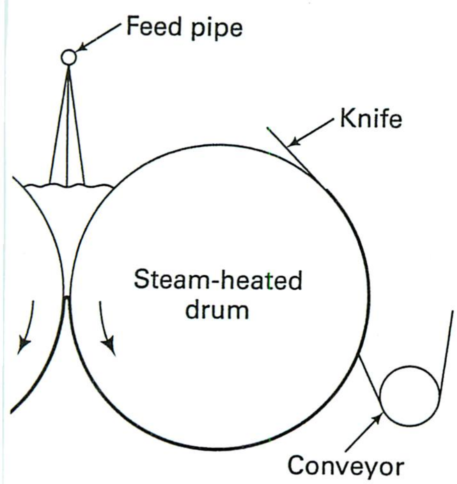
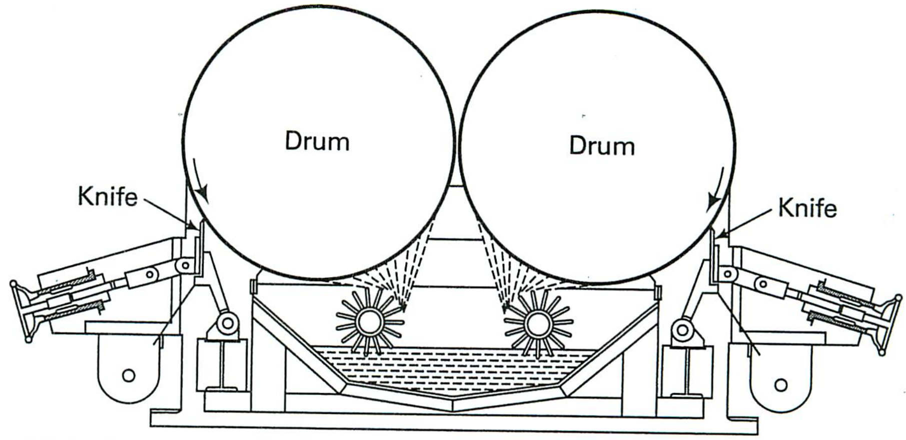
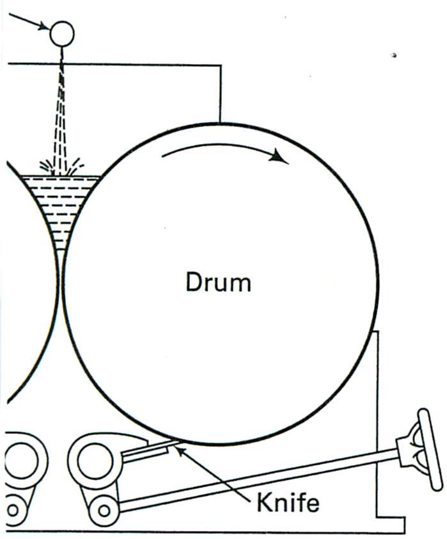
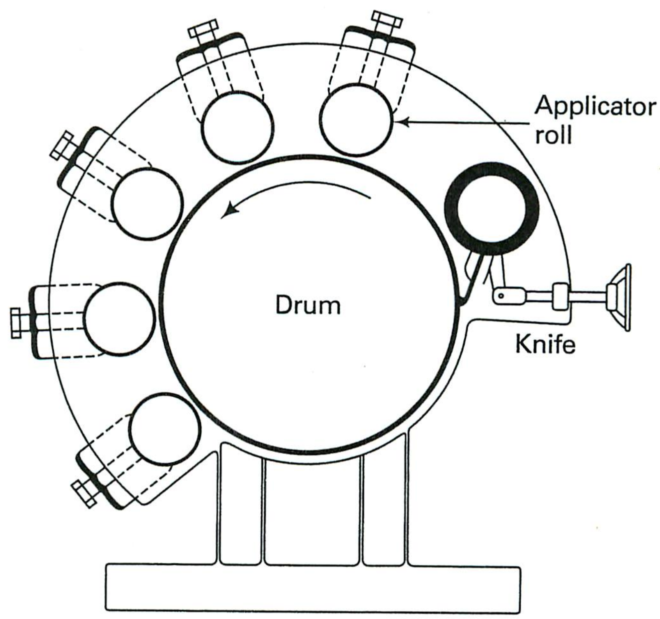




