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1. INTRODUCTION

The separation operation of drying converts a solid, semi-solid or liquid feedstock into a solid product by evaporation of the liquid into a vapor phase via application of heat. In the special case of freeze drying, which takes place below the triple point of the liquid being removed, drying occurs by sublimation of the solid phase directly into the vapor phase. This definition thus excludes conversion of a liquid phase into a concentrated liquid phase (evaporation), mechanical dewatering operations such as filtration, centrifugation, sedimentation, supercritical extraction of water from gels to produce extremely high porosity aerogels (extraction) or so-called drying of liquids and gases by use of molecular sieves (adsorption). Phase change and production of a solid phase as end product are essential features of the drying process. Drying is an essential operation in the chemical, agricultural, biotechnology, food, polymer, ceramics, pharmaceutical, pulp and paper, mineral processing, and wood processing industries.

Drying is perhaps the oldest, most common and most diverse of chemical engineering unit operations. Over four hundred types of dryers have been reported in the literature while over one hundred distinct types are commonly available. It competes with distillation as the most energy-intensive unit operation due to the high latent heat of vaporization and the inherent inefficiency of using hot air as the (most common) drying medium. Various studies report national energy consumption for industrial drying operations ranging from 10-15% for USA, Canada, France, and UK to 20-25% for Denmark and Germany. The latter figures have been obtained recently based on mandatory energy audit data supplied by industry and hence are more reliable.
Energy consumption in drying ranges from a low value of under five percent for the chemical process industries to thirty five percent for the papermaking operations. In the USA, for example, capital expenditures for dryers are estimated to be in the order of only $800 million per annum. Thus, the major costs for dryers are in their operation rather than in their initial investment costs.

Drying of various feedstocks is needed for one or several of the following reasons: need for easy-to-handle free-flowing solids, preservation and storage, reduction in cost of transportation, achieving desired quality of product, etc. In many processes, improper drying may lead to irreversible damage to product quality and hence a non-salable product.

Before proceeding to the basic principles, it is useful to note the following unique features of drying which make it a fascinating and challenging area for R&D:

- Product size may range from microns to tens of centimeters (in thickness or depth)
- Product porosity may range from zero to 99.9 percent
- Drying times range from 0.25 sec (drying of tissue paper) to five months (for certain hardwood species)
- Production capacities may range from 0.10 kg/h to 100 t/h
- Product speeds range from zero (stationary) to 2000 m/s (tissue paper)
- Drying temperatures range from below the triple point to above the critical point of the liquid
- Operating pressure may range from fraction of a millibar to 25 atmospheres
- Heat may be transferred continuously or intermittently by convection, conduction, radiation or electromagnetic fields

Clearly, no single design procedure that can apply to all or even several of the dryer variants is possible. It is therefore essential to revert to the fundamentals of heat, mass and momentum transfer coupled with a knowledge of the material properties (quality) when attempting design of a dryer or analysis of an existing dryer. Mathematically speaking, all processes involved, even in the simplest dryer, are highly nonlinear and hence scale-up of dryers is generally very difficult. Experimentation at laboratory and pilot scales coupled with field experience and know-how is essential to the development of a new dryer application. Dryer vendors are necessarily specialized and normally offer only a narrow range of drying equipment. The buyer must therefore be reasonably conversant with the basic knowledge of the wide assortment of dryers and be able to come up with an informal preliminary selection before going to the vendors with notable exceptions. In general, several different dryers may be able to handle a given application.

2. BASIC PRINCIPLES AND TERMINOLOGY

Drying is a complex operation involving transient transfer of heat and mass along with several rate processes, such as physical or chemical transformations, which, in turn,
may cause changes in product quality as well as the mechanisms of heat and mass transfer. Physical changes that may occur include: shrinkage, puffing, crystallization, glass transitions. In some cases, desirable or undesirable chemical or biochemical reactions may occur leading to changes in color, texture, odor or other properties of the solid product. In the manufacture of catalysts, for example, drying conditions can yield significant differences in the activity of the catalyst by changing the internal surface area.

Drying occurs by effecting vaporization of the liquid by supplying heat to the wet feedstock. As noted earlier, heat may be supplied by convection (direct dryers), by conduction (contact or indirect dryers), radiation or volumetrically by placing the wet material in a microwave or radio frequency electromagnetic field. Over 85 percent of industrial dryers are of the convective type with hot air or direct combustion gases as the drying medium. Over 99 percent of the applications involve removal of water. All modes except the dielectric (microwave and radio frequency) supply heat at the boundaries of the drying object so that the heat must diffuse into the solid primarily by conduction. The liquid must travel to the boundary of the material before it is transported away by the carrier gas (or by application of vacuum for non-convective dryers).

Transport of moisture within the solid may occur by any one or more of the following mechanisms of mass transfer:

- Liquid diffusion, if the wet solid is at a temperature below the boiling point of the liquid
- Vapor diffusion, if the liquid vaporizes within material
- Knudsen diffusion, if drying takes place at very low temperatures and pressures, e.g., in freeze drying
- Surface diffusion (possible although not proven)
- Hydrostatic pressure differences, when internal vaporization rates exceed the rate of vapor transport through the solid to the surroundings
- Combinations of the above mechanisms

Note that since the physical structure of the drying solid is subject to change during drying the mechanisms of moisture transfer may also change with elapsed time of drying.

2.1 Thermodynamic Properties of Air-Water Mixtures and Moist Solids

2.1.1 Psychrometry

As noted earlier, a majority of dryers are of direct (or convective) type. In other words, hot air is used both to supply the heat for evaporation and to carry away the evaporated moisture from the product. Notable exceptions are freeze and vacuum dryers, which are used almost exclusively for drying heat-sensitive products because they tend to be significantly more expensive than dryers operate near to atmospheric pressure. Another exception is the emerging technology of superheated steam drying (Mujumdar, 1995). In certain cases, such as the drum drying of pasty foods, some or all of the heat is supplied indirectly by conduction.
Drying with heated air implies humidification and cooling of the air in a well-insulated (adiabatic) dryer. Thus, hygrothermal properties of humid air are required for the design calculations of such dryers. Table 1 summarizes the essential thermodynamic and transport properties of the air-water system. In Table 2, a listing of brief definitions of various terms encountered in drying and psychrometry is given. It also includes several terms not explicitly discussed in the text.

**Table 1** Thermodynamic and transport properties of air-water system
(Mujumdar, 1995; Pakowski et al., 1991)

<table>
<thead>
<tr>
<th>Property</th>
<th>Expression</th>
</tr>
</thead>
<tbody>
<tr>
<td>( P_v )</td>
<td>( P_v = 100 \exp[27.0214 - (6887 / T_{ar}) - 532 \ln(T_{ar} / 273.16)] )</td>
</tr>
<tr>
<td>( Y )</td>
<td>( Y = 0.622 R H P / (P - R H P) )</td>
</tr>
<tr>
<td>( c_{pg} )</td>
<td>( c_{pg} = 1.00926 \times 10^3 - 4.0403 \times 10^{-2} T + 6.1759 \times 10^{-4} T^2 - 4.097 \times 10^{-6} T^3 )</td>
</tr>
<tr>
<td>( k_g )</td>
<td>( k_g = 2.425 \times 10^{-3} - 7.889 \times 10^{-5} T - 1.790 \times 10^{-7} T^2 - 8.570 \times 10^{-12} T^3 )</td>
</tr>
<tr>
<td>( \rho_g )</td>
<td>( \rho_g = PM_g / (R T_{ar}) )</td>
</tr>
<tr>
<td>( \mu_g )</td>
<td>( \mu_g = 1.691 \times 10^{-3} + 4.984 \times 10^{-3} T - 31.87 \times 10^{-11} T^2 + 1319 \times 10^{-14} T^3 )</td>
</tr>
<tr>
<td>( c_{pv} )</td>
<td>( c_{pv} = 1.883 - 1.6737 \times 10^{-4} T + 8.4386 \times 10^{-7} T^2 - 2.6966 \times 10^{-10} T^3 )</td>
</tr>
<tr>
<td>( c_{pw} )</td>
<td>( c_{pw} = 2.8223 + 1.1828 \times 10^{-2} T - 3.5043 \times 10^{-5} T^2 + 3.601 \times 10^{-8} T^3 )</td>
</tr>
</tbody>
</table>

Figure 1 is a psychrometric chart for the air-water system. It shows the relationship between the temperature (abscissa) and absolute humidity (ordinate, in kg water per kg dry air) of humid air at one atmosphere absolute pressure over 0\(^\circ\) to 130\(^\circ\)C. Line representing percent humidity and adiabatic saturation are drawn according to the thermodynamic definitions of these terms. Equations for the adiabatic saturation and wet-bulb temperature lines on the chart are as follows (Geankoplis, 1993):

\[
\frac{Y - Y_{ar}}{T - T_{ar}} = -\frac{c_s}{\lambda_{ar}} = -\frac{1.005 + 1.88Y}{\lambda_{as}} \tag{1}
\]

and

\[
\frac{Y - Y_{wb}}{T - T_{wb}} = -\frac{h / M_{ar} k_y}{\lambda_{wb}} \tag{2}
\]

The ratio \((h/M_{air}k_y)\), the psychrometric ratio, lies between 0.96 - 1.005 for air-water vapor mixtures; thus it is nearly equal to the value of humid heat \(c_s\). If the effect of humidity is neglected, the adiabatic saturation and wet-bulb temperatures \((T_{as} \text{ and } T_{wb})\) are almost equal for the air-water system. Note, however, that \(T_{as} \text{ and } T_{wb}\) are conceptually quite different. The adiabatic saturation temperature is a gas temperature and a thermodynamic entity while the wet-bulb temperature is a heat and mass transfer rate-based entity and refers to the temperature of the liquid phase. Under constant drying conditions, the surface of the drying material attains the wet-bulb temperature if heat transfer is by pure convection. The wet-bulb temperature is independent of surface geometry as a result of the analogy between heat and mass transfer.
### Table 2 Definitions of commonly encountered terms in psychrometry and drying

<table>
<thead>
<tr>
<th>Term/symbol</th>
<th>Meaning</th>
</tr>
</thead>
<tbody>
<tr>
<td>Adiabatic saturation temperature, $T_{as}$</td>
<td>Equilibrium gas temperature reached by unsaturated gas and vaporizing liquid under adiabatic conditions. (Note: For air/water system only, it is equal to the wet bulb temperature ($T_{wb}$))</td>
</tr>
<tr>
<td>Bound moisture</td>
<td>Liquid physically and/or chemically bound to solid matrix so as to exert a vapor pressure lower than that of pure liquid at the same temperature</td>
</tr>
<tr>
<td>Constant rate drying period</td>
<td>Under constant drying conditions, drying period when evaporation rate per unit drying area is constant (when surface moisture is removed)</td>
</tr>
<tr>
<td>Dew point</td>
<td>Temperature at which a given unsaturated air-vapor mixture becomes saturated</td>
</tr>
<tr>
<td>Dry bulb temperature</td>
<td>Temperature measured by a (dry) thermometer immersed in vapor-gas mixture</td>
</tr>
<tr>
<td>Equilibrium moisture content, $X^*$</td>
<td>At a given temperature and pressure, the moisture content of moist solid in equilibrium with the gas-vapor mixture (zero for non-hygroscopic solids)</td>
</tr>
<tr>
<td>Critical moisture content, $X_c$</td>
<td>Moisture content at which the drying rate first begins to drop (under constant drying conditions)</td>
</tr>
<tr>
<td>Falling rate period</td>
<td>Drying period (under constant drying conditions) during which the rate falls continuously in time</td>
</tr>
<tr>
<td>Free moisture, $X_f$</td>
<td>Moisture content in excess of the equilibrium moisture content (hence free to be removed) at given air humidity and temperature</td>
</tr>
<tr>
<td>Humid heat</td>
<td>Heat required to raise the temperature of unit mass of dry air and its associated vapor through one degree (J kg$^{-1}$ K$^{-1}$ or Btu lb$^{-1}$ °F$^{-1}$)</td>
</tr>
<tr>
<td>Humidity, absolute</td>
<td>Mass of water vapor per unit mass of dry gas (kg kg$^{-1}$ or lb lb$^{-1}$)</td>
</tr>
<tr>
<td>Humidity, relative</td>
<td>Ratio of partial pressure of water vapor in gas-vapor mixture to equilibrium vapor pressure at the same temperature</td>
</tr>
<tr>
<td>Unbound moisture</td>
<td>Moisture in solid which exerts vapor pressure equal to that of pure liquid at the same temperature</td>
</tr>
<tr>
<td>Water activity, $a_w$</td>
<td>Ratio of vapor pressure exerted by water in solid to that of pure water at the same temperature</td>
</tr>
<tr>
<td>Wet bulb temperature</td>
<td>Liquid temperature attained when large amounts of air-vapor</td>
</tr>
</tbody>
</table>
ture, $T_{wb}$ mixture is contacted with the surface. In purely convective drying, drying surface reaches $T_{wb}$ during the constant rate period

![Psychrometric chart for the air-water system](image)

**Figure 1** Psychrometric chart for the air-water system

Most handbooks of engineering provide more detailed psychrometric charts including additional information and extended temperature ranges. Mujumdar (1995) includes numerous psychrometric charts for several gas-organic vapor systems as well.

### 2.1.2 Equilibrium Moisture Content

The moisture content of a wet solid in equilibrium with air of given humidity and temperature is termed the equilibrium moisture content (EMC). A plot of EMC at a given temperature versus the relative humidity is termed sorption isotherm. An isotherm obtained by exposing the solid to air of increasing humidity gives the adsorption isotherm. That obtained by exposing the solid to air of decreasing humidity is known as the desorption isotherm. Clearly, the latter is of interest in drying as the moisture content of the solids progressively decreases. Most drying materials display “hysteresis” in that the two isotherms are not identical.
Figure 2 shows the general shape of the typical sorption isotherms. They are characterized by three distinct zones, A, B and C, which are indicative of different water binding mechanisms at individual sites on the solid matrix. In region A, water is tightly bound to the sites and is unavailable for reaction. In this region, there is essentially monolayer adsorption of water vapor and no distinction exists between the adsorption and desorption isotherms. In region B, the water is more loosely bound. The vapor pressure depression below the equilibrium vapor pressure of water at the same temperature is due to its confinement in smaller capillaries. Water in region C is even more loosely held in larger capillaries. It is available for reactions and as a solvent.

Numerous hypotheses have been proposed to explain the hysteresis. The reader is referred elsewhere (Bruin and Luyben, 1980; Fortes and Okos, 1980; Bruin, 1988) for more information on the topic.

Figure 3 shows schematically the shapes of the equilibrium moisture curves for various types of solids. Figure 4 shows the various types of moisture defined in Table 2. Desorption isotherms are also dependent on external pressure. However, in all practical cases of interest, this effect may be neglected.
According to Keey (1978), the temperature dependence of the equilibrium moisture content can be correlated by:

$$\frac{\Delta X^*}{\Delta T}_{\Psi=\text{constant}} = -\alpha X^*$$  \hspace{1cm} (3)

where $X^*$ is the dry-basis equilibrium moisture content, $T$ is the temperature and $\Psi^*$ is the relative humidity of air. The parameter $\alpha$ ranges from 0.005 to 0.01 K$^{-1}$. This correlation may be used to estimate the temperature dependence of $X^*$ if no data are available.

For hygroscopic solids, the enthalpy of the attached moisture is less than that of pure liquid by an amount equal to the binding energy, which is also termed the enthalpy of

**Figure 3** Equilibrium moisture content curves for various types of solids

**Figure 4** Various types of moisture content
wetting, $\Delta H_w$ (Keey, 1978). It includes the heat of sorption, hydration and solution and may be estimated from the following equation:

$$\frac{d(\ln \Psi)}{d(1/T)}|_{X=\text{constant}} = \frac{-\Delta H_w}{R_g T}$$

(4)

A plot of $\ln(\Psi)$ against $1/T$ is linear with a slope of $\Delta H_w/R_g$ where $R_g$ is the universal gas constant ($R_g = 8.314 \times 10^3$ kg kmol$^{-1}$ K$^{-1}$). Note that the total heat required to evaporate bound water is the sum of the latent heat of vaporization and the heat of wetting; the latter is a function of the moisture content $X$. The heat of wetting is zero for unbound water and increases with decreasing $X$. Since $\Delta H_w$ is responsible for lowering the vapor pressure of bound water, at the same relative humidity, $\Delta H_w$ is almost the same for all materials (Keey, 1978).

For most materials, the moisture binding energy is positive; generally it is a monotonically decreasing function of the moisture content, with a value of zero for unbound moisture. For hydrophobic food materials (e.g., peanut oil, starches at lower temperatures) the binding energy can, however, be negative.

In general, water sorption data must be determined experimentally. Some 80 correlations, ranging from those based on theory to those that are purely empirical, have appeared in the literature. Two of the most extensive compilations are due to Wolf et al. (1985) and Iglesias and Chirife (1982). Aside from temperature, the physical structure as well as composition of the material also affect water sorption. The pore structure and size as well as physical and/or chemical transformations during drying can cause significant variations in the moisture binding ability of the solid.

### 2.1.3 Water Activity

In drying of some materials, which require careful hygienic attention, e.g., food, the availability of water for growth of microorganisms, germination of spores, and participation in several types of chemical reaction becomes an important issue. This availability, which depends on relative pressure, or water activity, $a_w$, is defined as the ratio of the partial pressure, $p$, of water over the wet solid system to the equilibrium vapor pressure, $p_w$, of water at the same temperature. Thus, $a_w$, which is also equal to the relative humidity of the surrounding humid air, is defined as:

$$a_w = \frac{P}{P_w}$$

(5)

Different shapes of the $X$ versus $a_w$ curves are observed, depending on the type of material (e.g., high, medium or low hygroscopicity solids).

Table 3 lists the measured minimum $a_w$ values for microbial growth or spore germination. If $a_w$ is reduced below these values by dehydration or by adding water-binding agents like sugars, glycerol, or salt, microbial growth is inhibited. Such additives should not affect the flavor, taste, or other quality criteria, however. Since the amounts of soluble additives needed to depress $a_w$ even by 0.1 is quite large, dehydration becomes particularly attractive for high moisture foods as a way to reduce $a_w$. Figure 5 shows schematically the
water activity versus moisture content curve for different types of food. Rockland and Beuchat (1987) provide an extensive compilation of results on water activity and its applications.

**Table 3** Minimum water activity, $a_{w}$, for microbial growth and spore germination (adapted from Brockmann, 1973)

<table>
<thead>
<tr>
<th>Micro-organism</th>
<th>Water activity</th>
</tr>
</thead>
<tbody>
<tr>
<td>Organisms producing slime on meat</td>
<td>0.98</td>
</tr>
<tr>
<td><em>Pseudomonas, Bacillus cereus</em> spores</td>
<td>0.97</td>
</tr>
<tr>
<td><em>B. subtilis, C. botulinum</em> spores</td>
<td>0.95</td>
</tr>
<tr>
<td><em>C. botulinum, Salmonella</em></td>
<td>0.93</td>
</tr>
<tr>
<td>Most bacteria</td>
<td>0.91</td>
</tr>
<tr>
<td>Most yeast</td>
<td>0.88</td>
</tr>
<tr>
<td><em>Aspergillus niger</em></td>
<td>0.85</td>
</tr>
<tr>
<td>Most molds</td>
<td>0.80</td>
</tr>
<tr>
<td>Halophilic bacteria</td>
<td>0.75</td>
</tr>
<tr>
<td>Xerophilic fungi</td>
<td>0.65</td>
</tr>
<tr>
<td>Osmophilic yeast</td>
<td>0.62</td>
</tr>
</tbody>
</table>
Dry foods
Intermediate moisture foods (IMF)
e.g. syrups
cheese, sausages, candy

Figure 5 Water activity versus moisture content plot for different types of food

Figure 6 shows the general nature of the deterioration reaction rates as a function of $a_w$ for food systems. Aside from microbial damage, which typically occurs for $a_w > 0.70$, oxidation, non-enzymatic browning (Maillard reactions) and enzymatic reactions can occur even at very low $a_w$ levels during drying. Laboratory or pilot testing is essential to ascertain that no damage occurs in the selected drying process since this cannot, in general, be predicted.
2.2 Drying Kinetics

Consider the drying of a wet solid under fixed drying conditions. In the most general cases, after an initial period of adjustment, the dry-basis moisture content, \( X \), decreases linearly with time, \( t \), following the start of the evaporation. This is followed by a non-linear decrease in \( X \) with \( t \) until, after a very long time, the solid reaches its equilibrium moisture content, \( X^* \) and drying stops. In terms of free moisture content, defined as:

\[
X_f = (X - X^*)
\]

the drying rate drop to zero at \( X_f = 0 \).

By convention, the drying rate, \( N \), is defined as:

\[
N = \frac{M_s}{A} \frac{dX}{dt} \text{ or } \frac{M_s}{A} \frac{dX_f}{dt}
\]

under constant drying conditions. Here, \( N \) (kg m\(^{-2}\) h\(^{-1}\)) is the rate of water evaporation, \( A \) is the evaporation area (may be different from heat transfer area) and \( M_s \) is the mass of bone dry solid. If \( A \) is not known, then the drying rate may be expressed in kg water evaporated per hour.
A plot of $N$ versus $X$ (or $X_f$) is the so-called drying rate curve. This curve must always be obtained under constant drying conditions. Note that, in actual dryers, the drying material is generally exposed to varying drying conditions (e.g., different relative gas-solid velocities, different gas temperatures and humidities, different flow orientations). Thus, it is necessary to develop a methodology in order to interpolate or extrapolate limited drying rate data over a range of operating conditions.

Figure 7 shows a typical “textbook” drying rate curve displaying an initial constant rate period where $N = N_c = \text{constant}$. The constant rate period is governed fully by the rates of external heat and mass transfer since a film of free water is always available at the evaporating surface. This drying period is nearly independent of the material being dried. Many foods and agricultural products, however, do not display the constant rate period at all since internal heat and mass transfer rates determine the rate at which water becomes available at the exposed evaporating surface.

At the so-called critical moisture content, $X_c$, $N$ begins to fall with further decrease in $X$ since water cannot migrate at the rate $N_c$ to the surface due to internal transport limitations. The mechanism underlying this phenomenon depends on both the material and drying conditions. The drying surface becomes first partially unsaturated and then fully unsaturated until it reaches the equilibrium moisture content $X^*$. Detailed discussions of drying rate curves are given by Keey (1991), Mujumdar and Menon (1995). Approximate critical moisture content values for some selected materials are given in Table 4.

Note that a material may display more than one critical moisture contents at which the drying rate curve shows a sharp change of shape. This is generally associated with changes in the underlying mechanisms of drying due to structural or chemical changes. It is
also important to note that \( X_c \) is not solely a material property. It depends on the drying rate under otherwise similar conditions. It must be determined experimentally.

