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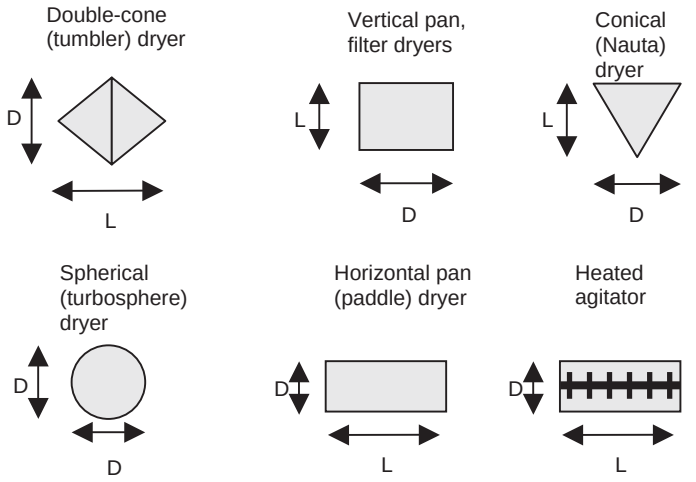

FIG. 12-55a Basic geometries for batch dryer calculations.

estimated heat-transfer coefficient for either the base case or the new dryer type is in error, the scaling factor will be wrong. All drying times have been shown in hours, as this is more convenient than seconds.

The paddle with heated agitator has the shortest drying time, and the filter dryer the longest (because the bottom plate is unheated). Other types are fairly comparable. The spherical dryer would usually have a higher heat-transfer coefficient and shorter drying time than shown.

Performance and Cost Data for Batch Vacuum Rotary Dryers Typical performance data for horizontal pan vacuum dryers are given in Table 12-27. Size and cost data for rotary agitator units are given in Table 12-28. Data for double-cone rotating units are in Table 12-29.

Continuous Agitated Dryers

Description These dryers, often known as paddle or horizontal agitated dryers, consist of one or more horizontally mounted shells with internal mechanical agitators, which may take many different forms. They are a continuous equivalent of the horizontal pan dryer and are similar in construction, but usually of larger dimensions. They have many similarities to continuous indirect rotary dryers and are sometimes classed as rotary dryers, but this is a misnomer because the outer shell does not rotate, although in some types there is an inner shell which does. Frequently, the internal agitator is heated, and a wide variety of designs exist. Often, two intermeshing agitators are used. There are important variants with high-speed agitator rotation and supplementary convective heating by hot air.

Classification Continuous; mechanical agitation and transport; layer; contact/conduction or convective (through-circulation).

The basic differences are in type of agitator, the two key factors being heat-transfer area and solids handling/stickiness characteristics. Unfortunately, the types giving the highest specific surface area (multiple tubes and coils) are often also the ones most liable to fouling and blockage and most difficult to clean. Figure 12-56 illustrates a number of different agitator types.

The most common problem with paddle dryers (and with their closely related cousins, steam-tube and indirect rotary dryers) is the buildup of sticky deposits on the surface of the agitator or outer jacket. This leads, first, to reduced heat-transfer coefficients and slower drying and, second, to blockages and stalling of the rotor. Also, thermal decomposition and loss of product quality can result. The problem is usually most acute at the feed end of the dryer, where the material is wettest and stickiest. A wide variety of different agitator designs have been devised to try to reduce stickiness problems and enhance cleanability while providing a high heat-transfer area. Many designs incorporate a high torque drive combined with rugged shaft construction to prevent rotor stall during processing, and stationary mixing elements are installed in the process housing which continually clean the heat-exchange surfaces of the rotor to minimize any crust buildup and ensure an optimum heat-transfer coefficient at all times. Another alternative is to use two parallel intermeshing shafts, as in the Nara paddle dryer (Fig. 12-57). Suitably designed continuous paddle and batch horizontal pan dryers can handle a wide range of product consistencies (dilute slurries, pastes, friable powders) and can be used for processes such as reactions, mixing, drying, cooling, melting, sublimation, distilling, and vaporizing. Bearing supports are usually provided at both ends of the unit for shaft support.

Design Methods for Paddle Dryers Product trials are conducted in small pilot dryers (8- to 60-L batch or continuous units) to determine material handling and process retention times. Variables such as drying temperature, pressure level, and shaft speed are analyzed during the test trials. For initial design purposes, the heat-transfer coefficient for paddle dryers is typically in the range of 10 W/(m^2·K) (light, free-flowing powders) up to 150 W/(m^2·K) (dilute slurries). However, it is preferable to scale up from the test results, finding the heat-transfer coefficient by backcalculation and scaling up on the basis of total area of heat-transfer surfaces, including heated agitators. Typical length/diameter ratios are between 5 and 8, similar to rotary dryers and greater than some batch horizontal pan dryers.

Continuous Rotary Dryers A rotary dryer consists of a cylinder that rotates on suitable bearings and that is usually slightly inclined to the horizontal. The cylinder length may range from 4 to more than 10 times

TABLE 12-26 Comparative Dimensions and Drying Times for Various Batch Contact Dryers

Dryer type	h, kW/(m^2·K)	L/D	D, m	L, m	A, m^2	t_{CR}, h	t_{FR}, h	t_{expt}, h
Tumbler/double-cone	0.05	1.5	3.71	5.56	38.91	2.78	5.98	11.0
Vertical pan	0.05	0.5	3.71	1.85	32.37	3.34	7.19	13.2
Spherical	0.05	1	3.37	3.37	35.63	3.04	6.54	12.0
Filter dryer	0.05	0.5	3.71	1.85	21.58	5.01	10.77	19.8
Conical agitated	0.05	1.5	3.71	5.56	34.12	3.17	6.82	12.5
Paddle (horizontal agitated)	0.05	5	1.72	8.60	46.50	2.33	5.01	9.2
Paddle, heated agitator	0.05	5	1.72	8.60	278.99	0.39	0.83	1.52

TABLE 12-27 Performance Data of Vacuum Rotary Dryers*

Material	Diameter × length, m	Initial moisture, % dry basis	Steam pressure, Pa × 10^3	Agitator speed, r/min	Batch dry weight, kg	Final moisture, % dry basis	Pa × 10^3	Time, h	Evaporation, kg/(h·m²)
Cellulose acetate	1.5 × 9.1	87.5	97	5.25	610	6	90–91	7	1.5
Starch	1.5 × 9.1	45–48	103	4	3630	12	88–91	4.75	7.3
Sulfur black	1.5 × 9.1	50	207	4	3180	1	91	6	4.4
Fuller's earth/mineral spirit	0.9 × 3.0	50	345	6	450	2	95	8	5.4

*Stokes Vacuum, Inc.

the diameter, which may vary from less than 0.3 to more than 3 m. Solids fed into one end of the drum are carried through it by gravity, with rolling, bouncing and sliding, and drag caused by the airflow either retarding or enhancing the movement, depending on whether the dryer is cocurrent or countercurrent. It is possible to classify rotary dryers into direct-fired, where heat is transferred to the solids by direct exchange between the gas and the solids, and indirect, where the heating medium is separated from physical contact with the solids by a metal wall or tube. Many rotary dryers contain flights or lifters, which are attached to the inside of the drum and which cascade the solids through the gas as the drum rotates.

For handling large quantities of granular solids, a cascading rotary dryer is often the equipment of choice. If the material is not naturally free-flowing, recycling of a portion of the final dry product may be used to precondition the feed, either in an external mixer or directly inside the drum. Hanging link chains and/or scrapper chains are also used for sticky feed materials.

Their operating characteristics when performing heat- and mass-transfer operations make them suitable for the accomplishment of drying, chemical reactions, solvent recovery, thermal decompositions, mixing, sintering, and agglomeration of solids. The specific types included are the following:

Direct cascading rotary dryer (cooler). This is usually a bare metal cylinder but with internal flights (shelves) which lift the material and drop it through the airflow. It is suitable for low- and medium-temperature operations, the operating temperature being limited primarily by the strength characteristics of the metal employed in fabrication.

Direct rotary dryer (cooler). As above but without internal flights.

Direct rotary kiln. This is a metal cylinder lined on the interior with insulating block and/or refractory brick. It is suitable for high-temperature operations.

Indirect steam-tube dryer. This is a bare metal cylinder provided with one or more rows of metal tubes installed longitudinally in the shell. It is suitable for operation up to available steam temperatures or in processes requiring water cooling of the tubes.

Indirect rotary calciner. This is a bare metal cylinder surrounded on the outside by a fired or electrically heated furnace. It is suitable for operation at medium temperatures up to the maximum that can be tolerated by the metal wall of the cylinder, usually 650 to 700 K for carbon steel and 800 to 1025 K for stainless steel.

Direct Roto-Louvre dryer. This is one of the more important special types, differing from the direct rotary unit in that true *through-circulation* of gas through the solids bed is provided. Like the direct rotary, it is suitable for low- and medium-temperature operation.

Direct heat rotary dryer. The direct heat units are generally the simplest and most economical in operation and construction, when the solids and gas can be permitted to be in contact. In design mode, the required gas flow rate can be obtained from a heat and mass balance. Bed cross-sectional area is found from a scoping design calculation (a typical gas velocity is 3 m/s for cocurrent and 2 m/s for countercurrent units). Length is normally between 5 and 10 times drum diameter (an *L/D* value of 8 can be used for initial estimation) or can be calculated by using an incremental model (see Examples 21 and 23).

A typical schematic diagram of a rotary dryer is shown in Fig. 12-58, while Fig. 12-59 shows typical lifting flight designs.

Classification Continuous; agitation and transport by rotation/gravity; layer (dispersion for cascading rotary dryers); convective (through-circulation) or contact/conduction.

Residence Time, Standard Configuration The residence time in a rotary dryer τ represents the average time that particles are present in the equipment, so it must match the required drying time.

Traditional approaches For rotary kilns, without lifting flights, Sullivan et al. (U.S. Bureau of Mines Tech. Paper 384) gave an early formula:

$$\tau = \frac{106.2L\sqrt{\gamma}}{N_m D \tan\alpha} \tag{12-88}$$

Here, the natural angle of repose of the material is γ, which increases as the material becomes more cohesive and less free-flowing, and the residence time τ is in seconds, but the rotation rate N_m is in revolutions per minute (rpm), not per second. The Friedman and Marshall equation [*Chem. Eng. Progr.* **45**(8): 482 (1949)] is derived from this, with an additional term to account for air drag on the solids:

$$\tau = \frac{13.8L}{N_m^{0.9} D\alpha} \pm \frac{590.6LG}{d_p^{0.5} G} \tag{12-89}$$

Here d_p is the particle size, in micrometers, while F and G are the mass flow rates of solids and gas, respectively. This formula has been frequently reported and includes a correction factor to the initial constant term to reflect actual experimental results. Friedman and Marshall took the angle of repose for the solids to be 40° and introduced a 0.9 power for the rotational speed, which had questionable justification within the accuracy of the data. The second term represents the airflow drag term and is negative for cocurrent flow and positive for countercurrent flow.

TABLE 12-28 Standard Rotary Vacuum Dryers*

Diameter, m	Length, m	Heating surface, m²	Working capacity, m³†	Agitator speed, r/min	Drive, kW	Weight, kg	Purchase price (1995)	
							Carbon steel	Stainless steel (304)
0.46	0.49	0.836	0.028	7½	1.12	540	$ 43,000	$ 53,000
0.61	1.8	3.72	0.283	7½	1.12	1,680	105,000	130,000
0.91	3.0	10.2	0.991	6	3.73	3,860	145,000	180,000
0.91	4.6	15.3	1.42	6	3.73	5,530	180,000	205,000
1.2	6.1	29.2	3.57	6	7.46	11,340	270,000	380,000
1.5	7.6	48.1	6.94	6	18.7	15,880	305,000	440,000
1.5	9.1	57.7	8.33	6	22.4	19,050	330,000	465,000

*Stokes Vacuum, Inc. Prices include shell, 50-lb/in²-gauge jacket, agitator, drive, and motor; auxiliary dust collectors, condensers.
†Loading with product level on or around the agitator shaft.

TABLE 12-29 Standard (Double-Cone) Rotating Vacuum Dryers*

Working capacity, m^3	Total volume, m^3	Heating surface, m^2	Drive, kW	Floor space, m^2	Weight, kg	Purchase cost (1995)	
						Carbon steel	Stainless steel
0.085	0.130	1.11	.373	2.60	730	$ 32,400	$ 38,000
0.283	0.436	2.79	.560	2.97	910	37,800	43,000
0.708	1.09	5.30	1.49	5.57	1810	50,400	57,000
1.42	2.18	8.45	3.73	7.15	2040	97,200	106,000
2.83	4.36	13.9	7.46	13.9	3860	198,000	216,000
4.25	6.51	17.5	11.2	14.9	5440	225,000	243,000
7.08	10.5	°38.7	11.2	15.8	9070	324,000	351,000
9.20	13.9	°46.7	11.2	20.4	9980	358,000	387,000
11.3	16.0	°56.0	11.2	26.0	10,890	378,000	441,000

°Stokes Vacuum, Inc. Price includes dryer, 15-lb/in^2 jacket, drive with motor, internal filter, and trunnion supports for concrete or steel foundations. Horsepower is established on 65 percent volume loading of material with a bulk density of 50 lb/ft^3. Models of 250 ft^3, 325 ft^3, and 400 ft^3 have extended surface area.

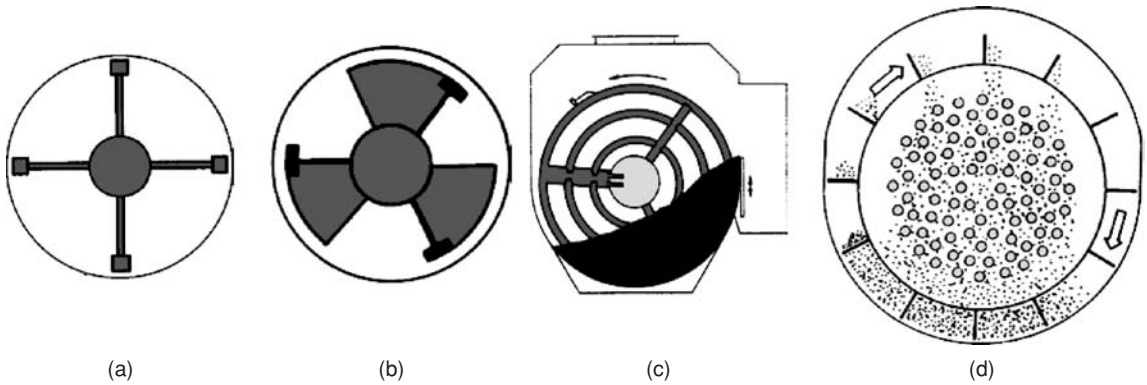

FIG. 12-56 Typical agitator designs for paddle (horizontal agitated) dryers. (*a*) Simple unheated agitator. (*b*) Heated cut-flight agitator. (*c*) Multicoil unit. (*d*) Tube bundle.

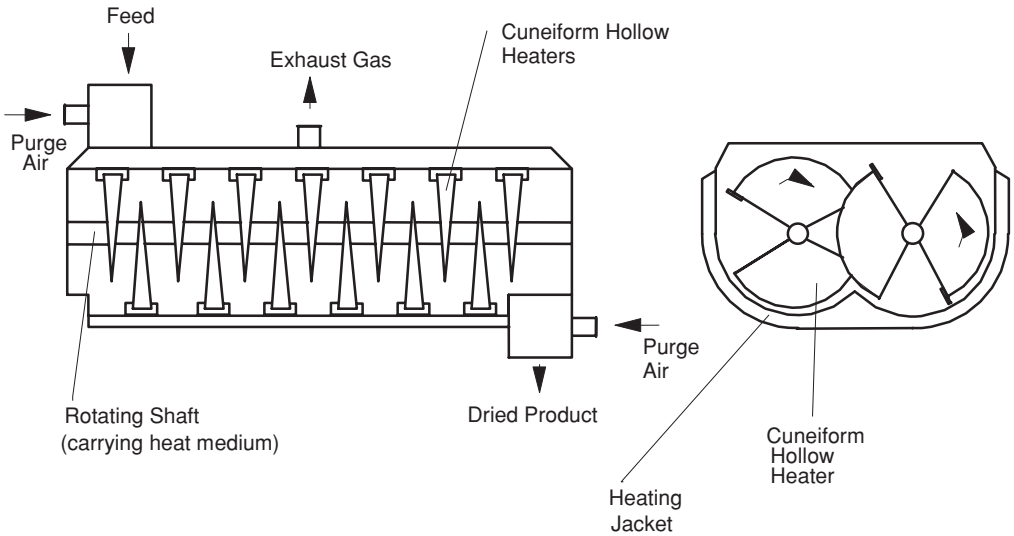

FIG. 12-57 Nara twin-shaft paddle dryer.

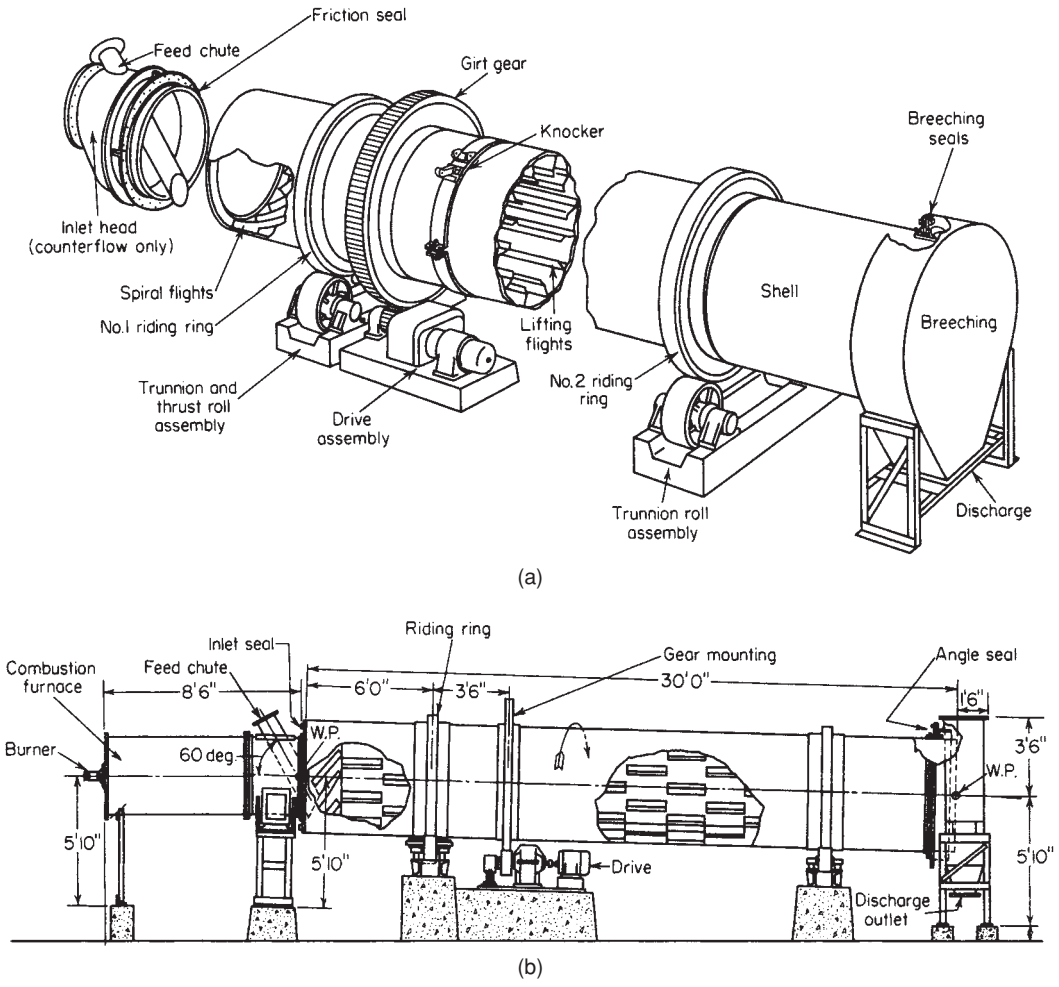

FIG. 12-58 Component arrangement (*a*) and elevation (*b*) of countercurrent direct-heat rotary dryer. (*Air Preheater Company, Raymond® & Bartlett Snow™ Products.*)

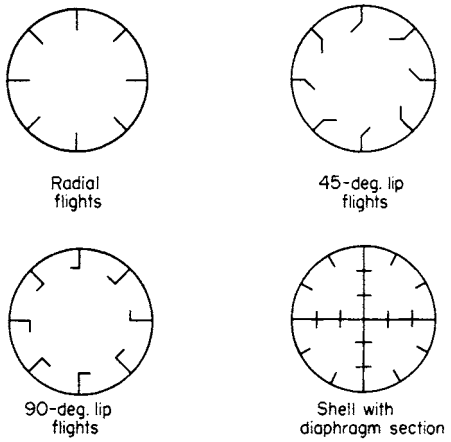

FIG. 12-59 Typical lifting flight designs.

Saeman and Mitchell [*Chem. Eng. Progr.* **50**(9):467 (1954)] proposed the following expression:

$$\tau = \frac{L}{f_H N D(\tan \alpha \pm k_m U_G)} \tag{12-90}$$

Here f_H is a *cascade factor*, with values typically between 2 and π, increasing as solids holdup increases, and k_m is an empirical constant (dimensional) for a given material. The superficial gas velocity through the empty drum is U_G. It was assumed that the airborne particle velocity was proportional to the air velocity. Two empirical constants f_H and k_m are also required to use the equation, and these are not generally available.

Schofield and Glikin [*Trans. IChemE* **40:**183 (1962)] analyzed particle motion from flights and airborne drag, obtaining

$$\tau = \frac{L}{\bar{y}\,\bar{\theta}\,N[\sin \alpha - (KU_G^2/g)]} \tag{12-91}$$

Here $\bar{y}$ is the mean distance of fall of the particles, $\bar{\theta}$ is the mean angle moved by particles in flights, and K is a dimensional drag constant. At

small angles α, $\sin\alpha \approx \tan\alpha$, and they noted that $\overline{y}\,\overline{\theta} \approx 2D$, so their final Eq. (12-91) is similar to that of Saeman and Mitchell (1954).

The above equations mainly differ in whether the drag term is additive or subtractive (as with Friedman and Marshall) or in the denominator (as with Saeman and Mitchell, and Schofield and Glikin). Some workers, including Sullivan et al. (1927), have neglected the effect of air drag completely. However, the general experience with rotary dryers is that the effect of air velocity and hence of air drag is very substantial, suggesting that neglecting air drag in any equation or analysis is unlikely to be sufficient unless the air velocity is very low. The formulas link L and τ, which is reasonably convenient for dryer performance assessment, but inconvenient for dryer design, where neither L nor τ is initially known.

Modern analysis Matchett and Baker [*J. Sep. Proc. Technol.* **8:**11 (1987)] provided a complete analysis of particle motion in rotary dryers. They considered both the airborne phase (particles falling through air) and the dense phase (particles in the flights or the rolling bed at the bottom). Typically, particles spend 90 to 95 percent of the time in the dense phase, but the majority of the drying takes place in the airborne phase. In the direction parallel to the dryer axis, most particle movement occurs through four mechanisms: by gravity and air drag in the airborne phase, and by bouncing, and sliding and rolling, in the dense phase. The combined particle velocity in the airborne phase is U_{P_1}, which is the sum of the gravitational and air drag components for cocurrent dryers and the difference between them for countercurrent dryers. The dense-phase velocity, arising from bouncing, sliding, and rolling, is denoted U_{P_2}.

Papadakis et al. [*Dry. Tech.* **12**(1&2):259 (1994)] rearranged the Matchett and Baker model from its original "parallel" form into a more computationally convenient "series" form. The sum of the calculated residence times in the airborne and dense phases, τ_G and τ_S, respectively, is the total solids residence time. The dryer length is simply the sum of the distances travelled in the two phases.

$$\tau = \tau_G + \tau_S \tag{12-92}$$

$$L = \tau_G U_{P_1} + \tau_S U_{P_2} \tag{12-93}$$

For airborne phase motion, the velocity U°_{P1} due to the gravitational component is given approximately by

$$U^{\circ}_{P_1} = \sqrt{\frac{gD_e}{2\cos\alpha}}\, K_{\text{fall}} \tan\alpha \tag{12-94}$$

where D_e is the effective diameter, which is the distance actually fallen by the particles. When one is designing a dryer, this parameter will not be known until the flight width is decided. And K_{fall} is a parameter that allows for particles falling from a number of positions, with different times of flight and lifting times, and is generally between 0.7 and 1. The velocity $U^{\circ}_{P_1}$ due to the gravitational component is most conveniently expressed as

$$U^{\circ}_{P_1} = \sqrt{\frac{gD}{2}}\tan\alpha\, K_{\text{fall}} \sqrt{\frac{D_e}{D\cos\alpha}} = \sqrt{\frac{gD}{2}}\tan\alpha K_K \tag{12-95}$$

The drag force gives a velocity component U^{d}_{P1} that must be obtained from experimental correlations, and combining these components gives U_{P_1}.

Bouncing, rolling, and sliding are not so easily analyzed theoretically. Matchett and Baker suggested that the dense-phase velocity could be characterized in terms of a dimensionless dense-phase velocity number a, through the equation

$$a = \frac{U_{P_2}}{N \cdot D \cdot \tan\alpha} \tag{12-96}$$

Other workers suggested that, in underloaded and design-loaded dryers, bouncing was a significant transport mechanism, whereas for overloaded dryers, rolling (kilning) was important. Bouncing mechanisms can depend on the airborne phase velocity U_{P_1}, since this affects the angle at which the particles hit the bottom of the kiln and the distance they move forward. Rolling mechanisms would be expected to depend on the depth of the bottom bed, and hence on the difference between the actual holdup H and the design-loaded holdup H^*.

As an example of the typical numbers involved, Matchett and Baker [*J. Sep. Proc. Technol.*, **9:**5 (1988)] used their correlations to assess the data of Saeman and Mitchell for an industrial rotary dryer with $D = 1.83$ m and $L = 10.67$ m, with a slope of 4°, 0.067 m/m. For a typical run with $U_G = 0.98$ m/s and $N = 0.08$ r/s, they calculated that $U^{\circ}_{P1} = 0.140$ m/s, $U^{d}_{P1} = -0.023$ m/s, $U_{P_1} = 0.117$ m/s, and $U_{P_2} = -0.02$ m/s. The dryer modeled was countercurrent and therefore had a greater slope and lower gas velocity than those of a cocurrent unit; for the latter, U°_{P1} would be lower and U^{d}_{P1} positive and larger. The ratio τ_S/τ_G is approximately 12 in this case, so that the distance traveled in dense-phase motion would be about twice that in the airborne phase.

Kemp and Oakley [*Dry. Tech.*, **20**(9):1699 (2002)] showed that the ratio τ_G/τ_S can be found by comparing the average time of flight from the top of the dryer to the bottom t_f to the average time required for the particles to be lifted by the flights t_d. They derived the following equation:

$$\frac{\tau_S}{\tau_G} = \frac{t_d}{t_f} = \frac{K_{fl}}{N}\sqrt{\frac{g}{D}} \tag{12-97}$$

Here all the unknowns have been rolled into a single dimensionless parameter K_{fl}, given by

$$K_{fl} = \frac{\theta}{\pi\sqrt{2\sin\theta}}\sqrt{\frac{D}{D_e}} \tag{12-98}$$

Here D_e is the effective diameter (internal diameter between lips of flights), and the solids are carried in the flights for an angle 2θ, on average, before falling. Kemp and Oakley concluded that K_{fl} can be taken to be 0.4 to a first (and good) approximation. For overloaded dryers with a large rolling bed, K_{fl} will increase. The form of Eq. (12-97) is very convenient for design purposes since it does not require D_e, which is unknown until a decision has been made on the type and geometry of the flights.

The model of Matchett and Baker has been shown by Kemp (Proc. IDS 2004, B, 790) to be similar in form to that proposed by Saeman and Mitchell:

$$\tau = \frac{1.1L}{ND\left[\tan\alpha\cdot(K_K/K_{fl}\sqrt{2}+a) + (1/K_{fl})\sqrt{\frac{1}{gD}}\cdot U^{d}_{P_1}\right]} \tag{12-99}$$

In Eq. (12-99), $K_K/(K_{fl}\sqrt{2})$ will typically be on the order of unity, and reported values of a are in the range of 1 to 4. The airborne gravity component is usually smaller than the dense-phase motion but is not negligible. The sum of these two terms is essentially equivalent to the factor f_H in Saeman and Mitchell's equation.

