerol. Concentrated solutions of salts, such as calcium chloride, are also effective. The process may be carried out either in a packed column or in a spray chamber. Regeneration of the liquid is an essential part of the process, and this is usually effected by evaporation.

**Solid adsorbents and absorbents.** The use of silica gel or solid calcium chloride to remove water vapour from gases is a common operation in the laboratory. Moderately large units can be made, although the volume of packed space required is generally large because of the comparatively small transfer surface per unit volume. If the particle size is too small, the pressure drop through the material becomes excessive. The solid desiccants are regenerated by heating.

Gas is frequently dried by using a calcium chloride liquor containing about 0.56 kg calcium chloride/kg solution. The extent of recirculation of the liquor from the base of the tower is governed by heating effects, since the condensation of the water vapour gives rise to considerable heating. It is necessary to install heat exchangers to cool the liquor leaving the base of the tower.

In the contact plant for the manufacture of sulphuric acid, sulphuric acid is itself used for drying the air for the oxidation of the sulphur. When drying hydrocarbons such as benzene, it is sometimes convenient to pass the material through a bed of solid caustic soda, although, if the quantity is appreciable, this method is expensive.

The great advantage of materials such as silica gel and activated alumina is that they enable the gas to be almost completely dried. Thus, with silica gel, air may be dried down to a dew point of 203 K. Small silica gel containers are frequently used to prevent moisture condensation in the low pressure lines of pneumatic control installations.

### 16.8. Further Reading


16.9. REFERENCES

## 16.10. NOMENCLATURE

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
<th>Units in SI System</th>
<th>Dimensions</th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>Area for heat transfer or evaporation</td>
<td>m²</td>
<td>in M, N, J, T, K, L, θ</td>
</tr>
<tr>
<td>Aₚ</td>
<td>Area of nozzle or jet normal to direction of flow</td>
<td>m²</td>
<td>L²</td>
</tr>
<tr>
<td>a</td>
<td>Surface area per unit volume</td>
<td>m²/m³</td>
<td>L⁻¹</td>
</tr>
<tr>
<td>B</td>
<td>Width of surface</td>
<td>m</td>
<td>L</td>
</tr>
<tr>
<td>C</td>
<td>Coefficient</td>
<td>kg²/m⁴ s²</td>
<td>M⁴ L⁻⁴ T⁻²</td>
</tr>
<tr>
<td>Cᵦ</td>