**Table 4** Approximate critical moisture contents for various materials

<table>
<thead>
<tr>
<th>Material</th>
<th>Critical moisture content (kg water/kg dry solid)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Salt crystals, rock salt, sand, wool</td>
<td>0.05-0.10</td>
</tr>
<tr>
<td>Brick clay, kaolin, crushed sand</td>
<td>0.10-0.20</td>
</tr>
<tr>
<td>Pigments, paper, soil, worsted wool fabric</td>
<td>0.20-0.40</td>
</tr>
<tr>
<td>Several foods, copper carbonate, sludges</td>
<td>0.40-0.80</td>
</tr>
<tr>
<td>Chrome leather, vegetables, fruits, gelatin, gels</td>
<td>&gt; 0.80</td>
</tr>
</tbody>
</table>

It is easy to see that \( N_c \) can be calculated using empirical or analytical techniques to estimate the external heat/mass transfer rates (Keey, 1978; Geankoplis, 1993). Thus,

\[
N_c = \sum \frac{q}{\lambda_v}
\]  

where \( \sum q \) represents the sum of heat fluxes due to convection, conduction and/or radiation and \( \lambda_v \) is the latent heat of vaporization at the solid temperature. In the case of purely convective drying, the drying surface is always saturated with water in the constant rate period and thus the liquid film attains the wet-bulb temperature. The wet-bulb temperature is independent of the geometry of the drying object due to the analogy between heat and mass transfer.

The drying rate in the falling rate period(s) is a function of \( X \) (or \( X_f \)) and must be determined experimentally for a given material being dried in a given type of dryer.

If the drying rate curve \( (N \text{ versus } X) \) is known, the total drying time required to reduce the solid moisture content from \( X_1 \) to \( X_2 \) can be simply calculated by:

\[
t_d = -\int_{X_1}^{X_2} \frac{M_s}{A} \frac{dX}{N}
\]  

Table 5 lists expressions for the drying time for constant rate, linear falling rates and a falling rate controlled by liquid diffusion of water in a thin slab. The subscripts \( c \) and \( f \) refer to the constant and falling rate periods, respectively. The total drying time is, of course, a sum of drying times in two succeeding periods. Different analytical expressions are obtained for the drying times \( t_f \) depending on the functional form of \( N \) or the model used to describe the falling rate, e.g., liquid diffusion, capillarity, evaporation-condensation. For some solids, a receding front model (wherein the evaporating surface recedes into the drying solid) yields a good agreement with experimental observations. The principal goal of all
falling rate drying models is to allow reliable extrapolation of drying kinetic data over various operating conditions and product geometries.

**Table 5** Drying times for various drying rate models (Mujumdar, 1997)

<table>
<thead>
<tr>
<th>Model</th>
<th>Drying time</th>
</tr>
</thead>
<tbody>
<tr>
<td>Kinetic model, ( N = -\frac{M_s}{A} \frac{dX}{dt} )</td>
<td>( t_d = ) Drying time to reach final moisture content ( X_2 ) from initial moisture content ( X_1 )</td>
</tr>
<tr>
<td>( N = N(X) ) (General)</td>
<td>( t_d = \frac{M_s}{A} \int_{X_1}^{X_2} \frac{dX}{N(X)} )</td>
</tr>
<tr>
<td>( N = N_c ) (Constant rate period)</td>
<td>( t_c = -\frac{M_s}{A} \frac{(X_2 - X_1)}{N_c} )</td>
</tr>
<tr>
<td>( N = aX + b ) (Falling rate period)</td>
<td>( t_f = \frac{M_s}{A} \frac{(X_1 - X_2)}{(N_1 - N_2)} \ln \frac{N_1}{N_2} )</td>
</tr>
<tr>
<td>( N = Ax ) ( X^* \leq X_2 \leq X_c ) (Falling rate period)</td>
<td>( t_f = \frac{M_s X_c}{A N_c} \ln \frac{X_c}{X_2} )</td>
</tr>
<tr>
<td>Liquid diffusion model</td>
<td>( t_f = \frac{4a^2}{\pi D_L} \ln \frac{8X_1}{\pi^2 X_2} )</td>
</tr>
<tr>
<td>( D_L = ) constant, ( X_2 = X_c ) Slab; one-dimensional diffusion, evaporating surface at ( X^* )</td>
<td>( X = ) average free moisture content ( a = ) half-thickness of slab</td>
</tr>
</tbody>
</table>

The expression for \( t_f \) in Table 5 using the liquid diffusion model (Fick's second law of diffusion form applied to diffusion in solids with no real fundamental basis) is obtained by solving analytically the following partial differential equation:

\[
\frac{\partial X_f}{\partial t} = D_L \frac{\partial^2 X_f}{\partial x^2}
\]

subject to the following initial and boundary conditions:

\[
\begin{align*}
X_f &= X_i, \text{ everywhere in the slab at } t = 0 \\
X_f &= 0, \text{ at } x = a \text{ (top, evaporating surface), and} \\
\frac{\partial X_f}{\partial x} &= 0, \text{ at } x = 0 \text{ (bottom, non-evaporating surface)}
\end{align*}
\]

The model assumes one-dimensional liquid diffusion with constant effective diffusivity, \( D_L \), and no heat (Soret) effects. \( X_2 \) is the average free moisture content at \( t = t_f \) obtained by integrating the analytical solution \( X_f(x,t_f) \) over the thickness of the slab, \( a \). The expression in Table 5 is applicable only for long drying times since it is obtained by retaining only the first term in the infinite series solution of the partial differential equation.
The moisture diffusivity in solids is a function of both temperature and moisture content. For strongly shrinking materials the mathematical model used to define $D_L$ must account for the changes in diffusion path as well. The temperature dependence of diffusivity is adequately described by the Arrhenius equation as follows:

$$D_L = D_{L0} \exp\left[-\frac{E_a}{R g T_{\text{abs}}}\right]$$  \hspace{1cm} (12)$$

where $D_L$ is the diffusivity, $E_a$ is the activation energy and $T_{\text{abs}}$ is the absolute temperature. Okos et al. (1992) have given an extensive compilation of $D_L$ and $E_a$ values for various food materials. Zogzas et al. (1996) provide methods of moisture diffusivity measurement and an extensive bibliography on the topic. Approximate ranges of effective moisture diffusivity for some selected materials are given in Table 6.

It should be noted that $D_L$ is not a true material property and care should be taken in applying effective diffusivity correlations obtained with simple geometric shapes (e.g., slab, cylinder or sphere) to the more complex shapes actually encountered in practice as this may lead to incorrect calculated results (Gong et al., 1997).

In addition to being dependent on geometric shapes, diffusivity depends as well on the drying conditions. At very high activity levels, no differences might be observed but at lower activity levels, the diffusivities may differ by an order-of-magnitude due to the inherently different physical structure of the dried product. Thus, the effective diffusivity is regarded as a lumped property that does not really distinguish between the transport of water by liquid or vapor diffusion, capillary or hydrodynamic flow due to pressure gradients set up in the material during drying. Further, the diffusivity values will show marked variations if the material undergoes glass transition during the drying process.

### Table 6 Approximate ranges of effective moisture diffusivity in some materials
(Zogzas et al. (1996), Marinos-Kouris and Marouris (1995) and other sources)

<table>
<thead>
<tr>
<th>Material</th>
<th>Moisture content (kg/kg, d.b.)</th>
<th>Temperature (°C)</th>
<th>Diffusivity (m$^2$/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Alfalfa stems</td>
<td>3.70</td>
<td>26</td>
<td>2.6 × $10^{-10}$ - 2.6 × $10^{-9}$</td>
</tr>
<tr>
<td>Animal feed</td>
<td>0.01 - 0.15</td>
<td>25</td>
<td>1.8 × $10^{-11}$ - 2.8 × $10^{-9}$</td>
</tr>
<tr>
<td>Apple</td>
<td>0.10 - 1.50</td>
<td>30 - 70</td>
<td>1.0 × $10^{-11}$ - 3.3 × $10^{-9}$</td>
</tr>
<tr>
<td>Asbestos cement</td>
<td>0.10 - 0.60</td>
<td>20</td>
<td>2.0 × $10^{-9}$ - 5.0 × $10^{-9}$</td>
</tr>
<tr>
<td>Banana</td>
<td>0.01 - 3.50</td>
<td>20 - 40</td>
<td>3.0 × $10^{-13}$ - 2.1 × $10^{-10}$</td>
</tr>
<tr>
<td>Biscuit</td>
<td>0.10 - 0.60</td>
<td>20 - 100</td>
<td>8.6 × $10^{-10}$ - 9.4 × $10^{-8}$</td>
</tr>
<tr>
<td>Carrot</td>
<td>0.01 - 5.00</td>
<td>30 - 70</td>
<td>1.2 × $10^{-9}$ - 5.9 × $10^{-9}$</td>
</tr>
<tr>
<td>Clay brick</td>
<td>0.20</td>
<td>25</td>
<td>1.3 × $10^{-8}$ - 1.4 × $10^{-8}$</td>
</tr>
<tr>
<td>Egg liquid</td>
<td>-</td>
<td>85 - 105</td>
<td>1.0 × $10^{-11}$ - 1.5 × $10^{-11}$</td>
</tr>
<tr>
<td>Fish muscles</td>
<td>0.05 - 0.30</td>
<td>30</td>
<td>8.1 × $10^{-11}$ - 3.4 × $10^{-10}$</td>
</tr>
<tr>
<td>Glass wool</td>
<td>0.10 - 1.80</td>
<td>20</td>
<td>2.0 × $10^{-9}$ - 1.5 × $10^{-8}$</td>
</tr>
<tr>
<td>Glucose</td>
<td>0.08 - 1.50</td>
<td>30 - 70</td>
<td>4.5 × $10^{-12}$ - 6.5 × $10^{-10}$</td>
</tr>
<tr>
<td>Kaolin clay</td>
<td>&lt; 0.50</td>
<td>45</td>
<td>1.5 × $10^{-8}$ - 1.5 × $10^{-7}$</td>
</tr>
</tbody>
</table>
Keey (1978) and Geankopolis (1993), among others, have provided analytical expressions for liquid diffusion and capillarity models of falling rate drying. Table 7 gives solution of the one-dimensional transient partial differential equations for cartesian, cylindrical and spherical coordinate systems. These results can be utilized to estimate the diffusivity from the falling rate drying data or to estimate the drying rate and drying time if the diffusivity value is known.

It is noteworthy that the diffusivity, $D_L$, is a strong function of $X_f$ as well as temperature and must be determined experimentally. Thus, the liquid diffusion model should be regarded purely as an empirical representation drying in the falling rate period. More advanced models are, of course, available but their widespread use in the design of dryers is hampered by the need for extensive empirical information required to solve the governing equations. Turner and Mujumdar (1997) provide a wide assortment of mathematical models of drying and dryers, and also discuss the application of various techniques for the numerical solution of the complex governing equations.

One simple approach to interpolating a given falling rate curve over a relatively narrow range of operating conditions is that first proposed by van Meel (1958). It is found that the plot of normalized drying rate $\nu = N/N_c$ versus normalized free moisture content $\eta = (X - X*)/(X_e - X*)$ was nearly independent of the drying conditions. This plot, called the characteristic drying rate curve, is illustrated in Figure 8. Thus, if the constant rate-drying rate, $N_c$, can be estimated and the equilibrium moisture content data are available, then the falling rate curve can be estimated using this highly simplified approach. Extrapolation over wide ranges is not recommended, however.

Waananen et al. (1993) have provided an extensive bibliography of over 200 references dealing with models for drying of porous solids. Such models are useful to describe drying processes for the purposes of engineering design, analysis and optimization. A mathematical description of the process is based on the physical mechanisms of internal heat and mass transfer that control the process resistances, as well as the structural and thermodynamic assumptions made to formulate the model. In the constant rate period, the
overall drying rate is determined solely by the heat and mass transfer conditions external to
the material being dried, such as the temperature, gas velocity, total pressure and partial
pressure of the vapor. In the falling rate period, the rates of internal heat and mass transfer
determine the drying rate. Modeling of drying becomes complicated by the fact that more
than one mechanism may contribute to the total mass transfer rate and the contributions
from different mechanisms may even change during the drying process.

\[ \eta = \frac{X - X^*}{X_c - X^*} \]

\[ v = \frac{R}{R_c} \]

**Figure 8** Characteristic drying rate curve

**Table 7** Solution to Fick's second law for some simple geometries
(Pakowski and Mujumdar, 1995)

<table>
<thead>
<tr>
<th>Geometry</th>
<th>Boundary conditions</th>
<th>Dimensionless average free M.C.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flat plate of thickness 2b</td>
<td>( t = 0; -b &lt; z &lt; b; X = X_0 ) ( t &gt; 0; z = \pm b; X = X^* )</td>
<td>( X = \frac{8}{\pi^2} ) ( \sum_{n=1}^{\infty} \frac{1}{(2n-1)} \exp\left[ - (2n-1)^2 \frac{\pi^2}{4b} \left( \frac{D_c t}{b} \right) \right] )</td>
</tr>
<tr>
<td>Infinitely long cylinder of radius ( R )</td>
<td>( t = 0; 0 &lt; r &lt; R; X = X_0 ) ( t &gt; 0; r = R; X = X^* )</td>
<td>( X = 4 \sum_{n=1}^{\infty} \frac{1}{R^2 \alpha_n} \exp(-D_c \alpha_n^2 t) ) where ( \alpha_n ) are positive roots of the equation ( J_0(R \alpha_n) = 0 )</td>
</tr>
<tr>
<td>Sphere of radius ( R )</td>
<td>( t = 0; 0 &lt; r &lt; R; X = X_0 ) ( t &gt; 0; r = R; X = X^* )</td>
<td>( X = \frac{6}{\pi^2} \sum_{n=1}^{\infty} \frac{1}{n^2} \exp\left[ - n^2 \frac{\pi^2}{R} \left( \frac{D_c t}{R} \right) \right] )</td>
</tr>
</tbody>
</table>
Diffusional mass transfer of the liquid phase, as discussed earlier, is the most commonly assumed mechanism of moisture transfer used in modeling drying that takes place at temperatures below the boiling point of the liquid under locally applied pressure. At higher temperatures, the pore pressure may rise substantially and cause a hydrodynamically driven flow of vapor, which, in turn, may cause a pressure driven flow of liquid in the porous material.

For solids with continuous pores, a surface tension driven flow (capillary flow) may occur as a result of capillary forces caused by the interfacial tension between the water and the solid. In the simplest model, a modified form of the Poiseuille flow can be used in conjunction with the capillary force equation to estimate the rate of drying. Geankoplis (1993) has shown that such a model predicts the drying rate in the falling rate period to be proportional to the free moisture content in the solid. At low solid moisture contents, however, the diffusion model may be more appropriate.

The moisture flux due to capillarity can be expressed in terms of the product of a liquid conductivity parameter and moisture gradient. In this case, the governing equation has, in fact, the same form as the diffusion equation.

For certain materials and under conditions such as those encountered in freeze drying, a “receding-front” model involving a moving boundary between “dry” and “wet” zones often describes the mechanism of drying much more realistically than does the simple liquid diffusion or capillarity model. Examination of the freeze drying of a thin slab indicates that the rate of drying is dependent on the rate of heat transfer to the “dry-wet” interface and the mass transfer resistance offered by the porous dry layer to permeation of the vapor which sublimes from the interface. Because of the low pressures encountered in freeze drying, Knudsen diffusion may be significant. Liapis and Marchello (1984) have discussed models of freeze drying involving both unbound and bound moisture.

When drying materials under intense drying conditions, diffusion or capillarity models generally do not apply. If evaporation can occur within the material there is a danger of the so-called “vapor-lock” that occurring within the capillary structure causing breaks in liquid-filled capillaries. This phenomenon can cause departure from the classical drying curve, e.g., no constant rate drying may appear under intense drying conditions but may do so under milder drying conditions (Zaharchuk, 1993).

**CLOSING REMARKS**

An attempt is made here to provide a concise overview of the fundamental principles and terminology used in the drying literature. Advanced models and calculation procedures for drying and dryers of various types can be found in the literature cited. It must be noted that the models and estimation methods given here are necessarily simplistic and caution must be exercised in applying them in practice. Almost without exception design and scale-up of most dryers must be preceded with appropriate laboratory and/or pilot scale experimentation. Although no mention is made here about quality considerations, it is important to recognize that drying involves both heat and mass transfer processes but also material science. Drying affects product quality in a decisive manner and hence must be an essential part of any dryer calculation and specification.
### NOMENCLATURE

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>$A$</td>
<td>evaporation area, m$^2$</td>
</tr>
<tr>
<td>$a_w$</td>
<td>water activity, -</td>
</tr>
<tr>
<td>$c_p$</td>
<td>specific heat, J kg$^{-1}$ K$^{-1}$</td>
</tr>
<tr>
<td>$c_s$</td>
<td>humid heat, J kg$^{-1}$ K$^{-1}$</td>
</tr>
<tr>
<td>$D_L$</td>
<td>effective diffusivity, m$^2$ s$^{-1}$</td>
</tr>
<tr>
<td>$D_{L0}$</td>
<td>effective diffusivity at reference temperature, m$^2$ s$^{-1}$</td>
</tr>
<tr>
<td>$E_a$</td>
<td>activation energy, J</td>
</tr>
<tr>
<td>$\Delta H_w$</td>
<td>enthalpy of wetting, J kg$^{-1}$</td>
</tr>
<tr>
<td>$h$</td>
<td>convective heat transfer coefficient, W m$^{-2}$ K$^{-1}$</td>
</tr>
<tr>
<td>$k_g$</td>
<td>thermal conductivity, W m$^{-1}$ K$^{-1}$</td>
</tr>
<tr>
<td>$k_v$</td>
<td>convective mass transfer coefficient, kg mol s$^{-1}$ m$^{-2}$ mol frac$^{-1}$</td>
</tr>
<tr>
<td>$M_{air}$</td>
<td>molar mass of air, kg mol$^{-1}$</td>
</tr>
<tr>
<td>$M_s$</td>
<td>mass of bone dry solid, kg</td>
</tr>
<tr>
<td>$N$</td>
<td>drying rate, kg m$^{-2}$ h$^{-1}$</td>
</tr>
<tr>
<td>$P_v$</td>
<td>vapor pressure of pure water, Pa</td>
</tr>
<tr>
<td>$p$</td>
<td>partial pressure, Pa</td>
</tr>
<tr>
<td>$p_w$</td>
<td>equilibrium vapor pressure of water, Pa</td>
</tr>
<tr>
<td>$R$</td>
<td>universal gas constant, 8.314 J mol$^{-1}$ K$^{-1}$</td>
</tr>
<tr>
<td>$RH$</td>
<td>relative humidity, decimal fraction, -</td>
</tr>
<tr>
<td>$T$</td>
<td>temperature, °C</td>
</tr>
<tr>
<td>$T_{abs}$</td>
<td>absolute temperature, K</td>
</tr>
<tr>
<td>$T_{wb}$</td>
<td>wet-bulb temperature, °C</td>
</tr>
<tr>
<td>$t$</td>
<td>time, s (or h)</td>
</tr>
<tr>
<td>$X$</td>
<td>total moisture content, kg water/kg dry solid, -</td>
</tr>
<tr>
<td>$X_c$</td>
<td>critical moisture content, kg water/kg dry solid, -</td>
</tr>
<tr>
<td>$X_f$</td>
<td>free moisture content, kg water/kg dry solid, -</td>
</tr>
<tr>
<td>$X^*$</td>
<td>equilibrium moisture content, kg water/kg dry solid, -</td>
</tr>
<tr>
<td>$Y$</td>
<td>absolute air humidity, kg water vapor/kg dry air</td>
</tr>
</tbody>
</table>

### Greek letters

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>$\eta$</td>
<td>normalized drying rate, -</td>
</tr>
<tr>
<td>$\lambda_v$</td>
<td>latent heat of vaporization, J kg$^{-1}$</td>
</tr>
<tr>
<td>$\mu_g$</td>
<td>dynamic viscosity, kg m$^{-1}$ s$^{-1}$</td>
</tr>
<tr>
<td>$\nu$</td>
<td>normalized drying rate, -</td>
</tr>
<tr>
<td>$\rho_g$</td>
<td>density, kg m$^{-3}$</td>
</tr>
</tbody>
</table>

### Subscripts

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
</tr>
</thead>
<tbody>
<tr>
<td>$c$</td>
<td>constant rate period</td>
</tr>
<tr>
<td>$f$</td>
<td>falling rate period</td>
</tr>
<tr>
<td>$g$</td>
<td>gas</td>
</tr>
<tr>
<td>$s$</td>
<td>solid</td>
</tr>
</tbody>
</table>
REFERENCES


1. INTRODUCTION

Dryer selection has long been practiced as an art rather than science depending more on prior experience and vendors’ recommendations. As drying technologies have evolved and become more diverse and complex, this has become an increasingly difficult and demanding task for the non-expert not conversant with the numerous types of equipment, their pros and cons, etc. Further, the task is exasperated by the need to meet stricter quality specifications, higher production rates, higher energy costs and stringent environmental regulations. In the absence of in-house experts in drying, there have been some attempts, albeit not fully successful, to develop expert systems for a non-expert to use. It is therefore necessary for an engineer responsible for selection of a dryer or, more appropriately, a drying system to be aware of what is available in the market, what the key criteria are in the selection process and thus arrive at alternative possibilities before going to vendors of such equipment for comparative quotes. It is time and effort well spent since the cost of incorrect selection can be very high.

This chapter is intended to give an introduction to this subject; the reader is referred to Mujumdar (1995) for further details. Note that over 80 percent of major chemical companies in Europe – each using over 1000 dryers in their production facilities – made errors in selecting dryers in the past year alone. What is optimal choice in one location at one point in time may be a wrong choice for another geographic location some years later. Prior use is a definite help but not the only criterion to be used in selecting drying systems.
As an example, concentrated nickel ore is dried in different parts of the world at very high production rates (20-75 t/h) using flash dryers, fluid bed dryers, rotary dryers as well as spray dryers. It is thus not a simple task to select a dryer for such applications based on what is done elsewhere.

Over 400 dryer types have been cited in the technical literature although only about 50 types are commonly found in practice. In this chapter, we will examine the key classification criteria for industrial dryers and then proceed to selection criteria with the explicit understanding that the latter is a complex process, which is not entirely scientific but also involves subjective judgment as well as considerable empiricism. It should also be noted that the pre-drying as well as post-drying stages have important bearing on the selection of appropriate dryer types for a given application. Indeed, for an optimal selection of process, one must examine the overall flowsheet as well as the “drying system.” This chapter will be confined, however, only to the classification and selection of dryers.

Another important point to note is that several dryer types (or drying systems) may be equally suited (technically and economically) for a given application. A careful evaluation of as many of the possible factors affecting the selection will help reduce the number of options. For a new application (new product or new process), it is important to follow a careful procedure leading to the choice of the dryers. Characteristics of different dryer types should be recognized when selecting dryers. Changes in operating conditions of the same dryer can affect the quality of the product. So, aside from the dryer type, it is also important to choose the right operating conditions for optimal quality and cost of thermal dehydration.

According to a very recent survey conducted by SPIN (Solids Processing Industrial Network, UK, founded by 14 large chemical companies based in Europe) selection of dryers is a key problem faced by all companies (Slangen, 2000). Over ninety percent of the companies had made errors in selection of their new dryers. Sometimes the selection is easy but when a new product is involved or the production capacity required for exceeds current practice, it is not always an easy task. New requirements on safety and environmental aspects can also make the selection more difficult. The SPIN report recommends development of user-friendly expert systems and better standardization to assist with this complex selection process. It should be noted that the selection process is further complicated by the fact that each category of dryers (e.g., fluid bed, flash, spray, rotary) has a wide assortment of sub-classes and, furthermore, each must be operated at optimal conditions to benefit from appropriate selection.