Heat- and Mass-Transfer Estimates Many rotary dryer studies have correlated heat- and mass-transfer data in terms of an overall volumetric heat-transfer coefficient $U_v a$ [W/(m$^3\cdot$K)], defined by

$$Q = U_v a \cdot V_{\text{dryer}} \cdot \Delta T_m \tag{12-100}$$

Here Q is the overall rate of heat transfer between the gas and the solids (W), V_{dryer} is the dryer volume (m^3), and ΔT_m is an average temperature driving force (K). When one is calculating the average temperature driving force, it is important to distinguish between the case of heat-transfer with dry particles, where the change in the particle temperature is proportional to the change in the gas temperature, and the case of drying particles, where the particle temperature does not change so significantly. Where the particles are dry, the average temperature difference is the logarithmic mean of the temperature differences between the gas and the solids at the inlet and outlet of the dryer, although Miller et al. (1942) took the logarithmic temperature difference as the average temperature difference even when the particles were drying. The volumetric heat-transfer coefficient itself consists of a heat-transfer coefficient U_v based on the effective area of contact between the gas and the solids, and the ratio a of this area to

TABLE 12-30 Values of the Index *n* in Correlations for the Volumetric Heat-Transfer Coefficient (after Baker, 1983)

Author(s)	Exponent *n*
Saeman and Mitchell (1954)	0
Friedman and Marshall (1949)	0.16
Aiken and Polsak (1982)	0.37
Miller et al. (1942)	0.46–0.60
McCormick (1962)	0.67
Myklestad (1963)	0.80

the dryer volume. Thus, this procedure eliminates the need to specify where most of the heat-transfer occurs (e.g., to material in the air, on the flights, or in the rolling bed). Empirical correlations are of the form

$$U_v a = \frac{K' U_{G\text{super}}^n}{D} \tag{12-101}$$

where K' depends on the solids properties, the flight geometry, the rotational speed, and the dryer holdup. Table 12-30 gives the values of n chosen by various authors.

McCormick (1962) reworked the data of Miller et al. (1942), Friedman and Marshall (1949), and Saeman and Mitchell (1954) with a view to obtaining a single correlation of the form of Eq. (12-101) for the volumetric heat-transfer coefficient. He demonstrated that all the data could be correlated with values of the exponent n from 0.46 to 0.67. Although the evidence was far from conclusive, he believed that a value of 0.67 for the exponent n was most reliable. Individual values of the constant K' were obtained from the results of each of the workers cited above. He found that it was a function of the solids properties, the flight geometry, the rotational speed, and the dryer holdup, but that there was insufficient evidence available to relate K' to these parameters.

A comparison between the correlations of various workers was made by Baker (1983), and this is given in Table 12-31. A 2-m-diameter dryer containing 16 flights was chosen as the basis for the comparisons. With the exception of the results of Myklestad (1963), the values of $U_v a$ were calculated by using the values of K' and a value of n of 0.67, as obtained by McCormick (1963). A 17-fold variation in the predicted values of $U_v a$ can be observed at both 1 and 3 m/s. The reason for this is not readily apparent. With the exception of the commercial data correlation of Miller et al. (1942), the results were all obtained in pilot-scale rigs having diameters ranging from 0.2 to 0.3 m. Differences in equipment size are therefore not likely to be the cause of the variation. Hence the variation must be attributed to a combination of experimental errors and differences in the experimental conditions which are unaccounted for in the correlations.

An alternative procedure is the use of a conventional film heat-transfer coefficient h_f[W/(m^2·K)]

$$Q = h_f \cdot A_s \cdot \Delta T \tag{12-102}$$

Here Q is the local heat-transfer rate (W), A_s is the total surface area of all the particles (m^2), and ΔT is the temperature difference between the gas and the solids (K). The method has the advantages that h_f can be determined by relatively simple tests (or calculated from appropriate correlations in the literature), variations in operating conditions can be allowed for, and analogies between heat and mass transfer allow the film coefficients for these processes to be related. However, the area for heat transfer must be estimated under the complex conditions of gas-solids interaction present in particle cascades. Schofield and Glikin (1962) estimated this area to be the surface area of particles per unit mass $6/(\rho_P d_P)$, multiplied by the fraction of solids in the drum that are cascading through the gas at any moment, which was estimated as the fraction of time spent by particles cascading through the gas:

$$A_s = \frac{6}{\rho_P d_P} \frac{t_f}{t_f + t_d} \tag{12-103}$$

Schofield and Glikin estimated the heat-transfer coefficient by using the correlation given by McAdams (1954), which correlates data for gas-to-particle heat transfer in air to about 20 percent over a range of Reynolds numbers (Re_P, defined in the previous section) between 17 and 70,000:

$$\mathrm{Nu}_P = 0.33 \cdot \mathrm{Re}_P^{1/2} \tag{12-104}$$

Here the particle Nusselt number is Nu_P, where $\mathrm{Nu}_P = h_f d_P / k_G$, and k_G is the thermal conductivity of the gas [W/(m·K)]. They stated that the heat-transfer rates predicted by this procedure were much larger than those measured on an industrial cooler, which is probably due to the particles on the inside of the cascades not experiencing the full gas velocity. Kamke and Wilson (1986) used a similar approach to model the drying of wood chips, but used the Ranz-Marshall (1952) equation to predict the heat-transfer coefficient:

$$\mathrm{Nu}_P = 2 + 0.6 \cdot \mathrm{Re}_P^{1/2} \cdot \mathrm{Pr}_G \tag{12-105}$$

where Pr_G is the Prandtl number of the gas.

Drying Time Estimates Sometimes, virtually all the drying takes place in the airborne phase. Under such circumstances, the airborne-phase residence time τ_G and the drying time are virtually the same, and the required drying time can be estimated from equivalent times in drying kinetics experiments, e.g., using a thin-layer test (Langrish, D.Phil. thesis, 1988).

An example of how to incorporate the concept of the characteristic drying curve into a design calculation is given in Example 23.

Example 23: Sizing of a Cascading Rotary Dryer The average gas velocity passing through a cocurrent, adiabatic, cascading rotary dryer is 4 m/s. The particles moving through the dryer have an average diameter of 5 mm, a solids density of 600 kg/m^{-3}, and a shape factor of 0.75. The particles enter with a moisture content of 0.50 kg/kg (dry basis) and leave with a moisture content of 0.15 kg/kg (dry basis). The drying kinetics may be assumed to be linear, with no unhindered (constant-rate) drying period. In addition, let us assume that the solids are nonhygroscopic (so that the equilibrium moisture content is zero; hygroscopic means that the equilibrium moisture content is nonzero).

The inlet humidity is 0.10 kg/kg (dry basis) due to the use of a direct-fired burner, and the ratio of the flow rates of dry solids to dry gas is unity. The gas temperature at the inlet to the dryer is 800°C, and the gas may be assumed to behave as a pure water vapor/air mixture.

The gas-phase residence time that is required was calculated in the fundamentals section to be 38.0 s.

How does this gas-phase residence time relate to the total residence time that is required and to the dryer dimensions?

Application of residence time calculations (practice): Suppose that this dryer has a slope α of 4° and a diameter D of 1.5 m, operating at a rotational speed N of 0.04 r/s. We already know that the gas velocity through the drum $U_{G\text{super}}$ is 4 m/s, and that the particles have a mean diameter d_P of 5 mm and a particle density ρ_P of 600 kg/m^3. As a first estimate, suppose that the gas density ρ_G is 1 kg/m^3 and the gas viscosity μ_G is 1.8×10^{-5} kg/(m·s).

Now $K_K/(K_{fl}\sqrt{2}) \approx 1$, $K_{fl} \approx 0.4$, $K_{fall} \approx 1$, and a is within the range of 1 to 4, say, 2.5, and $U_{P_1}^d$ is estimated by the following calculation, for Reynolds numbers up to 220.

$$\mathrm{U}_{P1}^d = 7.45 \times 10^{-4}\ \mathrm{Re}^{2.2}\ \frac{\mu U_{G\text{super}} t_a^*}{\rho_P d_P^2} \tag{12-106}$$

TABLE 12-31 Summary of the Predictions Using the Correlations for the Volumetric Heat-Transfer Coefficients of Various Authors (after Baker, 1983)

Author(s)	$U_v a$, W/(m^3·K)	
	$U_{G\text{super}_1} = 1$ m/s	$U_{G\text{super}_1} = 3$ m/s
Miller et al. (1942)		
Commercial data	248	516
Pilot-scale data	82	184
Friedman and Marshall (1949)	67	138
Saeman and Mitchell (1954)	495–1155	1032–2410
Myklestad (1963)	423	1019

Above this Reynolds number, the following equation was recommended by Matchett and Baker (1987):

$$U_{P1}^{l} = 125\,\frac{\mu U_{G\text{super}} t_a^*}{\rho_F d_P^2} \tag{12-107}$$

Here Re is the Reynolds number ($U_{G\text{super}} d_P \rho/\mu$) and t_f is the average time of flight of a particle in the airborne phase.

$$t_f = \left(\frac{2D}{g\cos\alpha}\right)^{1/2} K_{\text{fall}} \tag{12-108}$$

Substituting in the numbers gives

$$t_f = \left(\frac{2 \cdot 1.5\text{ m}}{9.81\text{ m/s}^2 \cdot \cos 4^\circ}\right)^{1/2} 1.0 = 0.554\text{ s}$$

$$\text{Re} = \frac{4\text{ m/s} \cdot 0.005\text{ m} \cdot 1\text{ kg/m}^3}{1.8 \times 10^5\,(\text{kg/m} \cdot \text{s})} = 1100$$

$$U_{P1}^{l} = 125\,\frac{1.8 \times 10^5\text{ kg/(m} \cdot \text{s)} \cdot 4\text{ m/s} \cdot 0.554\text{ s}}{(600\text{ kg/m}^3)(0.005\text{ m})^2}$$

$$= 0.332\text{ m/s}$$

$$\frac{\tau}{L} = \frac{1.1}{0.04\text{ s}^{-1} \cdot 1.5\text{ m}} \times \left[\frac{\tan 4^\circ \cdot (1 + 2.5) + \sqrt{\dfrac{1}{9.81\text{ m/s}^2 \cdot 1.5\text{ m}}}}{0.4} \cdot 0.332\text{ m/s}\right]$$

$$= 30\text{ s/m}$$

$$\frac{\tau_S}{\tau_G} = \frac{t_d}{t_f} = \frac{K_{fl}}{N}\sqrt{\frac{g}{D}} = \frac{0.4}{0.04\text{ s}^{-1}}\sqrt{\frac{9.81\text{ m/s}^2}{1.5\text{ m}}} = 25.6$$

Now, the required gas-phase residence time τ_G is 38.0 s. The ratio of solids to gas-phase residence times now gives us the required solids-phase residence time τ_S of 25.6 × 38.0 s = 972 s, and a total residence time of 972 + 38 = 1010 s. If the total residence time per unit length is 30 s/m, then the required dryer length is 1010 s/(30 s/m) = 34.2 m. The dryer length/diameter ratio is therefore 34.2 m/1.5 m = 22.8, which is significantly larger than the recommended ratio of between 5:1 and 10:1. The remedy would then be to use a larger dryer diameter and repeat these calculations. The larger dryer diameter would decrease the gas velocity, slowing the particle velocity along the drum, increasing the residence time per unit length, and hence decreasing the required drum length, to give a more normal length/diameter ratio.

Performance and Cost Data for Direct Heat Rotary Dryers Table 12-32 gives estimating-price data for direct rotary dryers employing steam-heated air. Higher-temperature operations requiring combustion chambers and fuel burners will cost more. The total installed cost of rotary dryers including instrumentation, auxiliaries, allocated building space, etc., will run from 150 to 300 percent of the purchase cost. Simple erection costs average 10 to 20 percent of the purchase cost.

Operating costs will include 5 to 10 percent of one worker's time, plus power and fuel required. Yearly maintenance costs will range from 5 to 10 percent of total installed costs. Total power for fans, dryer drive, and feed and product conveyors will be in the range of $0.5D^2$ to $1.0D^2$. Thermal efficiency of a high-temperature direct heat rotary dryer will range from 55 to 75 percent and, with steam-heated air, from 30 to 55 percent.

A representative list of materials dried in direct heat rotary dryers is given in Table 12-33.

Indirect Heat Rotary Steam-Tube Dryers Probably the most common type of indirect heat rotary dryer is the steam-tube dryer (Fig. 12-60). Steam-heated tubes running the full length of the cylinder are fastened symmetrically in one, two, or three concentric rows inside the cylinder and rotate with it. Tubes may be simple pipe with condensate draining by gravity into the discharge manifold or bayonet type. Bayonet-type tubes are also employed when units are used as water-tube coolers. When handling sticky materials, one row of tubes is preferred. These are occasionally shielded at the feed end of the dryer to prevent buildup of solids behind them. Lifting flights are usually inserted behind the tubes to promote solids agitation.

Wet feed enters the dryer through a chute or screw feeder. The product discharges through peripheral openings in the shell in ordinary dryers. These openings also serve to admit purge air to sweep moisture or other evolved gases from the shell. In practically all cases, gas flow is countercurrent to solids flow. To retain a deep bed of material within the dryer, normally 10 to 20 percent fillage, the discharge openings are supplied with removable chutes extending radially into the dryer. These, on removal, permit complete emptying of the dryer.

Steam is admitted to the tubes through a revolving steam joint into the steam side of the manifold. Condensate is removed continuously, by gravity through the steam joint to a condensate receiver and by means of lifters in the condensate side of the manifold. By employing simple tubes, noncondensables are continuously vented at the other ends of the tubes through Sarco-type vent valves mounted on an auxiliary manifold ring, also revolving with the cylinder.

Vapors (from drying) are removed at the feed end of the dryer to the atmosphere through a natural-draft stack and settling chamber or wet scrubber. When employed in simple drying operations with 3.5×10^5 to

TABLE 12-32 Warm-Air Direct-Heat Cocurrent Rotary Dryers: Typical Performance Data*

Dryer size, m × m	1.219 × 7.62	1.372 × 7.621	1.524 × 9.144	1.839 × 10.668	2.134 × 12.192	2.438 × 13.716	3.048 × 16.767
Evaporation, kg/h	136.1	181.4	226.8	317.5	408.2	544.3	861.8
Work, 10^8 J/h	3.61	4.60	5.70	8.23	1.12	1.46	2.28
Steam, kg/h at kg/m² gauge	317.5	408.2	521.6	725.7	997.9	131.5	2041
Discharge, kg/h	408	522	685	953	1270	1633	2586
Exhaust velocity, m/min	70	70	70	70	70	70	70
Exhaust volume, m³/min	63.7	80.7	100.5	144.4	196.8	257.7	399.3
Exhaust fan, kW	3.7	3.7	5.6	7.5	11.2	18.6	22.4
Dryer drive, kW	2.2	5.6	5.6	7.5	14.9	18.6	37.3
Shipping weight, kg	7700	10,900	14,500	19,100	35,800	39,900	59,900
Price, FOB Chicago	$158,000	$168,466	$173,066	$204,400	$241,066	$298,933	$393,333

*Courtesy of Swenson Process Equipment Inc.

NOTE:
Material: heat-sensitive solid
Maximum solids temperature: 65°C
Feed conditions: 25 percent moisture, 27°C
Product conditions: 0.5 percent moisture, 65°C
Inlet-air temperature: 165°C
Exit-air temperature: 71°C
Assumed pressure drop in system: 200 mm
System includes finned air heaters, transition piece, dryer, drive, product collector, duct, and fan.
Prices are for carbon steel construction and include entire dryer system (November, 1994).
For 304 stainless-steel fabrication, multiply the prices given by 1.5.

TABLE 12-33 Representative Materials Dried in Direct-Heat Rotary Dryers*

Material dried	Moisture content, % (wet basis)		Heat efficiency, %
	Initial	Final	
High-temperature:			
Sand	10	0.5	61
Stone	6	0.5	65
Fluorspar	6	0.5	59
Sodium chloride (vacuum salt)	3	0.04	70–80
Sodium sulfate	6	0.1	60
Ilmenite ore	6	0.2	60–65
Medium-temperature:			
Copperas	7	1 (moles)	55
Ammonium sulfate	3	0.10	50–60
Cellulose acetate	60	0.5	51
Sodium chloride (grainer salt)	25	0.06	35
Cast-iron borings	6	0.5	50–60
Styrene	5	0.1	45
Low-temperature:			
Oxalic acid	5	0.2	29
Vinyl resins	30	1	50–55
Ammonium nitrate prills	4	0.25	30–35
Urea prills	2	0.2	20–30
Urea crystals	3	0.1	50–55

*Taken from *Chem. Eng.,* June 19, 1967, p. 190, Table III.

10×10^5 Pa steam, draft is controlled by a damper to admit only sufficient outside air to sweep moisture from the cylinder, discharging the air at 340 to 365 K and 80 to 90 percent saturation. In this way, shell gas velocities and dusting are minimized. When used for solvent recovery or other processes requiring a sealed system, sweep gas is recirculated through a scrubber-gas cooler and blower.

Steam manifolds for pressures up to 10^6 Pa are of cast iron. For higher pressures, the manifold is fabricated from plate steel, staybolted, and welded. The tubes are fastened rigidly to the manifold faceplate and are supported in a close-fitting annular plate at the other end to permit expansion. Packing on the steam neck is normally graphite asbestos. Ordinary rotating seals are similar in design with allowance for the admission of small quantities of outside air when the dryer is operated under a slight negative internal pressure.

Steam-tube dryers are used for the continuous drying, heating, or cooling of granular or powdery solids which cannot be exposed to ordinary atmospheric or combustion gases. They are especially suitable for fine dusty particles because of the low gas velocities required for purging of the cylinder. Tube sticking is avoided or reduced by employing recycle, shell knockers, etc., as previously described; tube scaling by sticky solids is one of the major hazards to efficient operation. The dryers are suitable for drying, solvent recovery, and chemical reactions. Steam-tube units have found effective employment in soda ash production, replacing more expensive indirect-heat rotary calciners.

Special types of steam-tube dryers employ packed and purged seals on all rotating joints, with a central solids-discharge manifold through the steam neck to reduce the seal diameter. This manifold contains the product discharge conveyor and a passage for the admission of sweep gas. Solids are removed from the shell by special volute lifters and dropped into the discharge conveyor. Units have been fabricated for operation at 76 mm of water, internal shell pressure, with no detectable air leakage.

Design methods for indirect heat rotary steam-tube dryers Heat-transfer coefficients in steam-tube dryers range from 30 to 85 $W/(m^2 \cdot K)$. Coefficients will increase with increasing steam temperature because of increased heat transfer by radiation. In units carrying saturated steam at 420 to 450 K, the heat flux UT will range from 6300 W/m^2 for difficult-to-dry and organic solids to 1890 to 3790 W/m^2 for finely divided inorganic materials. The effect of steam pressure on heat-transfer rates up to 8.6×10^5 Pa is illustrated in Fig. 12-61.

Performance and cost data for indirect heat rotary steam-tube dryers Table 12-34 contains data for a number of standard sizes of steam-tube dryers. Prices tabulated are for ordinary carbon steel construction. Installed costs will run from 150 to 300 percent of purchase cost.

The thermal efficiency of steam-tube units will range from 70 to 90 percent, if a well-insulated cylinder is assumed. This does not allow for boiler efficiency, however, and is therefore not directly comparable with direct heat units such as the direct heat rotary dryer or indirect heat calciner.

Operating costs for these dryers include 5 to 10 percent of one person's time. Maintenance will average 5 to 10 percent of total installed cost per year.

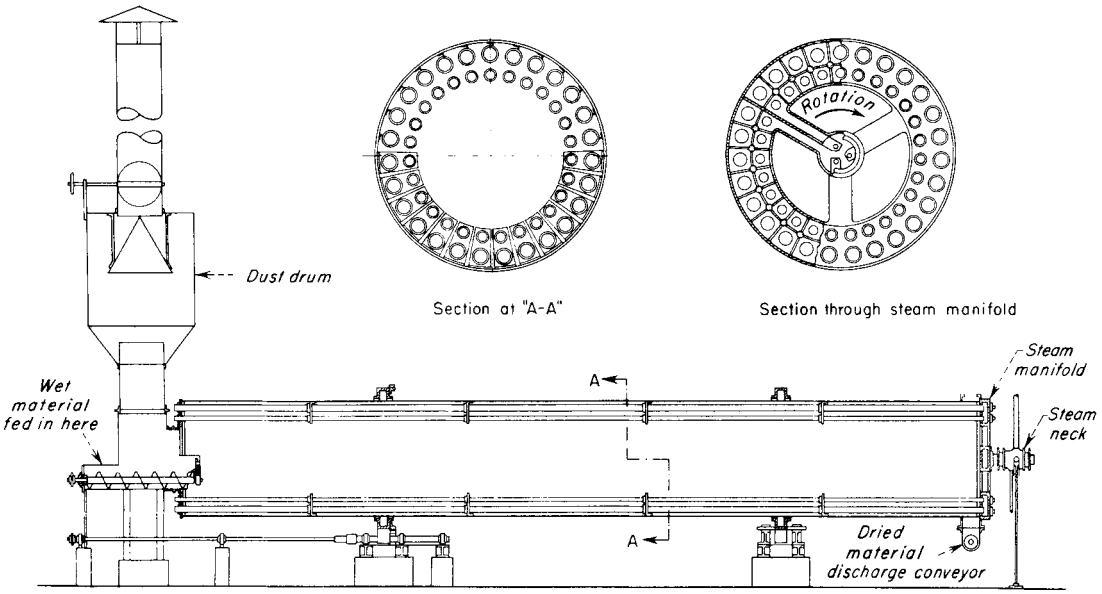

FIG. 12-60 Steam-tube rotary dryer.

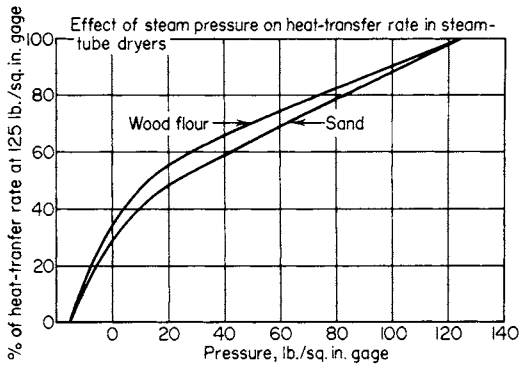

FIG. 12-61 Effect of steam pressure on the heat-transfer rate in steam-tube dryers.

Table 12-35 outlines typical performance data from three drying applications in steam-tube dryers.

Indirect Rotary Calciners and Kilns These large-scale rotary processors are used for very high temperature operations. Operation is similar to that of rotary dryers. For additional information, refer to Perry's 7th Edition, pages 12-56 to 12-58.

Indirect Heat Calciners Indirect heat rotary calciners, either batch or continuous, are employed for heat treating and drying at higher temperatures than can be obtained in steam-heated rotating equipment. They generally require a minimum flow of gas to purge the cylinder, to reduce dusting, and are suitable for gas-sealed operation with oxidizing, inert, or reducing atmospheres. Indirect calciners are widely utilized, and some examples of specific applications are as follows:

1. Activating charcoal
2. Reducing mineral high oxides to low oxides
3. Drying and devolatilizing contaminated soils and sludges
4. Calcination of alumina oxide-based catalysts
5. Drying and removal of sulfur from cobalt, copper, and nickel
6. Reduction of metal oxides in a hydrogen atmosphere
7. Oxidizing and "burning off" of organic impurities
8. Calcination of ferrites

This unit consists essentially of a cylindrical retort, rotating within a stationary insulation-lined furnace. The latter is arranged so that fuel combustion occurs within the annular ring between the retort and the furnace. The retort cylinder extends beyond both ends of the furnace. These end extensions carry the riding rings and drive gear. Material may be fed continuously at one end and discharged continuously at the other. Feeding and solids discharging are usually accomplished with screw feeders or other positive feeders to prevent leakage of gases into or out of the calciner.

In some cases in which it is desirable to cool the product before removal to the outside atmosphere, the discharge end of the cylinder is provided with an additional extension, the exterior of which is water-spray-cooled. In cocurrent flow calciners, hot gases from the interior of the heated portion of the cylinder are withdrawn through a special extraction tube. This tube extends centrally through the cooled section to prevent flow of gas near the cooled-shell surfaces and possible condensation. Frequently a separate cooler is used, isolated from the calciner by an air lock.

To prevent sliding of solids over the smooth interior of the shell, agitating flights running longitudinally along the inside wall are frequently provided. These normally do not shower the solids as in a direct heat vessel but merely prevent sliding so that the bed will turn over and constantly expose new surface for heat and mass transfer. To prevent scaling of the shell interior by sticky solids, cylinder scraper and knocker arrangements are occasionally employed. For example, a scraper chain is fairly common practice in soda ash calciners, while knockers are frequently utilized on metallic-oxide calciners.

Because indirect heat calciners frequently require close-fitting gas seals, it is customary to support all parts on a self-contained frame, for sizes up to approximately 2 m in diameter. The furnace can employ electric heating elements or oil and/or gas burners as the heat source for the process. The hardware would be zoned down the length of the furnace to match the heat requirements of the process. Process control is normally by shell temperature, measured by thermocouples or radiation pyrometers. When a special gas atmosphere must be maintained inside the cylinder, positive rotary gas seals, with one or more pressurized and purged annular chambers, are employed. The diaphragm-type seal is suitable for pressures up to 5 cm of water, with no detectable leakage.

In general, the temperature range of operation for indirect heat calciners can vary over a wide range, from 475 K at the low end to approximately 1475 K at the high end. All types of carbon steel, stainless, and

TABLE 12-34 Standard Steam-Tube Dryers*

Size, diameter × length, m	Tubes		m^2 of free area	Dryer speed, r/min	Motor size, hp	Shipping weight, kg	Estimated price
	No. OD (mm)	No. OD (mm)					
0.965 × 4.572	14 (114)		21.4	6	2.2	5,500	$152,400
0.965 × 6.096	14 (114)		29.3	6	2.2	5,900	165,100
0.965 × 7.620	14 (114)		36.7	6	3.7	6,500	175,260
0.965 × 9.144	14 (114)		44.6	6	3.7	6,900	184,150
0.965 × 10.668	14 (114)		52.0	6	3.7	7,500	196,850
1.372 × 6.096	18 (114)	18 (63.5)	58.1	4.4	3.7	10,200	203,200
1.372 × 7.620	18 (114)	18 (63.5)	73.4	4.4	3.7	11,100	215,900
1.372 × 9.144	18 (114)	18 (63.5)	88.7	5	5.6	12,100	228,600
1.372 × 10.668	18 (114)	18 (63.5)	104	5	5.6	13,100	243,840
1.372 × 12.192	18 (114)	18 (63.5)	119	5	5.6	14,200	260,350
1.372 × 13.716	18 (114)	18 (63.5)	135	5.5	7.5	15,000	273,050
1.829 × 7.62	27 (114)	27 (76.2)	118	4	5.6	19,300	241,300
1.829 × 9.144	27 (114)	27 (76.2)	143	4	5.6	20,600	254,000
1.829 × 10.668	27 (114)	27 (76.2)	167	4	7.5	22,100	266,700
1.829 × 12.192	27 (114)	27 (76.2)	192	4	7.5	23,800	278,400
1.829 × 13.716	27 (114)	27 (76.2)	217	4	11.2	25,700	292,100
1.829 × 15.240	27 (114)	27 (76.2)	242	4	11.2	27,500	304,800
1.829 × 16.764	27 (114)	27 (76.2)	266	4	14.9	29,300	317,500
1.829 × 18.288	27 (114)	27 (76.2)	291	4	14.9	30,700	330,200
2.438 × 12.192	90 (114)		394	3	11.2	49,900	546,100
2.438 × 15.240	90 (114)		492	3	14.9	56,300	647,700
2.438 × 18.288	90 (114)		590	3	14.9	63,500	736,600
2.438 × 21.336	90 (114)		689	3	22.4	69,900	838,200
2.438 × 24.387	90 (114)		786	3	29.8	75,300	927,100

*Courtesy of Swenson Process Equipment Inc. (prices from November, 1994). Carbon steel fabrication; multiply by 1.75 for 304 stainless steel.

TABLE 12-35 Steam-Tube Dryer Performance Data

	Class 1	Class 2	Class 3
Class of materials handled	High-moisture organic, distillers' grains, brewers' grains, citrus pulp	Pigment filter cakes, blanc fixe, barium carbonate, precipitated chalk	Finely divided inorganic solids, water-ground mica, water-ground silica, flotation concentrates
Description of class	Wet feed is granular and damp but not sticky or muddy and dries to granular meal	Wet feed is pasty, muddy, or sloppy; product is mostly hard pellets	Wet feed is crumbly and friable; product is powder with very few lumps
Normal moisture content of wet feed, % dry basis	233	100	54
Normal moisture content of product, % dry basis	11	0.15	0.5
Normal temperature of wet feed, K	310–320	280–290	280–290
Normal temperature of product, K	350–355	380–410	365–375
Evaporation per product, kg	2	1	0.53
Heat load per lb product, kJ	2250	1190	625
Steam pressure normally used, kPa gauge	860	860	860
Heating surface required per kg product, m^2	0.34	0.4	0.072
Steam consumption per kg product, kg	3.33	1.72	0.85

alloy construction are used, depending upon temperature, process, and corrosion requirements. Fabricated-alloy cylinders can be used over the greater part of the temperature range; however, the greater creep-stress abilities of cast alloys makes their use desirable for the highest calciner cylinder temperature applications.