<td>Coefficient of discharge</td>
<td>m³/s</td>
<td>L³ T⁻¹</td>
</tr>
<tr>
<td>Cₚ</td>
<td>Heat capacity of wet material</td>
<td>J/kg K</td>
<td>L² T⁻¹ θ⁻¹</td>
</tr>
<tr>
<td>Cₘ</td>
<td>Mean heat capacity of gas</td>
<td>J/kg K</td>
<td>L² T⁻¹ θ⁻¹</td>
</tr>
<tr>
<td>Cₛ</td>
<td>Heat capacity of dry solid</td>
<td>J/kg K</td>
<td>L² T⁻¹ θ⁻¹</td>
</tr>
<tr>
<td>Cₑ</td>
<td>Heat capacity of liquid evaporated</td>
<td>J/kg K</td>
<td>L² T⁻¹ θ⁻¹</td>
</tr>
<tr>
<td>D</td>
<td>Diameter of drum or disc</td>
<td>m</td>
<td>L</td>
</tr>
<tr>
<td>Dₑ</td>
<td>Diameter of evaporating drop at time t</td>
<td>m</td>
<td>L</td>
</tr>
<tr>
<td>dₒ</td>
<td>Initial diameter of drop</td>
<td>m</td>
<td>L</td>
</tr>
<tr>
<td>E</td>
<td>Wet solid feed rate</td>
<td>kg/s</td>
<td>M T⁻¹</td>
</tr>
<tr>
<td>E₀</td>
<td>Humidity</td>
<td>kg/kg</td>
<td>—</td>
</tr>
<tr>
<td>Eₛ</td>
<td>Humidity of saturated air at surface temperature</td>
<td>kg/kg</td>
<td>—</td>
</tr>
<tr>
<td>Eₘ</td>
<td>Humidity of saturated air at temperature θₘ</td>
<td>kg/kg</td>
<td>—</td>
</tr>
<tr>
<td>Eₖ</td>
<td>Heat transfer coefficient</td>
<td>W/m² K</td>
<td>M⁴ T⁻¹ θ⁻¹</td>
</tr>
<tr>
<td>Eₜ</td>
<td>Mass transfer coefficient</td>
<td>m/s</td>
<td>L T⁻¹</td>
</tr>
<tr>
<td>Eₘₗ</td>
<td>Mass transfer coefficient</td>
<td>m/s</td>
<td>L T⁻¹</td>
</tr>
<tr>
<td>Eₖₗ</td>
<td>Mass transfer coefficient</td>
<td>m/s</td>
<td>L T⁻¹</td>
</tr>
<tr>
<td>Eₜₗ</td>
<td>Mass transfer coefficient</td>
<td>m/s</td>
<td>L T⁻¹</td>
</tr>
<tr>
<td>j</td>
<td>Free moisture content</td>
<td>kg</td>
<td>M</td>
</tr>
<tr>
<td>jₑ</td>
<td>Free moisture content at critical condition</td>
<td>kg</td>
<td>M</td>
</tr>
<tr>
<td>jᵦ</td>
<td>Initial free moisture content</td>
<td>kg</td>
<td>M</td>
</tr>
<tr>
<td>F</td>
<td>Volumetric rate of feed per unit cross-section</td>
<td>m³/s</td>
<td>M³ T⁻¹</td>
</tr>
<tr>
<td>F₀</td>
<td>Friction head over a distance z₁ from surface</td>
<td>m</td>
<td>L</td>
</tr>
<tr>
<td>Fₛ</td>
<td>Suction potential immediately below meniscus</td>
<td>m</td>
<td>L</td>
</tr>
<tr>
<td>Fₘ</td>
<td>Theoretical suction potential of pore or waist</td>
<td>m</td>
<td>L</td>
</tr>
<tr>
<td>Fₚ</td>
<td>Suction potential at distance z₂ below meniscus</td>
<td>m</td>
<td>L</td>
</tr>
<tr>
<td>jₒ</td>
<td>Volumetric gas/liquid ratio</td>
<td>—</td>
<td>—</td>
</tr>
<tr>
<td>K</td>
<td>Coefficient</td>
<td>m³/kg</td>
<td>M⁻¹ L² T</td>
</tr>
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### DRYING