Baker (1997) has presented a “structural approach” for dryer selection, which is iterative. It includes the following steps:

- List all key process specifications
- Carry out preliminary selection
- Carry out bench scale tests including quality tests
- Make economic evaluation of alternatives
- Conduct pilot-scale trials
- Select most appropriate dryer types
Often, for same materials, a specific dryer type is indicated from the outset. If selection is based exclusively on past experience, it has some limitations:

- If the original selection is not optimal (although it works satisfactorily), the new choice will be less-than-optimal
- No new drying technologies are considered by default
- It is implicitly assumed the “old” choice was arrived at logically, which is often not the case

2. CLASSIFICATION OF DRYERS

There are numerous schemes used to classify dryers (Mujumdar, 1995; van't Land, 1991). Table 1 lists the criteria and typical dryer types. Types marked with an asterisk (*) are among the most common in practice.

Table 1 Classification of dryers

<table>
<thead>
<tr>
<th>Criterion</th>
<th>Types</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mode of operation</td>
<td>• Batch</td>
</tr>
<tr>
<td></td>
<td>• Continuous*</td>
</tr>
<tr>
<td>Heat input-type</td>
<td>• Convection*, conduction, radiation, electromagnetic fields, combination of heat transfer modes</td>
</tr>
<tr>
<td></td>
<td>• Intermittent or continuous*</td>
</tr>
<tr>
<td></td>
<td>• Adiabatic or non-adiabatic</td>
</tr>
<tr>
<td>State of material in dryer</td>
<td>• Stationary</td>
</tr>
<tr>
<td></td>
<td>• Moving, agitated, dispersed</td>
</tr>
<tr>
<td>Operating pressure</td>
<td>• Vacuum*</td>
</tr>
<tr>
<td></td>
<td>• Atmospheric</td>
</tr>
<tr>
<td>Drying medium (convection)</td>
<td>• Air*</td>
</tr>
<tr>
<td></td>
<td>• Superheated steam</td>
</tr>
<tr>
<td></td>
<td>• Flue gases</td>
</tr>
<tr>
<td>Drying temperature</td>
<td>• Below boiling temperature*</td>
</tr>
<tr>
<td></td>
<td>• Above boiling temperature</td>
</tr>
<tr>
<td></td>
<td>• Below freezing point</td>
</tr>
<tr>
<td>Relative motion between drying medium and drying solids</td>
<td>• Co-current</td>
</tr>
<tr>
<td></td>
<td>• Counter-current</td>
</tr>
<tr>
<td></td>
<td>• Mixed flow</td>
</tr>
<tr>
<td>Number of stages</td>
<td>• Single*</td>
</tr>
<tr>
<td></td>
<td>• Multi-stage</td>
</tr>
<tr>
<td>Residence time</td>
<td>• Short (&lt; 1 minute)</td>
</tr>
<tr>
<td></td>
<td>• Medium (1 – 60 minutes)</td>
</tr>
<tr>
<td></td>
<td>• Long (&gt; 60 minutes)</td>
</tr>
</tbody>
</table>

* Most common in practice
The above classification is rather coarse. Just the fluidized bed dryer can be subclassified into over thirty types depending on additional criteria.

Each type of dryer has specific characteristics, which make it suited or unsuitable for specific applications. Details can be found in Mujumdar (1995). Certain types are inherently expensive (e.g., freeze dryers) while others are inherently more efficient (e.g., indirect or conductive dryers). Thus, it is necessary to be aware of the wide variety of dryers available in the market as well as their special advantages and limitations. It should be noted that the aforementioned classification does not include most of the novel drying technologies, which are applicable for very specific applications. The reader is referred to Kudra and Mujumdar (1995) for details on novel drying technologies.

Following is a general scheme proposed by Baker (1997) for classification of batch and continuous dryers. Note that there is a more limited choice of batch dryers – only a few types can be operated in both batch and continuous modes.

**Batch Dryers: Classification (Baker, 1997)**

*(Particulate Solids)*

**Major Classes:** Layer (packed bed); Dispersion type

1. Layer type:
   a. Contact (conductive or indirect type), e.g., vacuum tray, agitated bed, rotary batch
   b. Convective (atmospheric tray)
   c. Special types (e.g., microwave, freeze, solar)

2. Dispersion type:
   a. Fluidized bed/spouted bed
   b. Vibrated bed dryer

**Continuous Dryers: Classification**

**Major Classes:** Layer; Dispersion type

1. Layer type:
   a. Conduction, e.g., drum, plate, vacuum, agitated bed, indirect rotary
   b. Convective, e.g., tunnel, spin-flash, throughflow, conveyor
   c. Special, e.g., microwave, RF, freeze, solar

2. Dispersion type:
   a. Fluid bed, vibrated bed, direct rotary, ring, spray, jet-zone

Classification of dryers on the basis of the mode of thermal energy input is perhaps the most useful since it allows one to identify some key features of each class of dryers.

**Direct dryers** – also known as convective dryers – are by far the most common. About 85 percent of industrial dryers are estimated to be of this type despite their relatively low thermal efficiency caused by the difficulty in recovering the latent heat of vaporization contained in the dryer exhaust in a cost-effective manner. Hot air produced
by indirect heating or direct firing is the most common drying medium although for some
special applications superheated steam has recently been shown to yield higher efficiency
and often higher product quality. Flue gases may be used when the product is not heat-
sensitive or affected by the presence of products of combustion. In direct dryers, the
drying medium contacts the material to be dried directly and supplies the heat required for
drying by convection; the evaporated moisture is carried away by the same drying
medium.

Drying gas temperatures may range from 50º C to 400º C depending on the
material. Dehumidified air may be needed when drying highly heat-sensitive materials.
An inert gas such as Nitrogen may be needed when drying explosive or flammable solids
or when an organic solvent is to be removed. Solvents must be recovered from the
exhaust by condensation so that the inert (with some solvent vapor) can be reheated and
returned to the dryer.

Because of the need to handle large volumes of gas, gas cleaning and product
recovery (for particulate solids) becomes a major part of the drying plant. Higher gas
temperatures yield better thermal efficiencies subject to product quality constraints.

**Indirect dryers** – involve supplying of heat to the drying material without direct
contact with the heat transfer medium, i.e., heat is transferred from the heat transfer
medium (steam, hot gas, thermal fluids, etc.) to the wet solid by conduction. Since no gas
flow is presented on the wet solid side it is necessary to either apply vacuum or use gentle
gas flow to remove the evaporated moisture so that the dryer chamber is not saturated
with vapor. Heat transfer surfaces may range in temperature from -40º C (as in freeze
drying) to about 300º C in the case of indirect dryers heated by direct combustion
products such as waste sludges. In vacuum operation, there is no danger of fire or
explosion. Vacuum operation also eases recovery of solvents by direct condensation thus
alleviating serious environmental problem. Dust recovery is obviously simpler so that
such dryers are especially suited for drying of toxic, dusty products, which must not be
entrained in gases. Furthermore, vacuum operation lowers the boiling point of the liquid
being removed; this allows drying of heat-sensitive solids at relatively fast rates.

Heat may also be supplied by radiation (using electric or natural gas-fired
radiator) or volumetrically by placing the wet solid in dielectric fields in the microwave
or radio frequency range. Since radiant heat flux can be adjusted locally over a wide
range it is possible to obtain high drying rates for surface-wet materials. Convection (gas
flow) or vacuum operation is needed to remove the evaporated moisture. Radiant dryers
have found important applications in some niche markets, e.g., drying of coated papers or
printed sheets. However, the most popular applications involve use of combined
convection and radiation. It is often useful to boost the drying capacity of an existing
convective dryer for sheets such as paper.

Microwave dryers are expensive both in terms of the capital and operating
(energy) costs. Only about 50 percent of line power is converted into the electromagnetic
field and only a part of it is actually absorbed by the drying solid. They have found
limited applications to date. However, they do seem to have special advantages in terms
of product quality when handling heat-sensitive materials. They are worth considering as
devices to speed up drying in the tail end of the falling rate period. Similarly, RF dryers
have limited industrial applicability. They have found some niche markets, e.g., drying of thick lumber and coated papers. Both microwave and RF dryers must be used in conjunction with convection or under vacuum to remove the evaporated moisture. Stand-alone dielectric dryers are unlikely to be cost-effective except for high value products in the next decade. See Schiffmann (1995) for detailed discussion of dielectric dryers.

It is possible, indeed desirable in some cases, to use combined heat transfer modes, e.g., convection and conduction, convection and radiation, convection and dielectric fields, to reduce the need for increased gas flow which results in lower thermal efficiencies. Use of such combinations increases the capital costs but these may be offset by reduced energy costs and enhanced product quality. No generalization can be made a priori without careful tests and economic evaluation. Finally, the heat input may be steady (continuous) or time varying. Also, different heat transfer modes may be deployed simultaneously or consecutively depending on individual application. In view of the significant increase in the number of design and operational parameters it is desirable to select the optimal operating conditions via a mathematical model. In batch drying intermittent energy input has great potential for reducing energy consumption and for improving quality of heat-sensitive products.

3. SELECTION OF DRYERS

In view of the enormous choices of dryer types one could possibly deploy for most products, selection of the best type is a challenging task that should not be taken lightly nor should it be left entirely to dryer vendors who typically specialize in only a few types of dryers. The user must take a proactive role and employ vendors' experience and bench-scale or pilot-scale facilities to obtain data, which can be assessed for a comparative evaluation of several options. A wrong dryer for a given application is still a poor dryer, regardless of how well it is designed. Note that minor changes in composition or physical properties of a given product can influence its drying characteristics, handling properties, etc., leading to a different product and in some cases severe blockages in the dryer itself. Tests should be carried out with the “real” feed material and not a “simulated” one where feasible.

Although here we will focus only on the selection of the dryer, it is very important to note that in practice one must select and specify a drying system which includes pre-drying stages (e.g., mechanical dewatering, evaporation, pre-conditioning of feed by solids backmixing, dilution or pelletization and feeding) as well as the post-drying stages of exhaust gas cleaning, product collection, partial recirculation of exhausts, cooling of product, coating of product, agglomeration, etc. The optimal cost-effective choice of dryer will depend, in some cases significantly, on these stages. For example, a hard pasty feedstock can be diluted to a pumpable slurry, atomized and dried in a spray dryer to produce a powder, or it may be pelletized and dried in a fluid bed or in a through circulation dryer, or dried as is in a rotary or fluid bed unit. Also, in some cases, it may be necessary to examine the entire flowsheet to see if the drying problem can be simplified or even eliminated. Typically, non-thermal dewatering is an order-of-magnitude less expensive than evaporation which, in turn, is many-fold energy efficient.
than thermal drying. Demands on product quality may not always permit one to select the least expensive option based solely on heat and mass transfer considerations, however. Often, product quality requirements have over-riding influence on the selection process (see Section 4).

As a minimum, the following quantitative information is necessary to arrive at a suitable dryer:

- Dryer throughput; mode of feedstock production (batch/continuous)
- Physical, chemical and biochemical properties of the wet feed as well as desired product specifications; expected variability in feed characteristics
- Upstream and downstream processing operations
- Moisture content of the feed and product
- Drying kinetics; moist solid sorption isotherms
- Quality parameters (physical, chemical, biochemical)
- Safety aspects, e.g., fire hazard and explosion hazards, toxicity
- Value of the product
- Need for automatic control
- Toxicological properties of the product
- Turndown ratio, flexibility in capacity requirements
- Type and cost of fuel, cost of electricity
- Environmental regulations
- Space in plant

For high value products like pharmaceuticals, certain foods and advanced materials, quality considerations override other considerations since the cost of drying is unimportant. Throughputs of such products are also relatively low, in general.

In some cases, the feed may be conditioned (e.g., size reduction, flaking, pelletizing, extrusion, back-mixing with dry product) prior to drying which affects the choice of dryers.

As a rule, in the interest of energy savings and reduction of dryer size, it is desirable to reduce the feed liquid content by less expensive operations such as filtration, centrifugation and evaporation. It is also desirable to avoid over-drying, which increases the energy consumption as well as drying time.

Drying of food and biotechnological products require adherence to GMP (Good Manufacturing Practice) and hygienic equipment design and operation. Such materials are subject to thermal as well as microbiological degradation during drying as well as in storage.

If the feed rate is low (< 100 kg/h), a batch-type dryer may be suited. Note that there is a limited choice of dryers that can operate in the batch mode.

In less than one percent of cases the liquid to be removed is a non-aqueous (organic) solvent or a mixture of water with a solvent. This is not uncommon in drying of pharmaceutical products, however. Special care is needed to recover the solvent and to avoid potential danger of fire and explosion.

Table 2 presents a typical checklist most dryer vendors use to select and quote an industrial dryer.
Table 2: Typical checklist for selection of industrial dryers

<table>
<thead>
<tr>
<th>Category</th>
<th>Details</th>
</tr>
</thead>
</table>
| Physical form of feed                       | • Granular, particulate, sludge, crystalline, liquid, pasty, suspension, solution, continuous sheets, planks, odd-shapes (small/large)  
• Sticky, lumpy                              |
| Average throughput                          | • kg/h (dry/wet); continuous  
• kg per batch (dry/wet)                         |
| Expected variation in throughput            | (turndown ratio)                                                                                                      |
| Fuel choice                                 | • Oil  
• Gas  
• Electricity                                                                                                      |
| Pre- and post-drying operations             | (if any)                                                                                                              |
| For particulate feed products               | • Mean particle size  
• Size distribution  
• Particle density  
• Bulk density  
• Rehydration properties                      |
| Inlet/outlet moisture content               | • Dry basis  
• Wet basis                                                                                                          |
| Chemical / biochemical / microbiological activity |                                                                                                                      |
| Heat sensitivity                            | • Melting point  
• Glass transition temperature                                                                                     |
| Sorption isotherms (equilibrium moisture content) |                                                                                                                      |
| Drying time                                 | • Drying curves  
• Effect of process variables                                                                                      |
| Special requirements                        | • Material of construction  
• Corrosion  
• Toxicity  
• Non-aqueous solution  
• Flammability limits  
• Fire hazard  
• Color/texture/aroma requirements (if any)                                                                           |
| Foot print of drying system                 | • Space availability for dryer and ancillaries                                                                       |
Drying kinetics play a significant role in the selection of dryers. Aside from simply deciding the residence time required, it limits the types of suitable dryers. Location of the moisture (whether near surface or distributed in the material), nature of moisture (free or strongly bound to solid), mechanisms of moisture transfer (rate limiting step), physical size of product, conditions of drying medium (e.g., temperature, humidity, flow rate of hot air for a convective dryer), pressure in dryer (low for heat-sensitive products), etc., have a bearing on the type of suitable dryer as well as the operating conditions. Most often, not more than one dryer type will likely meet the specified selection criteria.

We will not focus on novel or special drying techniques here for lack of space. However, it is worth mentioning that many of the new techniques use superheated steam as the drying medium or are simply intelligent combinations of traditional drying techniques, e.g., combination of heat transfer modes, multi-staging of different dryer types. Superheated steam as the convective drying medium offers several advantages, e.g., higher drying rates under certain conditions, better quality for certain products, lower net energy consumption if the excess steam produced in the dryer is used elsewhere in the process, elimination of fire and explosion hazard. Vacuum steam drying of timber, for example, can reduce drying times by a factor of up to four while enhancing wood quality and reducing net fuel and electricity consumption by up to 70 percent. The overall economics are also highly favorable.

4. SELECTION OF A DRYER BASED ON QUALITY

As the product quality requirements become increasingly stringent and as the environmental legislation becomes more and more demanding it is often found that we need to switch from one drying technology to the others. The rising cost of energy as well as the differences in the cost of fossil fuels versus electrical energy can also affect the choice of a dryer. Since up to 70 percent of the life cycle cost of a convective dryer is due to energy it is important to choose an energy-efficient dryer where possible even at a higher initial cost. Note that energy costs will continue to rise in the future so this will become increasingly important. Fortunately, improved efficiency also translates into better environmental implications in terms of reduced emissions of the greenhouse gas (CO$_2$) as well as NO$_x$ resulting from combustion.

Following is an example of how selection of the dryer is affected by quality of the dried product that may be used as raw material to produce different consumer products. Shah and Arora (1996) have surveyed the various possible dryers used for crystallization/drying of polyester chips from an initial moisture content of about 0.3-0.5% (w.b.) to under 50 ppm. Aside from low average moisture content it is also necessary to ensure uniform distribution of moisture, especially for some certain products, e.g., production of thin films. The uniformity constraint is less severe if the chips are to be used to make PET bottles. Figure 1 shows schematics of the crystallization/drying steps involved. Generally, it is a two-step process. The material is heat-sensitive. The initial crystallization/drying is faster than the drying step at low moisture levels. A two-stage dryer is indicated and is commonly used. It is possible to
use different dryer types for each stage as shown in Figure 2. A single dryer type (e.g., column or packed bed dryer with the chips moving downward slowly under gravity) is cheaper and hence recommended for the lower quality grade but a more expensive fluid bed followed by another fluid bed or column dryer may be needed for the higher quality grade. Note that numerous alternatives are possible in each case. It is also important to operate the dryers at the correct conditions of gas flow rate, temperature and humidity. Dehumidified air is needed to achieve low final moisture contents in accordance with the equilibrium moisture isotherms of the product.

**Figure 1** Schematic diagram of crystallization/drying steps in the production of polyester chips
Another example of dryer selection is related to the choice of a suitable atomizer for a spray dryer. A spray dryer is indicated when a pumpable slurry, solution or suspension is to be reduced to a free-flowing powder. With proper choice of atomizer, spray chamber design, gas temperature and flow rate it is possible to “engineer” powders of desired particle size and size distribution. Table 3 shows how the choice of the atomizer affects chamber design, size, as well as energy consumption for atomization and particle size distribution. The newly developed two-fluid sonic nozzles appear to be especially attractive choices when nearly monodisperse powders need to be produced from relatively moderate viscosity feeds (e.g., under 250 cp) at capacities up to 80 t/h by using multiple nozzles. More examples may be found in Masters (1985).

Table 3 Spray drying of emulsion-PVC. Effect of selection of atomizer on spray dryer performance: A Comparison between different atomizers

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Rotary disk</th>
<th>Two-fluid (sonic)</th>
<th>Two-fluid (standard)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Dryer geometry</td>
<td>Conical/cylindrical ( H/D \approx 1.2-1.5 ) ( )</td>
<td>Tall-form cylindrical ( H/D \approx 4 ) ( )</td>
<td>Tall-form cylindrical ( H/D \approx 5 ) ( )</td>
</tr>
<tr>
<td>Evaporation capacity (water)</td>
<td>1600 kg/h</td>
<td>1600 kg/h</td>
<td>1600 kg/h</td>
</tr>
<tr>
<td>Chamber ( D \times H )</td>
<td>6.5 m ( \times ) 8 m</td>
<td>3.5 m ( \times ) 15 m</td>
<td>3 m ( \times ) 18 m</td>
</tr>
<tr>
<td>Number of nozzles</td>
<td>1, 175-mm disk 15,000 rpm</td>
<td>16 nozzles 4 bar pressure</td>
<td>18 nozzles 4 bar pressure</td>
</tr>
<tr>
<td>Power for atomizer</td>
<td>25 W/kg slurry 20 W/kg slurry</td>
<td>80 W/kg slurry</td>
<td></td>
</tr>
</tbody>
</table>
New dryers are being developed continuously as a result of industrial demands. Over 250 US patents are granted each year related to dryers (equipment) and drying (process); in the European Community about 80 patents are issued annually on dryers. Kudra and Mujumdar (2000) have discussed a wide assortment of novel drying technologies, which are beyond the scope of this chapter. Suffice it to note that many of the new technologies (e.g., superheated steam, pulse combustion – newer gas-particle contactors as dryers) will eventually replace conventional dryers in the next decade or two. New technologies are inherently more risky and more difficult-to-scale-up. Hence there is natural reluctance to their adoption. Readers are encouraged to review the new developments in order to be sure their selection is the most appropriate one for the application at hand.

Some conventional and more recent drying techniques are listed in the Table 4.

**Table 4** Conventional versus innovative drying techniques

<table>
<thead>
<tr>
<th>Feed type</th>
<th>Dryer type</th>
<th>New techniques*</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid Suspension</td>
<td>• Drum</td>
<td>• Fluid/spouted beds of inert particles</td>
</tr>
<tr>
<td></td>
<td>• Spray</td>
<td>• Spray/fluid bed combination</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Vacuum belt dryer</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Pulse combustion dryers</td>
</tr>
<tr>
<td>Paste/sludge</td>
<td>• Spray</td>
<td>• Spouted bed of inerts</td>
</tr>
<tr>
<td></td>
<td>• Drum</td>
<td>• Fluid bed (with solid backmixing)</td>
</tr>
<tr>
<td></td>
<td>• Paddle</td>
<td>• Superheated steam dryers</td>
</tr>
<tr>
<td>Particles</td>
<td>• Rotary</td>
<td>• Superheated steam FBD</td>
</tr>
<tr>
<td></td>
<td>• Flash</td>
<td>• Vibrated bed</td>
</tr>
<tr>
<td></td>
<td>• Fluidized bed (hot air or combustion gas)</td>
<td>• Ring dryer</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Pulsated fluid bed</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Jet-zone dryer</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Yamato rotary dryer</td>
</tr>
<tr>
<td>Continuous sheets</td>
<td>• Multi-cylinder contact dryers</td>
<td>• Combined impingement/radiation dryers</td>
</tr>
<tr>
<td>(coated paper, paper, textiles)</td>
<td>• Impingement (air)</td>
<td>• Combined impingement and through dryers</td>
</tr>
<tr>
<td></td>
<td></td>
<td>(textiles, low basis weight paper)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Impingement and MW or RF</td>
</tr>
</tbody>
</table>

*New dryers do not necessarily offer better techno-economic performance for all products
CLOSING REMARKS

It is difficult to generate rules for both classification and selection of dryers because exceptions occur rather frequently. Often, minor changes in feed or product characteristics result in different dryer types being the appropriate choices. It is not uncommon to find different dryer types being used to dry apparently the same material. The choice is dependent on production throughput, flexibility requirements, cost of fuel as well as on the subjective judgment of the individual who specified the equipment.

We have not considered novel dryers in this chapter. Kudra and Mujumdar (2000) have discussed in detail most of the non-conventional and novel drying technologies reported in the literature. Most of them have yet to mature; a few have been commercialized successfully for certain products. It is useful to be aware of such advances so that the user can make intelligent decisions about dryer selection. Since dryer life is typically 25-40 years that effect of a poor “prescription” can have a long-term impact on the economic health of the plant. It is typically not a desirable option to depend exclusively on prior experience, reports in the literature or vendors’ recommendations. Each drying problem deserves its own independent evaluation and solution.