Design methods for calciners In indirect heat calciners, heat transfer is primarily by radiation from the cylinder wall to the solids bed. The thermal efficiency ranges from 30 to 65 percent. By utilization of the furnace exhaust gases for preheated combustion air, steam production, or heat for other process steps, the thermal efficiency can be increased considerably. The limiting factors in heat transmission lie in the conductivity and radiation constants of the shell metal and solids bed. If the characteristics of these are known, equipment may be accurately sized by employing the Stefan-Boltzmann radiation equation. Apparent heat-transfer coefficients will range from 17 $W/(m^2 \cdot K)$ in low-temperature operations to 85 $W/(m^2 \cdot K)$ in high-temperature processes.

Cost data for calciners Power, operating, and maintenance costs are similar to those previously outlined for direct and indirect heat rotary dryers. Estimating purchase costs for preassembled and frame-mounted rotary calciners with carbon steel and type 316 stainless-steel cylinders are given in Table 12-36 together with size, weight, and motor requirements. Sale price includes the cylinder, ordinary angle seals, furnace, drive, feed conveyor, burners, and controls. Installed cost may be estimated, not including building or foundation costs, at up to 50 percent of the purchase cost. A layout of a typical continuous calciner with an extended cooler section is illustrated in Fig. 12-62.

Small batch retorts, heated electrically or by combustion, are widely used as carburizing furnaces and are applicable also to chemical processes involving the heat treating of particulate solids. These are mounted on a structural-steel base, complete with cylinder, furnace, drive motor, burner, etc. Units are commercially available in diameters from 0.24 to 1.25 m and lengths of 1 to 2 m. Continuous retorts with helical internal spirals are employed for metal heat-treating purposes. Precise retention control is maintained in these operations. Standard diameters are 0.33, 0.5, and 0.67 m with effective lengths up to 3 m. These vessels are employed in many small-scale chemical process operations which require accurate control of retention. Their operating characteristics and applications are identical to those of the larger indirect heat calciners.

Direct Heat Roto-Louvre Dryer One of the more important special types of rotating equipment is the Roto-Louvre dryer. As illustrated in Fig. 12-63, hot air (or cooling air) is blown through louvres in a double-wall rotating cylinder and up through the bed of solids. The latter moves continuously through the cylinder as it rotates. Constant turnover of the bed ensures uniform gas contacting for heat and mass transfer. The annular gas passage behind the louvres is partitioned so that contacting air enters the cylinder only beneath the solids bed. The number of louvres covered at any one time is roughly 30 percent. Because air circulates through the bed, fillages of 13 to 15 percent or greater are employed.

Roto-Louvre dryers range in size from 0.8 to 3.6 m in diameter and from 2.5 to 11 m long. The largest unit is reported capable of evaporating 5500 kg/h of water. Hot gases from 400 to 865 K may be employed. Because gas flow is through the bed of solids, high pressure drop, from 7 to 50 cm of water, may be encountered within the shell. For this reason, both a pressure inlet fan and an exhaust fan are provided in most applications to maintain the static pressure within the equipment as closely as possible to atmospheric. This prevents excessive in-leakage or blowing of hot gas and dust to the outside. For pressure control, one fan is usually operated under fixed conditions, with an automatic damper control on the other, regulated by a pressure detector-controller.

In heating or drying applications, when cooling of the product is desired before discharge to the atmosphere, cool air is blown through

TABLE 12-36 Indirect-Heat Rotary Calciners: Sizes and Purchase Costs*

Diameter, ft	Overall cylinder length	Heated cylinder length	Cylinder drive motor hp	Approximate Shipping weight, lb	Approximate sale price in carbon steel construction†	Approximate sale price in No. 316 stainless construction
4	40 ft	30 ft	7.5	50,000	$275,000	$325,000
5	45 ft	35 ft	10	60,000	375,000	425,000
6	50 ft	40 ft	20	75,000	475,000	550,000
7	60 ft	50 ft	30	90,000	550,000	675,000

* ABB Raymond (Bartlett-Snow™).
† Prices for November, 1994.

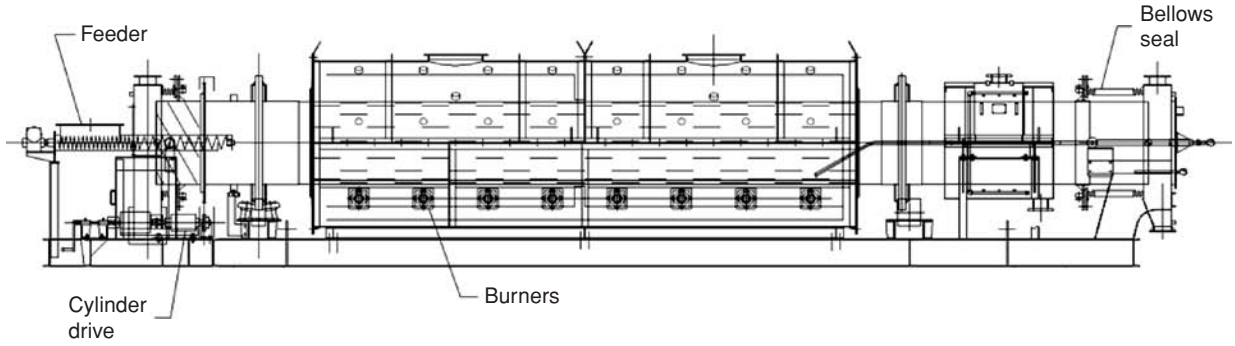

FIG. 12-62 Gas-fired rotary calciner with integral cooler. (*Air Preheater Company, Raymond® & Bartlett Snow™ Products.*)

a second annular space, outside the inlet hot-air annulus, and released through the louvres at the solids discharge end of the shell.

Roto-Louvre dryers are suitable for processing coarse granular solids which do not offer high resistance to airflow, do not require intimate gas contacting, and do not contain significant quantities of dust.

Heat transfer and mass transfer from the gas to the surface of the solids are extremely efficient; hence the equipment size required for a given duty is frequently less than that required when an ordinary direct heat rotary vessel with lifting flights is used. Purchase price savings are partially balanced, however, by the more complex construction of the Roto-Louvre unit. A Roto-Louvre dryer will have a capacity roughly 1.5 times that of a single-shell rotary dryer of the same size under equivalent operating conditions. Because of the cross-flow method of heat exchange, the average t is not a simple function of inlet and outlet t's. There are currently no published data which permit the sizing of equipment without pilot tests as recommended by the manufacturer. Three applications of Roto-Louvre dryers are outlined in Table 12-37. Installation, operating, power, and maintenance costs will be similar to those experienced with ordinary direct heat rotary dryers. *Thermal efficiency* will range from 30 to 70 percent.

Additional Reading

Aiken and Polsak, "A Model for Rotary Dryer Computation," in Mujumdar (ed.), *Drying '82*, Hemisphere, New York, 1982, pp. 32–35.
Baker, "Cascading Rotary Dryers," Chap. 1 in Mujumdar (ed.), *Advances in Drying*, vol. 2, pp. 1–51, Hemisphere, New York, 1983.
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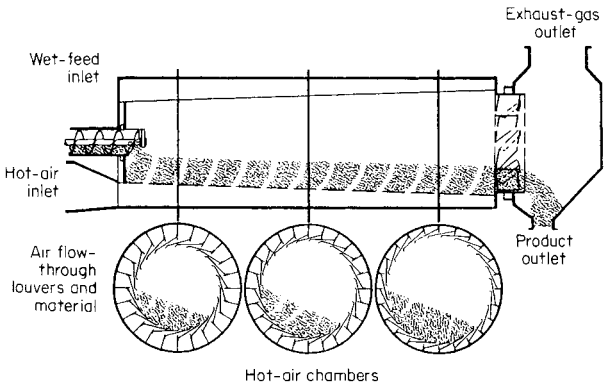

FIG. 12-63 FMC Link-Belt Roto-Louvre Dryer.

TABLE 12-37 Manufacturer's Performance Data for FMC Link-Belt Roto-Louvre Dryers*

Material dried	Ammonium sulfate	Foundry sand	Metallurgical coke
Dryer diameter	2 ft 7 in	6 ft 4 in	10 ft 3 in
Dryer length	10 ft	24 ft	30 ft
Moisture in feed, % wet basis	2.0	6.0	18.0
Moisture in product, % wet basis	0.1	0.5	0.5
Production rate, lb/h	2500	32,000	38,000
Evaporation rate, lb/h	50	2130	8110
Type of fuel	Steam	Gas	Oil
Fuel consumption	255 lb/h	4630 ft^3/h	115 gal/h
Calorific value of fuel	837 Btu/lb	1000 Btu/ft^3	150,000 Btu/gal
Efficiency, Btu, supplied per lb evaporation	4370	2170	2135
Total power required, hp	4	41	78

*Material Handling Systems Division, FMC Corp. To convert British thermal units to kilojoules, multiply by 1.06; to convert horsepower to kilowatts, multiply by 0.746.

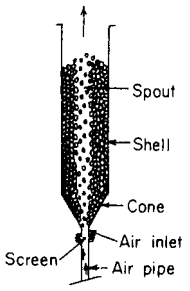

FIG. 12-64 Schematic diagram of sported bed. [*Mathur and Gishler,* Am. Inst. Chem. Eng. J., ***1,*** 2, *15 (1955).*]

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Fluidized and Spouted Bed Dryers

Spouted Beds The spouted bed technique was developed primarily for solids which are too coarse to be handled in fluidized beds.

Although their applications overlap, the methods of gas-solids mixing are completely different. A schematic view of a spouted bed is given in Fig. 12-64. Mixing and gas-solids contacting are achieved first in a fluid "spout," flowing upward through the center of a loosely packed bed of solids. Particles are entrained by the fluid and conveyed to the top of the bed. They then flow downward in the surrounding annulus as in an ordinary gravity bed, countercurrently to gas flow. The mechanisms of gas flow and solids flow in spouted beds were first described by Mathur and Gishler [*Am. Inst. Chem. Eng. J.* **1**(2): 157–164 (1955)]. Drying studies have been carried out by Cowan [*Eng. J.* **41:**5, 60–64 (1958)], and a theoretical equation for predicting the minimum fluid velocity necessary to initiate spouting was developed by Madonna and Lama [*Am. Inst. Chem. Eng. J.* **4**(4):497 (1958)]. Investigations to determine maximum spoutable depths and to develop theoretical relationships based on vessel geometry and operating variables have been carried out by Lefroy [*Trans. Inst. Chem. Eng.* **47**(5):T120–128 (1969)] and Reddy [*Can. J. Chem. Eng.* **46**(5):329–334 (1968)].

Gas flow in a spouted bed is partially through the spout and partially through the annulus. About 30 percent of the gas entering the system immediately diffuses into the downward-flowing annulus. Near the top of the bed, the quantity in the annulus approaches 66 percent of the total gas flow; the gas flow through the annulus at any point in the bed equals that which would flow through a loosely packed solids bed under the same conditions of pressure drop. Solids flow in the annulus is both downward and slightly inward. As the fluid spout rises in the bed, it entrains more and more particles, losing velocity and gas into the annulus. The volume of solids displaced by the spout is roughly 6 percent of the total bed.

On the basis of experimental studies, Mathur and Gishler derived an empirical correlation to describe the minimum fluid flow necessary for spouting, in 3- to 12-in-diameter columns:

$$u = \frac{D_p}{D_c}\left(\frac{D_o}{D_c}\right)^{0.33}\left[\frac{2gL(\rho_s - \rho_f)}{\rho_f}\right]^{0.5} \qquad (12\text{-}109)$$

where u = superficial fluid velocity through the bed, ft/s; D_p = particle diameter, ft; D_c = column (or bed) diameter, ft; D_o = fluid inlet orifice diameter, ft; L = bed height, ft; ρ_s absolute solids density, lb/ft^3; ρ_f fluid density, lb/ft^3; and g = 32.2 ft/s^2, gravity acceleration. To convert feet per second to meters per second, multiply by 0.305; to convert pounds per cubic foot to kilograms per cubic meter, multiply by 16. In SI units, g = 9.8 m/s^2. The inlet orifice diameter, air rate, bed diameter, and bed depth were all found to be critical and interdependent:

1. In a given-diameter bed, deeper beds can be spouted as the gas inlet orifice size is decreased. Using air, a 12-in-diameter bed containing 0.125- by 0.250-in wheat can be spouted at a depth of over 100 in with a 0.8-in orifice, but at only 20 in with a 2.4-in orifice.
2. Increasing bed diameter increases spoutable depth. By employing a bed/orifice diameter ratio of 12 for air spouting, a 9-in-diameter bed was spouted at a depth of 65 in while a 12-in-diameter bed was spouted at 95 in.
3. As indicated by Eq. (12-109), the superficial fluid velocity required for spouting increases with bed depth and orifice diameter and decreases as the bed diameter is increased.

Employing wood chips, Cowan's drying studies indicated that the volumetric heat-transfer coefficient obtainable in a spouted bed is at least twice that in a direct heat rotary dryer. By using 20- to 30-mesh Ottawa sand, fluidized and spouted beds were compared. The volumetric coefficients in the fluid bed were 4 times those obtained in a spouted bed. Mathur dried wheat continuously in a 12-in-diameter spouted bed, followed by a 9-in-diameter spouted bed cooler. A drying rate of roughly 100 lb/h of water was obtained by using 450 K inlet air. Six hundred pounds per hour of wheat was reduced from 16 to 26 percent to 4 percent moisture. Evaporation occurred also in the cooler by using sensible heat present in the wheat. The maximum drying bed temperature was 118°F, and the overall thermal efficiency of the system was roughly 65 percent. Some aspects of the spouted bed technique are covered by patent (U.S. Patent 2,786,280).

Cowan reported that significant size reduction of solids occurred when cellulose acetate was dried in a spouted bed, indicating its possible limitations for handling other friable particles.

Direct Heat Vibrating Conveyor Dryers Information on vibrating conveyors and their mechanical construction is given in Sec. 19, "Solid-Solid Operations and Equipment." The vibrating conveyor dryer is a modified form of fluidized-bed equipment, in which fluidization is maintained by a combination of pneumatic and mechanical forces. The heating gas is introduced into a plenum beneath the conveying deck through ducts and flexible hose connections and passes up through a screen, perforated, or slotted conveying deck, through the fluidized bed of solids, and into an exhaust hood (Fig. 12-65). If ambient air is employed for cooling, the sides of the plenum may be open and a simple exhaust system used; however, because the gas distribution plate may be designed for several inches of water pressure drop to ensure a uniform velocity distribution through the bed of solids, a combination pressure-blower exhaust-fan system is desirable to balance the pressure above the

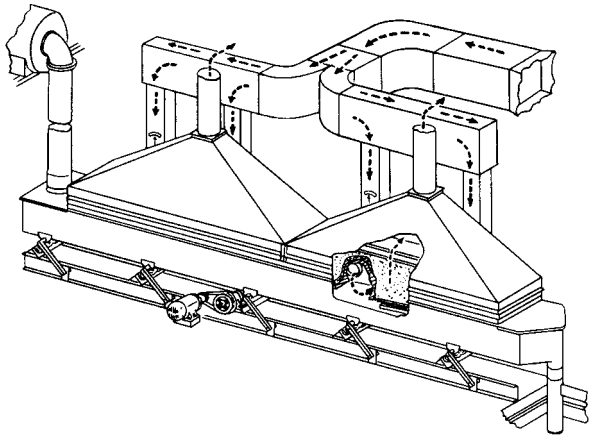

FIG. 12-65 Vibrating conveyor dryer. (*Carrier Vibrating Equipment, Inc.*)

TABLE 12-38 Table for Estimating Maximum Superficial Air Velocities through Vibrating-Conveyor Screens*

Mesh size	Velocity, m/s	
	2.0 specific gravity	1.0 specific gravity
200	0.22	0.13
100	0.69	0.38
50	1.4	0.89
30	2.6	1.8
20	3.2	2.5
10	6.9	4.6
5	11.4	7.9

*Carrier Vibrating Equipment, Inc.

deck with the outside atmosphere and prevent gas in-leakage or blowing at the solids feed and exit points.

Units are fabricated in widths from 0.3 to 1.5 m. Lengths are variable from 3 to 50 m; however, most commercial units will not exceed a length of 10 to 16 m per section. Power required for the vibrating drive will be approximately 0.4 kW/m^2 of deck.

In general, this equipment offers an economical heat-transfer area for first cost as well as operating cost. Capacity is limited primarily by the air velocity which can be used without excessive dust entrainment. Table 12-38 shows limiting air velocities suitable for various solids particles. Usually, the equipment is satisfactory for particles larger than 100 mesh in size. [The use of indirect heat conveyors eliminates the problem of dust entrainment, but capacity is limited by the heat-transfer coefficients obtainable on the deck (see Sec. 11)].

When a stationary vessel is employed for fluidization, all solids being treated must be fluidized; nonfluidizable fractions fall to the bottom of the bed and may eventually block the gas distributor. The addition of mechanical vibration to a fluidized system offers the following advantages:

1. Equipment can handle nonfluidizable solids fractions. Although these fractions may drop through the bed to the screen, directional-throw vibration will cause them to be conveyed to the discharge end of the conveyor. Prescreening or sizing of the feed is less critical than in a stationary fluidized bed.
2. Because of mechanical vibration, incipient channeling is reduced.
3. Fluidization may be accomplished with lower pressures and gas velocities. This has been evidenced on vibratory units by the fact that fluidization stops when the vibrating drive is stopped. Vibrating conveyor dryers are suitable for free-flowing solids containing mainly surface moisture. Retention is limited by conveying speeds which range from 0.02 to 0.12 m/s. Bed depth rarely exceeds 7 cm, although units are fabricated to carry 30- to 46-cm-deep beds; these also employ plate and pipe coils suspended in the bed to provide additional heat-transfer area. Vibrating dryers are not suitable for fibrous materials which mat or for sticky solids which may ball or adhere to the deck.

For estimating purposes for direct heat drying applications, it can be assumed that the average exit gas temperature leaving the solids bed will approach the final solids discharge temperature on an ordinary unit carrying a 5- to 15-cm-deep bed. Calculation of the heat load and selection of an inlet air temperature and superficial velocity (Table 12-38) will then permit approximate sizing, provided an approximation of the minimum required retention time can be made.

Vibrating conveyors employing direct contacting of solids with hot, humid air have also been used for the agglomeration of fine powders, chiefly for the preparation of agglomerated water-dispersible food products. Control of inlet air temperature and dew point permits the uniform addition of small quantities of liquids to solids by condensation on the cool incoming-particle surfaces. The wetting section of the conveyor is followed immediately by a warm-air-drying section and particle screening.

Fluidized-Bed Dryers The basic principles of fluid-bed technology are thoroughly described in Sec. 17, "Gas-Solid Operations and Equipment." Originally conceived as a heterogeneous chemical reactor, the use of this technology in connection with drying processes has increased considerably during the last several decades. The technology offers the following advantages when compared with other drying methods:

- It has no moving parts.
- It provides rapid heat and mass exchange between gas and particles.
- It provides high heat-transfer rates between the gas/particle bed and immersed objects such as heating panels.
- It provides intensive mixing of solids, leading to homogeneous conditions and reliable control of the drying process.

The fluid-bed technology can be applied to continuous as well as batch processes.

As described in Sec. 17, the process parameter of the highest importance is the gas velocity in the fluidized bed, referred to as the fluidizing velocity or the superficial gas velocity. This velocity is of nominal character since the flow field will be disturbed and distorted by the presence of the solid phase and the turbulent fluctuations created by the gas/solid interaction.

The fluid bed consists of a layer of particles suspended partly by a bed plate with perforations or a grid and partly by the fluidizing gas flowing through the bed plate. If the gas velocity is low, the gas will merely percolate through a bed of particles that appear to be fixed. At a higher velocity the particles will start to move under influence of the aerodynamic forces, and at the point where the pressure drop reaches the equivalent of the weight per unit of area, all the particles will tend to be moving in suspension, also called the incipiently fluidized state. The particle layer behaves as a liquid, and the bed volume expands considerably. At even higher velocities the motion will be stronger, and the excess gas flow will tend to appear as bubbles. In this state the particle layer will undergo vigorous mixing, while still appearing as a dense layer of fluidlike material or a boiling liquid. If the gas velocity is further increased, the solid phase will change into a slugging mode where gas bubbles throw lumps of solid material away from the bed surface. Even further increase will result in the solid phase being entrained by the gas flow and will appear as a lean phase undergoing transport or movement by the gas phase. Most fluid-bed drying processes are adjusted to operate safely below slugging conditions. Figure 12-66 shows a view into an operating drying fluid bed.

FIG. 12-66 Fluid bed for drying in operation. (*Niro A/S.*)

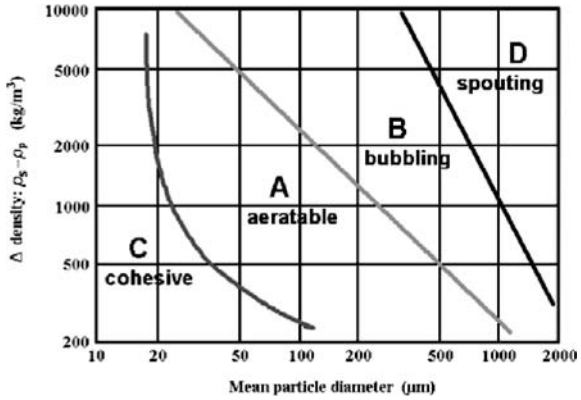

FIG. 12-67 Geldart diagram.

Proper design and operation of a fluid bed installation for drying requires consideration of several important topics. Among these are

- Ability of the material to be fluidized
- Drying characteristics of the material
- The fluidization velocity
- The design of the fluid-bed plate
- The operating conditions
- The mode of operation

Ability of the material to be fluidized has been investigated by Geldart, resulting in the well-known Geldart diagram, a version of which is shown in Fig. 12-67. The general knowledge to be derived from the Geldart diagram is that particulate material can be handled successfully in a fluid bed only if it is not too fine or too coarse. It must also have flowability. Fluid beds are best suited for particles that are regular in shape, not too sticky, and with a mean particle size between 20 μm and 10 mm. Particles of needle- or leaflike shape should be considered as nonfluidizable.

Drying characteristics of the material can be difficult to determine, but a test in a small batch fluid bed can reveal the drying curve of the material, as shown in Fig. 12-68. The drying curve clearly shows that the surface moisture is rapidly evaporated while the material is maintained at a low temperature close to the wet-bulb temperature of the drying gas. At a certain time the surface water has disappeared, and the so-called transition point has been reached. From here on the drying rate is controlled by internal diffusion inside the material, and the drying curve becomes characteristic for the individual material. While the moisture content of the material decreases, the bed temperature increases while approaching the inlet temperature of the drying air. The total drying time to reach the final moisture and the heat sensitivity of the material are important parameters for design of an industrial plant.

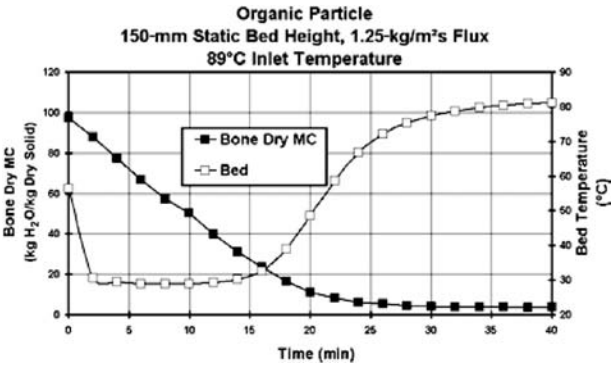

FIG. 12-68 Drying curve of organic material.

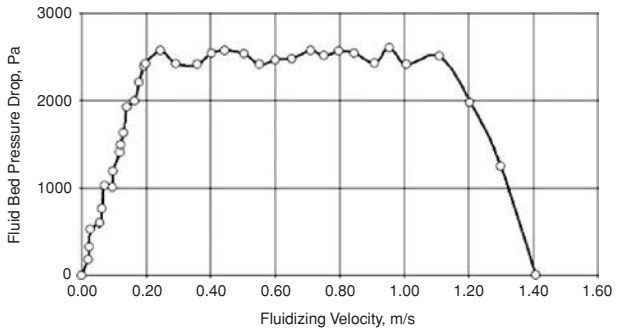

FIG. 12-69 Fluid-bed pressure drop versus fluidizing velocity. (*Niro A/S.*)

The fluidization velocity is of major importance, as indicated in the introduction. Each material will have individual requirements for the gas velocity and pressure drop to provide good fluidization. An investigation of the relationship between fluidization velocity and bed pressure drop for a given material may result in a diagram such as shown in Fig. 12-69. The results are illustrative and intended to give a clear picture of the relationship. The minimum fluidization velocity may be calculated from the Wen and Yu correlation given in Sec. 17.

At a fluidizing velocity below the value required for minimum or incipient fluidization, the pressure drop over the bed will increase proportionally with the velocity. Above a certain critical velocity, the pressure drop corresponds to the weight of the fluidized mass of material and remains roughly at this value even at higher velocities. The critical velocity for a given material may be estimated by methods mentioned in Sec. 17. At a much higher value of the fluidizing velocity, the material in the bed ceases to appear as a moving layer, and it is gradually carried away. Accordingly the pressure drop falls to zero. The fluidizing velocity value that will serve a drying task best cannot be derived exactly from the diagram. However, as a general recommendation, a value between the critical value and the value where the pressure drop falls off will be right. A first choice could be a factor of 2 to 5 times the minimum fluidization velocity. Further clarification must be derived from test work with the actual material.

The **design of the fluid-bed plate** is important for several reasons. First, the plate is responsible for the distribution of the drying or fluidization gas. This requires an even pattern of orifices in the plate and a sufficient pressure drop over the plate. As a general rule, the following guideline may be recommended:

$$\Delta P_{\text{plate}} = \tfrac{1}{2}\,\Delta P_{\text{powder}}$$

with the following limits: ΔP_{plate} minimum 500 Pa, ΔP_{plate} maximum 2500 Pa.

The estimation of the pressure drop in design situations may be difficult except for the case of the traditional perforated sheet with cylindrical holes perpendicular to the plane of the plate, as shown in Fig. 12-70. For this type of plate the formula of McAllister et al. may be useful. A calculation using this formula will show that a plate giving a required pressure drop of 1500 Pa and a typical fluidizing velocity of 0.35 m/s will need an open area of roughly 1 percent. Provided by a plate of 1-mm thickness and 1-mm-diameter holes, this requires approximately 12,500 holes per square meter.

FIG. 12-70 Traditional perforated plate for fluid-bed application.

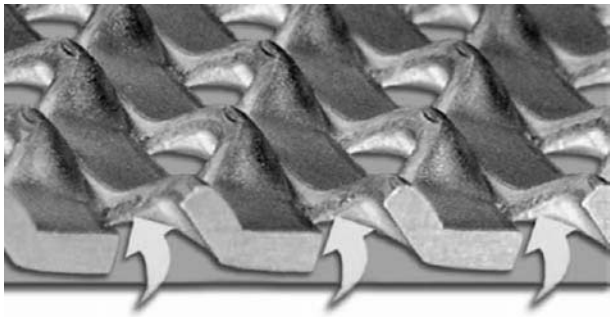

FIG. 12-71 Conidur® plate for fluid-bed application. (*Hein, Lehmann Trenn- und Fördertechnik GmbH.*)

However, this type of plate is being replaced in most fluid-bed applications due to its inherent disadvantages, which are caused by the difficulties of punching holes of smaller diameter than the thickness of the plate itself. The result is that the plates are weak and are prone to sifting back of the finer particles. The perpendicular flow pattern also means that the plate does not provide a transport capacity for lumps of powder along the plane of the plate.

This transport capacity is provided by plate of so-called gill types of which there are two distinct categories. One category is the type where plates are punched in a very fine regular pattern, not only to provide holes or orifices but also to deform the plate so that each orifice acquires a shape suited for acceleration of the gas flow in magnitude and direction. An example of this type is shown in Fig. 12-71, representing the Conidur® trademark.

The particular feature of Conidur® sheets is the specific hole shape which creates a directional airflow to help in discharging the product and to influence the retention time in the fluid bed. The special method of manufacturing Conidur® sheets enables finishing of fine perforations in sheets with an initial thickness many times over the hole width. Perforations of only 100 μm in an initial sheet thickness of 0.7 mm are possible. With holes this small 1 m^2 of plate may comprise several hundred thousand individual orifices.