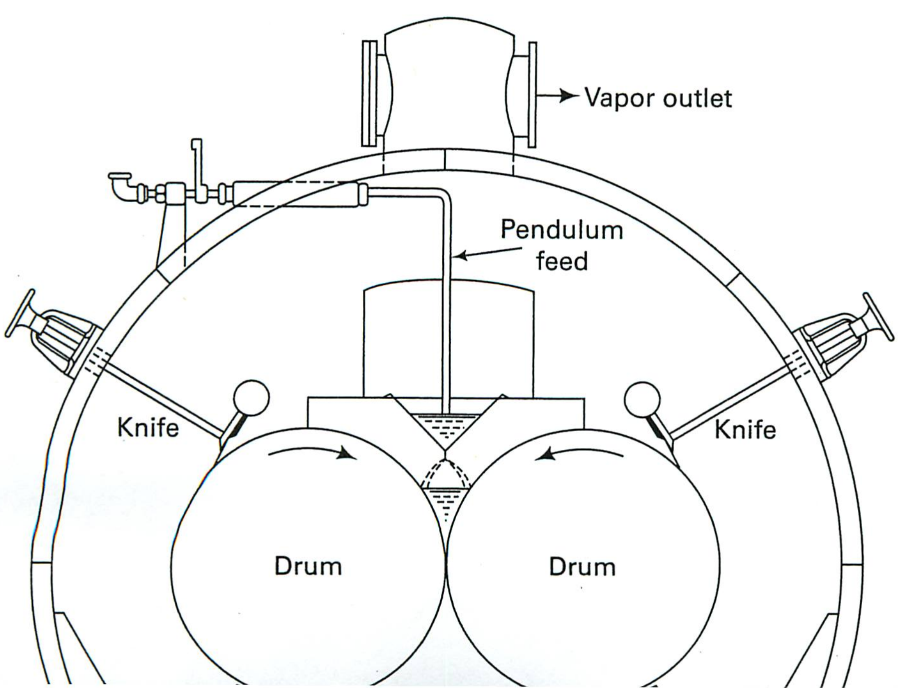
(c) Twin-drum dryer with splash feed



feed



(d) Single-drum dryer with applicator feed

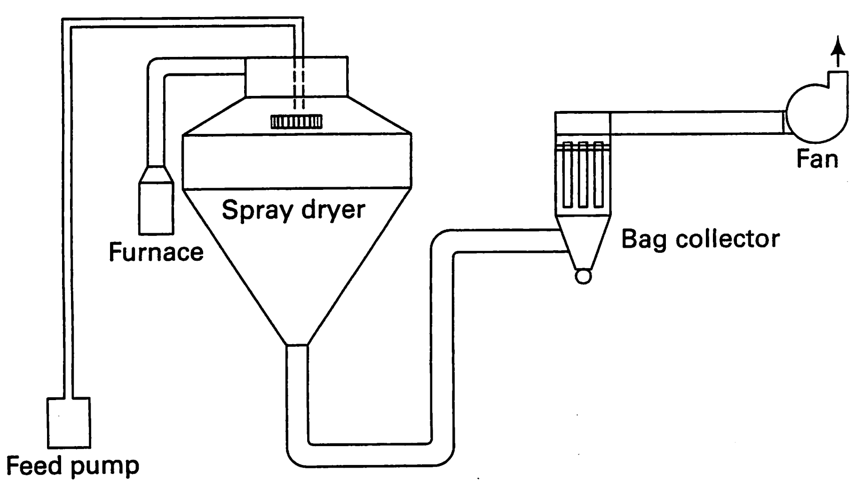


Large flash dryers are provided with pneumatic-conveying dryer ducts 1 m in diameter and 12 m high, with water-evaporation capacities up to 36,000 lb/h. Compared to many other dryers, they have small floor-area requirements and are used for drying filter cakes, centrifuge cakes and slurries, yeast cakes, whey, starch, sewage sludge, gypsum, fruit pulp, copper sulfate, clay, coal, chicken droppings, adipic acid, polystyrene beads, ammonium sulfate, and hexamethylene tetramine.