<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
<th>Units in SI System</th>
<th>Dimensions</th>
</tr>
</thead>
<tbody>
<tr>
<td>$K'$</td>
<td>Coefficient</td>
<td>$1/m^3$</td>
<td>$L^{-1} T^{1.8} K^{0.8}$</td>
</tr>
<tr>
<td>$K'$</td>
<td>Transfer coefficient $(hDpA)$</td>
<td>$kg/m^2 s$</td>
<td>$ML^{2} T^{-1}$</td>
</tr>
<tr>
<td>$k_D$</td>
<td>Mass transfer coefficient</td>
<td>$m/s$</td>
<td>$L^{-1} T$</td>
</tr>
<tr>
<td>$k_o$</td>
<td>Thermal conductivity of gas film at interface</td>
<td>$W/mK$</td>
<td>$MLT^{-3} K^{-1}$</td>
</tr>
<tr>
<td>$l$</td>
<td>Length of surface</td>
<td>$m$</td>
<td>$L$</td>
</tr>
<tr>
<td>$t$</td>
<td>Half thickness of slab</td>
<td>$m$</td>
<td>$L$</td>
</tr>
<tr>
<td>$M_o$</td>
<td>Molecular weight of air</td>
<td>$kg/kmol$</td>
<td>$MN^{-1}$</td>
</tr>
<tr>
<td>$M_w$</td>
<td>Molecular weight of water</td>
<td>$kg/kmol$</td>
<td>$MN^{-1}$</td>
</tr>
<tr>
<td>$m$</td>
<td>Ratio of rate of drying per unit area to moisture content</td>
<td>$m^2 s^{-3}$</td>
<td>$L^{-2} T^{1.8}$</td>
</tr>
<tr>
<td>$N$</td>
<td>Revolutions per unit time</td>
<td>$Hz$</td>
<td>$T^{-1}$</td>
</tr>
<tr>
<td>$n$</td>
<td>Index</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$n_f$</td>
<td>Number of flights</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$P$</td>
<td>Total pressure</td>
<td>$N/m^2$</td>
<td>$ML^{-1} T^{-2}$</td>
</tr>
<tr>
<td>$P_v$</td>
<td>Vapour pressure of water at surface of material</td>
<td>$N/m^2$</td>
<td>$ML^{-1} T^{-2}$</td>
</tr>
<tr>
<td>$P_e$</td>
<td>Partial pressure of water vapour</td>
<td>$N/m^2$</td>
<td>$ML^{-1} T^{-2}$</td>
</tr>
<tr>
<td>$P_{w,0}$</td>
<td>Partial pressure at surface of material at wet bulb temperature</td>
<td>$N/m^2$</td>
<td>$ML^{-1} T^{-2}$</td>
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<tr>
<td>$\Delta P$</td>
<td>Pressure drop across nozzle</td>
<td>$N/m^2$</td>
<td>$ML^{-1} T^{-2}$</td>
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<tr>
<td>$Q$</td>
<td>Rate of heat transfer</td>
<td>$W$</td>
<td>$ML^2 T^{-1}$</td>
</tr>
<tr>
<td>$q_f$</td>
<td>Rate of drying per unit area for constant rate period</td>
<td>$kg/m^2 s$</td>
<td>$ML^{-2} T^{-1}$</td>
</tr>
<tr>
<td>$R$</td>
<td>Universal gas constant</td>
<td>$8314 J/kmol K$</td>
<td>$MN^{-1} L^{2} T^{-2} K^{-1}$</td>
</tr>
<tr>
<td>$\rho_b$</td>
<td>Relative humidity (per cent)</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$r$</td>
<td>Radius of sphere</td>
<td>$m$</td>
<td>$L$</td>
</tr>
<tr>
<td>$r_c$</td>
<td>Radius of capillary</td>
<td>$m$</td>
<td>$L$</td>
</tr>
<tr>
<td>$\rho$</td>
<td>Slope of drum</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$T$</td>
<td>Absolute temperature</td>
<td>$K$</td>
<td>$\theta$</td>
</tr>
<tr>
<td>$T_b$</td>
<td>Bed temperature</td>
<td>$K$</td>
<td>$\theta$</td>
</tr>
<tr>
<td>$T_f$</td>
<td>Temperature of wet solids feed</td>
<td>$K$</td>
<td>$\theta$</td>
</tr>
<tr>
<td>$T_m$</td>
<td>Mean temperature of inlet gas</td>
<td>$K$</td>
<td>$\theta$</td>
</tr>
<tr>
<td>$\Delta T$</td>
<td>Temperature difference</td>
<td>$K$</td>
<td>$\theta$</td>
</tr>
<tr>
<td>$t$</td>
<td>Time</td>
<td>$s$</td>
<td>$T$</td>
</tr>
<tr>
<td>$t_c$</td>
<td>Time of constant rate period of drying</td>
<td>$s$</td>
<td>$T$</td>
</tr>