The capacity of contributing to the transport of powder in the plane of the plate due to the horizontal component of the gas velocity is also the present for the second category of plates of the gill-type. Figure 12-72 shows an example.

In this type of plate, the holes or orifices are large and the number of gills per square meter is just a few thousand. The gas flow through each of the gills has a strong component parallel to the plate, providing powder transport capacity as well as a cleaning effect. The gills are punched individually or in groups and can be oriented individually to provide a possibility of articulating the horizontal transport effect.

In certain applications in the food and pharmaceutical industries, the nonsifting property of a fluid-bed plate is particularly appreciated. This property of a gill-type plate can be enhanced as illustrated in Fig. 12-73, where the hole after punching is additionally deformed so that the gill overlaps the orifice.

The fifth and final type of fluid-bed plate to be mentioned here is the so-called bubble plate type. Illustrated in Fig. 12-74, it is in principle a gill-type plate. The orifice is cut out of the plate, and the bubble is subsequently pressed so that the orifice is oriented in a predominantly horizontal direction. A fluid-bed plate will typically have only 1600 holes per square meter. By this technology a combination of three key features is established. The plate is nonsifting, it has transport capacity that can be articulated through individual orientation of bubbles, and it is totally free of cracks that may compromise sanitary aspects of the installation.

FIG. 12-72 GILL PLATE™ for fluid-bed application. (*Niro A/S.*)

FIG. 12-73 NON-SIFTING GILL PLATE™. (*Patented by Niro A/S.*)

FIG. 12-74 BUBBLE PLATE™. (*Patented by Niro A/S.*)

The **operating conditions** of a fluid bed are to a high degree dictated by the properties of the material to be dried, as already indicated. One parameter can be chosen regardless of the fluidization process, namely, the fluidization air temperature. For most products, however, the temperature is of primary importance, since the fluidized state results in very high heat-transfer rates so that heat sensitivity may restrict temperature and thereby prolong process time.

To achieve the most favorable combination of conditions to carry out a fluid-bed drying process, it is necessary to consider the different modes of fluid-bed drying available.

Industrial fluid-bed drying The first major distinction between fluid-bed types is the choice of mode: batch or continuous.

Batch fluid beds may appear in several forms. The process chamber has a perforated plate or screen in the bottom and a drying gas outlet at the top, usually fitted with an internal filter. The drying gas enters the fluid bed through a plenum chamber below the perforated plate and leaves through the filter arrangement. The batch of material is enclosed in the process chamber for the duration of the process.

Figure 12-75 shows a sketch of a typical batch fluid-bed dryer as used in the food and pharmaceutical industries. The process chamber is conic in order to create a freeboard velocity in the upper part of the chamber that is lower than the fluidizing velocity just above the plate.

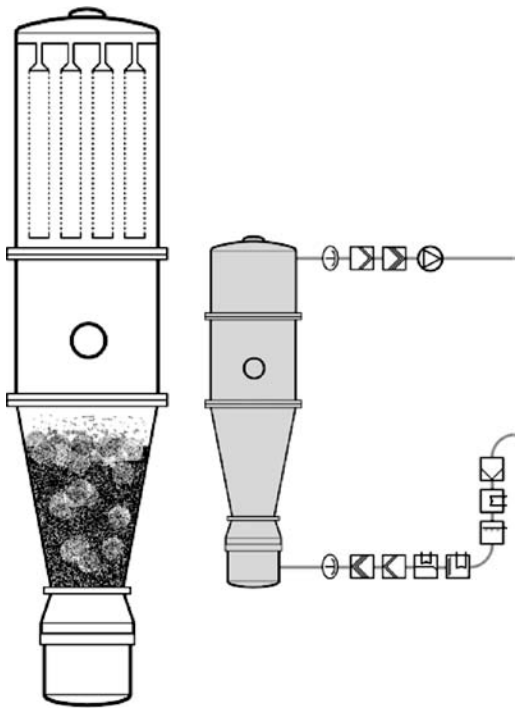

FIG. 12-75 Batch-type fluid-bed. (*Aeromatic-Fielder.*)

The enclosed product batch is prevented from escaping the process chamber and will therefore allow a freer choice of fluidizing velocity than is the case in a continuous fluid bed, as described later.

The right-hand side of Fig. 12-75 illustrates in symbol form the drying gas supply system comprising fan, filters of various grade, preheater, moisturizer, dehumidifier, final heater, and fast-closure valves. This arrangement is necessary for products with extreme quality requirements such as found in pharmaceutical production.

The drying can be carried out very like the process indicated in Fig. 12-68. The versatile drying gas supply system will allow the drying gas temperature and humidity to be controlled throughout the drying process to optimize process time and to minimize overheating of the product.

Continuous fluid beds may be even more varied than batch fluid beds. The main distinction between continuous fluid beds will be according to the solids flow pattern in the dryer. The continuous fluid bed will have an inlet point for moist granular material to be dried and an outlet for the dried material. If the moist material is immediately fluidizable, it can be introduced directly onto the plate and led through the bed in a plug-flow pattern that will enhance control of product residence time and temperature control. If the moist granular material is sticky or cohesive due to surface moisture and therefore needs a certain degree of drying before fluidization, it can be handled by a backmix fluid bed, to be described later.

Continuous plug-flow beds are designed to lead the solids flow along a distinct path through the bed. Baffles will be arranged to prevent or limit solids mixing in the horizontal direction. Thereby the residence time distribution of the solids becomes narrow. The bed may be of cylindrical or rectangular shape.

The temperature and moisture contents of the solids will vary along the path of solids through the bed and thereby enable the solids to come close to equilibrium with the drying gas. A typical plug-flow fluid bed is shown in Fig. 12-76.

Continuous plug-flow beds of stationary as well as vibrating type may benefit strongly from use of the gill-type fluid-bed plates with the capacity for controlling the movement of powder along the plate and around bends and corners created by baffles. Proper use of these means may make it possible to optimize the combination of fluidization velocity, bed layer height, and powder residence time.

Continuous backmix beds are used in particular when the moist granular material needs a certain degree of drying before it can fluidize. By distributing the material over the surface of an operating fluid bed arranged for total solids mixing, also called backmix flow, it will be absorbed by the dryer material in the bed, and lumping as well as sticking to the chamber surfaces will be avoided. The distribution of the feed can be arranged in different ways, among which a rotary thrower

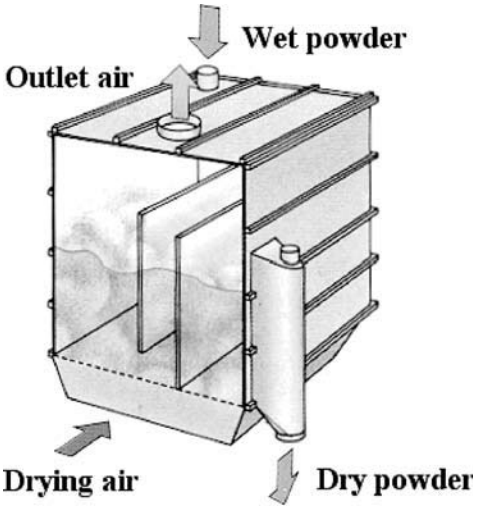

FIG. 12-76 Continuous plug-flow fluid bed. (*Niro A/S.*)

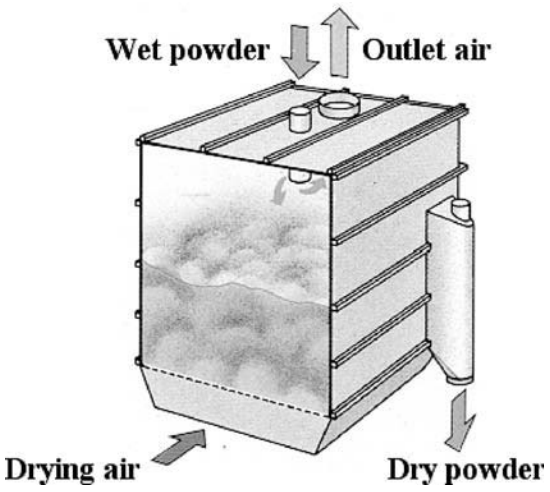

Figure 12-77 Continuous back-mix fluid bed. (*Niro A/S.*)

is an obvious choice. A typical backmix fluid bed is shown in Fig. 12-77. Backmix fluid beds can be of box-shaped design or cylindrical.

The whole mass of material in the backmix fluid bed will be totally mixed, and all powder particles in the bed will experience the same air temperature regardless of their position on the drying curve illustrated in Fig. 12-68. The residence time distribution becomes very wide, and part of the material may get a very long residence time while another part may get a very short time.

Continuous contact fluid beds are common in the chemical industry as the solution to the problem arising from materials requiring low fluidizing air temperature due to heat sensitivity and high energy input to complete the drying operation. An illustration of a Niro CONTACT FLUIDIZER™ is shown in Fig. 12-78.

The main feature of the contact fluid bed is the presence of heating panels, which are plate or tube structures submerged in the fluidized-bed layer and heated internally by an energy source such as steam, water, or oil. The fluidized state of the bed provides very high heat-transfer rates between the fluidizing gas, the fluidized material, and any objects submerged in the bed. The result is that a very significant portion of the required energy input can be provided by the heating panels without

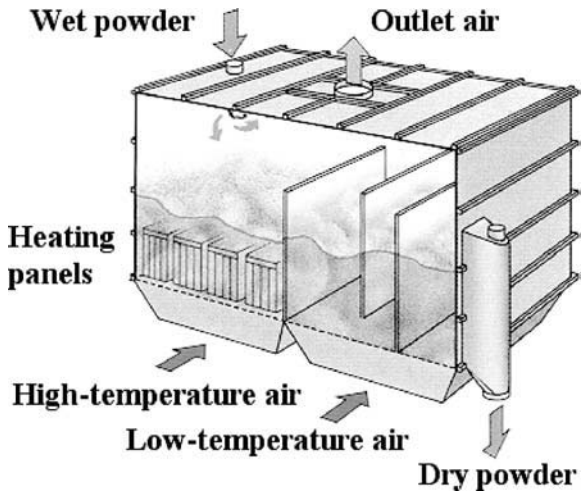

FIG. 12-78 Continuous CONTACT FLUIDIZER™. (*Niro A/S.*)

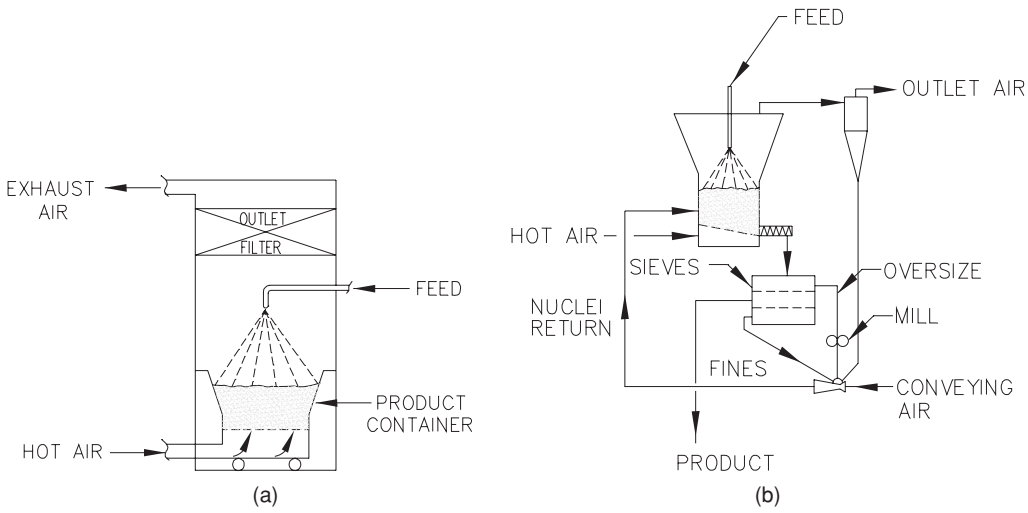

FIG. 12-79 Fluid-bed granulators. (*a*) Batch; (*b*) continuous.

risk of overheating the material. The fluidized state of the bed ensures that the material in the bed will flow with little restriction around the heating panels.

The CONTACT FLUIDIZER™ shown in Fig. 12-78 has a number of other features which in combination lead to compact design, high thermal efficiency, and low gas throughput: The first section of the bed is a backmix bed complete with rotary powder distributor and high-temperature fluidizing air supply. It takes care of the drying of the surface moisture, which is controlled mainly by heat supply. The heating panels are distributed over the whole bed volume of this section. The second section of the bed is a plug-flow bed with a fluidizing gas supply adjusted in both temperature and velocity to fit the requirements for the time-controlled diffusion drying of the powder present in this section.

The CONTACT FLUIDIZER™ is primarily used in the polymer industry for drying of polymer powders in high tonnages. Sizewise the individual units become very large, and units with a total fluid-bed area in excess of 60 m^2 are in operation.

Design methods for fluid beds When fluid-bed technology can be applied to drying of granular products, significant advantages compared to other drying processes can be observed. Design variables such as fluidizing velocity, critical moisture content for fluidization, and residence time required for drying to the specified residual moisture must, however, be established by experimental or pilot test before design steps can be taken. Reliable and highly integrated fluid-bed systems of either batch or continuous type can be designed, but only by using a combination of such pilot test and industrial experience. Scale-up rules are given by Kemp and Oakley (2002).

Additional Reading

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Grace, "Fluidized-Bed Hydrodynamics," Chap. 8.1 in *Handbook of Multiphase Systems*, McGraw-Hill, New York, 1982.
Gupta and Mujumdar, "Recent Developments in Fluidized Bed Drying," Chap. 5 in Mujumdar (ed.), *Advances in Drying*, vol. 2, Hemisphere, Washington, D.C., 1983, p. 155.
Kemp and Oakley, "Modeling of Particulate Drying in Theory and Practice," *Drying Technol.* **20**(9): 1699–1750 (2002).
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Poersch, *Aufbereitungs-Technik*, **4:** 205–218 (1983).
Richardson, "Incipient Fluidization and Particulate Systems," Chap. 2 in Davidson and Harrison (eds.), *Fluidization*, Academic Press, London, 1972.
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Dryers with Liquid Feeds If the feed is a liquid, paste, slurry, or solution, special equipment is required. The available choices are as follows:

Spray Dryers A pumpable feed is atomized into droplets by a rotary or nozzle atomizer, as described under "Entrainment Dryers." An integral fluid bed or belt may be added below the dryer to give longer residence time and some agglomeration. Semibatch and continuous operation is possible.

Fluidized-Bed Granulator A slurry or solution is sprayed onto a fluidized bed of particles, as shown in Fig. 12-79. The difference from the spray fluid bed is that the spray is still liquid when it contacts the particles, so that layered growth or surface agglomeration occurs, producing stronger large particles or agglomerates. Both batch and continuous forms exist (the latter involving continuous solids recycle with classification). In an older variant, the bed may be of inert balls, and the solid forming on the outside is periodically knocked off. Dryer construction and operation are largely as described under "Fluid-Bed Dryers."

Drum (Film-Drum) Dryers A film of liquid or paste is spread onto the outer surface of a rotating, internally heated drum. Drying occurs by conduction, and at the end of the revolution the dry product, which can be in the form of powder, flakes, or chips and typically is 100 to 300 μm thick, is removed by a doctor's knife. Drum dryers cannot handle feedstocks which do not adhere to metal, products which dry to a glazed film, or thermoplastics. The drum is heated normally by condensing steam or in vacuum drum dryers by hot water. Figure 12-80 shows three of the many possible forms. The dip feed system is the simplest and most common arrangement but is not suitable for viscous or pasty materials. The nip feed system is usually employed on double-drum dryers, especially for viscous materials, but it cannot handle lumpy or abrasive solids. The latter are usually applied by roller, and this is also effective for sticky and pasty materials. Spray and splash devices are used for feeding heat-sensitive, low-viscosity materials. Vacuum drum dryers are simply conventional units encased in a vacuum chamber with a suitable air lock for product discharge. Air impingement is also used as a secondary heat source on drum and can dryers, as shown in Fig. 12-81.

Contact Drying (Special thanks to R. B. Keey for the following example of contact drying.) In contact drying, the moist material covers a hot surface which supplies the heat required for the drying process.

Let us consider a moist material lying on a hot flat plate of infinite extent. Figure 12-82 illustrates the temperature profile for the fall in temperature from T_H in the heating fluid to T_C in the surrounding air. It is assumed that the temperatures remain steady, unhindered drying takes place, and there is no air-gap between the material being dried and the heating surface.

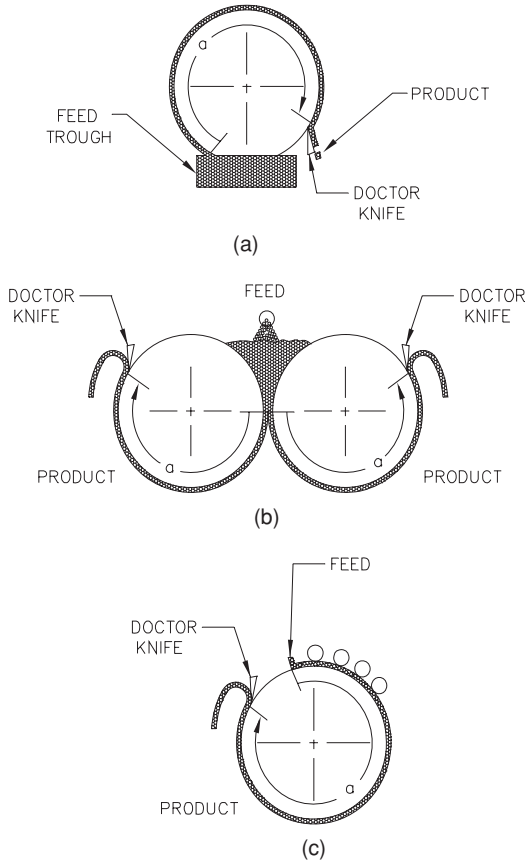

FIG. 12-80 Main types of drum dryers. (*a*) Dip; (*b*) nip; (c) roller.

The heat conducted through the wall and material is dissipated by evaporation of moisture and convection from the moist surface to the surrounding air. A heat balance yields

$$U(T_H - T_S) = N_W \Delta H_{VS} + h_C(T_S - T_G) \qquad (12\text{-}110)$$

where U is the overall heat-transfer coefficient. This coefficient is found from the reciprocal law of summing resistances in series:

$$\frac{1}{U} = \frac{1}{h_H} + \frac{b_B}{\lambda_B} + \frac{b_s}{\lambda_s} \qquad (12\text{-}111)$$

in which h_H is the heat-transfer coefficient for convection inside the heating fluid. If condensing steam is used, this coefficient is very large normally and the corresponding resistance $1/h_H$ is negligible. Rearrangement of Eq. (12-110) yields an expression for the maximum drying rate

$$N_W = \frac{U(T_H - T_S) - h_C\,(T_S - T_G)}{\Delta H_{VS}} \qquad (12\text{-}112)$$

Equation (12-112), as it stands, would give an overestimate of the maximum drying rate for the case of contact drying over heated rolls, when there are significant heat losses from the ends of the drum and only part of the drum's surface can be used for drying. In the roller drying arrangements shown in Fig. 12-80, only a fraction a of the drum's periphery is available from the point of pickup to the point where the solids are peeled off.

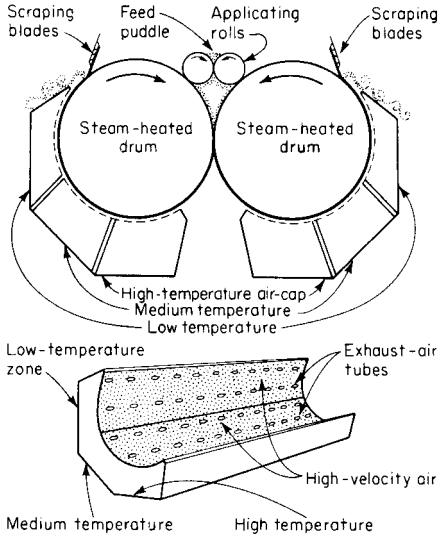

FIG. 12-81 Example of the use of air impingement in drying as a secondary heat source on a double-drum dryer. (Chem. Eng., ***197***, *June 19, 1967.*)

Let q_E be the heat loss per unit area from the ends. The ratio of the end areas to cylindrical surface, from a drum of diameter D and length L, is $2(\frac{1}{4}\pi D^2)/\pi DL$ or $D/2L$. Equation (12-112) for the maximum drying rate under roller drying conditions thus becomes

$$N_W = \frac{aU(T_H - T_S) - h_C\,(T_S - T_G) - Dq_E/2L}{\Delta H_{VS}} \qquad (12\text{-}113)$$

The total evaporation from the drum is $N_w a(\pi DL)$. Equation (12-113) could be refined further, as it neglects the effect caused by the small portion of the drum's surface being covered by the slurry in the feed trough, as well as thermal conduction through the axial shaft to the bearing mounts. The use of Eq. (12-113) to estimate the maximum drying rate is illustrated in Example 24.

Example 24: Heat-Transfer Calculations A single rotating drum of 1.250-m diameter and 3 m wide is internally heated by saturated steam at 0.27 MPa. As the drum rotates, a film of slurry 0.1 mm thick is picked up and dried. The dry product is removed by a knife, as shown in Fig. 12-80*a*. About three-quarters of the drum's surface is available for evaporating moisture. Estimate the maximum drying rate when the outside air temperature T_G is 15°C and the surface temperature 50°C, and compare the effectiveness of the unit with a dryer without end effects and in which all the surface could be used for drying.

Data:

Heat-transfer coefficient h_c	50 W/(m²·K)
Thickness of cylinder wall b_B	10 mm
Thermal conductivity of wall λ_B	40 W/(m·K)
Thermal conductivity of slurry film λ_s	0.10 W/(m·K)
Film transfer coefficient for condensing steam h_H	2.5 kW/(m²·K)

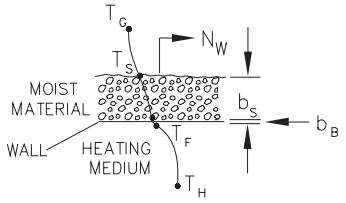

FIG. 12-82 Temperature profile in conductive drying.

Overall heat-transfer coefficient U: The thermal resistances are as follows:

Steamside $1/2.5 = 0.40\ m^2K/kW$

Wall $0.01/0.04 = 0.25\ m^2K/kW$

Filmside $0.0001/0.1 \times 10^{-3} = 1.0\ m^2K/kW$

$\therefore$ Overall resistance $= 0.40 + 0.25 + 1.0 = 1.65\ m^2K/kW$

$$U = 1/1.65 = 0.606\ kW/(m^2 \cdot K)$$

Wall temperature T_B: At 0.27 MPa, the steam temperature is 130°C. If it is assumed that the temperature drops between the steam and the film surface are directionally proportional to the respective thermal resistances, it follows that

$$\frac{T_H - T_B}{T_H - T_S} = \frac{0.40 + 0.25}{1.65} = 0.3939$$

$$\therefore T_B = T_H - 0.3939(T_H - T_S)$$
$$= 130 - 0.3939(130 - 50)$$
$$= 98.5°C$$

Heat losses from ends q_E: For an emissivity ~1 and an air temperature of 15°C with a drum temperature of 98.5°C, one finds [see Eq. (12-119)],

$$q_E = 1184\ W/m^2$$

Maximum drying rate N_W: From Eq. (12-113),

$$N_W = \frac{aU(T_H - T_S) - h_C(T_S - T_G) - Dq_E/2L}{\Delta H_{VS}}$$

$$= \frac{0.75 \times 0.606(130 - 50) - 0.05(50 - 15) - (1.25 \times 1.184)/6}{2382} \quad (12\text{-}114)$$

$$= 0.0144\ kg/(m^2 \cdot s)$$

The ideal maximum rate is given by Eq. (12-112) for an endless surface:

$$N_N = \frac{U(T_H - T_S) - h_c(T_S - T_G)}{\Delta H_{VS}}$$

$$= \frac{0.606(130 - 50) - 0.05(50 - 15)}{2382} \quad (12\text{-}115)$$

$$= 0.0196\ kg/(m^2 \cdot s)$$

Therefore the effectiveness of the dryer is 0.0144/0.0196 = 0.735.

The predicted thermal efficiency η is

$$\eta = 1 - \frac{h_c(T_S - T_G + Dq_E/2L}{aU(T_H - T_S)}$$

$$= 1 - \frac{0.05(50 - 15) + (1.25 \times 1.184)/6}{0.75 \times 0.606(130 - 50)} \quad (12\text{-}116)$$

$$= 0.945$$

These estimates may be compared with the range of values found in practice, as shown in Table 12-39 (Nonhebel and Moss, *Drying of Solids in the Chemical Industry,* Butterworths, London, 1971, p. 168).

The typical performance is somewhat less than the estimated maximum evaporative capacity, although values as high as 25 g/(m²·s) have been reported. As the solids dry out, so the thermal resistance of the film increases and the evaporation falls off accordingly. Heat losses through the bearing of the drum shaft have been neglected, but the effect of radiation is accounted for in the value of h_c taken. In the case of drying organic pastes, the heat losses have been determined to be 2.5 kW/m² over the whole surface, compared with 1.75 kW/m² estimated here for the cylindrical surface. The inside surface of the drum has been assumed to be clean, and scale would reduce the heat transfer markedly.

For constant hygrothermal conditions, the base temperature T_B is directly proportional to the thickness of the material over the hot surface. When the wet-bulb temperature is high and the layer of material is thick enough, the temperature T_B will reach the boiling point of the moisture. Under these conditions, a mixed vapor-air layer interposes between the material and the heating surface. This is known as the Leidenfrost effect, and the phenomenon causes a greatly increased thermal resistance to heat transfer to hinder drying.

TABLE 12-39 Operating Information

	This estimate	Typical range
Specific evaporation, g/(m²·s)	14.4	7–11
Thermal efficiency	0.945	0.4–0.7

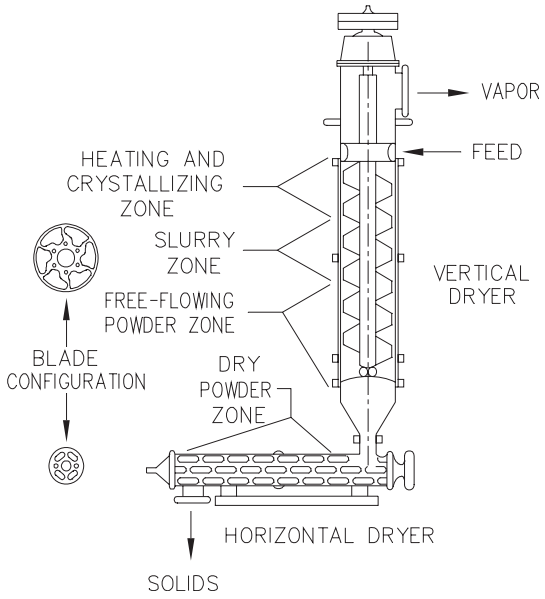

FIG. 12-83 Continuous thin-film dryer.

Thin-Film Dryers Evaporation and drying take place in a single unit, normally a vertical chamber with a vertical rotating agitator which almost touches the internal surface. The feed is distributed in a thin layer over the heated inner wall and may go through liquid, slurry, paste, and wet solid forms before emerging at the bottom as a dry solid. These dryers are based on wiped-film or scraped-surface (Luwa-type) evaporators and can handle viscous materials and deal with the "cohesion peak" experienced by many materials at intermediate moisture contents. They also offer good containment. Disadvantages are complexity, limited throughput, and the need for careful maintenance. Continuous or semibatch operation is possible. A typical unit is illustrated in Fig. 12-83.

Filter Dryers Basically this is a Nutsche filter (Sec. 18, "Liquid-Solid Operations and Equipment") followed by a batch dryer, usually of vertical pan type (see "Batch Agitated and Rotating Dryers" section). They are popular in the pharmaceutical and specialty chemicals industries as two unit operations are performed in the same piece of equipment without intermediate solids transfer, and containment is good.

Centrifuge Dryers Usually they are batch or continuous filtering centrifuges (Sec. 18, "Liquid-Solid Operations and Equipment") with hot air being blown over the solids in the discharge section. Manufacturers include Heinkel and Bird-Humboldt.

Pastelike feeds can be handled by some dryers for particulate materials, if either they do not require free-flowing feeds or some dry product can be backmixed with the wet feed to improve its handling.

Dryers for Films and Sheets The construction of dryers where both the feed and the product are in the form of a sheet, web, or film is markedly different from that for dryers used in handling particulate materials. The main users are the paper and textile industries. Almost invariably the material is formed into a very long sheet (often hundreds or thousands of meters long) which is dried in a continuous process. The sheet is wound onto a bobbin at the exit from the dryer; again, this may be 1 or 2 m in diameter. Alternatively, the sheet may be chopped into shorter sections.