Spray Dryers

When solutions, slurries, or pumpable pastes—containing more than 50 wt% moisture, at rates greater than 1,000 lb/h—are to be dried, a spray dryer should be considered. In the configuration in Figure 18.13a the drying chamber has a conical-shaped bottom section with a top diameter that may be nearly equal to the chamber height. Feed is pumped to the top center of the chamber, where it is dispersed into droplets or particles from 2 to 2,000 μm by any of three types of atomizers: (1) single-fluid pressure nozzles, (2) pneumatic nozzles, and (3) centrifugal disks or spray wheels. Hot gas enters the chamber, causing moisture in the atomized feed to rapidly evaporate. Gas flows cocurrently to the solids, and dried solids and gas are either partially separated in the chamber, followed by removal of dust from the gas by a cyclone separator, or, as shown in Figure 18.13a, are sent together to a cyclone separator, bag filter, or other gas–solid separator. The hot gas can be moved by a fan.

In many respects, spray dryers are similar in operating conditions to a pneumatic-conveyor dryer because particles



(a) Process system

are small, entering gas temperature can be h
time of the particles is short, mainly surfa
removed, and temperature-sensitive mate
handled. However, a unique feature of spray
ability, with some materials such as dyes, foo
gents, to produce, from a solution, rounded p
of fairly uniform size that can be rapidly dissol
in subsequent applications.

Although residence times are less than 5 s i
moisture is removed, residence times of up t
provided for evaporating internal moisture. S
also unique in that it combines, into one c
of equipment, evaporation, crystallization or
filtration or centrifugation, size reduction,
and drying.

A critical part of a spray dryer is the atomize
three atomizer types has advantages and d
Pneumatic (two-fluid) nozzles impinge the gas
relatively low pressures of up to 100 psig, but ar
at high capacities. Consequently, they find app
in pilot plants and low-capacity commercial p
tions are the dispersion of stringy and fibro
thick pastes, certain filter cakes, and polymer
cause with high atomizing gas-to-feed ratios, s
are produced.

Pressure (single-fluid) nozzles, with orifice
0.012–0.15 inch, require solution inlet press
4,000 psig to achieve breakup of the feed strea
zles can deliver the narrowest range of droplet
droplets are the largest delivered by the three ty
izers, and multiple nozzles are required in la
spray dryers. Also, orifice wear and plugging
lems with some feeds. Because the spray is la
ward, chambers are relatively slender and tall,
to-diameter ratios of 4–5.

The centrifugal disks (spray wheels) show
18.13b handle solutions or slurries, delivering t
feed that break up into small droplets in a nearly
tion at high capacities. Disks have the largest-di
pattern and therefore require the largest-diam
chambers to prevent particles from striking the c
while in a sticky state.

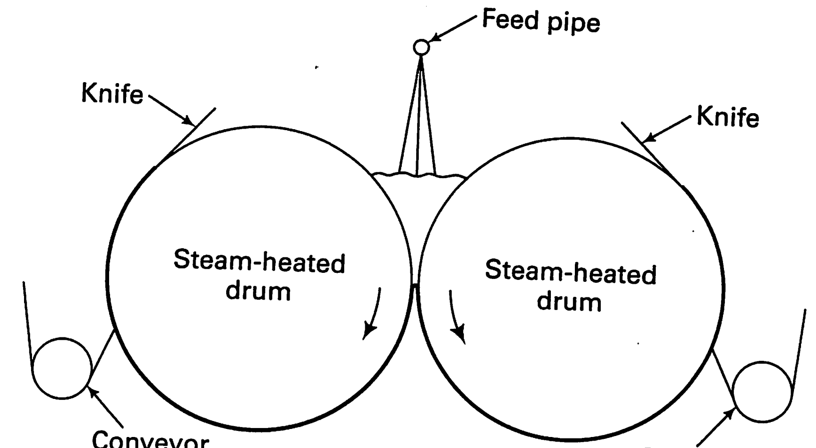
Centrifugal disks range in diameter from a fe
to 32 inches in large units. Disks spin at 3,000–
and can operate over a range of feed and rotatio
out significantly affecting the particle-size distr

to dry a wide variety of materials,
potatoes, skim milk, malted milk,
and vegetable glue; slurries of
$CaCO_3$; and solutions of sodium
$_4$, $CrSO_4$, and various organic com-
ng to a class of hot-cylinder dryers.
of cylinders in series and parallel
; sheets of woven fabrics and paper
f about 10 kg/h-m^2.