REFERENCES


ME5202 INDUSTRIAL TRANSPORT PROCESSES

Fluidized Bed drying-An Overview

Professor A. S. Mujumdar
ME Department, NUS
E-mail - mpeasm@nus.edu.sg

Contents

• Introduction to fluidized beds - Fundamentals
• Heat transfer in FBs
• Mass transfer in FBs
• An example calculation
• Application to drying
• Modeling a fluid bed dryer
• Key references (given at the end)
Fluidized beds: Particle suspended in an upward gas stream.
Pressure drop in fixed beds

Pressure drop through packed bed / fixed bed of uniform sized solids is correlated by Ergun equation

\[
\frac{\Delta P}{L} = \frac{150 (1 - \varepsilon_m)^2}{\varepsilon_m^3} \frac{\mu u_0}{(\varnothing_s d_p)^2} + 1.75 \frac{(1 - \varepsilon_m)}{\varepsilon_m^3} \frac{\rho_s u_0^2}{(\varnothing_s d_p)}
\]

It has two factors, viscous force and the kinetic energy force.

At low Reynolds number only viscous losses predominate

\[
\frac{\Delta P}{L} = \frac{150 (1 - \varepsilon_m)^2}{\varepsilon_m^3} \frac{\mu u_0}{(\varnothing_s d_p)^2} \quad \text{Re}_p \leq 20
\]

At high Reynolds number only kinetic energy losses need to be considered

\[
\frac{\Delta P}{L} = 1.75 \frac{(1 - \varepsilon_m)}{\varepsilon_m^3} \frac{\rho_s u_0^2}{(\varnothing_s d_p)} \quad \text{Re}_p \geq 1000
\]

Pressure drop & Minimum Fluidization Velocity

At the onset of fluidization, the gravity force on the particles in the bed must be balanced by the drag, buoyancy, and pressure forces.

\[
\text{Drag force by (upward moving) gas} = \text{(Weight of)} \frac{\text{Pressure drop}}{\text{Cross-sectional area of bed}} = \text{(Volume of)} \frac{\text{fraction of}}{\text{solids}} \text{(Specific weight of solids)}
\]
Pressure drop & Minimum Fluidization Velocity

\[\Delta P \cdot A_t = W = (A_t L_{mf}) (1 - \epsilon_{mf}) (\rho_s - \rho_g) g \]
\[\left(\frac{\Delta P}{L_{mf}}\right) = (1 - \epsilon_{mf}) (\rho_s - \rho_g) g\]

The \(u_{mf}\), the superficial velocity at minimum fluidizing condition is found by using expression for \(\Delta p/L\)

\[\frac{\Delta P}{L_{mf}} = \frac{150 (1 - \epsilon_{mf})^2 \mu u_{mf}^3}{(\phi_s \epsilon_{mf}^3 \rho_g)^2} + \frac{1.75 (1 - \epsilon_{mf}) \rho_g u_{mf}^2}{(\phi_s \epsilon_{mf}^3) \mu} = (1 - \epsilon_{mf}) (\rho_s - \rho_g) g\]

In a simplified form

\[\frac{150 (1 - \epsilon_{mf}) (d_p \rho_g u_{mf})}{\phi_s \epsilon_{mf}^3 \mu} + \frac{1.75 (d_p \rho_g u_{mf})}{\phi_s \epsilon_{mf}^3} = \frac{(\rho_s - \rho_g) g d_p^3 \rho_g}{\mu^2}\]

For small particles (\(Re_p < 20\))

\[\frac{150 (1 - \epsilon_{mf}) (d_p \rho_g u_{mf})}{\phi_s \epsilon_{mf}^3 \mu} = \frac{(\rho_s - \rho_g) g d_p^3 \rho_g}{\mu^2}\]

Or

\[u_{mf} = \frac{(\phi_s d_p)^2 (\rho_s - \rho_g) g}{150 \mu} \epsilon_{mf}^3 \]

Wen and Yu have found for variety of systems

\[\frac{1}{\phi_s \epsilon_{mf}^3} \approx 14 \quad \text{and} \quad \frac{1 - \epsilon_{mf}}{\phi_s \epsilon_{mf}^3} \approx 11\]

for small particles the simplified form is

\[u_{mf} = \frac{(d_p)^2 (\rho_s - \rho_g) g}{1650 \mu} \epsilon_{mf}^3 \]

For larger particles (\(Re_p > 20\))

\[u_{mf}^2 = \frac{d_p (\rho_s - \rho_g) g}{24.5 \mu} \epsilon_{mf}^3 \]

For whole range of Reynolds number

\[\frac{d_p \rho_g u_{mf}}{\mu} = \left[ (33.7)^2 + 0.0408 \frac{d_p^3 (\rho_s - \rho_g) g}{\mu} \right]^{1/2} - 33.7\]
**Geldart’s Classification**

<table>
<thead>
<tr>
<th>Group</th>
<th>Characteristics and Properties</th>
</tr>
</thead>
<tbody>
<tr>
<td>A: Aeratable ((U_{mb} &gt; U_{mf})) Material has significant deaeration time (FCC Catalyst)</td>
<td></td>
</tr>
<tr>
<td>B: Bubbles Above ((U_{mb} = U_{mf})) 500 micron sand</td>
<td></td>
</tr>
<tr>
<td>C: Cohesive (Flour, Fly Ash)</td>
<td></td>
</tr>
<tr>
<td>D: Spoutable (wheat, 2000 micron polyethylene pellets)</td>
<td></td>
</tr>
</tbody>
</table>

**Group Characteristics and Properties**

<table>
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<td>A</td>
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<td>B</td>
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<tr>
<td>C</td>
<td></td>
</tr>
<tr>
<td>D</td>
<td></td>
</tr>
</tbody>
</table>

**Geldart’s Classification** contd...

A: Aeratable (\(U_{mb} > U_{mf}\)) Material has significant deaeration time (FCC Catalyst)

- Good fluidization quality, aeratable, easily fluidized, smooth at low velocity and bubbling at higher velocity and slug at high velocity, bed expands, Good solids mixing
- Small mean particle size, low density, typically \(30 < \frac{d_p}{g} < 100 \mu m\) and \(p<1400 \text{ kgm}^{-3}\).

B: Bubbles Above (\(U_{mb} = U_{mf}\)) 500 micron sand

- Good fluidization quality, sand-like particles, vigorous bubbling, slug at high velocity, small bed expansion, good solids mixing in bubbling
- Typically \(40 \mu m < d_p < 500 \mu m\), \(1400 \text{ kgm}^{-3} < p < 4000 \text{ kgm}^{-3}\)

C: Cohesive (Flour, Fly Ash)

- Bad fluidization quality, cohesive due to strong interparticle force, severe slugging and agglomeration, may generates electrostatic charges, poor solids mixing.
- Fine and ultra-fine particles

D: Spoutable (wheat, 2000 micron polyethylene pellets)

- Poor fluidization quality, spoutable, difficult to fluidize in deep bed depth, large bubbles, severe channeling, relatively poor solids mixing
- Large and/or dense particles, typically \(d_p > 500 \mu m\), \(p > 1400 \text{ kgm}^{-3}\)
Geldart’s Classification contd...

Increasing size and density

<table>
<thead>
<tr>
<th>Group</th>
<th>C</th>
<th>A</th>
<th>B</th>
<th>D</th>
</tr>
</thead>
<tbody>
<tr>
<td>Most obvious characteristic</td>
<td>Cohesive, difficult to fluidize</td>
<td>Bubble-free range of fluidization</td>
<td>Start bubbling at ( U_{mf} )</td>
<td>Coarse solids</td>
</tr>
<tr>
<td>Typical solids</td>
<td>Flour, cement</td>
<td>Cracking catalyst</td>
<td>Building sand, table salt</td>
<td>Crushed limestone, coffee beans</td>
</tr>
<tr>
<td>1. Bed expansion</td>
<td>Low when bed channels, can be high when fluidized</td>
<td>High</td>
<td>Moderate</td>
<td>Low</td>
</tr>
<tr>
<td>2. Desorption rate</td>
<td>Initially fast, exponential</td>
<td>Slow, linear</td>
<td>Fast</td>
<td>Fast</td>
</tr>
<tr>
<td>3. Bubble properties</td>
<td>No bubbles, Channels, and cracks</td>
<td>Splitting, recoalescence predominates; large wake</td>
<td>No limit on size</td>
<td>No known upper size; small wake</td>
</tr>
<tr>
<td>4. Solids mixing ( a )</td>
<td>Very low</td>
<td>High</td>
<td>Moderate</td>
<td>Low</td>
</tr>
<tr>
<td>5. Gas backming ( a )</td>
<td>Very low</td>
<td>High</td>
<td>Moderate</td>
<td>Low</td>
</tr>
<tr>
<td>6. Slug properties</td>
<td>Solid slugs</td>
<td>Axisymmetric</td>
<td>Axisymmetric, asymmetric</td>
<td>Vertical, solid slugs, wall slugs</td>
</tr>
<tr>
<td>7. Spouting</td>
<td>No</td>
<td>No, except in very shallow beds</td>
<td>Shallow beds only</td>
<td>Yes, even in deep beds</td>
</tr>
</tbody>
</table>

Effect on properties 1 to 7 of:
- Mean particle size within group
- Cohesiveness increases as \( d_p \) decreases
- Properties improve as size decreases
- Properties improve as size decreases
- Not known
- Particle size distribution \( a \)
- Not known
- Increasing <45 \( d_p \) fraction improves properties
- None
- Increases segregation
- Increasing pressure, temperature, viscosity, density of gas
- Probably improves
- Definitely improves
- Uncertain, some possibly
- Uncertain, some possibly

Geldart’s Classification contd...

C
Cohesive
0-30 µm
flour

Bubbling
100-1000 µm
sand

A
Aeratable
30-100 µm
milk powder

D
Spoutable
>1000 µm
coffee beans

Drag
Attraction
Gravity
Fluidization Regimes

**Velocity range** | **Regime** | **Fluidization Features and Appearance**
--- | --- | ---
0 ≤ U ≤ U_{mf} | Fixed Bed | Particles are quiescent; gas flows through interstices
U_{mf} ≤ U ≤ U_{mb} | Particulate Regime | Bed expands smoothly and homogeneously with small-scale particle motion; bed surface is well defined
U_{mb} ≤ U ≤ U_{ms} | Bubbling Regime | Gas bubbles form above distributor, coalesce and grow; gas bubbles promote solids mixing during rise to surface and breakthrough
U_{ms} ≤ U ≤ U_{C} | Slug flow Regime | Bubble size approaches bed cross section; bed surface rises and falls with regular frequency with corresponding pressure fluctuation
U_{C} ≤ U ≤ U_{k} | Transition to turbulent Fluidization | Pressure fluctuation decrease gradually until turbulent fluidization regime is reached
U_{k} ≤ U ≤ U_{tr} | Turbulent Regime | Small gas voids and particle clusters and streamers dart to and fro; bed surface is diffused and difficult to distinguish
U ≥ U_{tr} | Fast Fluidization | Particles are transported out of the bed and need to be replaced and recycled; normally has a dense phase region at bottom coexisting with a dilute phase region on top; no bed surface
U ≫ U_{tr} | Pneumatic Conveying | Usually a once-through operation; all particles fed are transported out in dilute phase with concentration varying along the column height; no bed surface
Fluidization Regimes: Description

Transition between regimes

Equations have been published for transition lines between various regimes. This map can be used to identify the type of flow regime that will exist for a given particle under specific flow conditions.

\[ (Fr_{mf})(Re_{pf,mf}) \left( \frac{\rho_s - \rho_g}{\rho_g} \right) \left( \frac{L_{mf}}{d_c} \right) < 100 \]

**Smooth or particulate fluidization**

\[ (Fr_{mf})(Re_{pf,mf}) \left( \frac{\rho_s - \rho_g}{\rho_g} \right) \left( \frac{L_{mf}}{d_c} \right) > 100 \]

**Bubbling or aggregative fluidization**

**Slugging criteria**

\[ (U - u_{mf}) > 0.07 \sqrt{gD} + 1.6 \times 10^{-3} (60 D^{0.175} - H_{mf})^2 \]

Design of distributor in fluidized beds

- Uniform gas sparging is govern by the effective design of gas distribution system and is very important to have uniform heat transfer.

- Parameter affecting performance of Gas distribution system comprise of:
  - Gas Sparger geometry
  - Gas chamber geometry
  - Pressure drop across the gas distribution system

- A good distributor should;
  - Obtain a spatially uniform gas distribution, without stagnant zones
  - Prevent solids loss by leakage
  - Minimize solid erosion
  - Avoid choking of the distributor
  - Have a definite and non-changing (with time) pressure drop for the gas
Design of distributor in fluidized beds

Pressure Drop across the Distributor

The pressure drop across the distributor $\Delta P_d$ is used as the criterion for design, and $\Delta P_d$, values varying from 0.01 to 1.0 times the pressure drop across the bed $\Delta P_b$ have been suggested.

Siegel (1986):

$$\frac{\Delta P_d}{\Delta P_b} = 0.14 - 0.22 \text{ [Galileo number (1 - 10,000)]}$$

Kunii & Levenspiel (1991):

$$\frac{\Delta P_d}{\Delta P_b} = 0.1 – 0.3$$


Nozzle Position (Litz, 1972)

Side entry:

$$H = 0.2D + 0.5D_{noz} \quad , \text{when} \quad D_{noz} > D/100$$

$$H = 18D_{noz} \quad , \text{when} \quad D_{noz} < D/100$$

Bottom entry:

$$H = 3(D - D_{noz}) \quad , \text{when} \quad D_{noz} > D/36$$

$$H = 100D_{noz} \quad , \text{when} \quad D_{noz} < D/36$$

Where,

- $D$ - diameter of gas distribution chamber
- $D_{noz}$ - diameter of the nozzle
- $H$ - distance between the nozzle centerline and the distributor plate

A packed bed is composed of cubes 0.02 m on a side. The bulk density of the packed bed, with air, is 980 kg/m³. The density of the solid cubes is 1500 kg/m³.

- Calculate the void fraction (\( \varepsilon \)) of the bed.
- Calculate the effective diameter (\( D_p \)) where \( D_p \) is the diameter of a sphere having the equivalent volume.
- Determine the sphericity of the cubes.
- Estimate the minimum fluidization velocity using water at 38 C and a tower diameter of 0.15 m.

**Void Fraction**

*We know:* \( V_{\text{bed}} = V_{\text{fluid}} + V_{\text{solids}} \) and \( W_{\text{bed}} = W_{\text{fluid}} + W_{\text{solids}} \)

\[
\rho_{\text{bed}}V_{\text{bed}} = \rho_{\text{fluid}}V_{\text{fluid}} + \rho_{\text{solids}}V_{\text{solids}}
\]

\[
\rho_{\text{solids}}V_{\text{solids}} \gg \rho_{\text{fluid}}V_{\text{fluid}}
\]
\[ \therefore \rho_{\text{bed}} V_{\text{bed}} \equiv \rho_{\text{solids}} V_{\text{solids}} \quad \text{and} \quad \varepsilon = \frac{V_{\text{bed}} - \rho_{\text{bed}} V_{\text{bed}}}{\rho_{\text{solids}} V_{\text{solids}}} \]

\[ \varepsilon = 1 - \frac{\rho_{\text{bed}}}{\rho_{\text{solids}}} = 1 - \frac{980 \frac{\text{kg}}{\text{m}^3}}{1500 \frac{\text{kg}}{\text{m}^3}} = 0.35 \]

Effective diameter
\[ a^3 = \frac{\pi}{6} D_p^3 \]
\[ (0.02)^3 = \frac{\pi}{6} D_p^3 \quad \therefore D_p = 0.025 \text{m} \]

Minimum Fluidization Velocity
\[ (p - \rho_f) g = \rho_f u_{mf}^2 \left[ \frac{150 (1 - \varepsilon_{mf}) u}{\Phi_s D_p \varepsilon_{mf}^3} \right] + 1.75 \]
\[ \left( 1500 \frac{\text{kg}}{\text{m}^3} - 994 \frac{\text{kg}}{\text{m}^3} \right) \cdot 9.80 \frac{\text{m}}{\text{s}^2} = 4959 \frac{\text{kg}}{\text{m}^2 \text{s}^2} \]

Sphericity
\[ \Phi_s = \left( \frac{6}{\pi} \right)^{1/3} \frac{a}{\frac{1}{6} V^3} = \left( \frac{\pi}{6} \right)^{1/3} = 0.81 \]

\[ \Phi_s \varepsilon_{mf}^3 = \frac{1}{14} \quad \therefore \varepsilon_{mf} = 0.445 \]

\[ \frac{1.75 \cdot \rho_f u_{mf}^2}{\Phi_s D_p \varepsilon_{mf}^3} = \frac{1.75 \cdot 994 \frac{\text{kg}}{\text{m}}}{0.81 \cdot 0.025 \cdot (0.445)^3} = 9.748 \times 10^5 \frac{\text{kg}}{\text{m}^2 \text{s}^2} \cdot u_{mf}^2 \]

\[ \frac{150 \cdot (1 - \varepsilon_{mf}) u \cdot u_{mf}}{\Phi_s^2 D_p^2 \varepsilon_{mf}^3} = \frac{150 \cdot (1 - 0.445) \cdot (0.693 \text{ cm}) \cdot (0.001 \frac{\text{kg}}{\text{m} \text{s}}) \cdot u_{mf}}{(0.81)^3 \cdot (0.025 \text{ m})^3 \cdot (0.445)^3} \]

\[ = 1597 \frac{\text{kg}}{\text{m}^2 \text{s}^2} \cdot u_{mf} \]

\[ 0 = 9.748 \times 10^5 \frac{\text{kg}}{\text{m}^2 \text{s}^2} \cdot u_{mf}^2 + 1597 \frac{\text{kg}}{\text{m}^2 \text{s}^2} \cdot u_{mf} - 4959 \frac{\text{kg}}{\text{m}^2 \text{s}^2} \]

\[ u_{mf} = 0.071 \frac{\text{m}}{\text{s}} \]
Heat Transfer in fixed and Fluidized beds

- Main advantage of fluidized beds is the extremely large area of solid surface exposed to the fluidizing media.

- High solid surface area greatly facilitates solid-to-gas heat transfer.

- Because of the solids mixing generated within the bulk of a bubbling gas fluidized bed, temperature gradients are reduced to negligible proportions.

- High rates of heat transfer are obtainable between the fluidizing solids and the immersed transfer surface.

Particle-to-gas heat transfer

Bed-to-surface heat transfer

Use of immersed surfaces
Heat Transfer in Fixed Beds

Heat transfer in fixed bed consists of following mechanisms:

1. Conduction heat transfer between particles,
2. Convective heat transfer between particles and fluid,
3. Interaction of both (1) and (2),
4. Radiative heat transfer between particles and the flowing gas,
5. Heat transfer between bed wall and bed particles.

1) Particle to fluid heat transfer
   Heat transfer to single particle can be expressed as:
   \[ q_p = h_{sp} \alpha_p (T_s - T) \]
   Heat transfer coefficient can be evaluated as:
   \[ \frac{h_{sp} \alpha_p}{k_g} = 2 + 1.1 Pr^{0.33} (Re_p)^{0.6} \]

2) Heat transfer through wall (one dimensional model)
   \[ q = h_{w}A_t (T - T_w) \]
   For homogeneous model, temperature of fluid and of bed are assumed identical:
   \[ \frac{h_{w} \alpha_p}{k_g} = 2.26 (Re_p)^{0.6} Pr^{0.33} \exp \left( -\frac{6 \alpha_p}{D} \right) \]
   for \( 20 \leq Re \leq 7600 \) and \( 0.05 \leq \frac{d_p}{D} \leq 0.3 \)

3) Heat transfer through wall (two dimensional model)
   \[ q = h_{w}A_t (T_R - T_w) \]
   For homogeneous model, temperature of fluid and of bed are assumed identical:
   \[ \frac{h_{w} \alpha_p}{k_g} = 0.91 (Re_p)^{0.79} Pr^{0.33} \]
   for \( 20 \leq Re \leq 7600 \) and \( 0.05 \leq \frac{d_p}{D} \leq 0.3 \)

4) Effective radial thermal conductivity
   \[ \frac{q}{A_t} = k_g \frac{\partial T}{\partial r} |_{r=R} \]
   \[ \frac{h_{w} \alpha_p \varepsilon}{k_g (1 - \varepsilon)} = 0.27 \]
   for \( 500 \leq \frac{U d_p \rho_g}{\mu_g (1 - \varepsilon)} \leq 6000 \) and \( 0.05 \leq \frac{d_p}{D} \leq 0.15 \)
Heat Transfer in Fluid beds

Heat transfer in a bubbling fluidized bed

- Gas to particle heat transfer coefficients are typically small, of the order of 5 - 20 W/m²K

- However, because of the very large heat transfer surface area provided by a mass of small particles, the heat transfer between gas and particles is rarely limiting in fluid bed heat transfer

- One of the most commonly used correlations for gas-particle heat transfer coefficient is that of Kunii and Levenspiel (only for low particle Reynolds numbers)
  \[ \text{Nu}_{gp} = 0.03 \text{Re}_{p}^{1.3} \]
  
  for \( \text{Re}_{p} \leq 50 \)

- Gas to particle heat transfer is relevant where a hot fluidized bed is fluidized by cold gas

- While following are the correlation suggested based on experimental data*
  \[ \text{Nu}_{gp} = 0.0282 \text{Re}_{p}^{1.4} \text{Pr}_{g}^{0.33} \quad \text{for} \quad 0.1 \leq \text{Re}_{p} \leq 50 \]
  \[ \text{Nu}_{gp} = 1.01 \text{Re}_{p}^{0.48} \text{Pr}_{g}^{0.33} \quad \text{for} \quad 50 \leq \text{Re}_{p} \leq 10^4 \]

*Chen, J.C., Heat Transfer in handbook of fluidization and fluid systems, 2003

Analysis of Gas-particle heat transfer

The energy balance across the element gives

\[ -(C_{pg} U \rho_g) dT_g = h_{gp} a (T_g - T_p) dL \]

Integrating with boundary conditions \( T_g = T_{g0} \) at \( L = 0 \)

\[ \ln \left( \frac{T_g - T_p}{T_{g0} - T_p} \right) = - \left( \frac{h_{gp} a}{C_{pg} U \rho_g} \right) L \]

\[ \text{Nu}_{gp} = \frac{h_{gp} d_p}{k_g} = 0.03 \text{Re}_{p}^{1.3} = 0.03 \left( \frac{U d_p \rho_g}{\mu_g} \right)^{1.3} \]
Analysis of Gas-particle heat transfer

The distance $L_n$, in which the gas-to-particle temperature falls by a factor $n$, is given by:

$$n = \left(\frac{T_{g0} - T_p}{T_g - T_p}\right)$$

The distance over which the temperature distance is reduced to half its initial value, $L_{0.5}$, is then

$$L_{0.5} = \frac{5.5 \ln(0.5) \mu_g^{1.3} d_p^{0.7} C_{pg}}{\rho_g^{0.3} U^{0.3} k_g (1 - \varepsilon)}$$

Analysis of Gas-particle heat transfer

A bed of 450μm particles is operating at 150°C. The temperature and superficial velocity of the incoming gas are 550°C and 0.4 m/s, respectively. Approximately how far will the incoming gas have penetrated into the bed before it is cooled to 350°C?