Cylinder Dryers and Paper Machines The most common type of dryer in papermaking is the cylinder dryer (Fig. 12-84), which is a

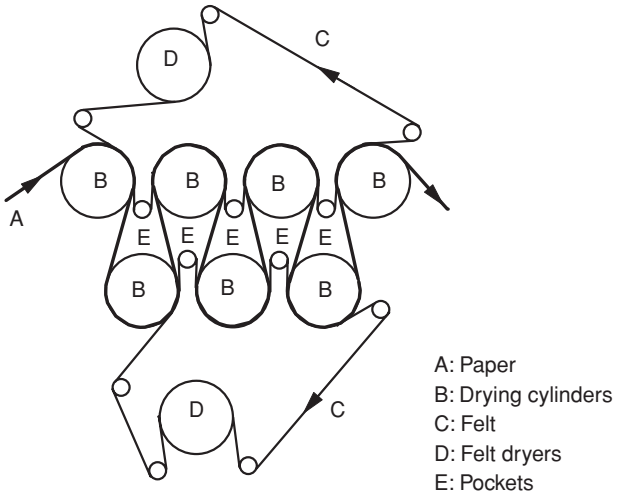

FIG. 12-84 Cylinder dryer (paper machine).

contact dryer. The paper web is taken on a convoluted path during which it wraps around the surface of cylinders which are internally heated by steam or hot water. In papermaking, the sheet must be kept taut, and a large number of cylinders are used, with only short distances between them and additional small unheated rollers to maintain the tension. Normally, a continuous sheet of felt is also used to hold the paper onto the cylinders, and this also becomes damp and is dried on a separate cylinder.

Most of the heating is conductive, through contact with the drums. However, infrared assistance is frequently used in the early stages of modern paper machines. This gets the paper sheet up to the wet-bulb temperature more rapidly, evaporates more surface moisture, and allows the number of cylinders to be reduced for a given throughput. Hot air jets (jet foil dryer) may also be used to supplement heating at the start of the machine. Infrared and dielectric heating may also be used in the later stages to assist the drying of the interior of the sheet.

Although paper is the most common application, multicylinder dryers can also be used for polymer films and other sheet-type feeds.

Convective dryers may be used as well in papermaking. In the Yankee dryer (Fig. 12-85), high-velocity hot airstreams impinging on the

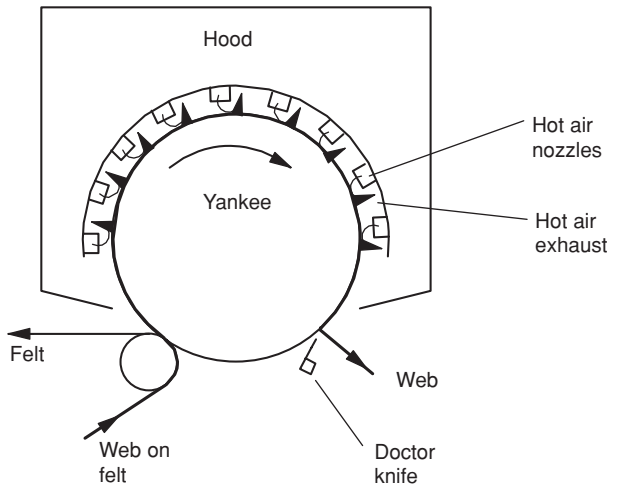

Fig. 12-85 Yankee dryer.

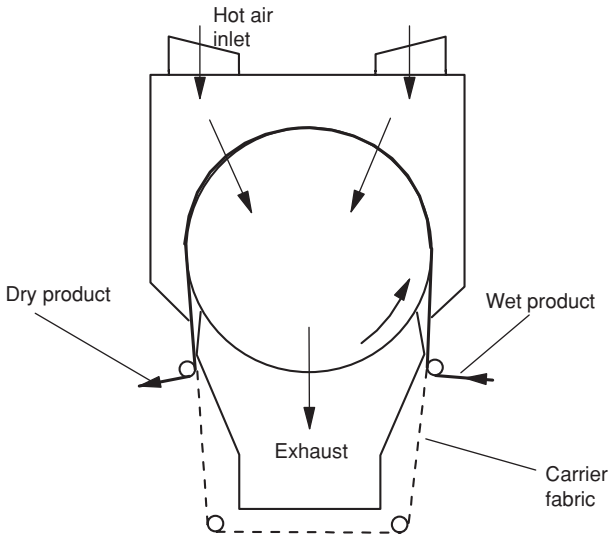

FIG. 12-86 Rotary through-dryer.

web surface give heating by cross-convection. The "Yankees" are barbs holding the web in place. Normally the cylinder is also internally heated, giving additional conduction heating of the lower bed surface. In the rotary through-dryer (Fig. 12-86), the drum surface is perforated and hot air passes from the outside to the center of the drum, so that it is a through-circulation convective dryer.

Another approach to drying of sheets has been to suspend or "float" the web in a stream of hot gas, using the Coanda effect, as illustrated in Fig. 12-87. Air is blown from both sides, and the web passes through as an almost flat sheet (with a slight "ripple"). The drying time is reduced because the heat transfer from the impinging hot air jets is faster than that from stagnant hot air in a conventional oven. It is essential to control the tension of the web very accurately. The technique is particularly useful for drying coated paper, as the expensive surface coating can stick to cylinder dryers.

Stenters (Tenters) and Textile Dryers These are the basic type of dryer used for sheets or webs in the textile industry. The sheet is held by its edges by clips (clip stenter) or pins (pin stenter), which not only suspend the sheet but also keep it taut and regulate its width—a vital consideration in textile drying. Drying is by convection; hot air is introduced from one or both sides, passes over the surface of the sheet, and permeates through it. Infrared panels may also be used to supply additional heat. A schematic diagram of the unit is shown in Fig. 12-88. A typical unit is 1.4 m wide and handles 2 to 4 t/h of material.

Heavy-duty textiles with thick webs may need a long residence time, and the web can be led up and down in "festoons" to reduce dryer length. Substantial improvements in drying rates have been obtained with radio-frequency heating assistance.

Air impingement dryers as in Fig. 12-87 may also be used for textiles.

Spray Dryers Spray drying is a drying process for transformation of a pumpable liquid feed in the form of a solution, dispersion, slurry, or paste into a particulate dried product in one single operation. The process comprises atomization of the feed followed by intense contact with hot air.

Due to the very large surface area created by the atomization of the feed, rapid evaporation occurs from the surface of each particle or droplet in the spray. The magnitude of the surface area can be illustrated by a simple calculation. Atomization of 1 L of water into a uniform spray of 100-μm droplets results in approximately 1.9×10^9 individual particles with a combined surface area of 60 m^2. A realistic spray with variation of the droplet size may have a substantially higher number of droplets and a somewhat higher surface area. The dry

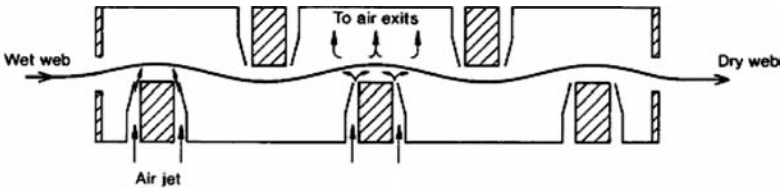

FIG. 12-87 Air flotation (impingement) dryer.

particulate product is formed while the spray droplets are still suspended in the hot drying air. The spray drying process is concluded by product recovery and separation from the drying air.

Spray drying belongs to the family of suspended particle processing (SPP) systems. Other members of this family are fluid-bed drying, flash drying, spray granulation, spray reaction, spray cooling, and spray absorption.

Drying Principles In the spray drying process or operation, the liquid to be removed by drying is predominantly water. Certain special products are produced with use of organic solvents, which are removed in a spray drying process. The drying principles involved for aqueous as well as nonaqueous systems are the same.

The liquid or moisture in a spray droplet is present in two basic forms: bound and unbound moisture. The nature of the solid and the liquid matter determines the drying characteristics of the product.

The category of bound moisture comprises water retained in small capillaries in the solid, water absorbed on solid surfaces, water bound as solutions in cells or fiber walls, and water bound as crystal water in chemical combination with the solid. Bound water exerts an equilibrium vapor pressure lower than that of pure water at the same temperature.

The category of unbound moisture can be described as the moisture in excess of the bound moisture. A hygroscopic material may contain bound as well as unbound moisture. A nonhygroscopic material contains unbound moisture only. The equilibrium vapor pressure of unbound water is equal to that of pure water at the same temperature.

The free moisture in a particle is the moisture in excess of the equilibrium moisture and may consist of unbound and some bound moisture. Only free moisture can be removed by evaporation during spray drying.

The mechanism of moisture flow in a droplet during spray drying is mainly diffusion supplemented by capillary flow. The drying characteristics of the droplet depend on the balance of bound and unbound as each category has distinct features.

The presence of unbound moisture in the droplet means that the drying proceeds at a constant high rate as long as the moisture diffusion within the droplet is able to maintain saturated surface conditions. When the diffusional and capillary flows can no longer maintain these conditions, a critical point is reached and the drying rate will decline until equilibrium moisture content is reached. The evaporation of bound moisture is strongly dependent on the nature of the solid matter in the spray droplet.

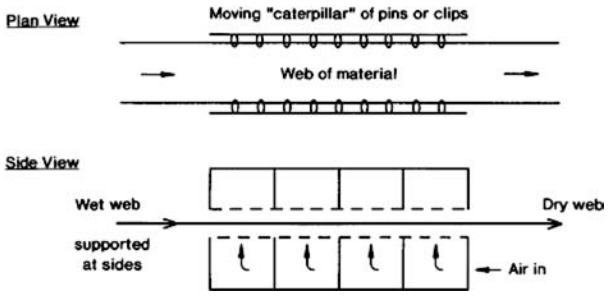

FIG. 12-88 Stenter or tenter for textile drying.

A spray drying plant comprises four process stages, as shown in Table 12-40.

Atomization Stage Spray drying is often used in industrial processes characterized by high production rates. Although the three different methods of atomization indicated in Table 12-40 are the same as those for many other atomization or spray forming processes, the relative weight of the methods is special for spray drying with rotary atomizers and hydraulic pressure nozzles having a very broad application, while two-fluid nozzles are only used to a smaller extent in specialized applications.

Rotary Atomizer Figure 12-89 shows a rotary atomizer in operation. The liquid feed is supplied to the atomizer by gravity or hydraulic pressure. A liquid distributor system leads the feed to the inner part of a rotating wheel. Since the wheel is mounted on a spindle supported by bearings in the atomizer structure, the liquid distributor is usually formed as an annular gap or a ring of holes or orifices concentric with the spindle and wheel. The liquid is forced to follow the wheel either by friction or by contact with internal vanes in the wheel. Due to the high centrifugal forces acting on the liquid, it moves rapidly toward the rim of the wheel, where it is ejected as a film or a series of jets or ligaments. By interaction with the surrounding air the liquid breaks up to form a spray of droplets of varying size. The spray pattern is virtually horizontal with a spray angle said to be 180°. The mean droplet size of the spray depends strongly on the atomizer wheel speed and to a much lesser degree on the feed rate and the feed physical properties such as viscosity. More details about spray characteristics such as droplet size distribution will be given below.

As indicated above, the atomizer wheel speed is the important parameter influencing the spray droplet size and thus the particle size of the final product. The atomizer machine will normally have the capability to operate the wheel at the required speed. More important for the atomization process is the selection of a wheel capable of handling a specific liquid feed with characteristic properties such as abrasiveness, high viscosity, nonnewtonian behavior, or tendency to coagulate.

The most common design of atomizer wheel has radial vanes, as shown in Fig.12-90. This wheel type is widely used in the chemical industry and is virtually blockage-free and simple to operate, even at very high speed. For high-capacity applications, the number and height of the vanes may be increased to maintain limited liquid film thickness conditions on each vane.

Wheels with radial vanes have one important drawback, i.e., their capacity for pumping large amounts of air through the wheel. This so-called air pumping effect causes unwanted product aeration, resulting in powders of low bulk density for some sensitive spray dried products.

TABLE 12-40 Stages of Spray Drying

Process stages of spray drying	Methods
1. Atomization	Rotary atomization Pressure nozzle atomization Two-fluid nozzle atomization
2. Spray/hot air contact	Cocurrent flow Countercurrent flow Mixed flow
3. Evaporation	Drying Particle shape formation
4. Product recovery	Drying chamber Dry collector Wet collectors

FIG. 12-89 Rotary atomizer operation. (*Niro*).

Unwanted air pumping effect and product aeration can be reduced through careful wheel design involving change of the shape of the vanes that may appear as forward-curved. This wheel type is used widely in the dairy industry to produce powders of high bulk density. The powder bulk density may increase as much as 15 percent when a curved vane wheel is replacing a radial vane wheel of standard design.

Another way of reducing the air pumping effect is to reduce the space between the vanes so that the liquid feed takes up a larger fraction of the available cross-sectional area. This feature is used with consequence in the so-called bushing wheels such as shown in Fig.12-91. This wheel combines two important design aspects. The air pumping effect is reduced by reducing the flow area to a number of circular orifices, each 5 to 10 mm in diameter. By placing these orifices or nozzles in replaceable bushings or inserts made of very hard materials such as technical ceramics, i.e., alumina or silicon carbide, a substantially abrasion-resistant atomizer wheel design is achieved. This feature is very important in a number of spray drying applications with abrasive feeds, which would wear down a standard vaned wheel in a matter of hours. With an abrasion-resistant wheel, almost unlimited lifetime can be expected for the atomizer wheel structure and several thousand hours for replaceable bushings.

The rotary atomizer machines are high-speed machines traditionally built with a step-up gear to increase the speed from the 3000 or 3600 rpm of the standard two-pole electric motors to 10,000 to 20,000 rpm normally required to achieve sufficiently fine atomization. Newer designs feature high-speed electric motors with frequency control of the atomizer speed. Table 12-41 gives the main operational parameters for three typical atomizers covering the wide range of capacity and size.

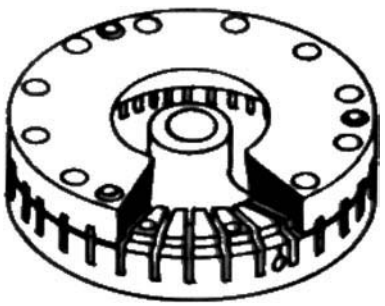

FIG. 12-90 Rotary atomizer wheel with radial vanes. (*Niro.*)

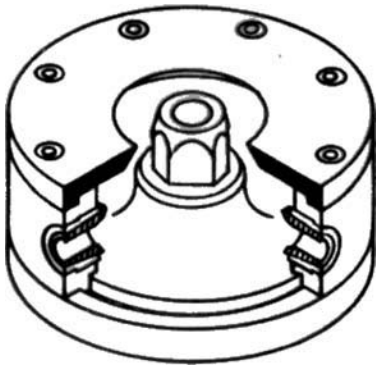

FIG. 12-91 Abrasion-resistant bushing atomizer wheel. (*Niro.*)

The F800 atomizer is the largest rotary atomizer offered to industry today. It has the capability of handling up to 200 t/h in one single atomizer. The capacity limit of an atomizer is normally its maximum power rating. As indicated above, the atomizer wheel speed is the important parameter influencing the spray droplet size. The wheel speed also determines the power consumption of the atomizer. It can be shown that the atomizer power consumption exclusive mechanical losses amount to

$$P_S = \frac{U^2}{3600}$$

where P_s = specific power consumption, kWh/t and U = peripheral velocity, m/s.

Since the atomizer wheel peripheral speed is proportional to the rotational speed, the maximum feed rate that can be handled by a rotary atomizer declines with the square of the rotational speed. The maximum feed rates indicated in Table 12-41 are therefore not available in the higher end of the speed ranges.

The rotary atomizer has one distinct advantage over other means of atomization. The degree or fineness of atomization achieved at a given speed is only slightly affected by changes in the feed rate. In other words, the rotary atomizer has a large turndown capability.

The larger atomizer machines cited in Table 12-41 represent a range of very large rotary atomizers available to industry. They are equipped with epicyclic-type gearboxes complete with a lubrication system. An extensive monitoring system is integrated in each machine.

Many atomization duties involve much lower capacities than foreseen for this range of atomizers. A full range of smaller rotary atomizers are available with nominal capacities down to less than 100 kg/h. Various designs may be seen with either belt drive or worm gears. Designs without gears are available with high-speed electric motor drive. Table 12-41 gives data for a smaller atomizer machine (FS1.5). It belongs to a family of high-speed machines without gears and lubrication systems capable of operating under the strictest requirements

TABLE 12-41 Operational Parameters for Atomizers (Niro)

Rotary atomizer designs				
Atomizer type		FS1.5	F160	F800
Nominal power rating	kW	1.5	160	1000
Maximum feed rate	t/h	0.3	50	200
Atomizer wheel diameter	mm	90	240	350
Typical gear ratio	#	1:1	4.4:1	2.9:1
Minimum speed	rpm	10,000	6,000	8,800
Maximum speed	rpm	30,000	18,200	11,500
Typical peripheral velocity	m/s	141	165	161
Typical specific power	kWh/t	5.5	7.6	7.2

for noncontamination of the product and in explosion-prone environments.

Hydraulic pressure nozzle In hydraulic pressure nozzle atomizers, the liquid feed is fed to the nozzle under pressure. In the nozzle orifice the pressure energy is converted to kinetic energy. The internal parts of the nozzle are normally designed to apply a certain amount of swirl to the feed flow so that it issues from the orifice as a high-speed film in the form of a cone with a desired vertex angle. This film disintegrates readily into droplets due to instability. The vertex or spray angle is normally on the order of 50° to 80°, a much narrower spray pattern than is seen with rotary atomizers. This means that spray drying chamber designs for pressure nozzle atomization differ substantially from designs used with rotary atomizers. The droplet size distribution produced by a pressure nozzle atomizer varies inversely with the pressure and to some degree with feed rate and viscosity. The capacity of a pressure nozzle varies with the square root of the pressure. To obtain a certain droplet size, the pressure nozzle must operate very close to the design pressure and feed rate. This implies that the pressure nozzle has very little turndown capability.

Hydraulic pressure nozzles cannot combine the capability for fine atomization with high feed capacity in one single unit. Many spray dryer applications, where pressure nozzles are applied, therefore require multinozzle systems with the consequence that start-up, operational control, and shutdown procedures become more complicated.

Two-fluid nozzle atomization In two-fluid nozzle atomizers, the liquid feed is fed to the nozzle under marginal or no pressure conditions. An additional flow of gas, normally air, is fed to the nozzle under pressure. Near the nozzle orifice, internally or externally, the two fluids (feed and pressurized gas) are mixed and the pressure energy is converted to kinetic energy. The flow of feed disintegrates into droplets during the interaction with the high-speed gas flow which may have sonic velocity.

The spray angle obtained with two-fluid nozzles is normally on the order of 10° to 20°, a very narrow spray pattern that is related to the spread of a free jet of gas. Spray drying chamber designs for two-fluid nozzle atomization are very specialized according to the application.

The droplet size produced by a two-fluid nozzle atomizer varies inversely with the ratio of gas to liquid and with the pressure of the atomization gas. The capacity of a two-fluid nozzle is not linked to its atomization performance. Therefore two-fluid nozzles can be attributed with some turndown capability.

Two-fluid nozzles share with pressure nozzles the lack of high feed capacity combined with fine atomization in one single unit. Many spray dryer applications with two-fluid nozzle atomization have a very high number of individual nozzles. The main advantage of two-fluid nozzles is the capability to achieve very fine atomization.

Choice of atomizer system The choice of atomizer system for a specific spray drying operation depends upon the particle size distribution required in the final dried product. It also depends upon the physical and chemical properties of the feed liquid.

In cases where the different types of atomizer means produce similar particle size distributions, the rotary atomizer may be preferred due to its greater flexibility and ease of operation. When one is comparing the atomizer types, the rotary atomizer has distinct advantages. (1) It can handle high feed rates in one single unit, (2) it can handle abrasive feeds with minimal wear, and (3) it has negligible blockage tendencies due to the large flow ports in the atomizer wheel. (4) It is a low-pressure system that can be served by a simple feed supply system, and (5) droplet size control is simple through wheel speed adjustment.

Although it lacks the flexibility of the rotary atomizer, the pressure nozzle is nevertheless widely used in spray drying applications. For many products the requirement for nondusty appearance calls for large mean particle size and lack of a fines fraction that cannot be met with a rotary atomizer. In the other end of the particle size range, some products require finer particles than are practically achievable with a rotary atomizer. This is the range where two-fluid nozzles are applied. The following guidelines may be used as an indication of the particle sizes obtainable in spray dryers:

- For spray dryers with rotary atomizer, the mean size of the dried product varies from 40 to 110 μm, although larger product mean sizes can be produced in large-diameter chambers.
- For spray dryers with pressure nozzle atomization, the mean particle size of the dried product varies in the range from 50 to 250 μm.
- For spray dryers with two-fluid nozzle atomization, the mean particle size of the dried product varies in the range from 15 to 50 μm.

The different means of atomization can also be compared in terms of energy power consumption. As indicated in Table 12-41, typical specific power figures for rotary atomizers are in the range of 5 to 11 kWh/t. Similar figures can be calculated for pressure and two-fluid nozzle systems, i.e., the pumping energy of the feed and the compression energy of the atomization gas. Any such calculation will show that similar median particle sizes are obtained for a given atomization energy independent of the means of atomization. None of the three types stand out as being energy-efficient. The hydraulic pressure nozzle is best suited for relatively coarse atomization, because pressures higher than 300 bar are impractical. Rotary atomizers are limited, because the wheel peripheral speeds required for very fine atomization put the wheel material under extreme tensile stress.

Droplet size distributions obtained with any means mentioned here are relatively well represented by a Rosin-Rammler distribution with an exponent of approximately 2. This means that approximately 80 percent of the droplet population mass is in the range of 0.39 to 1.82 times the median droplet size.

Theoretical prediction of mean particle sizes is difficult and of little practical importance, since the selection of spray drying operational parameters is based on experience and pilot-scale test work. The scientific literature, however, contains numerous estimation formulas to help predict the droplet sizes in sprays. Table 12-42 provides nomenclature for these estimation formulas.

For rotary atomizers median droplet sizes can be estimated from the following empirical equation of obscure origin:

$$d_{50} = K_r \times \dot{m}_L^{0.15} \times D^{-0.8} \times N^{-0.05} \times \omega^{-0.75} \times \mu_L^{0.07}$$

For hydraulic pressure nozzles the following formula proposed by Lefebvre may be used:

$$d_{50} = K_p \times \dot{m}_L^{0.25} \times \Delta P_L^{-0.5} \times (\sigma_L \times \mu_L)^{0.25} \times \rho_A^{-0.25}$$

Similarly, a range of equations or formulas are available for prediction of droplet size for sprays from two-fluid nozzles. The most widely cited in the literature is the Nukiyama-Tanasawa equation, which, however, is complicated and of doubtful validity at high flow rates. A much simpler equation has been proposed by Geng Wang et al.:

$$d_{50} = K_t \times \rho_A^{-0.325} \times \left(\frac{\dot{m}_L}{\dot{m}_L \times U_L + \dot{m}_A \times U_A} \right)^{0.55}$$

If any difference between the atomization means mentioned here were to be pointed out, it would be the tendency for two-fluid nozzles to have the wider particle size distribution and narrower pressure nozzles with rotary atomizers in between.

TABLE 12-42 Nomenclature for Atomization Equations

d_{50}	= mass median droplet size	m
K_r	= empirical factor	0.008
K_p	= empirical factor	4.0
K_t	= empirical factor	0.1
$\dot{m}_L$	= liquid feed rate	kg/s
D	= wheel diameter	m
N	= number of vanes	#
ω	= atomizer wheel speed	rad/s
μ_L	= liquid viscosity	Pa·s
ΔP_L	= atomization pressure	Pa
ρ_A	= air density	kg/m^3
σ_L	= liquid surface tension	N/m
$\dot{m}_A$	= atomization gas rate	kg/s
U_L	= liquid velocity	m/s
U_A	= atomization gas velocity	m/s

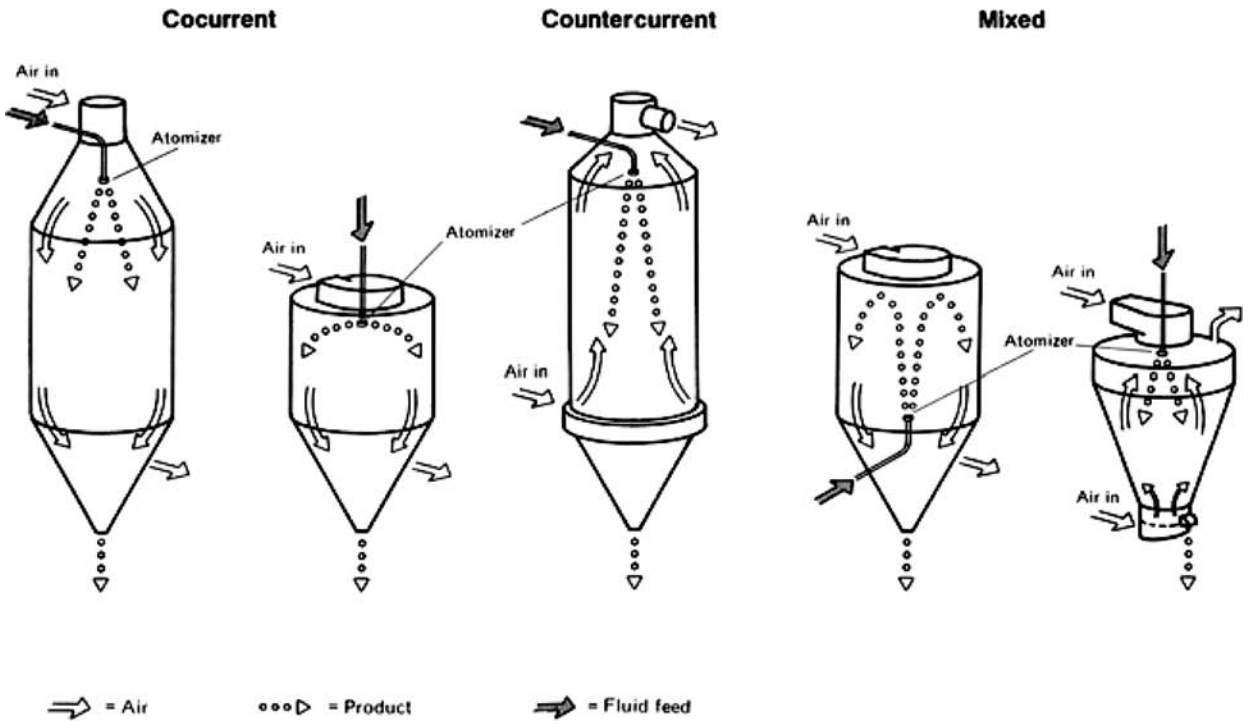

FIG. 12-92 Different forms of spray/hot air contact. (*Niro.*)

Spray/Hot Air Contact Atomization is first and most important process stage in spray drying. The final result of the process does, however, to a very large degree depend on the second stage, the spray/hot air contact. The way the spray of droplets is contacted by the hot air or gas carrying the thermal energy required to evaporate the moisture in the droplets is important for the quality of the product. In general terms three possible forms can be defined. These are as depicted in Fig. 12-92:

- Cocurrent flow
- Countercurrent flow
- Mixed flow

Different drying chamber forms and different methods of hot air introduction accompany the different flow pattern forms and are selected according to

- Required particle size in product specification
- Required particle form
- Temperature or heat sensitivity of the dried particle

In general terms, selection of chamber design and flow pattern form follows these guidelines:

- Use cocurrent spray drying for heat-sensitive products of fine as well as coarse particle size, where the final product temperature must be kept lower than the dryer outlet temperature.
- Use countercurrent spray drying for products which are not heat-sensitive, but may require some degree of heat treatment to obtain a special characteristic, i.e., porosity or bulk density. In this case the final powder temperature may be higher than the dryer outlet temperature.
- Use mixed-flow spray drying when a coarse product is required and the product can withstand short time exposure to heat without adverse effects on dried product quality.

Evaporation Stage Evaporation takes place from a moisture film which establishes on the droplet surface. The droplet surface temperature is kept low and close to the adiabatic saturation temperature of the drying air. As the temperature of the drying air drops off and the solids content of the droplet/particle increases, the evaporation rate is reduced. The drying chamber design must provide a sufficient residence time in suspended condition for the particle to enable completion of the moisture removal.

During the evaporation stage the atomized spray droplet size distribution may undergo changes as the droplets shrink, expand, collapse, fracture, or agglomerate.