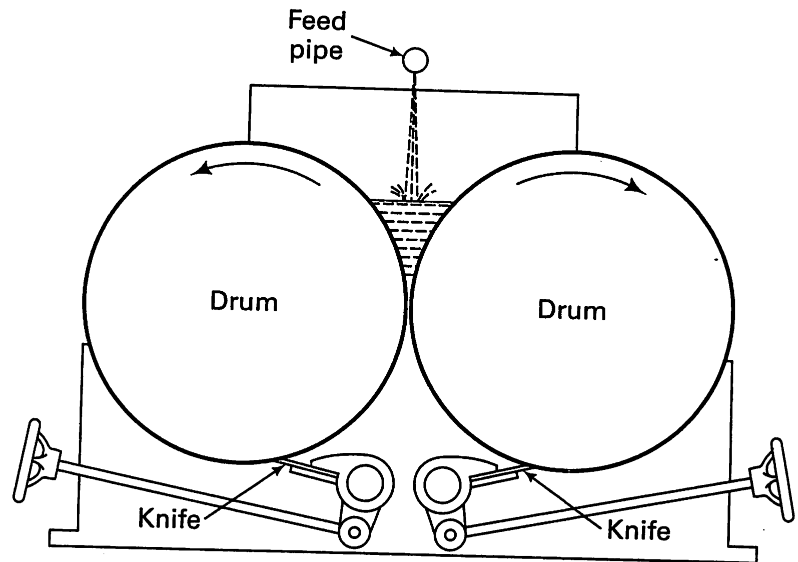
s

s have been developed for special
ared radiant energy, generation of
dielectric drying using radio or
nd freeze-drying by sublimation of

ransfer of heat by convection from
rial is often inadvertently supple-
on from hot surfaces that surround
heat contribution is usually minor,
g of certain films, sheets, and coat-
nal radiation as the major source of
y.
sed from matter as a result of oscil-
s electrons. For gases and transpar-
radiation can be emitted from
the matter. For opaque solids and
tion is quickly absorbed by adjoin-
et transfer of energy by radiation is
great importance in radiation heat
ansfer of radiation from a hot, opa-
nabsorbing gas or vacuum to the
transfer can be viewed as the prop-
ns (quanta) and/or as the propaga-
waves, consisting, as shown in
and magnetic fields that oscillate at
and to the direction in which the
n in the electromagnetic spectrum
elength, λ, of the radiation, which
n which it is generated, covers an
from gamma rays of 10^{-8} μm to
μm. Regardless of the wavelength

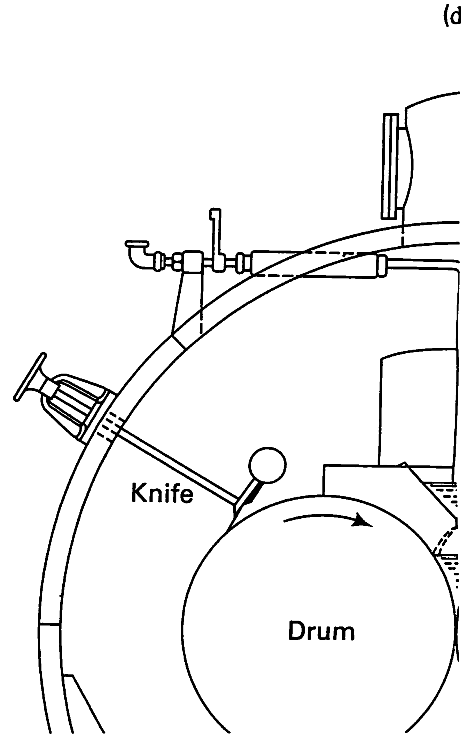


(a) Double-drum dryer



(b) Twin-drum dryer with top feed

(d



properties of a fluid. *Fluidization* is initiated when the gas velocity reaches the point where all the particles are suspended by the gas. As the gas velocity is increased further, the bed expands and bubbles of gas are observed to pass up through the bed. This regime of fluidization is referred to as *bubbling fluidization* and is the most desirable regime for most fluidized-bed operations, including drying. If the gas velocity is increased further, a transition to *slugging fluidization* eventually occurs, where bubbles coalesce and spread to a size that approximates the diameter of the vessel. To some extent, this behavior can be modified by placing baffles and low-speed agitators in the bed.