Gas physical properties can be estimated at average temperature over the specified range. In this particular case the average temperature is 450°C. Hence the physical properties of air:

- $\rho_g = 0.68 \text{ kg/m}^3$
- $k_g = 0.04 \text{ W/m K}$
- $C_{pg} = 1025 \text{ J/kg K}$
- $\mu_g = 2.8 \times 10^{-5} \text{ N s/m}^2$

Gas velocity at 450°C for an inlet velocity of 0.4m/s at 550°C will be 0.35m/s

$$n = \left(\frac{550 - 150}{350 - 150}\right) = 2$$

$$L_2 = \frac{5.5 \ln(2) (2.8 \times 10^{-5})^{1.3}(450 \times 10^{-6})^{0.7} 1025}{(0.68)^{0.3}(0.35)^{0.3} \times 0.04 (1 - 0.4)} \sim 1.4 \text{ mm}$$
Bed to surface heat transfer

In a bubbling fluidized bed the coefficient of heat transfer between bed and immersed surfaces (vertical bed walls or tubes) can be considered to be made up of three additive components.

The particle convective component $h_{cp}$, which is dependent upon heat transfer through particle exchange between the bulk of the bed and the region adjacent to the transfer surface (heat transfer due to the motion of packets of solids carrying heat to and from the surface).

The interphase gas convective component $h_{gc}$, by which heat transfer between particle and surface is augmented by interphase gas convective heat transfer.

The radiant component of heat transfer $h_r$.

Thus,

$$h = h_{pc} + h_{gc} + h_r$$

Approximate range of significance: 40µm → 1mm > 800 µm and at higher static pressure Higher temperatures (> 900 K) and difference

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Particle convective heat transfer

On a volumetric basis the solids in the fluidized bed have about one thousand times the heat capacity of the gas and so, since the solids are continuously circulating within the bed, they transport the heat around the bed. For heat transfer between the bed and a surface the limiting factor is the gas conductivity, since all the heat must be transferred through a gas film between the particles and the surface.

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Heat transfer from bed particles to an immersed surface
Heat Transfer in Fluid beds

Particle convective heat transfer

The particle-to-surface contact area is too small to allow significant heat transfer. Factors affecting the gas film thickness or the gas conductivity will therefore influence the heat transfer under particle convective conditions.

Decreasing particle size, for example, decreases the mean gas film thickness and so improves $h_{pc}$. However, reducing particle size into the Group C range will reduce particle mobility and so reduce particle convective heat transfer. Increasing gas temperature increases gas conductivity and so improves $h_{pc}$.

Particle convective heat transfer is dominant in Group A and B powders. Increasing gas velocity beyond minimum fluidization improves particle circulation and so increases particle convective heat transfer.

The heat transfer coefficient increases with fluidizing velocity up to a broad maximum $h_{max}$ and then declines as the heat transfer surface becomes blanketed by bubbles.

Bed to surface heat transfer

![Diagram of bed-to-surface heat transfer coefficients]

Range of fluidized bed-to-surface heat transfer coefficients
Heat Transfer in Fluid beds

Bed to surface heat transfer

Effect of fluidizing gas velocity on bed – surface heat transfer coefficient

Heat Transfer in Fluid beds

Particle convective heat transfer

The maximum in $h_{pc}$ occurs relatively closer to $U_{mf}$ for Group B and D powders since these powders give rise to bubbles at $U_{mf}$ and the size of these bubbles increase with increasing gas velocity.

Group A powders exhibit a non-bubbling fluidization between $U_{mf}$ and $U_{mb}$ and achieve a maximum stable bubble size.

Zabrodsky (1966) has given correlation for $h_{\text{max}}$ for group B particles

$$h_{\text{max}} = 35.8 \frac{k_g^{0.6} \mu_g^{0.2}}{d_p^{0.26}}$$

Khan (1978) has given correlation for $h_{\text{max}}$ for group A particles

$$Nu_{\text{max}} = 0.157 \text{ Ar}^{0.475}$$
Gas convective heat transfer

Gas convective heat transfer is not important in Group A and B powders where the flow of interstitial gas is laminar but becomes significant in Group D powders, which fluidize at higher velocities and give rise to transitional or turbulent flow of interstitial gas.

In gas convective heat transfer the gas specific heat capacity is important as the gas transports the heat around.

Gas specific heat capacity increases with increasing pressure and in conditions where gas convective heat transfer is dominant, increasing operating pressure gives rise to an improved heat transfer coefficient $h_{gc}$.

Baskakov and Suprun (1972) has given correlation for $h_{gc}$:

$$N_{u_{gc}} = 0.0175 \, A_r^{0.46} \, Pr^{0.33} \, [\text{for } U > U_m]$$

$$N_{u_{gc}} = 0.0175 \, A_r^{0.46} \, Pr^{0.33} \, \left(\frac{U}{U_m}\right)^{0.3} \, [\text{for } U_m < U < U_m]$$

where $U_m$ is the superficial velocity corresponding to the maximum overall bed heat transfer coefficient.

Gas convective heat transfer to immersed surfaces

Several approaches have been used to estimate $h_c$.

The most common approach assigns thermal resistance to a gaseous boundary layer at the heat transfer surface, the enhancement of heat transfer is then attributed to the scouring action of the solid particles on the gas film, decreasing the effective film thickness.

Lava’s correlation (1952) for vertical surfaces, for larger particles, is

$$N_{u_c} = \frac{h_c d_p}{k_g} = 0.525 \, \text{Re}_p^{0.75}$$

Wender and Cooper’s correlation (1958) for vertical tubes,

$$\frac{h_c d_p}{k_g} = 3.51 \times 10^{-4} C_R (1 - \epsilon) \text{Re}_p^{0.23} \left(\frac{C_{pg} \rho_g}{k_g}\right)^{0.43} \left(\frac{C_{ps}}{C_{pg}}\right)^{0.8} \left(\frac{\rho_s}{\rho_g}\right)^{0.66}$$

for $10^{-2} \leq \text{Re}_p \leq 10^2$

Where

$$C_R = 1.07 + 3.04 \left(\frac{r}{R_b}\right) - 3.29 \left(\frac{r}{R_b}\right)^2$$

Where $r$ is the radial position of the heat transfer tube and $R_b$ is the radius of the bed.
Gas convective heat transfer to immersed surfaces

Vreedenberg's correlation (1958) for horizontal tubes,

\[
\frac{h_cD_t}{k_g} = 420 \left( \frac{\rho_s}{\rho_g} \frac{Pr_g}{Pr} \frac{\mu_g^2}{g \rho_g^2 \delta_p^4} \right)^{0.3} Re_B^{0.3}
\]

for \( \frac{\rho_s}{\rho_g} Re_p \geq 2550 \)

and

\[
\frac{h_cD_t}{k_g} = 0.66 Pr_g^{0.3} \left( \frac{\rho_s (1 - \epsilon)}{\rho_g \epsilon} \right)^{0.44} Re_B^{0.44}
\]

for \( \frac{\rho_s}{\rho_g} Re_p \leq 2550 \)

Where \( Re_B = \frac{\mu_g d_t \rho_g}{U_g} \)

Radiative heat transfer

For temperatures beyond 600°C radiative heat transfer plays an increasing role and must be accounted for in calculations.

For rule of thumb estimate, the radiative heat transfer component can be estimated using absolute temperatures and an adaptation of the stefan-Boltzman equation in the form

\[
h_r = \frac{5.67 \times 10^{-8} \varepsilon_r (T_b^4 - T_s^4)}{T_b - T_s}
\]

where \( \varepsilon_r \) is the reduced emissivity to take into account the different emissivity properties of surface \( \varepsilon_s \) and bed \( \varepsilon_b \) and is given by

\[
\varepsilon_r = \frac{1}{\left( \frac{1}{\varepsilon_s} + \frac{1}{\varepsilon_b} \right) - 1}
\]

An alternative correlation given by Panov et al. (1978) for approximate estimate is

\[
h_r = 7.3 \sigma \ v_p \ e_s \ T_s^3
\]
Mass transfer

- Main applications of fluidized beds – Fluidized catalytic cracking; Combustion and Gasification; Drying; Granulation;

- The performance of fluidized beds in all the above processes is affected by the interface mass transfer

- Many times the mass transfer in fluidized beds is a potential rate controlling step in fluidized bed reactors

- Depending on the type gas-solid interactions, the rate controlling mechanism can be

1. Particle – gas mass transfer (Gas film diffusion) control
2. Pore diffusion control
3. Surface phenomenon control
Mass transfer in fluidized beds

There are two approaches that can be used for prediction of mass transfer rates in fluidized beds

• **Homogeneous bed approach** – considers fluidized bed behaving as a fixed bed reactor and correlate the mass transfer coefficient in fluidized bed in similar manner to that in a fixed bed based on plug-flow model

• **Bubbling bed approach** – Considers fluidized bed to consist of two phases, a bubble and an emulsion phase, the gas interchange between the two phases constitutes the rate of mass transfer

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**Useful Dimensionless Numbers**

**Sherwood Number** – ratio of convective to diffusive mass transport

\[
Sh = \frac{\text{Convective mass transfer coefficient}}{\text{Diffusive mass transfer coefficient}} \quad \text{or} \quad Sh = \frac{kL}{D}
\]

**Schmidt Number** – ratio of momentum to mass diffusivity

\[
Sc = \frac{\text{Momentum diffusivity}}{\text{Mass Diffusivity}} \quad \text{or} \quad Sc = \frac{\mu}{\rho D}
\]

**Reynolds Number** – Comparison of Inertial force to viscous force

\[
Re = \frac{\text{Inertial force}}{\text{Viscous force}} \quad \text{or} \quad Re_p = \frac{U_d \rho_g}{\mu_g}
\]

**Reynolds Number** – Comparison of gravitational and viscous force

\[
Ar = \frac{\text{Gravitational force}}{\text{Viscous force}} \quad \text{or} \quad Ar = \frac{g \rho_g (\rho_s - \rho_g)}{\mu_g^2} d_p^3
\]
Homogeneous Bed Approach

Transfer between single sphere and surrounding gas

- Rate of mass transfer between well-dispersed sphere and surrounding air can be written as

\[
\frac{dN_A}{dt} = k_{g,\text{single}} s_{ex,\text{single}} \left( C_A^2 - C_A \right)
\]

Transfer rate of A from particle to gas

Mass transfer coefficient of single particle

Concentration of A at particle-gas interphase

Concentration of A in the bulk gas stream

- The single particle mass transfer coefficient can be obtained from well established correlation (Froessling, 1938)

\[ Sh_{\text{single}} = \frac{k_{g,\text{single}} d_{p,\text{eff}} y}{D} = 2 + 0.6 (Re_{\text{ph}})^{0.5} (Sc)^{0.333} \]

Eq (2)

where particle Reynolds number and Schmidt number can be defined as

\[ Re_{\text{ph}} = \frac{\rho_g u d_p}{\mu} \quad \text{and} \quad Sc = \frac{\mu}{\rho_g D} \]

The mass transfer coefficient is proportional to diffusion coefficient of gas and inversely proportional to diameter of particle

Homogeneous Bed Approach

Transfer between single sphere and surrounding gas

For non-spherical particle the sieve diameter \((d_p)\) will replace the diameter of the particle and the equation becomes

\[ Sh_{\text{single}} = \frac{k_{g,\text{single}} d_p y}{D} = 2 + 0.6 (Re_p)^{0.5} (Sc)^{0.333} \]

where particle Reynolds number is

\[ Re_p = \frac{\rho_g u d_p}{\mu} \]
Homogeneous Bed Approach

Transfer between fixed bed particles and flowing gas

Rate of mass transfer between fixed bed of particles and surrounding gas can be written in the same fashion as the one for single particle

\[
\frac{dN_A}{dt} = k_{g,\text{bed}}S_{\text{ex,particles}}(C_A^i - C_A)
\]

- Combined mass transfer rate from all the particles
- Average mass transfer coefficient of particles
- Total exterior surface of all individual particles
- Concentration of A at particle-gas interphase
- Concentration of A in the bulk gas stream

The correlation for the average mass transfer coefficient \( k_{g,\text{bed}} \) was given by Ranz (1952)

\[
Sh_{\text{bed}} = \frac{k_{g,\text{bed}}d_p}{D} = 2 + 1.8(Re_p)^{0.5}(Sc)^{0.333}
\]

for \( Re_p > 80 \)

Total particle exterior surfaces \( S_{\text{ex,particles}} \)

Method of Kunni Levenspiel

\[
d' = (\text{surface of a particle/volume of a particle}) = \frac{\pi d_{\text{sph}}^2 \phi_s}{\pi d_{\text{sph}}^3/6} = \frac{6}{d_{\text{sph}} \phi_s}
\]

Hence the equation for total exterior particle surfaces becomes

\[
S_{\text{ex,particles}} = V_{\text{seg}} \frac{6(1 - \epsilon_{\text{bed}})}{\phi_s d_{\text{sph}}}
\]
Homogeneous Bed Approach

Transfer between fluidized bed particles and fluidizing gas

The rate of particle to gas mass transfer in the differential segment of fluidized bed can be written as

\[
\frac{dN_A}{dt} = k_{g,\text{bed}} S_{\text{ex}} \frac{C_A^i - C_A}{t}
\]

- Average mass transfer coefficient associated with fluidizing particles
- Total exterior surfaces of the fluidized particles in the segment of bed

\( k_{g,\text{bed}} \) for fluidized beds is always higher than that for fixed beds

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Homogeneous Bed Approach

Average mass transfer coefficient for fluidizing particles

The mass transfer coefficient for fluidized bed particles can be lower or higher than that of single particles. Generally for low Reynolds numbers the mass transfer coefficient for single particles is higher than for fluidizing particles; however, reverse is true for higher particle Reynolds numbers (>80).

Resnick and White (1949) reported the average mass transfer coefficient for the fluidizing particles (for air system with \( Sc = 2.35 \)):

1. For particles with size between mesh 14 and 20 (\( d_p = 1000 \mu m \)),
   \[
   Sh_{\text{bed}} = 0.200 \, Re_p^{0.937} \quad \text{for} \quad 30 < Re_p < 90
   \]

2. For particles with size between mesh 20 and 28 (\( d_p = 711 \mu m \)),
   \[
   Sh_{\text{bed}} = 0.274 \, Re_p^{0.709} \quad \text{for} \quad 15 < Re_p < 80
   \]
Homogeneous Bed Approach

Average mass transfer coefficient for fluidizing particles

3. For particles with size between mesh 28 and 35 ($d_p = 570 \mu m$),
   \[ Sh_{bed} = 0.773 \, Re_p^{1.107} \quad \text{for } 8 < Re_p < 60 \]

4. For particles with size between mesh 35 and 48 ($d_p = 410 \mu m$),
   \[ Sh_{bed} = 0.071 \, Re_p^{0.926} \quad \text{for } 6 < Re_p < 40 \]

5. For particles with size between mesh 48 and 65 ($d_p = 275 \mu m$),
   \[ Sh_{bed} = 0.041 \, Re_p^{1.036} \quad \text{for } 4 < Re_p < 15 \]

Homogeneous Bed Approach

Particle – gas mass transfer coefficient for fluidized bed particles

- It should be noted that the same correlations can be used for fluidized beds and fixed beds...
- “Sh” for same group of particles continues to increase with the particle Reynolds number even during the transition from fixed bed to fluidized bed operations

Comparison of mass transfer coefficient for fluidized bed and fixed bed

Using Sh equation for single particle Eq (2)
Homogeneous Bed Approach

Limitations of Homogeneous bed approach

The experimentally measured coefficient values for the bed particles under fixed bed or fluidized bed conditions can be lower or higher than the theoretically estimated values.

For fine particles the mass transfer coefficients were found to be well below the estimated values from correlations.

The measured mass transfer coefficient by this approach should be treated as empirical in nature.

Using Sh equation for single particle Eq (2)

Summary of particle-gas mass transfer coefficient

Bubbling Bed Approach

Takes in to account existence of a two phases: bubble phase and emulsion phase

Bubble phase – considered as spherical bubbles surrounded by spherical clouds

There are three different models available based on this approach

- Kunni and Levenspiel (bubble – emulsion transfer)
- Partridge and Rowe (cloud – emulsion transfer)
- Chavarie and Grace (Empirical correlation for bubble- emulsion transfer)
Model of Kunni and Levenspiel

Model for vaporization or sublimation of A from all particles in bed

Assumptions:
- Fresh gas enters the bed only as bubbles
- Equilibrium is established rapidly between $C_A$ at gas-particle interphase and its surroundings

These assumptions lead to the following equation in terms of bubble-emulsion mass transfer coefficient ($K_{GB}$) as follows:

$$\frac{dN_A}{dt} = u_b V_{bubble} \left( \frac{dC_{A,b}}{dz} \right) = K_{GB} V_{bubble} (C_{A,c} - C_{A,b})$$

In this approach both cloud and the emulsion phases are assumed perfectly mixed.

Model of Kunni and Levenspiel

Relation between $K_{GB}$ and $K_{g,bed}$

The equation reported in last slide for mass transfer by Kunni and Levenspiel can be derived using the mass transfer rate equation for homogeneous bed approach:

$$\frac{dN_A}{dt} = k_{g,bed} S_{ex,particles} (C_A^i - C_A)$$

With the assumption that fresh gas enters the bed only as bubbles the above equation becomes:

$$\frac{dN_A}{dt} = k_{g,bed} S_{ex,particles} (C_A^i - C_{A,b})$$

Where $C_{A,b}$ is the concentration of A in bubble phase.
Model of Kunni and Levenspiel

Relation between $K_{GB}$ and $k_{g,bed}$

The solutions of the previous equations finally result in the following relation between $K_{GB}$ and $k_{g,bed}$

$$k_{g,bed} = \frac{\delta}{(1 - \varepsilon_f) a'} K_{GB}$$

The definition of Sherwood's Number becomes

$$Sh_{bed} = \frac{k_{g,bed} d_p y}{D} = \frac{\delta}{(1 - \varepsilon_f) a'} K_{GB} d_p y$$

$$= \frac{y \phi_s d_p^2 \delta}{6D(1 - \varepsilon_f)} K_{GB}$$

$K_{GB}$ for Nonporous and Nonadsorbing Particles

The particles dispersed in bubble phase will not contribute to any additional mass transfer and hence; hence transfer across the bubble-cloud boundary therefore is the only source of mass transfer.

Hence

$$K_{GB} = K_{bc}$$

Where $K_{bc}$ is the bubble-cloud interchange coefficient derived by Davidson and Harrison as

$$K_{bc} = 4.5 \frac{U_{mf}}{d_b} + 5.85 \frac{D^{0.5} \varphi^{0.25}}{d_b^{1.25}}$$

$K_{GB}$ for highly adsorbing particles

$$K_{GB} = \gamma_b a' k_{g, single} + K_{bc} = \gamma_b \frac{6}{d_p \phi_s} k_{g, single} + K_{bc}$$

For single particle

$$k_{g, single} = \frac{Sh_{single} D}{d_{n,y}}$$

So the $K_{GB}$ equation becomes

$$K_{GB} = \gamma_b \left[ \frac{6}{(d_p \phi_s)} \frac{Sh_{single} D}{d_p y} \right] + K_{bc}$$

$$= \gamma_b \frac{6 Sh_{single} D}{\phi_s d_p^2 y} + K_{bc}$$
**$K_{GB}$ for highly adsorbing particles**

For highly adsorbing particles, both particles dispersed in bubble and the bubble-cloud gas interchange can contribute to the particle-gas mass transfer and the expression for $K_{GB}$ becomes

\[
K_{GB} = \gamma_{b}a'k_{g,\text{single}} + K_{bc} = \gamma_{b}\frac{6}{d_{p}\phi_{s}}k_{g,\text{single}} + K_{bc}
\]

For single particle

So the $K_{GB}$ equation becomes

\[
K_{GB} = \gamma_{b}\left[\frac{6}{(d_{p}\phi_{s})}\frac{Sh_{\text{single}}D}{d_{p}y}\right] + K_{bc}
\]

Hence the Sherwood number for bed becomes

\[
Sh_{\text{bed}} = \gamma_{b}\phi_{s}d_{p}^{2}\delta \frac{6Sh_{\text{single}}D}{6D(1-\varepsilon_{f})} + K_{bc}
\]

\[
= \frac{\delta}{(1-\varepsilon_{f})}\left\{\gamma_{b}Sh_{\text{single}} + \phi_{s}d_{p}^{2}y\frac{6D}{6D}K_{bc}\right\}
\]

---

**$K_{GB}$ for Porous or Partially Adsorbing Particles**

For porous or partially adsorbing particles, Kunni and Levenspiel derived the following equation

\[
Sh_{\text{bed}} = \frac{\delta}{(1-\varepsilon_{f})}\left[\gamma_{b}Sh_{\text{single}}\eta_{d} + \phi_{s}d_{p}^{2}yK_{bc}\right]
\]

Where,

\[
\eta_{d} = \frac{1}{1+\alpha/m}
\]

And

\[
\alpha = \frac{6k_{g,\text{single}}t_{p,\text{mean}}}{\phi_{s}d_{p}^{2}}
\]

$m$ is the adsorption equilibrium constant defined as

\[
C_{A_{s}} = mC_{A}^{2}
\]

$C_{A_{s}}$ is the concentration of tracer A within the particle in equilibrium with the concentration $C_{A}$ of tracer gas at the gas-particle interphase.
The cloud surrounding the bubble is considered as the primary mass transfer boundary and the bubble and the cloud phases are considered as perfectly mixed single phase as shown in the following figure.