Dry Product Recovery Product recovery is the last stage of the spray drying process. Two distinct systems are used:

- In two-point discharge, primary discharge of a coarse powder fraction is achieved by gravity from the base of the drying chamber. The fine fraction is recovered by secondary equipment downstream of the chamber air exit.
- In single-point discharge, total recovery of dry product is accomplished in the dryer separation equipment.

Collection of powder from an airstream is a large subject of its own. In spray drying, dry collection of powder in a nondestructive way is achieved by use of cyclones, filters with textile bags or metallic cartridges, and electrostatic precipitators.

With the current emphasis on environmental protection, many spray dryers are equipped with additional means to collect even the finest fraction. This collection is often destructive to the powder. Equipment in use are wet scrubbers, bag or other kinds of filters, and in a few cases incinerators.

Industrial Designs and Systems Thousands of different products are processed in spray dryers representing a wide range of feed and product properties as well as drying conditions. The flexibility of the spray drying concept, which is the main reason for this wide application, is described by the following systems.

Plant Layouts Figure 12-93*a* shows a standard cocurrent cone-based chamber with roof gas disperser. The chamber can have either single- or two-point discharge and can be equipped with rotary or

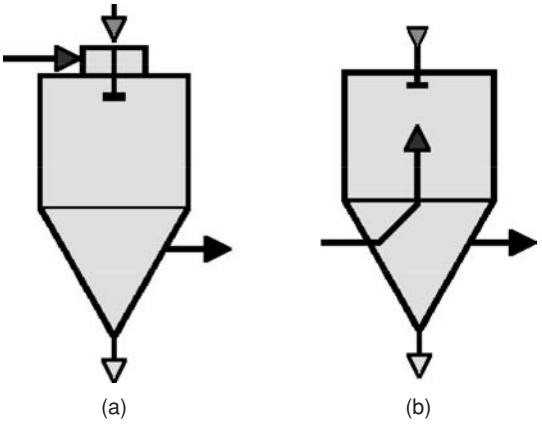

FIG. 12-93 (*a*) Standard cocurrent and (*b*) high-temperature chambers.

nozzle atomization. Fine or moderately coarse powders can be produced. This type of dryer finds application in dairy, food, chemical, pharmaceutical, agrochemical, and polymer industries.

Figure 12-93*b* shows a high-temperature chamber with the hot gas distributor arranged internally on the centerline of the chamber. The atomizer is rotary. Inlet temperature in the range of 600 to 1000°C can be utilized in the drying of non-heat-sensitive products in the chemical and mining industries. Kaolin and mineral flotation concentrates are typical examples.

Figure 12-94*a* shows a cocurrent cone-based tall form chamber with roof gas disperser. This chamber design is used primarily with pressure nozzle atomization to produce powders of large particle sizes with a minimum of agglomeration. The chamber can be equipped with an oversize cone section to maximize powder discharge from the chamber bottom. This type of dryer is used for dyestuffs, baby foods, detergents, and instant coffee powder.

Figure 12-94*b* shows a countercurrent flow chamber with pressure nozzle atomization. This design is in limited use because it cannot produce heat-sensitive products. Detergent powder is the main application.

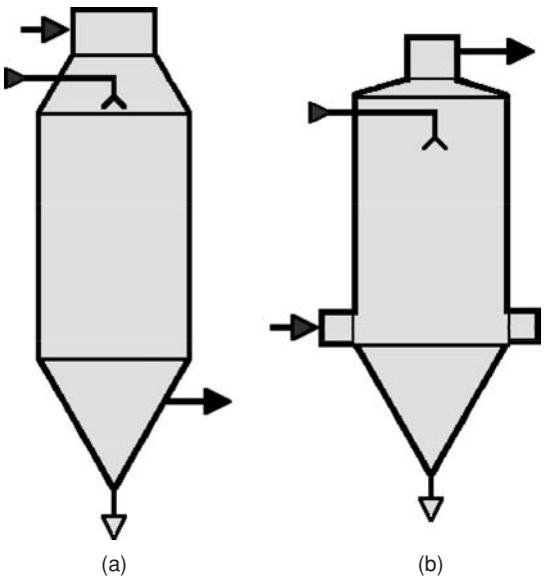

FIG. 12-94 Tall form: (*a*) cocurrent and (*b*) countercurrent chambers.

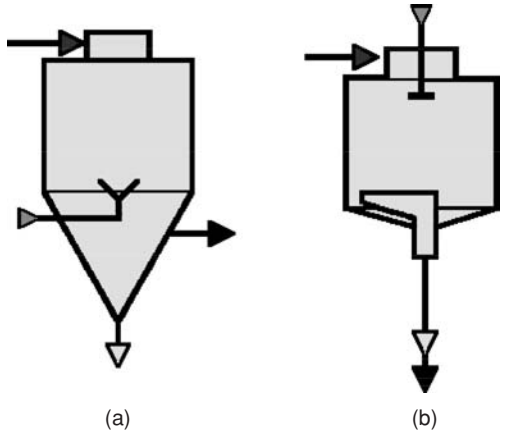

FIG. 12-95 (*a*) Mixed-flow and (*b*) flat chambers.

Figure 12-95*a* shows a mixed-flow chamber with pressure nozzle atomization arranged in so-called fountain mode. This design is ideal for producing a coarse product in a limited-size low-cost drying chamber. This type of dryer is used extensively for ceramic products. Figure 12-95*b* shows a flat-based cocurrent chamber as used with limited building height. Powder removal requires a sweeping suction device. One of few advantages is ease of access for manual cleaning. These are widely used in production of flavoring materials.

Figure 12-96*a* shows an integrated fluid-bed chamber which represents the latest development in spray dryer design. The final stage of the drying process is accomplished in a fluid bed located in the lower cone of the chamber. This type of operation allows lower outlet temperatures to be used, leading to fewer temperature effects on the powder and higher energy efficiency. Figure 12-96*b* shows an integrated belt chamber where product is sprayed onto a moving belt, which also acts as the air exhaust filter. It is highly suitable for slowly crystallizing and high-fat products. Previous operational difficulties derived from hygienic problems on the belt have been overcome, and the integrated belt dryer is now moving the limits of products that can be dried by spray drying.

Atomization/Gas Disperser Arrangement Some of the abovementioned layouts allow a choice of atomization means while others are restricted to a particular choice. The arrangement of the gas distributor means will be closely related to the choice of atomizer. A

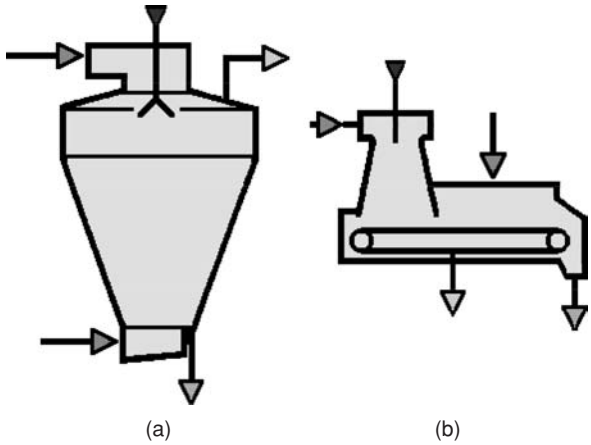

FIG. 12-96 (*a*) Integrated fluid-bed and (*b*) belt designs.

rotary atomizer will generally be arranged in a roof gas disperser as suited for the chambers in Figs. 12-93 and 12-95. The hot gas or air enters through a scroll-shaped housing which distributes the air evenly into an annular gap entry with adjustable guide vanes. The geometry and adjustment of the entry gap may determine the success of the drying process. Figure 12-95*b* shows an alternative arrangement of a rotary atomizer with a central gas disperser such as suited for the high-temperature spray dryer layout.

Hot Air Supply System All the above-mentioned chamber layouts can be used in open-cycle, partial recycle, or closed-cycle layouts. The selection is based on the needs of operation, feed, and powder specification and on environmental considerations.

An open-cycle layout is by far the most common in industrial spray drying. The open layout involves intake of drying air from the atmosphere and discharge of exhaust air to the atmosphere. Drying air can be supplemented by a waste heat source to reduce overall fuel consumption. The heater may be direct, i.e., natural gas burner, or indirect by steam-heated heat exchanger.

A closed-cycle layout is used for drying inflammable or toxic solvent feedstocks. The closed-cycle layout ensures complete solvent recovery and prevents explosion and fire risks. The reason for the use of a solvent system is often to avoid oxidation/degradation of the dried product. Consequently closed-cycle plants are gastight installations operating with an inert drying medium, usually nitrogen. These plants operate at a slight gauge pressure to prevent inward leakage of air.

Partial recycle is used in a plant type applied for products of moderate sensitivity toward oxygen. The atmospheric drying air is heated in a direct fuel-burning heater. Part of the exhaust air, depleted of its oxygen content by the combustion, is condensed in a condenser and recycled to the heater. This type of plant is also designated self-inertizing.

Industrial Applications As mentioned above, thousands of products are spray dried. The most common products may be classified as follows:

- Agrochemicals
- Catalysts
- Ceramics
- Chemicals
- Dyestuffs
- Foodstuffs
- Pharmaceuticals

Table 12-43 shows some of the operational parameters associated with specific and typical products. For each of these product groups and any other product, successful drying depends on the proper selection of a plant concept and proper selection of operational parameters, in particular inlet and outlet temperatures and the atomization method. These parameters are traditionally established through pilot-scale test work, and leading suppliers on the spray drying market often have extensive test stations to support their sales efforts.

Table 12-43 shows the variety of process parameters used in practical applications of spray drying. The air temperatures are traditionally established through experiments and test work. The inlet temperatures reflect the heat sensitivity of the different products, and the outlet temperatures the willingness of the products to release moisture. The percent water in feed parameter is an indication of feed viscosity and other properties that influence the pumpability and behavior under atomization of the individual feeds.

As a consequence, the amount of drying air or gas required for drying one unit of feed or product varies considerably. Table 12-43 shows for the individual products the ratio of drying gas to evaporation as well as the ratio of drying gas to product on a mass basis. The calculation behind the table neglects the variation of thermodynamic properties with temperature and the variation of residual moisture in each product.

A quick scoping estimate of the size of an industrial spray dryer can be made on this basis. The required evaporation rate or product rate can be multiplied by the relevant ratio from the table to give the mass flow rate of the drying gas. The next step would be to calculate the size of a spray drying chamber to allow the drying gas at outlet conditions approximately 25 s of residence time. A cylindrical chamber with diameter D and height H equal to D and a 60° conical bottom has a nominal volume of

$$V_{\text{chamber}} = \frac{\pi}{4} D^2 \times \left(H + \frac{\sqrt{3}}{2} D \right) = 1.47 \times D^3$$

Accordingly a zinc sulfate spray dryer with a drying capacity of 2 t/h would require a drying gas flow rate of approximately 8.45 kg/s. With an outlet gas density of 0.89 kg/m^3 and the above-mentioned gas residence time, this results in a required chamber volume of

$$V_{\text{chamber}} = 8.44 \text{ kg/s} / 0.89 \text{ kg/m}^3 \times 25 \text{ s} = 237 \text{ m}^3$$

The chamber size now becomes

$$D = \sqrt[3]{\frac{237}{1.47}} = 5.5 \text{ m}$$

A similar calculation for the other products based on a powder capacity of 2 t/h would reveal a variation of gas flow rates from 8.4 to 114 kg/s and chamber diameters from 5.5 to 12.7 m.

The selection of the plant concept involves the drying modes illustrated in Figs. 12-93 through 12-96. For different products a range of plant concepts are available to secure successful drying at the lowest cost. Three different concepts are illustrated in Figs. 12-97, 12-98, and 12-99.

Figure 12-97 shows a traditional spray dryer layout with a cone-based chamber and roof gas disperser. The chamber has two-point discharge and rotary atomization. The powder leaving the chamber bottom as well as the fines collected by the cyclone is conveyed pneumatically to a conveying cyclone from where the product discharges. A bag filter serves as the common air pollution control system.

Figure 12-98 shows closed-cycle spray dryer layout used to dry certain products with a nonaqueous solvent in an inert gas flow. The background for this may be product sensitivity to water and oxygen or severe explosion risk. Typical products can be tungsten carbide or pharmaceuticals.

Figure 12-99 shows an integrated fluid-bed chamber layout of the type used to produce agglomerated product. The drying process is accomplished in several stages, the first being a spray dryer with atomization. The second stage is an integrated static fluid bed located in the lower cone of the chamber. The final stages are completed in external

TABLE 12-43 Some Products That Have Been Successfully Spray Dried

Product	Air temperature, K		Water in feed, %	Air/evap. ratio, kg/kg	Air/prod. ratio, kg/kg	Product	Air temperature, K		Water in feed, %	Air/evap. ratio, kg/kg	Air/prod. ratio, kg/kg
	In	Out					In	Out			
Animal blood	440	345	65	27.6	51.3	Detergent A	505	395	50	25.4	25.4
Yeast	500	335	86	15.7	96.2	Detergent B	510	390	63	22.8	38.8
Zinc sulfate	600	380	55	12.4	15.2	Detergent C	505	395	40	25.8	17.2
Lignin	475	365	63	24.3	41.4	Manganese sulfate	590	415	50	16.3	16.3
Aluminum hydroxide	590	325	93	9.7	128.4	Aluminum sulfate	415	350	70	40.5	94.4
Silica gel	590	350	95	10.9	206.5	Urea resin A	535	355	60	14.8	22.1
Magnesium carbonate	590	320	92	9.5	108.7	Urea resin B	505	360	70	18.3	42.7
Tanning extract	440	340	46	26.4	22.5	Sodium sulfide	500	340	50	16.5	16.5
Coffee extract	420	355	70	40.6	94.8	Pigment	515	335	73	14.4	39.0

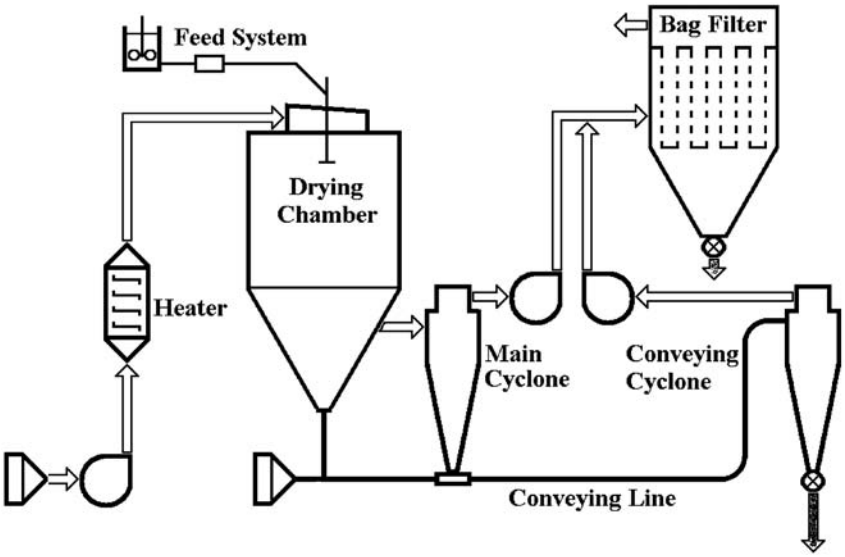

FIG. 12-97 Spray dryer with rotary atomizer and pneumatic powder conveying. (*Niro.*)

fluid beds of the vibrating type. This type of operation allows lower outlet temperatures to be used, leading to fewer temperature effects on the powder and higher energy efficiency. The chamber has a mixed-flow concept with air entering and exiting at the top of the chamber. This chamber is ideal for heat-sensitive, sticky products. It can be used with pressure nozzle as well as rotary atomization. An important feature is the return of fine particles to the chamber to enhance the agglomeration effect. Many products have been made feasible for spray drying by the development of this concept, which was initially aimed at the food and dairy industry. Recent applications have, however, included dyestuffs, agrochemicals, polymers, and detergents.

Additional Reading

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Pneumatic Conveying Dryers A gas-solids contacting operation in which the solids phase exists in a dilute condition is termed a *dispersion system.* It is often called a pneumatic system because, in most cases, the quantity and velocity of the gas are sufficient to lift and convey the solids against the forces of gravity and friction. (These systems are sometimes incorrectly called flash dryers when in fact the moisture is not actually "flashed" off. True flash dryers are sometimes used for soap drying to describe moisture removal when pressure is

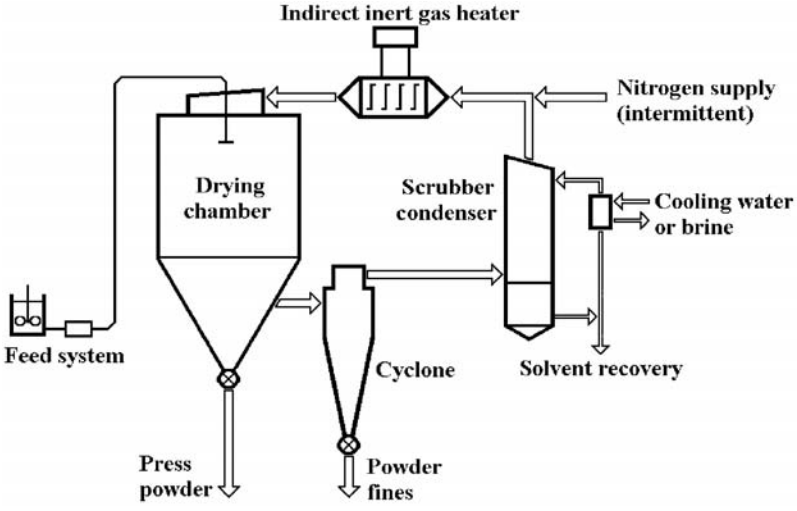

FIG. 12-98 Spray dryer with rotary atomizer and closed-cycle layout. (*Niro.*)

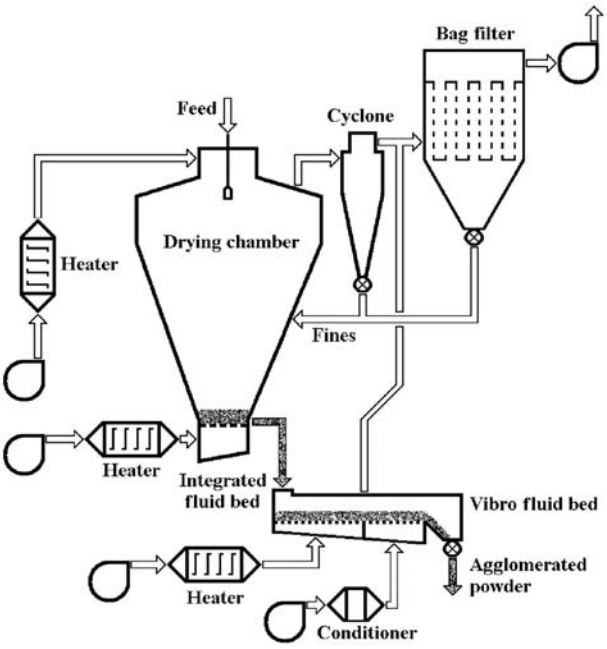

FIG. 12-99 Spray dryer with nozzle atomizer and integrated fluid bed. (*Niro.*)

quickly reduced.) Pneumatic systems may be distinguished by two characteristics:

1. Retention of a given solids particle in the system is on the average very short, usually no more than a few seconds. This means that any process conducted in a pneumatic system cannot be diffusion-controlled. The reaction must be mainly a surface phenomenon, or the solids particles must be so small that heat transfer and mass transfer from the interiors are essentially instantaneous.

2. On an energy-content basis, the system is balanced at all times; i.e., there is sufficient energy in the gas (or solids) present in the system at any time to complete the work on all the solids (or gas) present at the same time. This is significant in that there is no lag in response to control changes or in starting up and shutting down the system; no partially processed residual solids or gas need be retained between runs.

It is for these reasons that pneumatic equipment is especially suitable for processing heat-sensitive, easily oxidized, explosive, or flammable materials which cannot be exposed to process conditions for extended periods.

Gas flow and solids flow are usually cocurrent, one exception being a countercurrent flow spray dryer. The method of gas-solids contacting is best described as through-circulation; however, in the dilute condition, solids particles are so widely dispersed in the gas that they exhibit apparently no effect upon one another, and they offer essentially no resistance to the passage of gas among them.

Pneumatic Conveyor Dryers Pneumatic conveyor dryers, often also referred to as flash dryers, comprise a long tube or duct carrying a gas at high velocity, a fan to propel the gas, a suitable feeder for addition and dispersion of particulate solids in the gas stream, and a cyclone collector or other separation equipment for final recovery of solids from the gas.

The solids feeder may be of any type: Screw feeders, venturi sections, high-speed grinders, and dispersion mills are employed. For pneumatic conveyors, selection of the correct feeder to obtain thorough initial dispersion of solids in the gas is of major importance. For example, by employing an air-swept hammer mill in a drying operation, 65 to 95 percent of the total heat may be transferred within the mill itself if all the drying gas is passed through it. Fans may be of the induced-draft or the forced-draft type. The former is usually preferred because the system can then be operated under a slight negative pressure. Dust and hot gas will not be blown out through leaks in the equipment. Cyclone separators are preferred for low investment. If maximum recovery of dust or noxious fumes is required, the cyclone may be followed by a wet scrubber or bag collector.

In ordinary heating and cooling operations, during which there is no moisture pickup, continuous recirculation of the conveying gas is frequently employed. Also, solvent recovery operations employing continuously recirculated inert gas with intercondensers and gas reheaters are carried out in pneumatic conveyors.

Pneumatic conveyors are suitable for materials which are granular and free-flowing when dispersed in the gas stream, so they do not stick on the conveyor walls or agglomerate. Sticky materials such as filter cakes may be dispersed and partially dried by an air-swept disintegrator in many cases. Otherwise, dry product may be recycled and mixed with fresh feed, and then the two dispersed together in a disintegrator. Coarse material containing internal moisture may be subjected to fine grinding in a hammer mill. The main requirement in all applications is that the operation be instantaneously completed; internal diffusion of moisture must not be limiting in drying operations, and particle sizes must be small enough that the thermal conductivity of the solids does not control during heating and cooling operations. Pneumatic conveyors are rarely suitable for abrasive solids. Pneumatic conveying can result in significant particle size reduction, particularly when crystalline or other friable materials are being handled. This may or may not be desirable but must be recognized if the system is selected. The action is similar to that of a fluid-energy grinder.

Pneumatic conveyors may be single-stage or multistage. The former is employed for evaporation of small quantities of surface moisture. Multistage installations are used for difficult drying processes, e.g., drying heat-sensitive products containing large quantities of moisture and drying materials initially containing internal as well as surface moisture.

Typical single- and two-stage drying systems are illustrated in Figs. 12-100, 12-101, and 12-102. Figure 12-100 illustrates the flow diagram of a single-stage dryer with a paddle mixer, a screw conveyor followed by a rotary disperser for introduction of the feed into the airstream at the throat of a venturi section. The drying takes place in the drying column after which the dry product is collected in a cyclone. A diverter introduces the option of recycling part of the product into the mixer in order to handle somewhat sticky products. The environmental requirements are met with a wet scrubber in the exhaust stream.

Figure 12-101 illustrates a two-stage dryer where the initial feed material is dried in a flash dryer by using the spent drying air from the second stage. This semidried product is then introduced into the second-stage flash dryer for contact with the hottest air. This concept is in use in the pulp and paper industry. Its use is limited to materials that are dry enough on the surface after the first-stage to avoid plugging of the first-stage cyclone. The main advantage of the two-stage concept is the heat economy which is improved considerably over that of the single-stage concept.

Figure 12-102 is an elevation view of an actual single-stage dryer, employing an integral coarse-fraction classifier, used to separate undried particles for recycle.

Several typical products dried in pneumatic conveyors are described in Table 12-44.

Design methods for pneumatic conveyor dryers Depending upon the temperature sensitivity of the product, inlet air temperatures between 125 and 750°C are employed. With a heat-sensitive solid, a high initial moisture content should permit use of a high inlet air temperature. Evaporation of surface moisture takes place at essentially the wet-bulb air temperature. Until this has been completed, by which time the air will have cooled significantly, the surface-moisture film prevents the solids temperature from exceeding the wet-bulb temperature of the air. Pneumatic conveyors are used for solids having initial moisture contents ranging from 3 to 90 percent, wet basis. The air quantity required and solids-to-gas loading are fixed by the moisture load, the inlet air temperature, and, frequently, the exit air humidity. If the last is too great to permit complete drying, i.e., if the

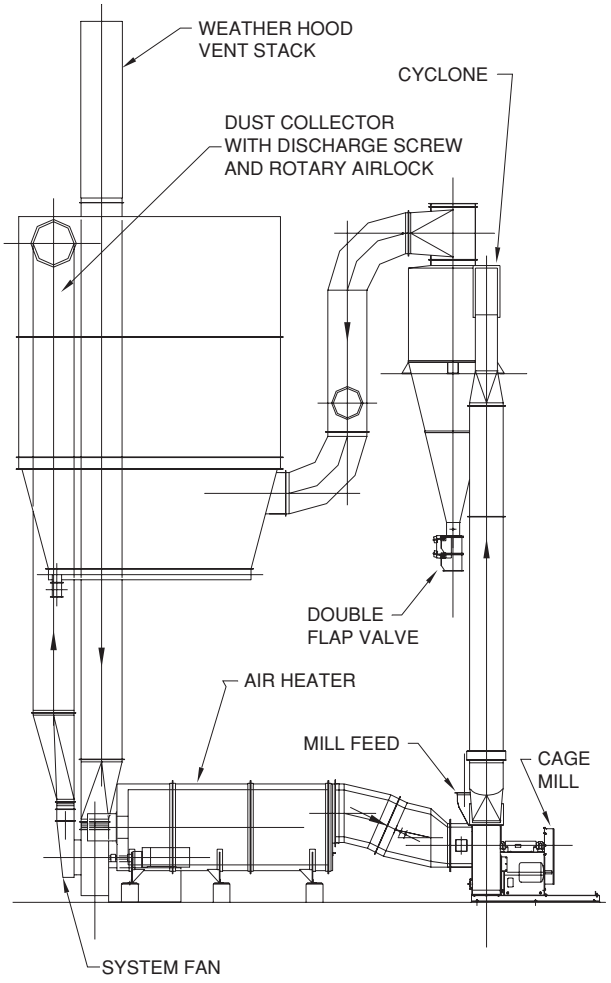

FIG. 12-100 Flow diagram of single-stage flash dryer. (*Air Preheater Company, Raymond® & Bartlett Snow™ Products.*)

exit air humidity is above that in equilibrium with the product at required dryness, then the solids/gas ratio must be reduced together with the inlet air temperature.

The gas velocity in the conveying duct must be sufficient to convey the largest particle. This may be calculated accurately by methods given in Sec. 17, "Gas-Solids Operations and Equipment." For estimating purposes, a velocity of 25 m/s, calculated at the exit air temperature, is frequently employed. If mainly surface moisture is present, the temperature driving force for drying will approach the log mean of the inlet and exit gas wet-bulb depressions. (The exit solids temperature will approach the exit gas dry-bulb temperature.)

Observation of operating conveyors indicates that the solids are rarely uniformly dispersed in the gas phase. With infrequent exceptions, the particles move in a streaklike pattern, following a streamline along the duct wall where the flow velocity is at a minimum. Complete or even partial diffusion in the gas phase is rarely experienced even with low-specific-gravity particles. Air velocities may approach 20 to 30 m/s. It is doubtful, however, that even finer and lighter materials reach more than 80 percent of this speed, while heavier and larger fractions may travel at much slower rates [Fischer, *Mech. Eng.*, **81**(11): 67–69 (1959)]. Very little information and few operating data on pneumatic conveyor dryers which would permit a true theoretical basis for design have been published.

Therefore, firm design always requires pilot tests. It is believed, however, that the significant velocity effect in a pneumatic conveyor is the difference in velocities between gas and solids, which is strongly linked to heat- and mass-transfer coefficients and is the reason why a major part of the total drying actually occurs in the feed input section.

For estimating purposes, the conveyor cross-section is fixed by the assumed air velocity and quantity. The standard scoping design method is used, obtaining the required gas flow rate from a heat and mass balance, and the duct cross-sectional area and diameter from the gas velocity (if unknown, a typical value is 20 m/s). An incremental mode may be used to predict drying conditions along the duct. However, several parameters are hard to obtain, and conditions change rapidly near the feed point. Hence, for reliable estimates of drying time and duct length, pilot-plant tests should always be used. A conveyor length larger than 50 diameters is rarely required. The length of the full-scale dryer should always be somewhat larger than required in pilot-plant tests, because wall effects are higher in small-diameter ducts. This gives greater relative velocity (and thus higher heat transfer) and lower particle velocity in the pilot-plant dryer, both effects giving a shorter length than the full-scale dryer for a given amount of drying. If desired, the length difference on scale-up can be predicted by using the incremental model and using the pilot-plant data to back-calculate the uncertain parameters; see Kemp, *Drying Technol.* **12**(1&2):279 (1994) and Kemp and Oakley (2002).