Before fluidization occurs, when the bed of solids is fixed, the pressure drop across the bed for gas flow, ΔP_b, is predicted by the Ergun [30] equation, discussed in §6.8.2:

$$\frac{\Delta P_b}{L_b} = 150\frac{(1-\epsilon_b)^2}{\epsilon_b^3}\frac{\mu u_s}{(\phi_s d_p)^2} + 1.75\frac{(1-\epsilon_b)}{\epsilon_b^3}\frac{\rho_g u_s^2}{\phi_s d_p} \tag{18-85}$$

where L_b = bed height, u_s = superficial-gas velocity, and ϕ_s = particle sphericity. The first term on the RHS is dominant at low-particle Reynolds numbers where streamline flow exists, and the second term dominates at high-particle Reynolds numbers where turbulent flow exists.

The onset of fluidization occurs when the drag force on the particles by the upward-flowing gas becomes equal to the weight of the particles (accounting for displaced gas):

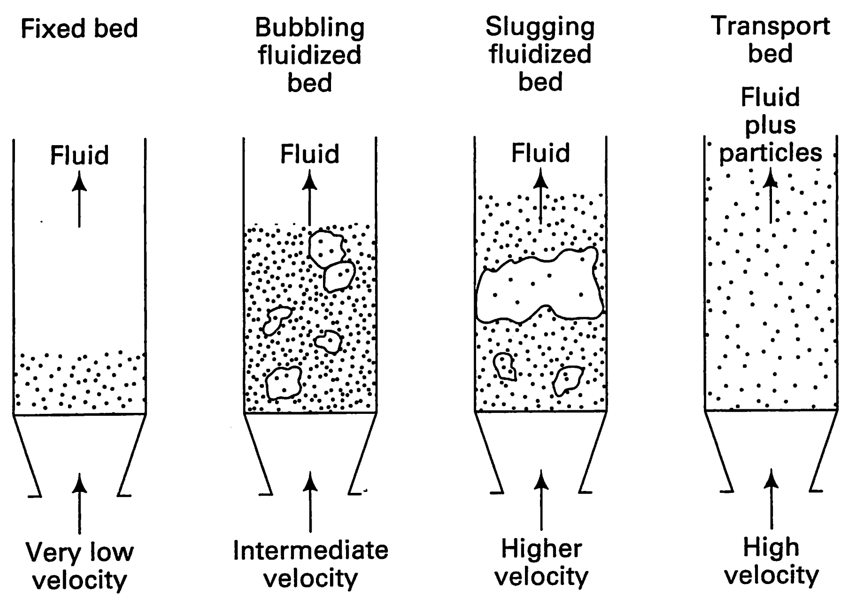


Figure 18.38 Regimes of fluidization of a bed of particles by a gas.

The minimum gas-fluidization sup
obtained by solving (18-85) and (18
$u = u_{mf}$. For $N_{\mathrm{Re},p} = d_p u_{mf} \rho_g/\mu <$
contribution to (18-85) is negligible a

$$u_{mf} = \frac{d_p^2\left(\rho_p - \rho_g\right)g}{150\mu}\left(\frac{\epsilon_b^3}{1-\ldots}\right.$$

For operation in the bubbling
superficial-gas velocity of $u_s = 2u_{mf}$
At this velocity, the bed will be ex
with no further increase in pressure
this regime, the solid particles are
temperature is so uniform that fluidi
trially to calibrate thermocouples a
fluidized bed is operated continuous
tions rather than batchwise with res
particles will have a residence-time d
fluid in a continuous-stirred-tank rea
ticles will be in the dryer for only a v
and will experience almost no decre
Other particles will be in the dryer f
come to equilibrium before that tim
dried solids will have a distribution c
is in contrast to a batch-fluidization
cles have the same residence time a
final moisture content. This is an
because continuous, fluidized-bed d
up from data obtained in small, bat
Therefore, it is important to have
batch drying time and continuous dry

The distribution of residence time
fectly mixed vessel operating at conti
ditions is given by Fogler [31] as

$$E\{t\} = \exp(-t/\ldots$$

where τ is the average residence ti
such that $E\{t\}dt$ = the fraction of e
time between t and $t + dt$. Thus, $\int_0^{t_1}$
effluent with a residence time less tha
average particle-residence time is 10
cles will have a residence time of les
particles are small and nonporous su
place in the constant-rate period, and
plete drying, then

$$\frac{X}{X_o} = 1 - \frac{t}{\theta}, \quad t \ldots$$