The mass transfer equation is

\[
\frac{dN_A}{dt} = V_{\text{cloud}} \left( \frac{dC_{A,e}}{dt} \right) = k_{e,c} S_{e,\text{cloud}} (C_{A,e} - C_{A,c})
\]

Partidge and Rowe have given a correlation for mass transfer coefficient in terms of Sherwood number as follows

\[
Sh_c = \frac{k_{e,c} d_c}{D} = 2 + 0.69 Sc^{0.33} Re_c^{0.5}
\]

where $Sc$ is the Schmidt number as defined earlier while $Re_c$ is defined as

\[
Re_c = \frac{\rho g d_c \rho_e}{\mu}
\]
Measurement of mass transfer rates for bubble containing ozone injected into an air-fluidized two dimensional bed and proposed following empirical equation,

\[
\frac{dN_A}{dt} = V_{\text{bubble}} \left( \frac{d(C_{A,b})}{dt} \right) = k_{gt} S_{ex,bubble} (C_{A,e} - C_{A,b})
\]

Table shows the models for \( k_{gc} \) given by various researchers

<table>
<thead>
<tr>
<th>Model</th>
<th>Reference</th>
<th>Coefficient</th>
<th>( k_{gt} ) equivalent</th>
<th>Calculated ( k_{gc} ) (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1. Penetration</td>
<td>Chavarie (1973)</td>
<td>( k_{gc} = \frac{4D_{inlet}u_b}{\pi d_e^2} )</td>
<td>( k_{gc} )</td>
<td>0.0051</td>
</tr>
<tr>
<td>2. Cloud two film</td>
<td>Walker (1970)</td>
<td>( k_{gc} = \frac{0.93D_{inlet}U_{mf}/\sqrt{x - 1}}{d_e} )</td>
<td>( \frac{2}{3} k_{gc} )</td>
<td>0.0017</td>
</tr>
<tr>
<td>3. Partridge and Rowe</td>
<td>Partridge and Rowe</td>
<td>( k_{gc} = \frac{0.26D_{inlet}e^{0.8}Sc^{0.33}}{d_e} )</td>
<td>( \delta_m k_{gc} )</td>
<td>0.0020</td>
</tr>
<tr>
<td>4. Chiba and Kobayashi</td>
<td>Chiba and Kobayashi</td>
<td>( k_{gc} = \sqrt{\frac{4D_{inlet}u_b(x - 1)}{\pi d_e x}} )</td>
<td>( k_{gc} )</td>
<td>0.0030</td>
</tr>
</tbody>
</table>

Table showing some more models for \( k_{gc} \) given by various researchers

\[
\frac{dN_A}{dt} = V_{\text{bubble}} \left( \frac{d(C_{A,b})}{dt} \right) = k_{gt} S_{ex,bubble} (C_{A,e} - C_{A,b})
\]

<table>
<thead>
<tr>
<th>Model</th>
<th>Reference</th>
<th>Coefficient</th>
<th>( k_{gt} ) equivalent</th>
<th>Calculated ( k_{gc} ) (m/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>5. Kunnii and Levenspiel</td>
<td>Kunnii and Levenspiel (1969)</td>
<td>( k_{gb} = \frac{0.6D^{1/2}(g/d_h)^{3/4}}{q} ) ( q = 2U_{mf}d_hw )</td>
<td>( \frac{k_{gc}Q}{k_{gc}S_h + Q} )</td>
<td>0.0051</td>
</tr>
<tr>
<td></td>
<td></td>
<td>( Q = q + k_{gb}S_b )</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td></td>
<td>( k_{gc} = \frac{4D_{inlet}u_b}{\pi d_e} )</td>
<td></td>
<td></td>
</tr>
<tr>
<td>6. Toei et al.</td>
<td>Toei et al. (1969)</td>
<td>( k_{gc} = \frac{1.02S_{mf}/(x - 1)}{d_h \sqrt{x - 1} \sqrt{y + 1}} ) ( k_{gb} = 0 ) ( q = U_{mf}d_hw )</td>
<td>( k_{gc} + k_{gb} )</td>
<td>0.0056</td>
</tr>
<tr>
<td>7. Murray through-flow, no diffusion</td>
<td>Murray (1965)</td>
<td>( k_{gb} = 0 ) ( q = U_{mf}d_hw )</td>
<td>( q/S_b )</td>
<td>0.0160</td>
</tr>
<tr>
<td>Empirical value</td>
<td>from Chavarie and Grace, 1976a</td>
<td></td>
<td></td>
<td>0.0160</td>
</tr>
</tbody>
</table>
The correlations for particle to gas mass and heat transfer coefficient are closely related when the Sherwood number (Sh) is equivalent to Nusselt number (Nu) and the Schmidt number (Sc) is equivalent to Prandtl number (Pr).

For example, the correlation for Nusselt number for particles in a fixed bed is expressed as

\[
\text{Nu}_{\text{bed}} = \frac{h d_p}{k_g} = 2 + 1.8(Re_p)^{0.5}(Pr)^{0.333}
\]

for \(Re_p > 80\)

The corresponding correlation equation for the Sherwood number has the form

\[
\text{Sh}_{\text{bed}} = \frac{k_{g,\text{bed}} d_p y}{D} = 2 + 1.8(Re_p)^{0.5}(Sc)^{0.333}
\]

for \(Re_p > 80\)

Hence the correlations obtained for Nu (heat transfer) can be converted to Sherwood number to obtain mass transfer coefficient, provided \(Sh \rightarrow Nu \) and \(Sc \rightarrow Pr\)

---

**Some Concluding Remarks on Mass Transfer in FB**

Fluidized beds are very important for solid-fluid contact for various important industrial applications.

Heat as well as mass transfer is important for design considerations.

When kinetic processes such as mass transfer are carried out with fluidized beds the particles dispersed in bubble phase should be taken in to account.

Mass transfer coefficient measured for the bed as a whole \((k_{g,\text{bed}})\) is model dependent.

For large particles (cloudless bubble bed) the plug flow model closely matches the bed conditions in the bed and the mass transfer coefficient for bed should be equal to single particle mass transfer coefficient.

For fine particles \(k_{g,\text{bed}} \ll k_{g,\text{single}}\)
Application to drying

Fluidized bed drying

Fluidizing with hot air is an attractive means for drying many moist powders and granular products.

The technique has been used industrially for drying crushed minerals, sand, polymers, fertilizers, pharmaceuticals, crystalline materials and many other products.

The main reason for its popularity is

- Efficient gas-solids contacting leads to compact units and relatively low capital cost combined with high thermal efficiency
- Very high heat and mass transfer and hence reduced drying times
- The absence of moving parts, low maintenance cost and possibility of using continuous mode

The main limitation is the material to be dried should be fluidizable and should have narrow particle size distribution.
Fluid Bed Dryer

**Air suction system:**
- Fan
- Blower
- Compressor

**Fluidized bed system:**
- Conventional
- Modified
- Hybrid / combined

**Heating system:**
- Electricity
- Fuel
- Steam

**Dust separation system:**
- Cyclone
- Bag filter
- Precipitator
- Scrubber

Fluidized Bed Drying

Different designs of fluid bed dryers

(a) Well-mixed fluidized bed dryer
(b) Plug flow fluidized bed dryer
(c) Vibrated fluidized bed dryer
(d) Spouted bed dryer
Fluidized Bed Drying

**Well mixed fluid bed dryer**
- common FBD used in industry.
- bed temperature uniform, equal to the product and exhaust gas temperatures.
- particle residence time distribution is wide
- wide range of product moisture content.
- feed is continuously charged into FB of relatively dry particles, this enhances fluidization quality.
- a series of well-mixed continuous dryers may be used with variable operating parameters.

**Plug flow fluid bed dryer**
- vertical baffles are inserted to create a narrow particle flow path.
- narrow particle residence time distribution.
- nearly equal residence time for all particles regardless of their size
- uniform product moisture content.
- length-to-width ratio from 5:1 to 30:1.
- inlet region may be agitated or apply back-mixing, or use a flash dryer to remove the surface moisture.
Developments in Fluidized Bed Drying

- Conventional FBD’s use steady flow rate, constant temperature and operate in batch or continuous mode at near atmospheric pressure with air as the drying/fluidizing medium.
- Modified FBDs may use pulsed flow, variable temperature, vibration to assist fluidization, use superheated steam as drying medium, operate at reduced pressure etc.
- FBDs may be used to dry slurries or continuous sheets (e.g. leather in a bed of adsorbent particles).
- Fluidized beds compete with rotary dryers, conveyor dryers and moving bed or packed bed dryers due to their advantages such as higher efficiency, better heat/mass transfer rates, smaller floor area; ability to operate in batch or continuous modes etc.
- Limitations include high power consumption, attrition, need to have fluidizable material etc.
- Often a second stage after flash drying or spray drying where it is also used as an agglomerate.

FBD Classification

- FB Dryers
  - Batch
    - Conventional (Well Mix)
    - Modified (Plug Flow (narrow RTD))
  - Continuous
    - Conventional
      - Single stage
    - Modified
      - Multistage
      - Hybrid
  - Hybrid

- Input
  - Continuous input
  - Variable input

- Well Mix (wide RTD)
Modeling - Fluid Bed Dryer

Diffusion model
This model assumes that the drying of single particle in the fluidized bed is controlled by the diffusion of moisture inside the particle

Empirical model
• In this approach the drying process is divided into different periods where drying mechanisms in each drying period are different;
• The solution of Fick's law of diffusion expresses the moisture content in terms of drying time by exponential function;
• Experimental data obtained from fluid bed drying experiments can be correlated using exponential function

Kinetic model

Single phase model - Explained later

Two-phase model - Explained later

---

Modeling - Fluid Bed Dryer

Single phase model
The fluidized bed is regarded essentially as a continuum
Heat and mass balances are applied over the fluidized bed
Assumption - particles in the bed are perfectly mixed

Mass balance

\[ -M_s \frac{dY}{dr} = G_e (Y_{out} - Y_{in}) \]

Energy balance

\[ M_s c_p \frac{dT}{dt} = G_e (c_p + Y_{in} c_{p, v}) (T_{in} - T_{out}) - G_e (Y_{out} - Y_{in}) \lambda \]
Two phase model

Two-phase model of fluidized bed drying treats the fluidized bed to be composed of a bubble phase (dilute phase) and an emulsion phase (dense phase)

gas in excess of minimum fluidization velocity, $u_{mf}$, flows through the bed as bubbles whereas the emulsion phase stays stagnant at the minimum fluidization conditions.

Mass balance

Modeling - Fluid Bed Dryer

Two phase model

Mass balance of liquid in the bubble phase

$$
p_b \varepsilon_{bb} \frac{dY_{bb}}{dt} + p_g \frac{V_{gb}}{V_{bt}} (Y_{bb} - Y_{in}) = 6 K_c p_g \varepsilon_{bb} (Y_d - Y_{bb})
$$

Mass balance of liquid in the interstitial gas in the dense phase gives the following equation

$$
\frac{6 K_c p_g \varepsilon_{bb}}{d_{bb}} (Y_{bb} - Y_d) - \rho_g \frac{V_{gd}}{V_{bt}} (Y_d - Y_{in}) + m = \rho_e \varepsilon_{mf} (1 - \varepsilon_{bb}) \frac{dY_d}{dt}
$$

Mass balance of liquid in the dense-phase particles

$$
m = -\rho_p (1 - \varepsilon_{mf}) (1 - \varepsilon_{bb}) \frac{dV}{dt}
$$

The coupled mass and energy balance in dense phase that consists of particles and interstitial gas phases

$$
\rho_p (1 - \varepsilon_{mf}) (1 - \varepsilon_{bb}) (c_m - c_{dpV}) \frac{dT_p}{dt} = \rho_g \frac{V_{gd}}{V_{bt}} (c_{pg} + Y_{in \epsilon_{pv}} (T_{g_{in}} - T_p) - \Delta H_{evap})
$$

$$
\times \left[ \rho_a \frac{V_{gd}}{V_{bt}} (Y_d - Y_{in}) - \frac{6 K_c p_g \varepsilon_{bb}}{d_{bb}} (Y_{bb} - Y_d) \right]
$$

Modeling - Fluid Bed Dryer

Two phase model

Mass balance of liquid in the bubble phase

$$
p_b \varepsilon_{bb} \frac{dY_{bb}}{dt} + p_g \frac{V_{gb}}{V_{bt}} (Y_{bb} - Y_{in}) = 6 K_c p_g \varepsilon_{bb} (Y_d - Y_{bb})
$$

Mass balance of liquid in the interstitial gas in the dense phase gives the following equation

$$
\frac{6 K_c p_g \varepsilon_{bb}}{d_{bb}} (Y_{bb} - Y_d) - \rho_g \frac{V_{gd}}{V_{bt}} (Y_d - Y_{in}) + m = \rho_e \varepsilon_{mf} (1 - \varepsilon_{bb}) \frac{dY_d}{dt}
$$

Mass balance of liquid in the dense-phase particles

$$
m = -\rho_p (1 - \varepsilon_{mf}) (1 - \varepsilon_{bb}) \frac{dV}{dt}
$$

The coupled mass and energy balance in dense phase that consists of particles and interstitial gas phases

$$
\rho_p (1 - \varepsilon_{mf}) (1 - \varepsilon_{bb}) (c_m - c_{dpV}) \frac{dT_p}{dt} = \rho_g \frac{V_{gd}}{V_{bt}} (c_{pg} + Y_{in \epsilon_{pv}} (T_{g_{in}} - T_p) - \Delta H_{evap})
$$

$$
\times \left[ \rho_a \frac{V_{gd}}{V_{bt}} (Y_d - Y_{in}) - \frac{6 K_c p_g \varepsilon_{bb}}{d_{bb}} (Y_{bb} - Y_d) \right]
$$
Some questions for self study

For drying of certain product, the ambient air of 80% relative humidity is heated to 70°C before entering the fluid bed dryer of uniform particle size.

• Discuss qualitatively, what will be the effect if the air is cooled down to 5°C and reheated back to same temperature of 70°C before entering the dryer. During which drying rate period will this be more useful
• Discuss what will be the effect on performance if the bed particles have wide size distribution, Suggest improvements to tackle this situation.
• What will be the effect on drying if heated inert spherical particles of same size are added in a bed of particles to be dried?
• How can one enhance the drying of a non-spherical particles in a fluidized bed?

Key References

• Yang, W.C.; Handbook of Fluidization and Fluid –Particle Systems, Marcel Dekker, USA, 2003
• Kunni, D.; Levenspiel, O. Fluidization Engineering, 1969
Drying Technology - An Overview

Supplementary Notes for students

ME5202 – Industrial Transport Processes

Professor A. S. Mujumdar
National University of Singapore

Contents

Introduction to Drying
  Basic terms in drying
  Why so many dryers (complex process)
  Key criteria for classification
  Criteria for dryer selection
  Different dryer types
  Energy related issues in drying
  Special/Innovative techniques
  Closure
Introduction to drying

Removal of a liquid from a solid/semi-solid/liquid to produce solid product by thermal energy input causing phase change (Sometimes converts solid moisture into vapor by sublimation eg. Freeze drying with application of heat.)

Needed for the purposes of preservation and storage, reduction in cost of transportation, etc.

Most common and diverse operation with over 100 types of dryers in industrial use

Competes with distillation as the most energy-intensive operation

Basics about drying

Energy Input by
- Conduction
- Convection
- Radiation
- Microwave and Radio Frequency Fields
- Combined mode

Moisture Output by
- Liquid diffusion
- Vapor diffusion
- Capillary flow (Permeability)
- Knudsen diffusion (Mean free path < pore dia.)
- Surface diffusion
- Poiseuille flow
- Combination of above
Drying a Complex Process

Multicomponent Moisture transport
Change of physical structure
Coupled with mass transfer
Input Continuous/intermittent
Chemical/biochemical reactions
Phase change
Change in quality
Transient
Shrinkage

Drying as a complex thermal process

Drying based on heat input

I. Direct (Convective)

Hot gas
Wet product
Direct Dryer
Humid gas
Dry product

Drying medium directly contacts material to be dried and carries evaporated moisture.

II. Indirect (Contact, Conduction)

Gas flow (low)
Wet product
Vacuum or low gas flow
Dry product

Heat supplied by heat exchanger (through metal wall)
Drying based on heat input

III. Radiant

<table>
<thead>
<tr>
<th>Heater (radiant)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Wet feed</td>
</tr>
<tr>
<td></td>
</tr>
<tr>
<td>Dry product</td>
</tr>
</tbody>
</table>

Vacuum or low gas flow to carry evaporated moisture away.

IV. Microwave or RF

Electromagnetic energy absorbed selectively by water (volumetric heating).

Typically less than 50% of total heat supplied in most direct dryers is used for evaporation. Water is the most common solvent removed in dryers.

---

**Summarization of Basic Terms**

<table>
<thead>
<tr>
<th>Terms/Symbol</th>
<th>Meaning</th>
</tr>
</thead>
<tbody>
<tr>
<td>Adiabatic saturation</td>
<td>Equilibrium gas temperature reached by unsaturated gas and vaporizing liquid under adiabatic conditions. Only for air/water system, it is equal to the wet bulb temperature</td>
</tr>
<tr>
<td>temperature, ( T_{ad} )</td>
<td></td>
</tr>
<tr>
<td>Bound moisture</td>
<td>Liquid physically and/or chemically bound to solid matrix so as to exert a vapor pressure lower than that of pure liquid at the same temperature</td>
</tr>
<tr>
<td>Constant rate drying period,</td>
<td>Under constant drying conditions, drying period when evaporation rate per unit drying area is constant (when surface moisture is removed)</td>
</tr>
<tr>
<td>( N_c )</td>
<td></td>
</tr>
<tr>
<td>Dew point, ( T_d )</td>
<td>Temperature at which a given unsaturated air-vapor mixture becomes saturated</td>
</tr>
<tr>
<td>Dry bulb temperature, ( T_{db} )</td>
<td>Temperature measured by a (dry) thermometer immersed in vapor-gas mixture.</td>
</tr>
<tr>
<td>Equilibrium moisture content,</td>
<td>At a given temperature and pressure, the moisture content of moist solid in equilibrium with the gas-vapor mixture (zero for non-hygroscopic materials)</td>
</tr>
<tr>
<td>( X_e )</td>
<td></td>
</tr>
<tr>
<td>Critical moisture content, ( X_c )</td>
<td>Moisture content at which the drying rate first begins to drop (under constant drying conditions)</td>
</tr>
<tr>
<td>Falling rate period, ( N_f )</td>
<td>Drying period under constant drying conditions during which the rate false continuously with time</td>
</tr>
<tr>
<td>Free moisture,</td>
<td>Moisture content in excess of the equilibrium moisture content (hence free to be removed) at given air humidity and temperature.</td>
</tr>
</tbody>
</table>

---
Summarization of Basic Terms

<table>
<thead>
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</tr>
</thead>
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<tr>
<td>Humid heat</td>
<td>Heat required to raise the temperature of unit mass of dry air and its associated vapor through one degree (J kg(^{-1}) K(^{-1}))</td>
</tr>
<tr>
<td>Humidity, absolute, Y</td>
<td>Mass of water vapor per unit mass of dry air (kg kg(^{-1}))</td>
</tr>
<tr>
<td>Humidity, relative</td>
<td>Ratio of partial pressure of water vapor in gas-vapor mixture to equilibrium vapor pressure at the same temperature.</td>
</tr>
<tr>
<td>Unbound moisture</td>
<td>Moisture in solid which exerts vapor pressure equal to that of pure liquid at the same temperature.</td>
</tr>
<tr>
<td>Water activity, (a_w)</td>
<td>Ratio of vapor pressure exerted by water in solid to that of pure water at the same temperature</td>
</tr>
<tr>
<td>Wet bulb temperature, (T_{wb})</td>
<td>Liquid temperature attained when large amount of air-vapor mixture is contacted with the surface. In purely convective drying, drying surface reaches (T_{wb}) during constant rate period</td>
</tr>
</tbody>
</table>

Basic Terms

Various types of moisture content
Basic Terms

Sorption Isotherm

Equilibrium moisture content curves for various types of solids
Basic Terms

**dX/dt = constant**

\[ X \text{ (d.b.)} = \frac{\text{mass of water}}{\text{mass of dry solid}} \]

\[ X \text{ (w.b.)} = \frac{\text{mass of water}}{\text{mass of wet solid}} \]

Typical drying curve

Typical textbook batch drying rate curve under constant drying conditions
By convention, the drying rate, \( N \), is defined as

\[
N = -\frac{M_s}{A} \frac{dX}{dt} \quad \text{or} \quad -\frac{M_s}{A} \frac{dX_f}{dt}
\]

an initial constant rate period where \( N = N_c = \text{constant} \).

The constant rate period is governed fully by the rates of external heat and mass transfer since a film of free water is always available at the evaporating surface.

\( N_c \) can be calculated using empirical or analytical techniques to estimate the external heat/mass transfer rates

\[
N_c = \sum q \lambda \frac{1}{\lambda}
\]

The drying rate in the falling rate period(s) is a function of \( X \) (or \( X_f \)) and must be determined experimentally for a given material being dried in a given type of dryer.

---

### Basic Terms

If the drying rate curve (\( N \) versus \( X \)) is known, the total drying time required to reduce the solid moisture content from \( X_1 \) to \( X_2 \) can be simply calculated by

\[
t_d = \int \frac{x_1}{M_s} \frac{dX}{N}
\]

<table>
<thead>
<tr>
<th>Model</th>
<th>Drying time</th>
</tr>
</thead>
<tbody>
<tr>
<td>Kinetic model, ( N = \frac{M_s}{A} \frac{dX}{dt} )</td>
<td>( t_d = ) Drying time to reach final moisture content ( X_f ) from initial moisture content ( X_i )</td>
</tr>
<tr>
<td>( N = N(X) ) (General)</td>
<td>( t_d = \frac{M_s}{A} \int_{x_i}^{x_f} \frac{dX}{N} )</td>
</tr>
<tr>
<td>( N = N_c ) (Constant rate period)</td>
<td>( t_c = \frac{M_s}{A} \frac{X_2 - X_1}{N_c} )</td>
</tr>
<tr>
<td>( N = aX + b ) (Falling rate period)</td>
<td>( t_f = \frac{M_s}{A} \frac{(X_1 - X_2)}{N_f} \ln \frac{N_f}{N_2} )</td>
</tr>
<tr>
<td>( N = \frac{AX_{X^<em>}}{X^</em>} \leq X_2 \leq X_c )</td>
<td>( t_f = \frac{M_s}{A N_c} \ln \frac{X_1}{X_2} )</td>
</tr>
</tbody>
</table>
**Basic Terms**

**Unusual Drying Rate Curves**

- RDF
- vapor-lock
- Through/impingement drying
- Casehardening
- Textbook DRC
- SHD

**Basic Terms**

**Unusual Drying Rate Curves**

- Crystallization
- Melting
- skinning
- shrinkage
- glass transition
- Puffing
- precipitation
- Physical structure
- Chemical changes
- Heating procedures
- Drying medium
- SHS
- air

**Reasons for non-textbook shapes**

- change of mass
- change of physical structure

**Heating procedures**

- boundary heating
- volumetric heating

* Constant drying conditions
**Basic Terms**

**Diffusion during drying of solids**

Fick’s law

\[
\frac{\partial X_f}{\partial t} = D_L \frac{\partial^2 X_f}{\partial x^2}
\]

Solution subject to the following initial and boundary conditions

\[X_l = X_0, \text{ everywhere in the slab at } t = 0\]
\[X_l = 0, \text{ at } x = a \text{ (top, evaporating surface), and} \]
\[\frac{\partial X_f}{\partial x} = 0 \text{ at } x=0 \text{ (bottom, non-evaporating surface)}\]

<table>
<thead>
<tr>
<th>Liquid diffusion model</th>
<th>[t_r = \frac{4a^2}{\pi D_L} \frac{8X_i}{\pi^2 X_z}]</th>
</tr>
</thead>
</table>
| Slab; one-dimensional diffusion, evaporating surface at \(X^*\) | \[X = \text{average free moisture content}\]
| \(a = \text{half-thickness of slab}\) |

---

**Diffusion (Contd)**

<table>
<thead>
<tr>
<th>Geometry</th>
<th>Boundary conditions</th>
<th>Dimensionless average free M.C.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flat plate of thickness (2b)</td>
<td>[t = 0; -b &lt; z &lt; b; X = X_0] [t &gt; 0; z = \pm b; X = X^*]</td>
<td>[X = \frac{8}{\pi^2} \sum_{n=1}^{\infty} \frac{1}{(2n-1)} \exp\left[-(2n-1)^2 \pi^2 X_0 \frac{D_L}{b}\right]]</td>
</tr>
<tr>
<td>Infinitely long cylinder of radius (R)</td>
<td>[t = 0; 0 &lt; r &lt; R; X = X_0] [t &gt; 0; r = R; X = X^*]</td>
<td>[X = 4 \sum_{n=1}^{\infty} \frac{1}{R^2 \alpha_n^2} \exp(-D_L \alpha_n^2 t)]</td>
</tr>
<tr>
<td>Sphere of radius (R)</td>
<td>[t = 0; 0 &lt; r &lt; R; X = X_0] [t &gt; 0; r = R; X = X^*]</td>
<td>[X = \frac{6}{\pi} \sum_{n=1}^{\infty} \frac{1}{n^2} \exp\left[-n^2 \pi^2 \left(\frac{D_L t}{R}\right)\right]]</td>
</tr>
</tbody>
</table>
**Basic Terms (useful mainly for food and biologicals)**

**WATER ACTIVITY (\(a_w\))**:

\[
a_w = \frac{\text{Partial pressure of water over wet solid}}{\text{Equilibrium vapor pressure of water at same temp.}}
\]

State of water in bio-product:

- **Free water** - intra-cellular water; nutrients and dissolved solids needed for living cells
- **Bound water** - water built into cells or biopolymer structures
  Needs additional energy to break "bonds" with solid. Bound water also resists freezing

For safe storage, bio-products must be dried to appropriate levels and stored under appropriate conditions

---

**Basic Terms (water activity)**

![Water activity versus moisture content plot for different food materials](image)

**Water activity versus moisture content plot for different food materials**
Why so many dryer types?

- Over 500 reported in literature studies; over 100 commercially available
- Over 50,000 materials are dried commercially at rates of a few kg/hr to 30 T/hr or more
- Drying times (residence times within drying chamber) can range from 1/3 sec. to months
- Temperature and pressure range from below triple point to super-critical
- Numerous constraints on physical/chemical properties of feed as well as dried product require a bewildering array of dryer designs
- Wide range of feeds (liquid, solid, semi-solid, particulate, pasty; sludge-like; sticky etc); wide specs on dried product

Why so many dryer types?