An alternative method of estimating dryer size very roughly is to estimate a volumetric heat-transfer coefficient [typical values are around 2000 $J/(m^3 \cdot s \cdot K)$] and thus calculate dryer volume.

Pressure drop in the system may be computed by methods described in Sec. 6, "Fluid and Particle Dynamics." To prevent excessive leakage into or out of the system, which may have a total pressure drop of 2000 to 4000 Pa, rotary air locks or screw feeders are employed at the solids inlet and discharge.

The conveyor and collector parts are thoroughly insulated to reduce heat losses in drying and other heating operations. Operating control is maintained usually by control of the exit gas temperature, with the inlet gas temperature varied to compensate for changing feed conditions. A constant solids feed rate must be maintained.

Ring Dryers The ring dryer is a development of flash, or pneumatic conveyor, drying technology, designed to increase the versatility of application of this technology and overcome many of its limitations.

One of the great advantages of flash drying is the very short retention time, typically no more than a few seconds. However, in a conventional flash dryer, residence time is fixed, and this limits its application to materials in which the drying mechanism is not diffusion-controlled and where a range of moisture within the final product is acceptable. The ring dryer offers two advantages over the flash dryer. First, residence time is controlled by the use of an adjustable internal classifier that allows fine particles, which dry quickly, to leave while larger particles, which dry slowly, have an extended residence time within the system. Second, the combination of the classifier with an internal mill can allow simultaneous grinding and drying with control of product particle size and moisture. Available with a range of different feed systems to handle a variety of applications, the ring dryer provides wide versatility.

The essential difference between a conventional flash dryer and the ring dryer is the manifold centrifugal classifier. The manifold provides classification of the product about to leave the dryer by using differential centrifugal force. The manifold, as shown in Fig. 12-103, uses the centrifugal effect of an airstream passing around the curve to concentrate the product into a moving layer, with the dense material on the outside and the light material on the inside.

This enables the adjustable splitter blades within the manifold classifier to segregate the denser, wetter material and return it for a further circuit of drying. Fine, dried material is allowed to leave the dryer with the exhaust air and to pass to the product collection system. This selective extension of residence time ensures a more evenly dried material than is possible from a conventional flash

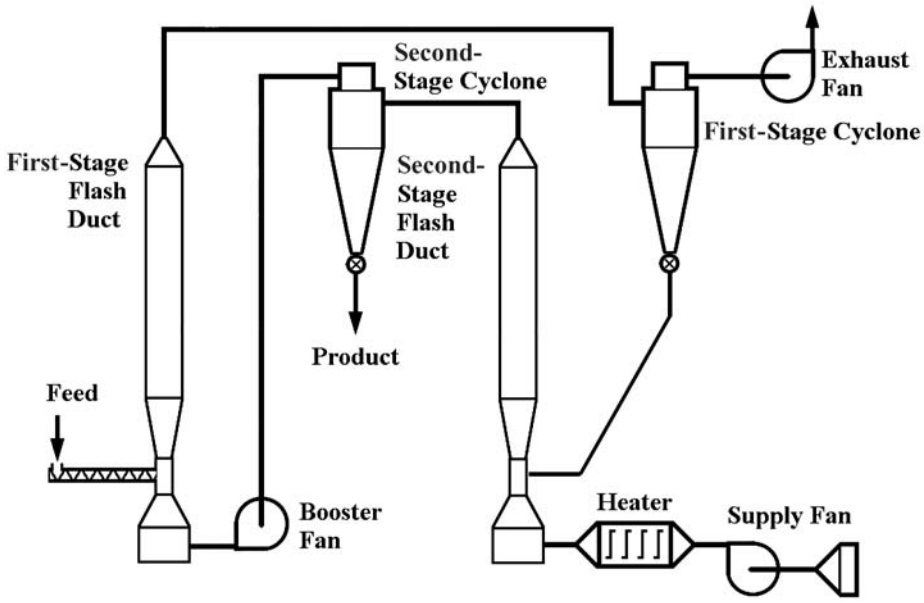

FIG. 12-101 Flow diagram of countercurrent two-stage flash dryer. (*Niro.*)

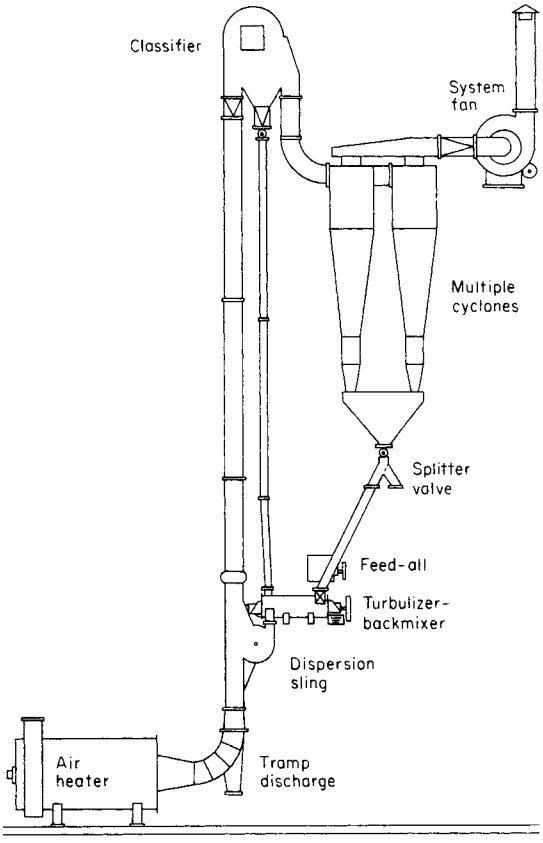

FIG. 12-102 Flow diagram of Strong Scott flash dryer with integral coarse-fraction classifier. (*Bepex Corp.*)

dryer. Many materials that have traditionally been regarded as difficult to dry can be processed to the required moisture content in a ring dryer. The recycle requirements of products in different applications can vary substantially depending upon the scale of operation, ease of drying, and finished-product specification. The location of reintroduction of undried material back into the drying medium has a significant impact upon the dryer performance and final-product characteristics.

Three configurations of the ring dryer have been developed to offer flexibility in design and optimal performance:

1. *Single-stage manifold-vertical configuration* The feed ring dryer (see Fig. 12-104) is similar to a flash dryer but incorporates a single-stage classifier, which diverts 40 to 60 percent of the product back to the feed point. The feed ring dryer is ideally suited for materials which neither are heat-sensitive nor require a high degree of classification. An advantage of this configuration is that it can be manufactured to very large sizes to achieve high evaporative capacities.

2. *Full manifold-horizontal configuration* The full ring dryer (see Fig. 12-105) incorporates a multistage classifier which allows much higher recycle rates than the single-stage manifold. This configuration usually incorporates a disintegrator which provides adjustable amounts of product grinding depending upon the speed and manifold setting. For sensitive or fine materials, the disintegrator can be omitted. Alternative feed locations are available to suit the material sensitivity and the final-product requirements. The full ring configuration gives a very high degree of control of both residence time and particle size, and is used for a wide variety of applications from small production rates of pharmaceutical and fine chemicals to large production rates of food products, bulk chemicals, and minerals. This is the most versatile configuration of the ring dryer.

3. *P-type manifold-vertical configuration* The P-type ring dryer (see Fig. 12-106) incorporates a single-stage classifier and was developed specifically for use with heat-sensitive materials. The undried material is reintroduced into a cool part of the dryer in which it recirculates until it is dry enough to leave the circuit.

An important element in optimizing the performance of a flash or ring dryer is the degree of dispersion at the feed point. Maximizing the product surface area in this region of highest evaporative

TABLE 12-44 Typical Products Dried in Pneumatic Conveyor Dryers (Barr-Rosin)

Material	Initial moisture, wet basis, %	Final moisture, wet basis, %	Plant configuration
Expandable polystyrene beads	3	0.1	Single-stage flash
Coal fines	23	1.0	Single-stage flash
Polycarbonate resin	25	10	Single-stage flash
Potato starch	42	20	Single-stage flash
Aspirin	22	0.1	Single-stage flash
Melamine	20	0.05	Single-stage flash
Com gluten meal	60	10	Feed-type ring dryer
Maize fiber	60	18	Feed-type ring dryer
Distillers dried grains (DDGs)	65	10	Feed type ring dryer
Vital wheat gluten	70	7	Full-ring dryer
Casein	50	10	Full-ring dryer
Tricalcium phosphate	30	0.5	Full-ring dryer
Zeolite	45	20	Full-ring dryer
Orange peels	82	10	Full-ring dryer
Modified com starch	40	10	P-type ring dryer
Methylcellulose	45	25	P-type ring dryer

driving force is a key objective in the design of this type of dryer. Ring dryers are fed using similar equipment to conventional flash dryers. Ring dryers with vertical configuration are normally fed by a flooded screw and a disperser which propels the wet feed into a high-velocity venturi, in which the bulk of the evaporation takes place. The full ring dryer normally employs an air-swept disperser or mill within the drying circuit to provide screenless grinding when required. Together with the manifold classifier this ensures a product with a uniform particle size. For liquid, slurry, or pasty feed materials, backmixing of the feed with a portion of the dry product will be carried out to produce a conditioned friable material. This further increases the versatility of the ring dryer, allowing it to handle sludge and slurry feeds with ease.

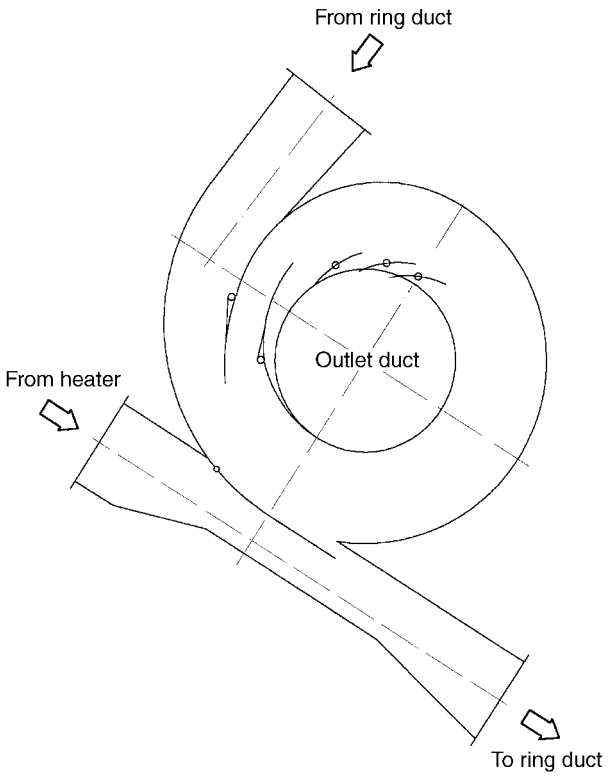

FIG. 12-103 Full manifold classifier for ring dryer. (*Barr-Rosin.*)

Dried product is collected in either cyclones or bag filters depending upon the product-particle properties. When primary collection is carried out in cyclones, secondary collection in a bag filter or scrubber is usually necessary to comply with environmental regulations. A rotary valve is used to provide an air lock at the discharge point. Screws are utilized to combine product from multiple cyclones or large bag filters. If required, a portion of the dried product is separated from the main stream and returned to the feed system for use as backmix.

Design methods for ring dryers Depending on the temperature sensitivity of the material to be processed, air inlet temperatures as high as 750°C can be utilized. Even with heat-sensitive solids, high feed moisture content may permit the use of high air inlet temperature since evaporation of surface moisture takes place at the wet-bulb air temperature. Until the surface moisture has been removed, it will prevent the solids temperature from exceeding the air wet-bulb temperature, by which time the air will generally have cooled significantly. Ring dryers have been used to process materials with feed moisture contents between 2 and 95 percent, weight fraction. The product moisture content has been controlled to values from 20 percent down to less than 1 percent.

The air velocity required and air/solids ratio are determined by the evaporative load, the air inlet temperature, and the exhaust air humidity. Too high an exhaust air humidity would prevent complete drying, so then a lower air inlet temperature and air/solids ratio would be required. The air velocity within the dryer must be sufficient to convey the largest particle, or agglomerate. The air/solids ratio must be high enough to convey both the product and backmix, together with internal recycle from the manifold. For estimating purposes a velocity of 25 m/s, calculated at dryer exhaust conditions, is appropriate both for pneumatic conveyor and ring dryers.

Agitated Flash Dryers Agitated flash dryers produce fine powders from feeds with high solids contents, in the form of filter cakes, pastes, or thick, viscous liquids. Many continuous dryers are unable to dry highly viscous feeds. Spray dryers require a pumpable feed. Conventional flash dryers often require backmixing of dry product to the feed in order to fluidize. Other drying methods for viscous pastes and filter cakes are well known, such as contact, drum, band, and tray dryers. They all require long processing time, large floor space, high maintenance, and aftertreatment such as milling.

The agitated flash dryer offers a number of process advantages, such as ability to dry pastes, sludges, and filter cakes to a homogeneous, fine powder in a single-unit operation; continuous operation; compact layout; effective heat- and mass-transfer short drying times; negligible heat loss and high thermal efficiency; and easy access and cleanability.

The agitated flash dryer (Fig. 12-107) consists of four major components: feed system, drying chamber, heater, and exhaust air system. Wet feed enters the feed tank, which has a slow-rotating impeller to break up large particles. The level in the feed tank is maintained by a

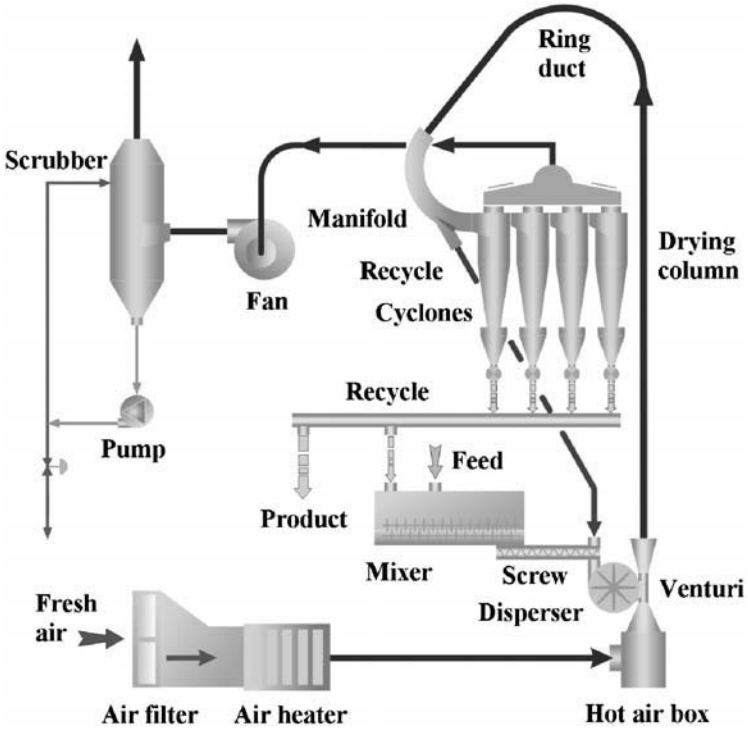

FIG. 12-104 Flow diagram of feed-type ring dryer. (*Barr-Rosin.*)

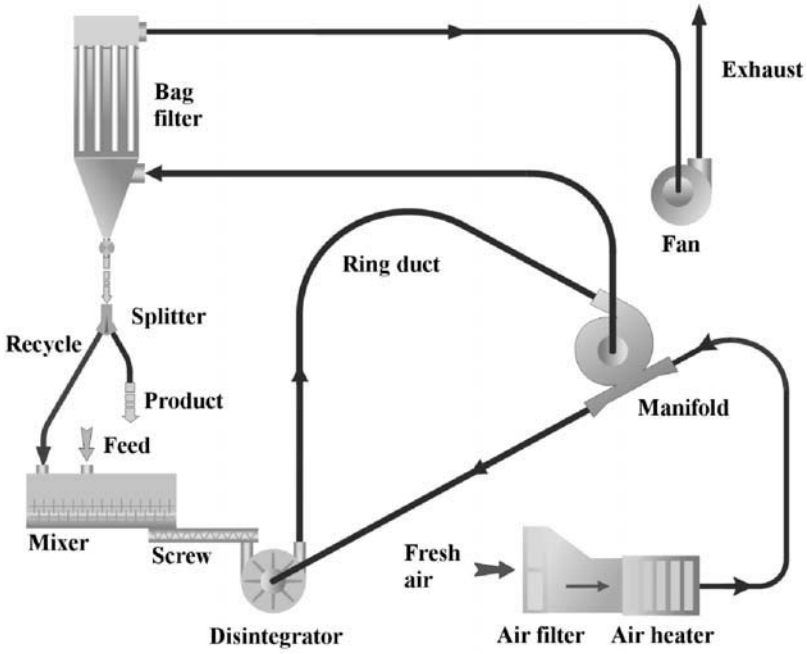

FIG. 12-105 Flow diagram of full manifold-type ring dryer. (*Barr-Rosin.*)

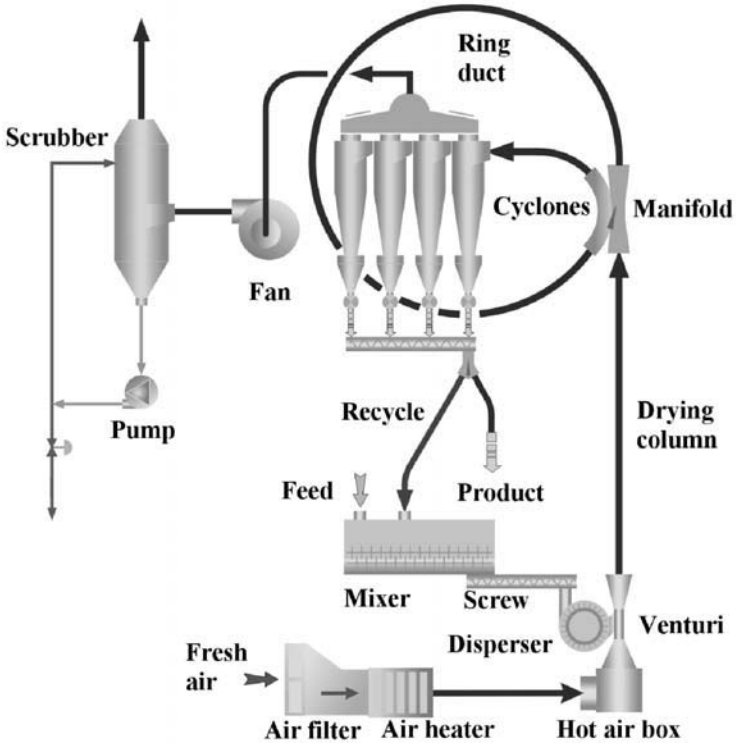

FIG. 12-106 Flow diagram of P-type ring dryer. (*Barr-Rosin.*)

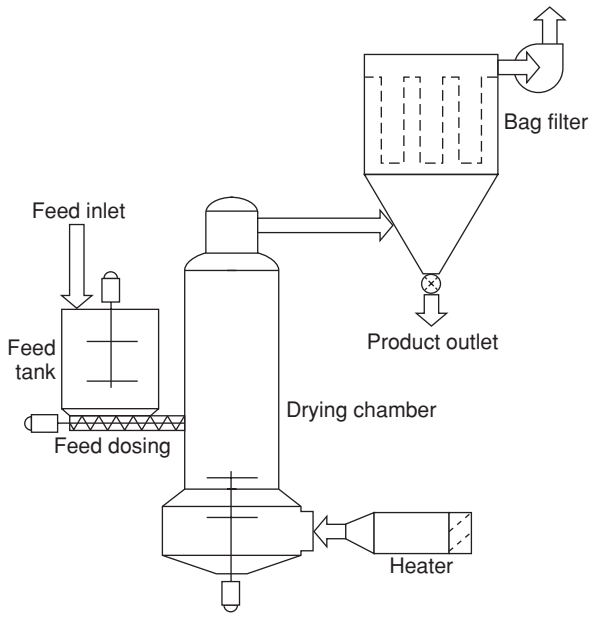

FIG. 12-107 Agitated flash dryer with open cycle. (*Niro, Inc.*)

level controller. The feed is metered at a constant rate into the drying chamber via a screw conveyor mounted under the feed tank. If the feed is shear thinning and can be pumped, the screw feeder can be replaced by a positive displacement pump.

The drying chamber is the heart of the system consisting of three important components: air disperser, rotating disintegrator, and drying section. Hot, drying air enters the air disperser tangentially and is introduced into the drying chamber as a swirling airflow. The swirling airflow is established by a guide-vane arrangement. The rotating disintegrator is mounted at the base of the drying chamber. The feed, exposed to the hot, swirling airflow and the agitation of the rotating disintegrator, is broken up and dried. The fine dry particles exit with the exhaust air and are collected in the bag filter. The speed of the rotating disintegrator controls the particle size. The outlet air temperature controls the product moisture content.

The drying air is heated either directly or indirectly, depending upon the feed material, powder properties, and available fuel source. The heat sensitivity of the product determines the drying air temperature. The highest possible value is used to optimize thermal efficiency. A bag filter is usually recommended for collecting the fine particles produced. The exhaust fan maintains a slight vacuum in the dryer, to prevent powder leakage into the surroundings. The appropriate process system is selected according to the feed and powder characteristics, available heating source, energy utilization, and operational health and safety requirements.

Open systems use atmospheric air for drying. In cases where products pose a potential for dust explosion, plants are provided with pressure relief or suppression systems. For recycle systems, the drying system medium is recycled, and the evaporated solvent recovered as condensate. There are two alternative designs. In the self-inertizing mode, oxygen content is held below 5 percent by combustion control at the heater. This is recommended for products with serious dust

explosion hazards. In the inert mode, nitrogen is the drying gas. This is used when an organic solvent is evaporated or product oxidation during drying must be prevented.

Design methods The size of the agitated flash dryer is based on the evaporation rate required. The operating temperatures are product-specific. Once established, they determine the airflow requirements. The drying chamber is designed based on air velocity (approximately 3 to 4 m/s) and residence time (product-specific).

Other Dryer Types

Freeze Dryer Industrial freeze drying is carried out in two steps:

1. Freezing of the food or beverage product
2. Freeze drying, i.e., sublimation drying of the ice content and desorption drying of the bound or crystal water content

Freeze drying differs from conventional drying in that when ice is sublimated, only water vapor is transported within the product, causing no displacement of soluble substances such as sugars, salts, and acids. In all conventional drying systems in which water is dried, the water containing the soluble substances is transported to the product surface by capillary action. The water will evaporate from the surface, leaving the soluble substances displaced on the product surface. The major advantages of freeze drying are therefore

- Preservation of original flavor, aroma, color, shape, and texture
- Very little shrinkage, resulting in excellent and instant rehydration characteristics
- Negligible product loss
- Minimal risk of cross-contamination

The freeze drying process is today used widely for a number of products including vegetables, fruits, meat, fish, and beverage products, such as

- Instant coffee for which excellent flavor and aroma retention are of special importance
- Strawberries for which excellent color preservation is of special importance
- Chives for which shape preservation is of special importance

Freezing The freezing methods applied for solid products are all conventional freezing methods such as blast freezing, individual quick freezing (IQF), or similar.

The products maintain their natural cell structure, and the aim is to freeze the free water to pure ice crystals, leaving the soluble substances as high concentrates or even crystallized. To ensure good stability of the product during storage, a product temperature of –20 to –30°C should be achieved to ensure that more than 95 percent of the free water is frozen.

Liquid products have no cell structure, thus the structure of the freeze dried products is formed by the freezing process. The intercrystalline matrix of the concentrated product giving the structure of the freeze dried product is formed around the ice crystals. The size of the ice crystals is a function of the freezing time. Quick freezing results in small ice crystals, slow freezing in large ice crystals. The structure of the matrix determines the freeze drying performance as well as the appearance, mechanical strength, and solubility rate. Small ice crystals lead to light color (high surface reflection of light), diffusion restrictions for vapor transport inside the product, and a good mechanical strength of the freeze dried product. Large ice crystals lead to the opposite results.

Thus the freezing method must be carefully adapted to the quality criteria of the finished product. The preferred methods are

- Drum freezing, by which a thin slab of 1.5 to 3 mm is frozen within 1.5 to 3 min
- Belt freezing, by which a slab of 6 to10 mm passing through different freezing zones is frozen during 10 to 20 min
- Foaming, used to influence the structure and mainly to control the density of the freeze dried product

Freeze drying Freeze drying of foods takes place in a freeze dryer at vacuum levels of 0.4 to 1.3 mbar absolute, corresponding to sublimation temperatures from –30 to –17°C depending on the product requirements. The main components of the freeze dryer are

- The vacuum chamber, heating plates, and vapor traps, all built into the freeze dryer

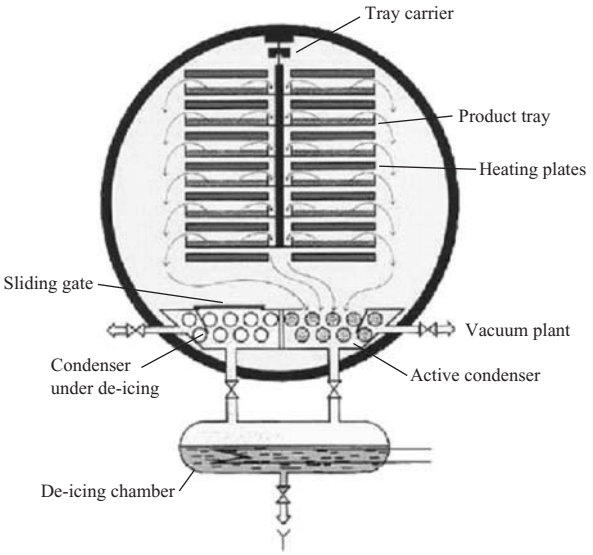

FIG. 12-108 Cross-section of RAY™ batch freeze dryer. (*Niro A/S.*)

- The external systems, such as the transport system for the product trays, the deicing system, and the support systems for supply of heat, vacuum, and refrigeration

Batch freeze drying The frozen product is carried in trays, and the trays are carried in tray trollies suspended in an overhead rail system for easy transport and quick loading and unloading. The freeze dryer as illustrated in Fig. 12-108 is charged with 3 to 6 trolley loads depending on the size of the freeze dryer. The trollies place the trays between the heating plates for radiation heat transfer. Radiation is preferred to ensure an even heat transfer over the large heating surface, typically 2× (70 to 140 m^2). The distribution of the heating medium (water or thermal oil) to the heating plates and the flow rate inside the plates are very important factors. To avoid uneven drying, the surface temperature difference of the heating plates should not exceed 2 to 3°C at maximum load.

When the loading is completed, the freeze dryer is closed and vacuum applied.

The operation vacuum should be reached quickly (within 10 min) to avoid the risk of product melting. For the same reason, the heating plates are cooled to approximately 25°C. When the operation vacuum is achieved, the heating plate temperature is raised quickly to the maximum drying temperature restricted by the capacity of the vapor traps, to perform the sublimation drying as quickly as possible for capacity reasons. During this period, the product is kept cool by the sublimation, and approximately 75 to 80 percent of the free water is sublimated.

The capability of the freeze drying plant to perform during this period is vital for efficient operation. To maintain the required sublimation temperature, the surface temperature of the ice layer on the vapor trap condenser must compensate for the pressure loss of the vapor flow from the sublimation front to the condenser.

The evaporation temperature of the refrigerant must further compensate for the temperature difference through the ice layer to the evaporating refrigerant.

With the flow rate at 1 mbar of approximately 1 $m^3/(s \cdot m^2$ of tray area), the thermodynamic design of the vapor trap is the main issue for a well-designed freeze dryer.

A built-in vapor trap allowing a large opening for the vapor flow to the condenser and a *c*ontinuous *dei*cing (CDI) system, reducing the ice layer on the condenser to a maximum of 6 to 8 mm, are important features of a modern freeze drying plant. Approximately 75 percent of the energy costs relate to the refrigeration plant, and if the requirement

TABLE 12-45 Freeze Dryer, Performance Data, Niro RAY™ and CONRAD™ Types

	Tray area, m^2	Typical sublimation capacity		Electricity consumption, kWh/kg, sublimated	Steam consumption, kg/kg sublimated
		Flat tray, kg/h	Ribbed tray, kg/h		
RAY Batch Plant—1 mbar					
RAY 75	68	68	100	1.1	2.2
RAY 100	91	91	136	1.1	2.2
RAY 125°	114	114	170	1.1	2.2
CONRAD Continuous Plant—1 mbar					
CONRAD 300	240	240	360	1.0	2.0
CONRAD 400	320	320	480	1.0	2.0
CONRAD 500°	400	400	600	1.0	2.0

°Other sizes available.

of the evaporation temperature is 10°C lower than optimum, the energy consumption of the refrigeration plant will increase by approximately 50 percent.