<tr>
<td>$t_f$</td>
<td>Time of drying in falling rate period</td>
<td>$s$</td>
<td>$T$</td>
</tr>
<tr>
<td>$U$</td>
<td>Overall heat transfer coefficient</td>
<td>$W/m^2 K$</td>
<td>$MT^{-3} \theta^{-1}$</td>
</tr>
<tr>
<td>$u$</td>
<td>Gas velocity</td>
<td>$m/s$</td>
<td>$L T^{-1}$</td>
</tr>
<tr>
<td>$u_f$</td>
<td>Fluidising velocity</td>
<td>$m/s$</td>
<td>$L T^{-1}$</td>
</tr>
<tr>
<td>$u^L$</td>
<td>Liquid velocity in jet or spray</td>
<td>$m/s$</td>
<td>$L T^{-1}$</td>
</tr>
<tr>
<td>$u_r$</td>
<td>Velocity of gas relative to liquid</td>
<td>$m/s$</td>
<td>$L T^{-1}$</td>
</tr>
<tr>
<td>$V$</td>
<td>Volume</td>
<td>$m^3$</td>
<td>$L^3$</td>
</tr>
<tr>
<td>$W$</td>
<td>Mass rate of evaporation</td>
<td>$kg/s$</td>
<td>$MT^{-3}$</td>
</tr>
<tr>
<td>$w_o$</td>
<td>Total-moisture</td>
<td>$kg$</td>
<td>$M$</td>
</tr>
<tr>
<td>$w_{c1}$</td>
<td>Critical moisture content</td>
<td>$kg$</td>
<td>$M$</td>
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<tr>
<td>$w_e$</td>
<td>Equilibrium moisture content</td>
<td>$kg$</td>
<td>$M$</td>
</tr>
<tr>
<td>$w_i$</td>
<td>Initial moisture content</td>
<td>$kg$</td>
<td>$M$</td>
</tr>
<tr>
<td>$X$</td>
<td>Hold-up of drum</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$X_p$</td>
<td>Hold-up of drum with air flow</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$\phi$</td>
<td>Moisture content of wet feed</td>
<td>$kg/kg$</td>
<td></td>
</tr>
<tr>
<td>$\delta$</td>
<td>Factor depending on type of packing</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$y$</td>
<td>Distance in direction of diffusion</td>
<td>$m$</td>
<td>$L$</td>
</tr>
<tr>
<td>$z_d$</td>
<td>Distance below meniscus</td>
<td>$m$</td>
<td>$L$</td>
</tr>
<tr>
<td>$\alpha$</td>
<td>Angle of contact</td>
<td></td>
<td></td>
</tr>
<tr>
<td>$\Delta$</td>
<td>Coefficient</td>
<td></td>
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<tr>
<td>$\Theta$</td>
<td>Gas temperature</td>
<td>$K$</td>
<td>$\theta$</td>
</tr>
<tr>
<td>$\theta$</td>
<td>Surface temperature</td>
<td>$K$</td>
<td>$\theta$</td>
</tr>
<tr>
<td>$\theta_w$</td>
<td>Wet bulb temperature</td>
<td>$K$</td>
<td>$\theta$</td>
</tr>
<tr>
<td>$\delta_h$</td>
<td>Latent heat of vaporisation per unit mass</td>
<td>$kJ/kg$</td>
<td>$L^2 T^{-3}$</td>
</tr>
<tr>
<td>Symbol</td>
<td>Description</td>
<td>Units in SI System</td>
<td>Dimensions in M, N, L, T, θ</td>
</tr>
<tr>
<td>--------</td>
<td>-----------------------------------------------------------</td>
<td>--------------------</td>
<td>----------------------------</td>
</tr>
<tr>
<td>$\lambda_b$</td>
<td>Mean latent heat vaporisation per unit mass of $T_b$</td>
<td>kJ/kg</td>
<td>L$^2$ T$^{-2}$</td>
</tr>
<tr>
<td>$\lambda_{opt}$</td>
<td>Optimum wavelength for jet disruption</td>
<td>m</td>
<td>L</td>
</tr>
<tr>
<td>$\mu$</td>
<td>Viscosity</td>
<td>Ns/m$^2$</td>
<td>ML$^{-1}$ T$^{-1}$</td>
</tr>
<tr>
<td>$\rho$</td>
<td>Density</td>
<td>kg/m$^3$</td>
<td>ML$^{-3}$</td>
</tr>
<tr>
<td>$\rho_A$</td>
<td>Density of air at its mean partial pressure</td>
<td>kg/m$^3$</td>
<td>ML$^{-3}$</td>
</tr>
<tr>
<td>$\sigma$</td>
<td>Surface tension</td>
<td>J/m$^2$</td>
<td>MT$^{-2}$</td>
</tr>
<tr>
<td>$\phi$</td>
<td>Half-angle of spray cone or sheet</td>
<td>—</td>
<td>—</td>
</tr>
</tbody>
</table>