- Different sources of energy input (conduction, convection, radiation, MW, RF etc)
- Energy input continuous or intermittent
- Batch, continuous or semi-continuous operation
- Quality is key parameter for many products
- Limited number used in pharma industry
- Need to reduce the cost
- Need to consider drying system rather than dryer, i.e. Pre- and post-drying stages are important and often cost more than dryer
- Environmental regulations demand new drying techniques
Criterion for selection of dryers

• Numerous criteria, with different weights
• Many dryers can typically meet specs; hence several dryers can do a given job in general.
• Choice depends on mode of operation, physical form of feed and dried product desired; heat sensitivity; quality requirements; production rate; whether non-aqueous solvents are present in feed; whether material is toxic/inflammable or friable etc
• Key criterion- dryer must be able to handle the product- move it from feed to exit! Other criteria follow
• For pharma products - quality is NO 1 criterion!
Criterion for selection of dryers

- **Dryer Selection: A black art or science?**
- Little published work on subject
- In view of tremendous diversity of dryers, buyer must know more about dryers and drying
- Most vendors specialize in selected dryer types; so buyer needs to make choice
- Multiple choices are possible and can make selection process complex
- Expertise needed to make right choice!
- Energy, environment, safety and cost are important considerations in selection.
- **Special care needed when handling nonaqueous solvents in wet material**

Why select dryer carefully?

- Can affect bottom-line..
- Product quality, energy usage affected by choice
- Choose right drying system—not just dryer
- Weakest link decides ultimate goodness of system choice
- Survey of 10 largest pharma and chemical companies in Europe in 1990’s identified dryer selection as main problem facing industry!
- Expert systems exist for selection. Different expert systems give different selections
- Know product and process well before choosing drying system; imitation can cause problems!
- Simple decision trees suggested (SPS)
Some notes for dryer selection

- Must examine **drying system** cost rather than **dryer** cost for final selection.
- Largely untested in industrial practice – trend is to “repeat history”
- **Do not copy dryer** or dryer system used elsewhere without critical evaluation from square 1!
- Nickel ore concentrate is dried in different places using spray, fluid bed, rotary and flash dryers/ Which one do you COPY?
- Local fuel availability and relative costs of different energy sources, environmental requirements as well as legislation can change selection of dryer for same application

Main dryer classification criteria

<table>
<thead>
<tr>
<th>Criterion</th>
<th>Types</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mode of operation</td>
<td>• Batch</td>
</tr>
<tr>
<td></td>
<td>• Continuous*</td>
</tr>
<tr>
<td>Heat input-type</td>
<td>• Convection*, conduction, radiation, electromagnetic fields, combination of heat transfer modes</td>
</tr>
<tr>
<td></td>
<td>• Intermittent or continuous*</td>
</tr>
<tr>
<td></td>
<td>• Adiabatic or non-adiabatic</td>
</tr>
<tr>
<td>State of material in dryer</td>
<td>• Stationary</td>
</tr>
<tr>
<td></td>
<td>• Moving, agitated, dispersed</td>
</tr>
<tr>
<td>Operating pressure</td>
<td>• Vacuum*</td>
</tr>
<tr>
<td></td>
<td>• Atmospheric</td>
</tr>
<tr>
<td>Drying medium (convection)</td>
<td>• Air*</td>
</tr>
<tr>
<td></td>
<td>• Superheated steam</td>
</tr>
<tr>
<td></td>
<td>• Flue gases</td>
</tr>
</tbody>
</table>
### Main dryer classification criteria

<table>
<thead>
<tr>
<th>Criterion</th>
<th>Types</th>
</tr>
</thead>
<tbody>
<tr>
<td>Drying temperature</td>
<td>• Below boiling temperature*</td>
</tr>
<tr>
<td></td>
<td>• Above boiling temperature</td>
</tr>
<tr>
<td></td>
<td>• Below freezing point</td>
</tr>
<tr>
<td>Relative motion between drying medium and</td>
<td>• Co-current</td>
</tr>
<tr>
<td>drying solids</td>
<td>• Counter-current</td>
</tr>
<tr>
<td></td>
<td>• Mixed flow</td>
</tr>
<tr>
<td>Number of stages</td>
<td>• Single*</td>
</tr>
<tr>
<td></td>
<td>• Multi-stage</td>
</tr>
<tr>
<td>Residence time</td>
<td>• Short (&lt; 1 minute)</td>
</tr>
<tr>
<td></td>
<td>• Medium (1 – 60 minutes)</td>
</tr>
<tr>
<td></td>
<td>• Long (&gt; 60 minutes)</td>
</tr>
</tbody>
</table>

* Most common in practice

### Typical checklist for selection of industrial dryers

<p>| Physical form of feed                          | • Granular, particulate, sludge, crystalline, liquid, pasty,         |
|                                                | suspension, solution, continuous sheets, planks, odd-shapes         |
|                                                | (small/large)                                                       |
|                                                | • Sticky, lumpy                                                     |
| Average throughput                             | • kg/h (dry/wet); continuous                                        |
|                                                | • kg per batch (dry/wet)                                            |
| Expected variation in throughput (turndown     | • Oil                                                                |
| ratio)                                        | • Gas                                                                |
|                                                | • Electricity                                                       |
| Pre- and post-drying operations (if any)       | • Mean particle size                                                |
| For particulate feed products                 | • Size distribution                                                 |
|                                                | • Particle density                                                  |
|                                                | • Bulk density                                                       |
|                                                | • Rehydration properties                                            |</p>
<table>
<thead>
<tr>
<th>Typical checklist for selection of industrial dryers</th>
</tr>
</thead>
<tbody>
<tr>
<td>Chemical/biochemical/microbiological activity</td>
</tr>
<tr>
<td>Heat sensitivity</td>
</tr>
<tr>
<td>• Melting point</td>
</tr>
<tr>
<td>• Glass transition temperature</td>
</tr>
<tr>
<td>Inlet/outlet moisture content</td>
</tr>
<tr>
<td>• Dry basis</td>
</tr>
<tr>
<td>• Wet basis</td>
</tr>
<tr>
<td>Sorption isotherms (equilibrium moisture content)</td>
</tr>
<tr>
<td>Drying time</td>
</tr>
<tr>
<td>• Drying curves</td>
</tr>
<tr>
<td>• Effect of process variables</td>
</tr>
<tr>
<td>Special requirements</td>
</tr>
<tr>
<td>• Material of construction</td>
</tr>
<tr>
<td>• Corrosion</td>
</tr>
<tr>
<td>• Toxicity</td>
</tr>
<tr>
<td>• Non-aqueous solution</td>
</tr>
<tr>
<td>• Flammability limits</td>
</tr>
<tr>
<td>• Fire hazard</td>
</tr>
<tr>
<td>• Color/texture/aroma requirements (if any)</td>
</tr>
</tbody>
</table>

More guidelines for Dryer Selection
## Principal Data Needed

Include as much relevant data as possible

<table>
<thead>
<tr>
<th>Principal Data Needed</th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Solids throughput</td>
<td>Mass flow $W_s$</td>
</tr>
<tr>
<td></td>
<td>Turndown ratio</td>
</tr>
<tr>
<td>Moisture content</td>
<td>Inlet X1,</td>
</tr>
<tr>
<td></td>
<td>Outlet X0, variation</td>
</tr>
<tr>
<td>Particle properties</td>
<td>Size, size distribution</td>
</tr>
<tr>
<td></td>
<td>Density, $r_p$, $r_s$</td>
</tr>
<tr>
<td>Drying kinetics</td>
<td>Drying curves</td>
</tr>
<tr>
<td></td>
<td>E.M.C. data</td>
</tr>
<tr>
<td>Temperature limits</td>
<td>long-term</td>
</tr>
<tr>
<td></td>
<td>Instantaneous</td>
</tr>
<tr>
<td>Gas and solvent</td>
<td>Identity</td>
</tr>
<tr>
<td></td>
<td>Physical properties</td>
</tr>
<tr>
<td>Other features</td>
<td>Safety, ease of handling, attrition, etc.</td>
</tr>
<tr>
<td></td>
<td>Quality aspects</td>
</tr>
<tr>
<td></td>
<td>Toxicity, flammability</td>
</tr>
</tbody>
</table>

## Additional Qualitative Data Needed

- Fires and dust explosions
- Toxicity
- Potential for environmental damage
- Product value
- Need for containment
- Capital cost
- Attrition, hardness and friability
- Cohesion, adhesion, agglomeration
- Operating time
- Need for size reduction/enlargement
- Post-drying operations and Pre-drying factors
Small Scale Lab Tests

Small scale tests give valuable information:

• Drying kinetics – drying rates (parametric effects)
• Equilibrium moisture content – effect of T, humidity
• Microscopic examination – surface, agglomeration
• Lab-scale rotary evaporator – overheating, balling, adhesion
• Rotating drum tester – attrition, dustiness
• Cohesion and adhesion – handling, sticky point
  • Vital to have a representative sample of final material
  • Not necessary to carry out all of above tests in all cases

Basic Choice: Form of Feed

Feed and product can be in one of these main basic forms:

• Particulate solids (bed/layer/or dispersed)
• Sheet or film
• Block or slab
• Slurry or solution (feed only) or paste

  • Mostly require completely different types of dryer
  • Widest choice available for particulate solids
  • Specification of final product also critical in selection
**Basic Choice: Batch or Continuous**

**Batch dryers favored by:**
- Low throughput (under 50 kg/h)
- Long residence time (i.e. mainly falling rate drying)
- Batch equipment upstream and downstream
- Requirement for batch integrity

**Continuous dryers favored by**
- opposite conditions

**Match production made of feed where possible**

---

**Basic Choice: Information From Kinetic Data**

**Interpretation of drying curves assists choice:**
- Unhindered drying period – favors convective/displacement
- Long hindered drying period – favors contact drying
- Estimate of required solids residence time
- Maximum likely drying rate
- Indication of mechanisms controlling drying
- Difference between initial and final drying rates *

* (If high, favors well-mixed, parallel flow or two-stage)
## Dryers: Solid Exposure to Heat Conditions

<table>
<thead>
<tr>
<th>Dryers</th>
<th>Typical residence time within dryer</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>0-10 sec</td>
</tr>
<tr>
<td><strong>Convection</strong></td>
<td></td>
</tr>
<tr>
<td>Belt conveyor dryer</td>
<td></td>
</tr>
<tr>
<td>Flash dryer</td>
<td>X</td>
</tr>
<tr>
<td>Fluid bed dryer</td>
<td></td>
</tr>
<tr>
<td>Rotary dryer</td>
<td></td>
</tr>
<tr>
<td>Spray dryer</td>
<td>X</td>
</tr>
<tr>
<td>Tray dryer (batch)</td>
<td></td>
</tr>
<tr>
<td>Tray dryer (continuous)</td>
<td></td>
</tr>
<tr>
<td><strong>Conduction</strong></td>
<td></td>
</tr>
<tr>
<td>Drum dryer</td>
<td>X</td>
</tr>
<tr>
<td>Steam jacket rotary dryer</td>
<td></td>
</tr>
<tr>
<td>Steam tube rotary dryer</td>
<td></td>
</tr>
<tr>
<td>Tray dryer (batch)</td>
<td></td>
</tr>
<tr>
<td>Tray dryer (continuous)</td>
<td></td>
</tr>
</tbody>
</table>

## Product Classification and Dryer Types

<table>
<thead>
<tr>
<th>Dryers</th>
<th>Evap. Rate (kg/m²/hr)</th>
<th>Fluid, liquid suspension</th>
<th>Pastes</th>
<th>Powders</th>
<th>Granules, pellets</th>
<th>Operation</th>
</tr>
</thead>
<tbody>
<tr>
<td>Forced Convection (through flow)</td>
<td>7.5</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>Good</td>
<td>Batch</td>
</tr>
<tr>
<td>Double Cone</td>
<td>10</td>
<td>-</td>
<td>Poor</td>
<td>Fair</td>
<td>Poor</td>
<td>Batch</td>
</tr>
<tr>
<td>FBD</td>
<td>130</td>
<td>-</td>
<td>-</td>
<td>Good</td>
<td>Good</td>
<td>Continuous</td>
</tr>
<tr>
<td>Band</td>
<td>30</td>
<td>-</td>
<td>Fair</td>
<td>-</td>
<td>Good</td>
<td>Continuous</td>
</tr>
<tr>
<td>Film Drum</td>
<td>22</td>
<td>Good</td>
<td>Fair</td>
<td>-</td>
<td>-</td>
<td>Continuous</td>
</tr>
<tr>
<td>Flash</td>
<td>750</td>
<td>-</td>
<td>Fair</td>
<td>Good</td>
<td>Fair</td>
<td>Continuous</td>
</tr>
<tr>
<td>Rotary (indirect)</td>
<td>33</td>
<td>-</td>
<td>Poor</td>
<td>Good</td>
<td>Fair</td>
<td>Continuous</td>
</tr>
<tr>
<td>Spin Flash</td>
<td>185</td>
<td>-</td>
<td>Good</td>
<td>Good</td>
<td>Fair</td>
<td>Continuous</td>
</tr>
<tr>
<td>Spray</td>
<td>15</td>
<td>Good</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>Continuous</td>
</tr>
</tbody>
</table>
**Different Industrial Dryer Types**

**Turbo Tray Dryers**

- Suitable for granular feeds, operate with rotating shelves and force convection of air above the shelves.
- The Dryer can have 30+ trays and provide large residence time.
- Hermetic sealing is possible for solvent recovery.
Rotary Dryer

Direct-Heat Rotary Drying

- Combined cascade motion with heat & mass transfer.
- Large capital & operating cost.
- Used in fertilizers, pharmaceutical, lead & zinc concentrate for smelting, cement.

- Size 0.3 to 5 m diameter & 2 to 90 m length.

Steam Tube Rotary Dryer

Indirect-heat, steam-tube, rotary dryer
Fluid Bed Dryers - Variations

(a) Well-mixed fluidized bed dryer
(b) Plug flow fluidized bed dryer
(c) Vibrated fluidized bed dryer
(d) Spouted bed dryer

Fluid Bed Dryers - Modifications

Homogeneous FB without channeling or bubbles; high gas velocity possible
Deeper bed depth is possible if the bed is agitated - not commonly used

- Centrifugal / rotating FB - flowing gas radially - rotating cylindrical perforated distributor:
- promising contacting $U_{mf}$ and $U_t$ can be controlled
Rotocone Dryers (Batch)

- Drying of pharmaceuticals - tableting formulation
- Maximum capacity 10 m³.
- Evaporation rate 2-7 kg/hr.m²

Microwave Dryers

Used in ceramics industries, foods & pharmaceuticals to drive off final traces of moisture.
**Vacuum Dryers – Heat Sensitive Materials**

- **Vacuum tray dryer**
- **Agitated vacuum dryer** (About 10 min⁻¹)
- **Agitated vacuum dryer** (About 75 min⁻¹)
- **Trommel**

**Paddle Dryer**

- Provides drying time upto several hours.
- Suitable for pastelike & granular material.
- Evap. rate upto 10 kg/hr.m²
Screw Conveyor dryer

- Use to dry variety of materials
- Can be use as conveyor + dryer
- Possibility of conductive heating, use of sub atmospheric pressure

Yamato TACO Rotary Dryer

- Modified rotary dryer — instead of axial air flow and particle air contact in the freeboard region, this dryer injects hot air directly into the rolling bed of particulates.
- Volumetric heat transfer coefficient about two times that of conventional rotary dryer
- Not suited for all materials — unlike conventional rotary dryer
Superheated Steam Drying

Typical SSD set-up

Superheated Steam Drying

- **Saturated Steam Feed**
  - Assume 100°C, 1 bar; \( H = 2,690 \text{ kJ/kg} \)

- **Steam Superheater**
  - Assume 110°C, 1 bar; \( H = 2,720 \text{ kJ/kg} \)

- **Drying chamber**
  - Bleeding off for other uses

- **Saturated Steam Exhaust**
  - Back to 100°C, 1 bar; \( H = 2,690 \text{ kJ/kg} \)

- **Recycled steam**

- **Fan/blower**

- **Closed steam drying system**

- **Direct use of steam**

- **Energy recovery via heat exchanger**

- **Removal of condensate**

- **Steam from boiler**

- **Heater**

- **Purged steam**
Superheated Steam Drying

Classification of SSD

- **Low pressure**
  - Example: products which trend to melt or undergo glass transition

- **Near atmospheric pressure**
  - Example: potato chip, tortilla chip, shrimp, paddy, soybean, noodles

- **High Pressure**
  - Example: Beet pulp

Possible Types of SSD

- Flash dryers with or without indirect heating of walls
- FBDs with or without immersed heat exchangers
- Spray dryers
- Impinging jet dryers
- Conveyor dryers
- Rotary dryers
- Impinging stream dryers
Selection of Fluid Bed Dryer

Fluidization Regimes

1. Depends on types of solids, fluidization can appear in different types of regimes

2. Smooth fluidization and bubbling fluidization are preferable

3. Any material that can be fluidized in these two regimes can be dried in a FBD

4. By modifications of FBD, some solids / powders / pastes / solutions / sheets can also be fluidized well

Increasing gas velocity
Developments in Fluidized beds

- Conventional FBD's use steady flow rate, constant temperature and operate in batch or continuous mode at near atmospheric pressure with air as the drying/fluidizing medium.
- Modified FBDs may use pulsed flow, variable temperature, vibration to assist fluidization, use superheated steam as drying medium, operate at reduced pressure etc.
- FBDs may be used to dry slurries or continuous sheets (e.g. leather in a bed of adsorbent particles).
- Fluidized beds compete with rotary dryers, conveyor dryers and moving bed or packed bed dryers due to their advantages such as higher efficiency, better heat/mass transfer rates, smaller floor area; ability to operate in batch or continuous modes etc.
- Limitations include high power consumption, attrition, need to have fluidizable material etc.
- Often a second stage after flash drying or spray drying where it is also used as an agglomerate.

Comparison of FBD with other dryers

<table>
<thead>
<tr>
<th>Criterion</th>
<th>Rotary</th>
<th>Flash</th>
<th>Conveyor</th>
<th>FBD</th>
<th>Modified FBD</th>
</tr>
</thead>
<tbody>
<tr>
<td>Particle size</td>
<td>Large range</td>
<td>Fines</td>
<td>500μm-10mm</td>
<td>100-2000μm</td>
<td>10μm-10mm</td>
</tr>
<tr>
<td>PS Distribution</td>
<td>Flexible</td>
<td>Limited</td>
<td>Flexible</td>
<td>Limited</td>
<td>Wide</td>
</tr>
<tr>
<td>Drying time</td>
<td>Up to 60min</td>
<td>10-30sec</td>
<td>Up to 120min</td>
<td>Up to 60min</td>
<td>Up to 60min</td>
</tr>
<tr>
<td>Floor area</td>
<td>Large</td>
<td>Large length</td>
<td>Large</td>
<td>Small</td>
<td>Small</td>
</tr>
<tr>
<td>Turndown ratio</td>
<td>Large</td>
<td>Small</td>
<td>Small</td>
<td>Small</td>
<td>Small</td>
</tr>
<tr>
<td>Attrition</td>
<td>High</td>
<td>High</td>
<td>Low</td>
<td>High</td>
<td>High</td>
</tr>
<tr>
<td>Power consumption</td>
<td>High</td>
<td>Low</td>
<td>Low</td>
<td>Medium</td>
<td>Medium</td>
</tr>
<tr>
<td>Maintenance</td>
<td>High</td>
<td>Medium</td>
<td>Medium</td>
<td>Medium</td>
<td>Medium</td>
</tr>
<tr>
<td>Energy efficiency</td>
<td>Medium</td>
<td>Medium</td>
<td>High</td>
<td>High</td>
<td>High</td>
</tr>
<tr>
<td>Ease of control</td>
<td>Low</td>
<td>Medium</td>
<td>High</td>
<td>High</td>
<td>High</td>
</tr>
<tr>
<td>Capacity</td>
<td>High</td>
<td>Medium</td>
<td>Medium</td>
<td>Medium</td>
<td>High</td>
</tr>
<tr>
<td>Criterion</td>
<td>Types of FB Dryer</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>---------------------------------</td>
<td>----------------------------------------------------------------------------------</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Mode</td>
<td>Batch, continuous, semi continuous</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Flow regime</td>
<td>Well mixed, plug flow, circulating, hybrid</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Pressure</td>
<td>Low (heat sensitive), atmospheric, high (steam)</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Temperature</td>
<td>Constant, time dependant</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas flow</td>
<td>Continuous, pulsed</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Heat supply</td>
<td>Convection/conduction; continuous/pulsed</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fluidization action</td>
<td>Gas flow, jet flow, mechanical, external field</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fluidizing material</td>
<td>Particulate solids, paste slurry (inert solids bed)</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Fluidizing medium</td>
<td>Heated air, s. steam, dehumidified air, freeze</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**FB Dryers**

- **Batch**
  - Conventional
    - Continuous input
  - Modified
    - Variable input
- **Continuous**
  - Conventional
    - Single stage
  - Modified
    - Multistage
    - Hybrid

**Batch (Well Mix)**

**Plug Flow (narrow RTD)**

**Well Mix (wide RTD)**
**Well Mixed and Plug Flow Fluid Bed dryer**

- common FBD used in industry.
- bed temperature uniform, equal to the product and exhaust gas temperatures.
- particle residence time distribution is wide
- wide range of product moisture content.
- feed is continuously charged into FB of relatively dry particles, this enhances fluidization quality.
- a series of well-mixed continuous dryers may be used with variable operating parameters.

- vertical baffles are inserted to create a narrow particle flow path.
- narrow particle residence time distribution.
- nearly equal residence time for all particles regardless of their size
- uniform product moisture content.
- length-to-width ratio from 5:1 to 30:1.
- inlet region may be agitated or apply back-mixing, or use a flash dryer to remove the surface moisture.