At the end of the sublimation drying, the product surface temperature reaches the maximum allowable product temperature, requiring that the temperature of the heating plates be lowered gradually, and the drying will change to desorption drying. The temperature will finally be kept constant at the level of the maximum allowable product temperature until the residual moisture has been reduced to 2 to 3 percent, which is a typical level for a freeze dried product.

Continuous freeze drying From the description of batch freeze drying, it can be seen that the utility requirements vary considerably. During sublimation drying the requirements are 2 to 2.5 times the average requirement. To overcome this peak load and to meet the market request for high unit capacities, continuous freeze dryer designs have been developed. The special features are twofold:

- The tray transport system is a closed-loop system in which the trays pass one by one under the tray filler, where frozen product is automatically filled into the trays at a preset weight. The full tray is charged to the vacuum lock which is then evacuated to the drier vacuum level. Then the tray is pushed into the dryer and grabbed by an elevator which is filled stepwise with a stack of trays. Next a full stack of trays is pushed into the drying area whereby each of the stacks inside the drying area will move one step forward. Thus the last stack containing the finished, freeze dried product will be pushed out of the drying area to an outlet elevator which will be emptied stepwise by discharge of the trays through the outlet vacuum lock. From the outlet vacuum lock the trays are pushed to the emptying station for emptying and then returned to the tray filler.
- As the tray stacks are pushed forward through the freeze dryer, they pass through various temperature zones. The temperature zones form the heating profile, high temperatures during the sublimation drying, medium temperatures during the transition period toward desorption drying, and low temperatures during the final desorption drying. The temperature profile is selected so that overheating of the dry surface is avoided.

Design methods The size of the freeze drying plant is based on the average sublimation capacity required as well as on the product type and form. The external systems for batch plants must be designed for a peak load of 2 to 2.5 times the average capacity in the case of a single plant. Further, a batch plant is not available for drying all the time. A modern batch freeze dryer with the CDI system loses approximately 30 min per batch. Typically, 2 to 3 batches will be freeze dried per day. The evaporation temperature of the refrigeration plant depends on the required vacuum. At 1 mbar it will be –35 to –40°C depending on the vapor trap performance. Sample data are shown in Table 12-45.

Field Effects Drying—Drying with Infrared, Radio-Frequency, and Microwave Methods

Dielectric Methods (Radio-Frequency and Microwave) Schiffmann (1995) defines dielectric (radio-frequency) frequencies as covering the range of 1 to 100 MHz, while microwave frequencies range from 300 MHz to 300 GHz. The devices used for generating microwaves are called magnetrons and klystrons.

Water molecules are dipolar (i.e., they have an asymmetric charge center), and they are normally randomly oriented. The rapidly changing polarity of a microwave or radio-frequency field attempts to pull these dipoles into alignment with the field. As the field changes polarity, the dipoles return to a random orientation before being pulled the other way. This buildup and decay of the field, and the resulting stress on the molecules, causes a conversion of electric field energy to stored potential energy, then to random kinetic or thermal energy. Hence dipolar molecules such as water absorb energy in these frequency ranges. The power developed per unit volume P_v by this mechanism is

$$P_v = kE^2 f\varepsilon' \tan \delta = kE^2 f\varepsilon'' \qquad (12\text{-}117)$$

where k is a dielectric constant, depending on the units of measurement, E is the electric field strength (V/m^3), f is the frequency, ε' is the relative dielectric constant or relative permeability, tan δ is the loss tangent or dissipation factor, and ε'' is the loss factor.

The field strength and the frequency are dependent on the equipment, while the dielectric constant, dissipation factor, and loss factor are material-dependent. The electric field strength is also dependent on the location of the material within the microwave/radio-frequency cavity (Turner and Ferguson, 1995), which is one reason why domestic microwave ovens have rotating turntables (so that the food is exposed to a range of microwave intensities). This mechanism is the major one for the generation of heat within materials by these electromagnetic fields.

There is also a heating effect due to ionic conduction, since the ions (sodium, chloride, and hydroxyl) in the water inside materials are accelerated and decelerated by the changing electric field. The collisions which occur as a result of the rapid accelerations and decelerations lead to an increase in the random kinetic (thermal) energy of the material. This type of heating is not significantly dependent on either temperature or frequency, and the power developed per unit volume P_v from this mechanism is

$$P_v = E^2 qn\mu \qquad (12\text{-}118)$$

where q is the amount of electric charge on each of the ions, n is the charge density ($ions/m^3$), and μ is the level of mobility of the ions.

Schiffmann (1995) indicates that the dielectric constant of water is over an order of magnitude higher than that of most underlying materials, and the overall dielectric constant of most materials is usually nearly proportional to moisture content up to a critical moisture content, often around 20 to 30 percent. Hence microwave and radio-frequency methods preferentially heat and dry wetter areas in most materials, a process which tends to give more uniform final moisture contents. The dielectric constant of air is very low compared with that of water, so lower density usually means lower heating rates. For water and other small molecules, the effect of increasing temperature is to decrease the heating rate slightly, hence leading to a self-limiting effect.

Other effects (frequency, conductivity, specific heat capacity, etc.) are discussed by Schiffmann (1995), but are less relevant because the range of available frequencies (which do not interfere with radio transmissions) is small (2.45 GHz, 910 MHz). Lower frequencies lead

to greater penetration depths into material than higher frequencies, with 2.45-GHz frequencies sometimes having penetration depths as low as 1 in. For in-depth heating ("volumetric heating"), radio frequencies, with lower frequencies and longer wavelengths, are often used.

Infrared Methods Infrared radiation is commonly used in the dehydration of coated films and to even out the moisture content profiles in the drying of paper and boards. The mode of heating is essentially on the material surface, and IR sources are relatively inexpensive compared with dielectric sources.

The heat flux obtainable from an IR source is given by

$$q = F\alpha\varepsilon\,(T^4_{\text{source}} - T^4_{\text{drying material}}) \qquad (12\text{-}119)$$

where q = heat flux, W/m^2; α = Stefan-Boltzmann constant = 5.67×10^{-8} W/(m$^2 \cdot$ K^4); ε = emissivity; F = view factor; and T = absolute temperature of the source or drying material.

The emissivity is a property of the material. The limiting value is 1 (blackbody); shiny surfaces have a low value of emissivity. The view factor is a fractional value that depends on the geometric orientation of the source with respect to the heating object.

It is very important to recognize the T^4 dependence on the heat flux. IR sources need to be very hot to give appreciable heat fluxes. Therefore, IR sources should not be used with flammable materials. Improperly designed IR systems can also overheat materials and equipment.

OPERATION AND TROUBLESHOOTING

Troubleshooting Dryer troubleshooting is not extensively covered in the literature, but a systematic approach has been proposed by Kemp and Gardiner (2001). The main steps of the algorithm are as follows:

- *Problem definition*—definition of the dryer problem to be solved.
- *Data gathering*—collection of relevant information, e.g., plant operating data
- *Data analysis*—e.g., heat and mass balance—and *identification of the cause of the problem*
- *Conclusions and actions*—selection and implementation of a solution in terms of changes to process conditions, equipment, or operating procedures
- *Performance auditing*—monitoring to ensure that the problem was permanently solved

There is often a danger in practice that the pressure to get the plant back into production as soon as possible may lead to some of these stages being omitted. Even if a short-term fix has been found, it is highly desirable to make sure what the problem really was, to see whether there are better ways of solving it in the long term, and to check that the problem really has been solved (sometimes it reappears later, e.g., when a temporarily cleaned heat exchanger becomes fouled again, or climatic conditions return to previous values).

The algorithm might also be considered as a "plant doctor." The doctor collects data, or symptoms, and makes a diagnosis of the cause or causes of the problem. Then alternative solutions, or treatments, are considered and a suitable choice is made. The results of the treatment are reviewed (i.e., the process is monitored) to ensure that the "patient" has returned to full health. See Fig. 12-109.

The algorithm is an excellent example of the "divergent-convergent" (brainstorming) method of problem solving. It is important to list all possible causes and solutions, no matter how ridiculous they may initially seem; there may actually be some truth in them, or they may lead to a new and better idea.

Problem Categorization In the problem definition stage, it is extremely useful to categorize the problem, as the different broad groups require different types of solution. Five main categories of dryer problems can be identified:

1. Drying performance (outlet moisture content too high, throughput too low)
2. Materials handling (dried material too sticky to get out of dryer, causing blockage)
3. Product quality (too many fines in product or bulk density too low)

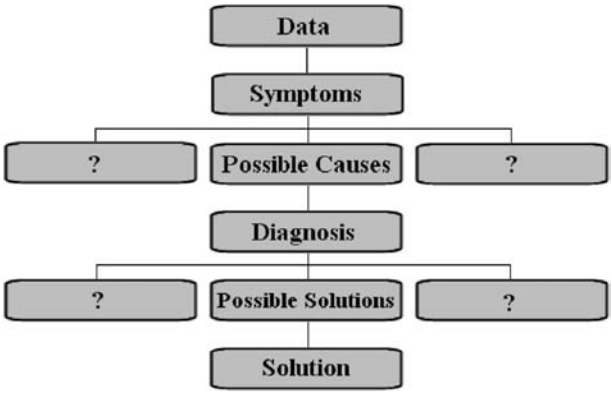

FIG. 12-109 Schematic diagram of algorithm for dryer troubleshooting.

4. Mechanical breakdown (catastrophic sudden failure)
5. Safety, health, and environmental (SHE) issues

Experience suggests that the majority of problems are of the first three types, and these are about equally split over a range of industries and dryer types. Ideally, unforeseen SHE problems will be rare, as these will have been identified in the safety case before the dryer is installed or during commissioning. Likewise, major breakdowns should be largely avoided by a planned maintenance program.

Drying Performance Problems Performance problems can be further categorized as

1. Heat and mass balance deficiencies (not enough heat input to do the evaporation)
2. Drying kinetics (drying too slowly, or solids residence time in dryer too short)
3. Equilibrium moisture limitations (reaching a limiting value, or regaining moisture in storage)

For the heat and mass balance, the main factors are

- Solids throughput
- Inlet and outlet moisture content
- Temperatures and heat supply rate
- Leaks and heat losses

As well as problem-solving, these techniques can be used for performance improvement and debottlenecking.

Drying kinetics, which are affected by temperature, particle size, and structure, are limited by external heat and mass transfer to and from the particle surface in the early stages, but internal moisture transport is the main parameter at lower moisture.

Equilibrium moisture content increases with higher relative humidity, or with lower temperature. Problems that depend on the season of the year, or vary between day and night (both suggesting a dependence on ambient temperature and humidity), are often related to equilibrium moisture content.

Materials Handling Problems The vast majority of handling problems in a dryer concern sticky feedstocks. Blockages can be worse than performance problems as they can close down a plant completely, without warning. Most stickiness, adhesion, caking, and agglomeration problems are due to *mobile liquid bridges* (surface moisture holding particles together). These are extensively described in particle technology textbooks. Unfortunately, these forces tend to be at a maximum when the solid forms the continuous phases and surface moisture is present, which is the situation for most filter and centrifuge cakes at discharge. By comparison, slurries (where the liquid forms the continuous phase) and dry solids (where all surface moisture has been eliminated) are relatively free-flowing and give fewer problems.

Other sources of problems include electrostatics (most marked with fine and dry powders) and *immobile liquid bridges*, the so-called sticky-point phenomenon. This latter is sharply temperature-dependent, with only a weak dependence on moisture content, in contrast to mobile

liquid bridges. It occurs for only a small proportion of materials, but is particularly noticeable in amorphous powders and foods and is often linked to the glass transition temperature.

Product Quality Problems (These do not include moisture level of the main solvent.) Many dryer problems either concern product quality or cannot be solved without considering the effect of any changes on product quality. Thus it is a primary consideration in most troubleshooting, although product quality measurements are specific to the particular product, and it is difficult to generalize. However, typical properties may include color, taste (not easily quantifiable), bulk density, viscosity of a paste or dispersion, dispersibility, or rate of solution. Others are more concerned with particle size, size distribution (e.g., coarse or fine fraction), or powder handling properties such as rate of flow through a standard orifice. These property measurements are nearly always made off-line, either by the operator or by the laboratory, and many are very difficult to characterize in a rigorous quantitative manner. (See also "Fundamentals" Section.)

Storage problems, very common in industry, result if the product from a dryer is free-flowing when packaged, but has caked and formed solid lumps when received by the customer. Sometimes, the entire internal contents of a bag or drum have welded together into a huge lump, making it impossible to discharge.

Depending on the situation, there are at least three different possible causes:

1. *Equilibrium moisture content*—hygroscopic material is absorbing moisture from the air on cooling.
2. *Incomplete drying*—product is continuing to lose moisture in storage.
3. *Psychrometry*—humid air is cooling and reaching its dew point.

The three types of problem have some similarities and common features, but the solution to each one is different. Therefore, it is essential to understand which mechanism is actually occurring.

Option 1: The material is hygroscopic and is absorbing moisture back from the air in storage, where the cool air has a higher relative humidity than the hot dryer exhaust. *Solution:* Pack and seal the solids immediately on discharge in tough impermeable bags (usually double- or triple-lined to reduce the possibility of tear and pinholes), and minimize the ullage (airspace above the solids in the bags) so that the amount of moisture that can be absorbed is too low to cause any significant problem. Dehumidifying the air to the storage area is also possible, but often very expensive.

Option 2: The particles are emerging with some residual moisture, and continue to dry after being stored or bagged. As the air and solids cool down, the moisture in the air comes out as dew and condenses on the surface of the solids, causing caking by mobile liquid bridges. *Solution:* If the material is meeting its moisture content specification, cool the product more effectively before storage, to stop the drying process. If the outlet material is wetter than specification, alter dryer operating conditions or install a postdryer.

Option 3: Warm, wet air is getting into the storage area or the bags, either because the atmosphere is warm with a high relative humidity (especially in the tropics) or because dryer exhaust air has been allowed to enter. As in option 2, when the temperature falls, the air goes below its dew point and condensation occurs on the walls of the storage area or inside the bags, or on the surface of the solids, leading to caking. *Solution:* Avoid high-humidity air in the storage area. Ensure the dryer exhaust is discharged a long way away. If the ambient air humidity is high, consider cooling the air supply to storage to bring it below its dew point and reduce its absolute humidity.

See Kemp and Gardiner, "An Outline Method for Troubleshooting and Problem-Solving in Dryers," *Drying Technol.* **19**(8):1875–1890 (2001).

Dryer Operation

Start-up Considerations It is important to start up the heating system before introducing product into the dryer. This will minimize condensation and subsequent product buildup on dryer walls. It is also important to minimize off-quality production by not overdrying or underdrying during the start-up period. Proper control system design can aid in this regard. The dryer turndown ratio is also an important consideration during start-up. Normally the dryer is started up at the lowest end of the turndown ratio, and it is necessary to match heat input with capacity load.

Shutdown Considerations The sequence for dryer shutdown is also very important and depends on the type of dryer. The sequence must be thoroughly thought through to prevent significant off-quality product or a safety hazard. The outlet temperature during shutdown is a key operating variable to follow.

Energy Considerations The first consideration is to minimize moisture content of the dryer feed, e.g., with dewatering equipment, and to establish as high an outlet product moisture target as possible. Other energy considerations vary widely by dryer type. In general, heating with gas, fuel oil, and steam is significantly more economical than heating with electricity. Hence RF, microwave, and infrared drying is energy-intensive. Direct heating is more efficient than indirect in most situations. Sometimes air recycle (direct or indirect) can be effective to reduce energy consumption. And generally operating at high inlet temperatures is more economical.

Recycle In almost all situations, the process system must be able to accommodate product recycle. The question is, How to handle it most effectively, considering product quality, equipment size, and energy?

Improvement Considerations The first consideration is to evaluate mass and energy balances to identify problem areas. This will identify air leaks and excessive equipment heat losses and will enable determination of overall energy efficiency.

A simplified heat balance will show what might need to be done to debottleneck a convective (hot gas) dryer, i.e., increase its production rate F.

$$F(X_I - X_O)\lambda_{ev} \approx GC_{PG}(T_{GI} - T_{GO}) - Q_{wl}$$

Before proceeding along this line, however, it is necessary to establish that the dryer is genuinely heat and mass balance limited. If the system is controlled by kinetics or equilibria, changing the parameters may have undesirable side effects, e.g., increasing the product moisture content.

The major alternatives are then as follows (assuming gas specific heat capacity C_{PG} and latent heat of evaporation λ_{ev} are fixed):

1. Increase gas flow rate G—usually increases pressure drop, so new fans and gas cleaning equipment may be required.
2. Increase inlet gas temperature T_{GI}—usually limited by risk of thermal damage to product.
3. Decrease outlet gas temperature T_{GO}—but note that this increases NTUs, outlet humidity, and relative humidity, and reduces both temperature and humidity driving forces. Hence it may require a longer drying time and a larger dryer, and may also increase equilibrium and outlet moistures X_E and X_O.
4. Reduce inlet moisture content X_I, say, by dewatering by gas blowing, centrifuging, vacuum or pressure filtration, or a predryer.
5. Reduce heat losses Q_{Wl} by insulation, removing leaks, etc.

Dryer Safety This section discusses some of the key considerations in dryer safety. General safety considerations are discussed in Sec. 23, "Safety and Handling of Hazardous Materials," and should be referred to for additional guidance.

Fires, explosions, and, to a lesser extent, runaway decompositions are the primary hazards associated with drying operations. The outbreak of fire is a result of ignition which may or may not be followed by an explosion. A hazardous situation is possible if

1. The product is combustible
2. The product is wetted by a flammable solvent
3. The dryer is direct-fired

An explosion can be caused by dust or flammable vapors, both of which are fires that rapidly propagate, causing a pressure rise in a confined space.

Dust Explosions Dispersion dryers can be more hazardous than layer-type dryers if we are drying a solid combustible material which is then dispersed in air, particularly if the product is a fine particle size. If this finely dispersed product is then exposed to an ignition source, an explosion can result. The following conditions (van't Land, *Industrial Drying Equipment,* Marcel Dekker, New York, 1991) will be conducive to fire and explosion hazard:

1. Small particle sizes, generally less than 75 μm, which are capable of propagating a flame
2. Dust concentrations within explosive limits, generally 10 to 60 g/m^3
3. Ignition source energy of 10 to 1000 mJ or as low as 5 mJ for highly explosive dust sources
4. Atmosphere supporting combustion

Since most product and hence dust compositions vary widely, it is generally necessary to do quantitative testing in approved test equipment.

Flammable Vapor Explosions This can be a problem for products wetted by flammable solvents if the solvent concentration exceeds 0.2% v/v in the vapor phase. The ignition energy of vapor-air mixtures is lower (< 1 mJ) than that of dust-air suspensions. Many of these values are available in the literature, but testing may sometimes be required.

Ignition Sources There are many possible sources of an ignition, and they need to be identified and addressed by both designers and operators. A few of the most common ignition sources are

1. Spontaneous combustion
2. Electrostatic discharge
3. Electric or frictional sparks
4. Incandescent solid particles from heating system

Safety hazards must be addressed with proper dryer design specifications. The following are a few key considerations in dryer design.

Inert system design The dryer atmosphere is commonly inerted with nitrogen, but superheated steam or self-inertized systems are also possible. Self-inertized systems are not feasible for flammable solvent systems. These systems must be operated with a small overpressure to ensure no oxygen ingress. And continuous on-line oxygen concentration monitoring is required to ensure that oxygen levels remain well below the explosion hazard limit.

Relief venting Relief vents that are properly sized relieve and direct dryer explosions to protect the dryer and personnel if an explosion does occur. Normally they are simple pop-out panels with a minimum length of ducting to direct the explosion away from personnel or other equipment.

Suppression systems Suppression systems typically use an inert gas such as carbon dioxide to minimize the explosive peak pressure rise and fire damage. Dryer operating pressure must be properly monitored to detect the initial pressure rise followed by shutdown of the dryer operating systems and activation of the suppression system.

Clean design Care should be taken in the design of both the dryer and dryer ancillary (cyclones, filters, etc.) equipment to eliminate ledges, crevices, and other obstructions which can lead to dust and product buildup. Smooth drying equipment walls will minimize deposits. This can go a long way in prevention. No system is perfect, of course, and a routine cleaning schedule is also recommended.

Start-up and shutdown Start-up and shutdown situations must be carefully considered when designing a dryer system. These situations can create higher than normal dust and solvent concentrations. This coupled with elevated temperatures can create a hazard well beyond normal continuous operation.

Environmental Considerations Environmental considerations are continuing to be an increasingly important aspect of dryer design and operation as environmental regulations are tightened. The primary environmental problems associated with drying are particulate and volatile organic compound (VOC) emissions. Noise can be an issue with certain dryer types.

Environmental Regulations These vary by country, and it is necessary to know the specific regulations in the country in which the dryer will be installed. It is also useful to have some knowledge of the direction of regulations so that the environmental control system is not obsolete by the time it becomes operational.

Particulate emission problems can span a wide range of hazards. Generally there are limits on both toxic and nontoxic particles in terms of annual and peak emissions limits. Particles can present toxic, bacterial, viral, and other hazards to human, animal, and plant life.

Likewise, VOC emissions can span a wide range of hazards and issues from toxic gases to smelly gases.

Environmental Control Systems We should consider environmental hazards before the drying operation is even considered. The focus should be on minimizing the hazards created in the upstream processing operations. After potential emissions are minimized, these hazards must be dealt with during dryer system design and then subsequently with proper operational and maintenance procedures.

Particle Emission Control Equipment The four most common methods of particulate emissions control are as follows:

1. *Cyclone separators* The advantage of cyclones is they have relatively low capital and operating costs. The primary disadvantage is that they become increasingly ineffective as the particle size decreases. As a general rule of thumb, we can say that they are 100 percent efficient with particles larger than 20 μm and 0 percent efficient with particles smaller than 1 μm. Cyclones can also be effective precleaning devices to reduce the load on downstream bag filters.
2. *Scrubbers* The more general classification is wet dedusters, the most common of which is the wet scrubber. The advantage of wet scrubbers is that they can remove fine particles that the cyclone does not collect. The disadvantages are they are more costly than cyclones and they can turn air contamination into water contamination, which may then require additional cleanup before the cleaning water is put to the sewer.
3. *Bag filters* The advantages of filters are that they can remove very fine particles and bag technologies continue to improve and enable ever-smaller particles to be removed without excessive pressure drops or buildup. The primary disadvantages are higher cost relative to cyclones and greater maintenance costs, especially if frequent bag replacement is necessary.
4. *Electrostatic precipitators* The capital cost of these systems is relatively high, and maintenance is critical to effective operation.

VOC Control Equipment The four most prevalent equipment controls are

1. *Scrubbers* Similar considerations as above apply.
2. *Absorbers* These systems use a high-surface-area absorbent, such as activated carbon, to remove the VOC absorbate.
3. *Condensers* These systems are generally only feasible for recovering solvents from nonaqueous wetted products.
4. *Thermal and catalytic incinerators* These can be quite effective and are generally a low capital and operating cost solution, except in countries with high energy costs.

Noise Noise analysis and abatement is a very specialized area. Generally, the issue with dryers is associated with the fans, particularly for systems requiring fans that develop very high pressures. Noise is a very big issue that needs to be addressed with pulse combustion dryers and can be an issue with very large dryers such as rotary dryers and kilns.

Additional considerations regarding environmental control and waste management can be found in Secs. 22, "Waste Management," and 23, "Process Safety."

Control and Instrumentation The purpose of the control and instrumentation system is to provide a system that enables the process to produce the product at the desired moisture target and that meets other quality control targets discussed earlier (density, particle size, color, solubility, etc.). This segment discusses key considerations for dryer control and instrumentation. Additional more detailed information can be found in Sec. 8, "Process Control."

Proper control of product quality starts with the dryer selection and design. Sometimes two-stage or multistage systems are required to meet product quality targets. Multistage systems enable us to better control temperature and moisture profiles during drying. Assuming the proper dryer design has been selected, we must then design the control and instrumentation system to ensure we meet all product quality targets.

Manual versus Automatic Control Dryers can be controlled either manually or automatically. Generally lab-, pilot-, and small-scale production units are controlled manually. These operations are usually batch systems, and manual operation provides lower cost and greater flexibility. The preferred mode for large-scale, continuous dryers is automatic.

Key Control Variables Product moisture and product temperature are key control variables. Ideally both moisture and temperature measurement are done on-line, but frequently moisture measurement is done off-line and temperature (or exhaust air temperature) becomes the primary control variable. And generally, inlet temperature

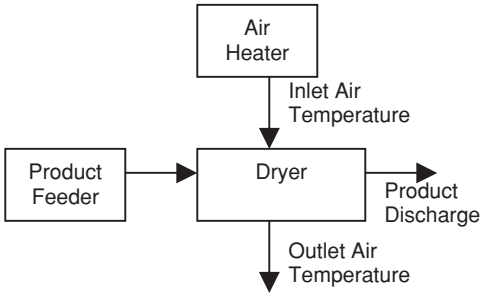

FIG. 12-110 Typical dryer system.

will control the rate of production and outlet temperature will control the product moisture and other product quality targets.

Common Control Schemes Two relatively simple, but common control schemes in many dryer systems (Fig. 12-110) are as follows:

1. Outlet air temperature is controlled by feed rate regulation with inlet temperature controlled by gas heater regulation.
2. Outlet air temperature is controlled by heater regulation with feed rate held constant.

Alternatively, product temperatures can replace air temperatures with the advantage of better control and the disadvantage of greater maintenance of the product temperature sensors.

Other Instrumentation and Control

Pressure Pressure and equipment pressure drops are important to proper dryer operation. Most dryers are operated under vacuum. This prevents dusting to the environment, but excess leakage in decreases dryer efficiency. Pressure drops are especially important for stable fluid-bed operation.

Air (gas) flow rate Obviously gas flows are another important parameter for proper dryer operation. Pitot tubes are useful when a system has no permanent gas flow sensors. Averaging pitot tubes work well in permanent installations. The devices work best in straight sections of ductwork which are sometimes difficult to find and make accurate measurement a challenge.

Product feed rate It's important to know that product feed rates and feed rate changes are sometimes used to control finished product moistures. Weigh belts are common for powdered products, and there is a wide variety of equipment available for liquid feeds. Momentum devices are inexpensive but less accurate.

Humidity The simplest method is sometimes the best. Wet- and dry-bulb temperature measurement to get air humidity is simple and works well for the occasional gas humidity measurement. The problem with permanent humidity measurement equipment is the difficulty of getting sensors robust enough to cope with a hot, humid, and sometimes dusty environment.

Interlocks Interlocks are another important feature of a well-designed control and instrumentation system. Interlocks are intended to prevent damage to the dryer system or to personnel, especially during the critical periods of start-up and shutdown. The following are a few key interlocks to consider in a typical dryer system.

Drying chamber damage This type of damage can occur when the chamber is subjected to significant vacuum when the exhaust fans are started up before the supply fans.

Personnel injury This interlock is to prevent injury due to entering the dryer during operation, but more typically to prevent dryer start-up with personnel in the main chamber or inlet or exhaust air ductwork on large dryers. This typically involves microswitches on access doors coupled with proper door lock devices and tags.

Assurance of proper startup and shutdown These interlocks ensure, e.g., that the hot air system is started up before the product feed system and that the feed system is shut down before the hot air system.

Heater system There are a host of important heater system interlocks to prevent major damage to the entire drying system. Additional details can be found in Sec. 23, "Process Safety."

Drying Software Several software programs for psychrometric charts and calculations are available and are described in the "Psychrometry" section.

Dryers are included as modules in standard process simulators such as Aspen Plus and HYSYS (Aspen Technology), Pro/II (Simsci/Invensys) and Unisim (Honeywell), and the prototype Solidsim solids process simulator. These are confined (as of 2006) to heat and mass balances or, at most, simple scoping design.

Many higher-level dryer models have been produced by researchers and universities, but they are not commercially available. Windows-based drying programs are available in the Process Tools (Aspen Technology), including a psychrometric chart, dryer selection expert system, dryer scoping design, and fluid-bed dryer simulation. Some CFD programs (e.g., Fluent, CFX) include a module for spray dryers.

In addition to textbooks, a detailed online knowledge base, the Process Manual, is available from Aspen Technology (see www.processmanual.com). This covers equipment, scientific background, design, and operation for drying and 10 other technical areas in solids and separation processes. Company and university licenses are available.

Detailed reviews of drying software packages given by Menshutina and Kudra, *Drying Technol.* **19**(8):1825–1850 (2001); and by Kemp, Chap. 7 in *Modern Drying Technology*, vol. 1, Wiley–VCH (2007), and *Drying Technol.* **25** (2007).