---

**Multi-stage / Multi Process FBD**

- Fluidized bed drying: heated air
- Fluidized bed cooling: cold air

- Well-mix fluidized bed - plug flow fluidized bed
- Drying: Well-mix - pre-dryer; Plug flow - second stage drying.

**Fluidized Bed Dryer/Cooler**

Combines FB drying and cooling.
Cooling to eliminate condensation
Multi-stage / Multi Process FBD

upper stage drying - lower stage cooling well-mixed FB - plug flow FB

Spray drying
FB drying,
FB cooling

spray drying - well-mixed
FB drying/cooling - plug flow

FBD Variations

Pulsating Fluidized Bed

- Variable operating parameters: pulsating gas flow, variable temperature, adjustable heat input or periodic fluidization
- energy and cost effectiveness
- Pulsating – provides vertical vibration, improving the fluidization quality
- Bed temperature - batch FB - constant - adjustable heat input
- Pulsating FB –rotated hot air inlet – intermittent drying times

- low energy consumption due to high specific moisture extraction rate (SMER)
- high coefficient of performance (COP)
- wide range of drying temperature (-20°C to 110°C)
- environmental friendly
- high product quality.
- suitable for heat sensitive products (food, bio-origin products.)

Fluid Bed Heat Pump Dryer
Selection of Spray Dryer

Components of Spray Dryer

A conventional spray drying process consists of the following four stages:
1. Atomization of feed into droplets
2. Heating of hot drying medium
3. Spray-air contact and drying of droplets
4. Product recovery and final air treatment
**Types of atomizers: Rotary atomizer**

Advantages:
- Handles large feed rates with single wheel or disk
- Suited for abrasive feeds with proper design
- Has negligible clogging tendency
- Change of wheel rotary speed to control the particle size distribution
- More flexible capacity (but with changes in powder properties)

Limitations:
- Higher energy consumption compared to pressure nozzles
- More expensive
- Broad spray pattern requires large drying chamber diameter

---

**Types of atomizers: Pressure nozzle**

Advantages:
- Simple, compact and cheap
- No moving parts
- Low energy consumption

Limitations:
- Low capacity (feed rate for single nozzle)
- High tendency to clog
- Erosion can change spray characteristics
Types of atomizers: Pressure nozzle

Advantages:
- Simple, compact and cheap
- No moving parts
- Handle the feedstocks with high-viscosity
- Produce products with very small size particle

Limitations:
- High energy consumption
- Low capacity (feed rate)
- High tendency to clog

Types of Spray Dryers-flow patterns

Co-current flow  Counter-current flow  Mixed-current flow
Selection Tree for Spray Drying System

Energy Aspects in Drying
Why Energy Saving Needed?

- Rising energy costs - may even end up with carbon tax
- Greenhouse gas emission related directly to energy consumed
- Industrial dryers operate at 20-70% efficiency - lot of scope to improve operation
- Globalization will force competition
- This part includes global energy status, Energy in Drying....
- possible options for energy savings in general way (Not specific criteria)

General observations

- **Drying** - a highly energy-intensive operation due to
  1. High vaporization latent of water
  2. Affinity of solids to water
  3. Difficulty in getting heat to where water is; resistances to heat/mass transfer, etc

- **Some obvious options to save energy**
  1. Avoid drying - not feasible!
  2. Displace water with solvent with low heat of vaporization and high vapor presence
  3. Use mechanical means (filtration, ultra-centrifugation, etc)
  4. Use evaporation (for dilute liquids)
Some indices used to define dryer efficiency

- **Energy efficiency** $\eta$

\[ \eta = \frac{E_{ev}}{E_t} = \frac{\text{Energy used for moisture evaporation}}{\text{Total energy given to dryer}} \]

- **Energy efficiency** $\eta_T$ (range from 0 to 1)

\[ \eta_T = \frac{T_1 - T_2}{T_1 - T_0} \]

- $T_1 =$ Inlet air temperature; $T_2 =$ Outlet air temperature; $T_0 =$ Ambient temperature

  - **Max $\eta_T$ is when** $T_2 = T_{wb}$

    $T_{wb} =$ wet bulb temperature at exit conditions

Think Drying System instead of Dryer!!

- Always look at full flowsheet - and the whole drying system - not just dryer!
- Pre-drying (mechanical dewatering etc) and post-drying (cooling, granulation, blending etc) should be considered in overall strategy to save energy
- Mechanical dewatering is 10 times cheaper than evaporation which is 10 times cheaper than drying! (Rough rules of thumb!)
- Not always possible to use minimum energy option as it may not be optimal for quality etc.
Techniques for energy savings

- How energy is produced for drying- direct firing is most efficient; convective drying can be more rapid
- Electric heating technologies- some new technologies have emerged to be cost-competitive
- Heat recovery- various options
- Control of dryer
- New developments e.g. superheated steam pulsed combustion, adsorption drying etc.

Direct firing-why?

- CDDET recommends direct drying with natural gas to reduce chain of heat transfer equipment. Claims advantages such as:
  - Reduced maintenance; Ease of installation
  - Uniform heating, better control of temperature
  - Potential for integration into existing control system.

**Limitations**

- Loss of energy in exhaust air
- Heat recovery from low temperature exhaust is not inexpensive
- Product maybe contaminated by combustion gas; not allowed in some countries for foods
- Flammable solids/vapors require self-inertizing (low oxygen atmosphere).
Electric Heating

- Higher grade energy—typically an expensive source for low grade heat; use only if there is quality advantage
- MW drying can be more energy-efficient than solar energy in energy consumed, not cost-effective always look at bottom-line!
- IR heating, MW/RF, induction heating can deliver energy direct to product—no air needed—hence more efficient thermally.

Advantages

- Precise temperature control
- Reduced dryer size
- Clean, easy-to-maintain
- Better quality, no contamination
- Short start-up, shut-down times.

Heat recovery

- Recycle ratio of exhaust air, if drying rates are high due to high air velocity and short contact time in dryer e.g. Yankee dryer for tissue paper- 80-90% of exhaust needs to be recycled!
- Consider cost associated with recycle ducting and some reduction in driving potential for drying
- Can cause problems of condensation etc as weather changes significantly.
Use of Process Control

- Refer to chapter on Control of Dryers in Handbook of Industrial Drying, 3rd Edition for details of control strategies
- Here we look only at potential for energy savings
- % Energy savings depends on the quality of base case used!!.
Some heat recovery methods

- Use of heat pumps
- Heat Pipe heat exchangers
- Gas/gas plate heat exchangers; fin-tube exchangers
- Heat Wheels- can have short paybacks
- Control- improve quality, reliability, energy consumption etc
- Good housekeeping; good insulation etc.

Summary of case studies reported (Energy Recovery)

<table>
<thead>
<tr>
<th>Sector</th>
<th>Title</th>
<th>Energy savings</th>
<th>Payback</th>
<th>Comments</th>
</tr>
</thead>
<tbody>
<tr>
<td>Wood</td>
<td>Hardwood manufacturer uses electric driven heat pump to kiln dry lumber</td>
<td>33%</td>
<td>Not given</td>
<td>Electric heat pump, lower inventory costs due to elimination of air drying step, improved product quality</td>
</tr>
<tr>
<td>Wood</td>
<td>Lumber dry kiln heat recovery system</td>
<td>30%</td>
<td>2 years</td>
<td>Heat pipe heat exchanger system. No remixing of exhaust and inlet</td>
</tr>
<tr>
<td>Ceramics</td>
<td>Saving energy with heat pump at a pottery</td>
<td>45%(cost saving)</td>
<td>Not given</td>
<td>Electric heat pump dehumidifier</td>
</tr>
<tr>
<td>Malt</td>
<td>Electric heat pump dehumidification halved energy costs</td>
<td>50%</td>
<td>2 years</td>
<td>Electric heat pump dehumidifier, improved product quality</td>
</tr>
</tbody>
</table>
### Summary of case studies reported (Energy Recovery)

<table>
<thead>
<tr>
<th>Sector</th>
<th>Title</th>
<th>Energy savings</th>
<th>Payback</th>
<th>Comments</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ceramics</td>
<td>Heat recovery from spray dryer using a glass tube heat exchanger</td>
<td>20%</td>
<td>2 years</td>
<td>Spray dryer exhaust gas heat recovery</td>
</tr>
<tr>
<td>Food</td>
<td>Heat recovery using a run-around coil in the air and exhaust system</td>
<td>25%</td>
<td>Not given</td>
<td>PVC coating on exchangers to prevent corrosion</td>
</tr>
<tr>
<td>Ceramics</td>
<td>Waste heat recovery in drying towers in manufacture of detergents</td>
<td>2 years</td>
<td></td>
<td>Exhaust heat recovery, glass tube heat exchanger, reduced fouling</td>
</tr>
</tbody>
</table>

### General guidelines for saving energy

- Use indirect heating (Conduction, IR) to avoid loss of energy in exhaust air
- Use intermittent, multi-mode heating, where possible - for batch drying
- Use less air, higher temperature, where possible
- Use waste heat, renewable energy (solar, wind, etc), where feasible
- Recover exhaust heat or recycle
- Optimize operating conditions; model based control
- Some options involve extra capital costs
# Energy Efficiency for Different Dryers

<table>
<thead>
<tr>
<th>Dryer type</th>
<th>η range%</th>
<th>Dryer type</th>
<th>η range%</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Direct continuous</td>
<td>Indirect continuous</td>
<td></td>
</tr>
<tr>
<td>Tower</td>
<td>20-40</td>
<td>Drum</td>
<td>60-75</td>
</tr>
<tr>
<td>flash</td>
<td>50-75</td>
<td>Rotary</td>
<td>50-60</td>
</tr>
<tr>
<td>Sheeting (stenters)</td>
<td>50-60</td>
<td>Cylinder</td>
<td>60-70</td>
</tr>
<tr>
<td>conveyor</td>
<td>40-60</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Rotary (bundle)</td>
<td>40-70</td>
<td>Agitated pan</td>
<td>60-70</td>
</tr>
<tr>
<td>Spray</td>
<td>40-60</td>
<td>Vacuum rotary</td>
<td>Up to 70</td>
</tr>
<tr>
<td>Tunnel</td>
<td>35-40</td>
<td>Vacuum tray</td>
<td>60-70</td>
</tr>
<tr>
<td>Fluidized bed</td>
<td>40-70</td>
<td>Infrared</td>
<td>50-80</td>
</tr>
<tr>
<td>Batch</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Tray</td>
<td>85</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

## Methods to improve energy efficiency

<table>
<thead>
<tr>
<th>Methods</th>
<th>Potential</th>
</tr>
</thead>
<tbody>
<tr>
<td>Use of heat exchangers including heat pipes</td>
<td>M</td>
</tr>
<tr>
<td>Use of heat pumps</td>
<td>M-L</td>
</tr>
<tr>
<td>Model-based control</td>
<td>M-L</td>
</tr>
<tr>
<td>Optimized operation</td>
<td>H-M</td>
</tr>
<tr>
<td>Superheated steam drying with utilization of excess steam</td>
<td>L</td>
</tr>
<tr>
<td>Multi-stage drying</td>
<td>M-L</td>
</tr>
<tr>
<td>More efficient heat source (e.g. pulse combustion)</td>
<td>M-L</td>
</tr>
</tbody>
</table>

H: high  M: medium  L: low
How to save energy in drying?

- Do regular energy audits- find out where heat is being lost. Temperature and humidity measurements at inlet, outlet; IR thermography can help detect leaks.
- Typically 20-30% energy can be saved at no cost!
- Using advanced/modified drying systems
  1. Superheated steam drying
  2. Multi-stage drying system
  3. Low pressure spray drying
  4. Pulse combustion drying

Introduction to pulse combustion drying

- Pulse combustion is intermittent; can be subsonic or supersonic Mach Number >1.0 (supersonic)

<table>
<thead>
<tr>
<th>Features</th>
<th>Steady</th>
<th>Pulsed</th>
</tr>
</thead>
<tbody>
<tr>
<td>Combustion intensity (kW/m³)</td>
<td>100-1000</td>
<td>10000-50000</td>
</tr>
<tr>
<td>Efficiency of burning (%)</td>
<td>80-96</td>
<td>90-99</td>
</tr>
<tr>
<td>Temperature level (K)</td>
<td>2000-2500</td>
<td>1500-2000</td>
</tr>
<tr>
<td>CO concentration in exhaust (%)</td>
<td>0-2</td>
<td>0-1</td>
</tr>
<tr>
<td>NOx concentration in exhaust (mg/m³)</td>
<td>100-7000</td>
<td>20-70</td>
</tr>
<tr>
<td>Convective heat transfer coefficient (W/m²K)</td>
<td>50-100</td>
<td>100-500</td>
</tr>
<tr>
<td>Time of reaction (s)</td>
<td>1-10</td>
<td>0.01-0.5</td>
</tr>
<tr>
<td>Excess air ratio</td>
<td>1.01-1.2</td>
<td>1.00-1.01</td>
</tr>
</tbody>
</table>

http://blastwavejet.com/pulsejet.htm
More about pulse combustion drying

- High drying rates
  - Increased turbulence and flow reversal in the drying zone promote gas/materials mixing
  - Decreased boundary layer thickness of materials
  - Increased heat and mass transfer rates
  - High driving force because of high gas temperature

- Short contact time
  - Suitable for some heat sensitive materials

- High energy efficiency and economic use of fuels

- Environmentally friendly operation

---

Energy consumption between PC and conventional dryer

<table>
<thead>
<tr>
<th>Dryers</th>
<th>Typical evaporation capacity</th>
<th>Typical consumption (kJ/kgH₂O)</th>
</tr>
</thead>
<tbody>
<tr>
<td>PC dryers</td>
<td>250-2000 kg H₂O/h</td>
<td>3000-3500</td>
</tr>
<tr>
<td>Tunnel dryer</td>
<td></td>
<td>5500-6000</td>
</tr>
<tr>
<td>Impingement dryer</td>
<td>50 kg H₂O/hm²</td>
<td>5000-7000</td>
</tr>
<tr>
<td>Rotary dryer</td>
<td>30-80 kg H₂O/hm²</td>
<td>4600-9200</td>
</tr>
<tr>
<td>Fluid bed dryer</td>
<td></td>
<td>4000-6000</td>
</tr>
<tr>
<td>Flash dryer</td>
<td>5-100 kg H₂O/hm³</td>
<td>4500-9000</td>
</tr>
<tr>
<td>Spray dryer</td>
<td>1-30 kg H₂O/hm³</td>
<td>4500-11500</td>
</tr>
<tr>
<td>Drum dryer (pastes)</td>
<td>6-20 kg H₂O/hm²</td>
<td>3200-6500</td>
</tr>
</tbody>
</table>
Advanced Drying Methods

Heat Pump Dryer (Concept)
**Advanced Drying Methods**

**Typical Heat Pump Dryer**

![Diagram of a heat pump dryer]

**Advantages and Limitations of Heat Pump Dryer**

<table>
<thead>
<tr>
<th>Advantages</th>
<th>Limitations</th>
</tr>
</thead>
<tbody>
<tr>
<td>• High Energy Efficiency</td>
<td>• Higher initial cost and maintenance cost due to need to maintain compressor, refrigerant filters and charging of refrigerant</td>
</tr>
<tr>
<td>• Better quality products especially for heat sensitive materials.</td>
<td>• Leakage of refrigerant.</td>
</tr>
<tr>
<td>• Wide range of operating conditions.</td>
<td>• Marginally complex operation relative to simple convection dryer.</td>
</tr>
<tr>
<td>• Low temperature drying.</td>
<td>• Additional floor space requirement.</td>
</tr>
<tr>
<td>• Better solvent recovery in closed loop system (useful for recovering solvent during drying of pharma products).</td>
<td></td>
</tr>
</tbody>
</table>
**Advanced Drying Methods**

**Atmospheric Freeze Drying**

Scope:
1. AFD with fixed bed
2. AFD with OD
3. AFD with vibrated bed and absorbent

---

**Atmospheric Freeze Drying (Advantages)**

- Significant reduction in energy costs - due to the absence of a vacuum chamber and ancillary equipment.
- Continuous system - higher productivity and lower operating cost.
- Decrease energy consumption and drying time – due to application of heat-pump system and different process temperature elevating modes.
- Minimize product degradation – by using inert gas drying environment.
- High heat transfer co-efficient - about 20-40 times greater than that in vacuum dryer.

**Atmospheric Freeze Drying (Limitations)**

- Long drying time – lower diffusivity of water vapor with increasing pressure in the chamber.
- Bulky system – Require more space.
- Two mechanical agents are required - does not seem economical from the energy point of view.
- It also takes time to setup, de-humidify and cool the drying chamber.
Advanced Drying Methods

**Spray freeze drying**

- Method which combines processing steps common to freeze drying and spray drying
- The protein drug is dissolved
- Solution is nebulized into a cryogenic medium (e.g., liquid nitrogen) which generates a dispersion of shock frozen droplets
- Dispersion is then dried in lyophilizer
- So far mainly used for highly specialized food and pharma products

![Diagram of spray freeze drying process](image)

Advanced Drying Methods

**Microwave Dryer**

- Microwave dryers are expensive both in terms of the capital and operating (energy) costs.
- They have found limited applications to date.
- Seem to have special advantages in terms of product quality when handling heat-sensitive materials. They are worth considering as devices to speed up drying in the tail end of the falling rate period.
- Similarly, RF dryers have limited industrial applicability
- Applications mainly in food/pharma
Advanced Drying Methods

Intermittent Drying

Energy Savings & Quality Enhancement
Intermittent Drying

Batch - temporal

Cyclic or time-varying heat input by convection, conduction, radiation, dielectric fields, etc.

Continuous - spatial

Inherent

Rotary Dryers
Spouted Beds
Multi-cylinder paper dryers

Imposed

Freeze Dryers
Wood Drying Kilns
Pulsed Fluid Beds

Some examples of Intermittent Drying

- Rotating Jet Spouted Bed dryer
- Pulsed bed - intermittent fluidization
- Vibrated bed with tempering periods
- Intermittent IR/MW in a batch heat pump dryer
- Conveyor (Apron) dryer with parts of the dryer unheated

Aside from reduced energy/air consumption, product quality may be better for heat-sensitive and/or fragile solids. Slight increases in drying time are expected.
## Advanced Drying Methods

### Comparison of Conventional and Innovative Drying Techniques

<table>
<thead>
<tr>
<th>Feed type</th>
<th>Dryer type</th>
<th>New techniques*</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid Suspension</td>
<td>- Drum</td>
<td>• Fluid/spouted beds of inert particles</td>
</tr>
<tr>
<td></td>
<td>- Spray</td>
<td>• Spray/fluid bed combination</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Vacuum belt dryer</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Pulse combustion dryers</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Spray freeze drying</td>
</tr>
<tr>
<td>Paste/sludge</td>
<td>- Spray</td>
<td>• Spouted bed of inert particles</td>
</tr>
<tr>
<td></td>
<td>- Drum</td>
<td>• Fluid bed (with solid backmixing)</td>
</tr>
<tr>
<td></td>
<td>- Paddle</td>
<td>• Superheated steam dryers</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Screw conveyor dryer</td>
</tr>
</tbody>
</table>

### Advanced Drying Methods

<table>
<thead>
<tr>
<th>Feed type</th>
<th>Dryer type</th>
<th>New techniques*</th>
</tr>
</thead>
<tbody>
<tr>
<td>Particles</td>
<td>- Rotary</td>
<td>• Superheated steam FBD</td>
</tr>
<tr>
<td></td>
<td>- Flash</td>
<td>• Vibrated bed</td>
</tr>
<tr>
<td></td>
<td>- Fluidized bed (hot air or combustion gas)</td>
<td>• Ring dryer</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Pulsated fluid bed</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Jet-zone dryer</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Yamato rotary dryer</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Screw conveyor dryer</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Immers heat exchanged dryer</td>
</tr>
<tr>
<td>Continuous sheets</td>
<td>- Multi-cylinder contact dryers</td>
<td><strong>Combined impingement/radiation dryers</strong></td>
</tr>
<tr>
<td></td>
<td>- Impingement (air)</td>
<td>• Combined impingement and through dryers (textiles, low basis weight paper)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>• Impingement and MW or RF</td>
</tr>
</tbody>
</table>
### Advanced Drying Methods

#### Developments in fluid bed drying (comparison innovative and conventional fluidized beds)

<table>
<thead>
<tr>
<th>Variant</th>
<th>Conventional</th>
<th>Innovative</th>
</tr>
</thead>
<tbody>
<tr>
<td>Mode of heat transfer</td>
<td>Only convection</td>
<td>Convection + conduction (immersed heaters in bed) + radiative heat transfer (MW assisted fluid beds)</td>
</tr>
<tr>
<td>Gas flow</td>
<td>Steady</td>
<td>Pulsating; on/off</td>
</tr>
<tr>
<td>Mode of fluidization</td>
<td>Pneumatic</td>
<td>Mechanically agitated / vibrations</td>
</tr>
<tr>
<td>Drying media</td>
<td>Air / flue gases</td>
<td>Superheated steam / heat pump assisted (even using inert media)</td>
</tr>
<tr>
<td>Type of material dried</td>
<td>Particulate material</td>
<td>Drying of pastes / slurries using bed of inert particles</td>
</tr>
</tbody>
</table>

#### Advanced Drying Methods

#### Hybrid Drying Technologies

- **Fluidized bed dryers**
  - Combined mode of heat transfer
    - Convective / conduction (Agitated fluid bed dryer)
  - Convective with MW/RF/IR
    - Continuous or intermittent
  - Each stage with same dryer type (Two-stage fluid bed)
  - Different dryers at each stage
    - (Spray and fluid bed dryer or flash and fluid bed dryer)
  - Different drying technologies at each stage (superheated steam drying / by air drying)

- **Multi-stage drying systems**
  - Filter cum dryer
  - Drying & cooling (in plug flow fluid bed dryer)
  - Drying and agglomeration (spray followed by fluid bed)

- **Multiprocessing dryers**
Closure

- Very important step (after establishing need to dry and optimal flow sheet for non thermal dewatering)
- Wrong choice leads to severe penalties – start-up costs, downtime and need to replace
- User must do “homework” fist; vendors valuable thereafter
- Several dryers may do the job – same quality, cost etc.
- Selection does depend on cost of fuel, relative cost of different energy sources; geographical location; legislative regulations; emission control; safety, etc.
- Consider new technologies as well – when available and proven
- Expert systems now available (e.g. SPS) to aid in selection – still a combination of art (experience) and science!
- Selection may be dominated by just one criterion in some cases e.g. quality for pharma products
- Several different dryers can do same job at same cost in some cases
- Choice can depend on geographic location, cost of energy etc

Closure

- Energy is a important point to be considered in drying which is highly energy intensive unit operation
- Different routes can be used to minimize the energy losses
- Carbon foot prints can be minimized by making the drying system energy efficient
- Developing energy intensive methods with sustainability – need to develop innovative drying techniques
- Some of the advanced/innovative ways of drying are discussed
Books /monographs by Prof. Arun S. Mujumdar
During 2010
Visit us at:
http://serve.me.nus.edu.sg/arun

Selected books and journals in drying and transport phenomena edited / authored by Prof. Arun S. Mujumdar,
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Thank You for your Attention !!!

Q&A

Am S. M. -