Continuous crystallizers serve a wide range of applications in the chemical process industries. Optimize system performance by heeding these practical design and operating principles.

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Continuous crystallizers are frequently employed in the chemical industry for the production of commodity and specialty chemicals. The evaporative mode of these crystallizers is also utilized for batch processing of pharmaceuticals and low-volume specialty chemicals (1).

The engineer has a number of options when choosing continuous units, each of which has its own merits. The configurations include the forced circulation (FC), draft-tube-baffle (DTB) and Oslo or fluidized-suspension (FS) units. This article discusses the characteristics of these crystallizers along with guidelines for selection.

The evaporation can be achieved either adiabatically or isothermally by means of indirect heat input via a heat exchanger. The choice depends on the slope of the solubility curve and the feed conditions, including concentration and temperature. For example, sodium chloride (NaCl), with its near-horizontal solubility curve, requires indirect steam heating to achieve evaporation, while adipic acid, with its upward sloping solubility curve in the operating region, can be produced by adiabatic crystallization. For the latter case, the percent solvent evaporation is often 3–7%.

Continuous units are normally operated either in a mixed-suspension, mixed-product removal (MSMPR) mode, in which case they are also called circulating magna crystallizers, or as configurations that modify the crystal size distribution (CSD) via fines destruction, clear liquid advance (deliberate removal of mother liquor (ML)) to change the slurry density and/or product classification. Classification is a separation process based on the principle that the residence time (RT) of the crystals depends on their sizes. Both the CSD and average particle sizes can be changed by varying the crystal RT and slurry density. Additional improvements can be achieved by fines destruction and classified product removal.

The goal of most evaporative crystallizations is to produce a product with a consistently large narrow CSD. All configurations are designed to ensure that crystals are present at the boiling surface or where the supersaturation is being relieved. This encourages growth on existing crystals, reduces nucleation and extends the run time for the units by reducing encrustation or scaling. The CSD is influenced by the kinetics of nucleation and growth, along with the level of supersaturation. The mechanical design of the crystallizer can greatly impact the nucleation rate by means of secondary nucleation caused by crystal-to-crystal, crystal-to-impeller and crystal-to-wall contact (2, 3). The energy input required to produce a pound of crystals via water evaporation (a function of the solubility) is typically around 2,000 Btu/lb for material with 33% solubility, such as NaCl.

Mechanical vapor compression can also be used to reduce energy requirements. In this case, water vapor is compressed by a high-efficiency compressor to a pressure at which it can be condensed in the heat exchanger that
supplies heat to the crystallizer. The Btu equivalent of the power required to compress 1 lb of water at atmospheric pressure, with a heat of vaporization of approximately 1,000 Btu, is around 84 Btu (4). Vapor compression can be used for compounds with a low boiling-point elevation, and when a large amount of heat is required to produce the evaporation for crystallization. Thermocompression is another option used to reduce the heat load.

**Forced-circulation crystallizer**

The Table below describes the characteristics of the most commonly used continuous evaporative crystallizers. The FC crystallizer is a MSMPR unit and has the lowest capital cost per pound of product (Figure 1). The feed typically enters the downpipe, leaving the bottom of the crystallizer body at a point following the recovery of product slurry through a conduit leading to a thickener or separating device, such as a hydroclone, centrifuge or filter. A combined stream consisting of fresh feed and recirculated slurry is pumped through the circulating pipe to a vertical or horizontal heat exchanger, where it is heated by condensing steam. This stream is introduced to the crystallizer below the liquid surface in the vapor body (where flashing occurs), mixes with the slurry near the point of feed entry, and raises the local temperature enough to cause flashing at the liquid surface. Subsequent cooling at the surface results in supersaturation that is relieved either as crystal growth or the birth of new nuclei in the active volume.

FC units operate with a high rate of circulation, which limits scaling, lending this unit to the evaporation of solutions with components that will scale, such as calcium sulfate (CaSO₄).

Some crystallizers can be designed for a classified product discharge, usually by suspending an elutriation leg beneath the crystallizer body. Introduction of the ML to the lower portion of the leg fluidizes the particles prior to discharge and selectively returns the finest particles to the crystallizer body for further growth. In a true FC crystallizer, the liquid has the same RT as the solids (i.e., there is no

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**Table. Characteristics of continuous evaporative crystallizers.**

<table>
<thead>
<tr>
<th>Crystallizer Type</th>
<th>Typical Product Size †</th>
<th>Examples of Products</th>
<th>Characteristic Solubility Curves ‡</th>
</tr>
</thead>
<tbody>
<tr>
<td>Forced-Circulation</td>
<td>105–500 µm U.S. mesh 35–140</td>
<td>LiCl, Li(OH)₂H₂O, NaCl, C₆H₈O₇H₂O (citric acid) Na₂SO₄, Na₂CO₃H₂O</td>
<td>Normal, flat or inverted</td>
</tr>
<tr>
<td>Draft-Tube Baffle</td>
<td>300–4,000 µm U.S. mesh 5–50</td>
<td>K₂SO₄, KCl, (NH₄)₂SO₄ H₂BO₃ (boric acid), CH₄N₂O (urea), MgSO₄.7H₂O, C₆H₁₀O₄ (adipic acid)</td>
<td>Normal, flat or steep</td>
</tr>
<tr>
<td>Fluidized Suspension</td>
<td>180–4,000 µm U.S. mesh 5–80</td>
<td>KCl, (NH₄)₂SO₄ MgSO₄.7H₂O, Na₂B₂O₅·H₂O, CuSO₄·5H₂O, C₆H₁₂O₆, C₇H₁₅NO₃S (saccharin)</td>
<td>Normal, flat or steep</td>
</tr>
</tbody>
</table>

† The ASTM E11 standard for particle measurement is compatible with the international ISO 565 scale. The U.S mesh values are a close, but not exact, match of the particle sizes indicated. This is because different mesh standards specify different wire diameters for the screen apertures.

‡ In normal solubility curves, solubility increases with temperature in a linear fashion. In steep solubility curves, the slope is higher, and may even reflect an logarithmic increase of solubility with temperature.
Crystallization

elutriation leg) and there is no deliberate attempt to control the RT of any crystal size. Thus, the product stream should have the same CSD as the body of the crystallizer. Occasionally, a FC crystallizer will achieve a classified product discharge by means of suspending an elutriation leg, such as in the manufacture of NaCl.

To obtain predictable results, one must pay special attention to the quantity and velocity of the recirculation stream, the size of the body, and the type and speed of the circulating pump. To this end, the FC unit uses a low-shear, low-head, and high-volume axial flow pump that turns over the active volume (which includes the body, piping and heat exchanger) about 1–4 times/min. Despite its low-shear design, the circulating pump can still be a source of crystal breakage and secondary nucleation, which contributes to the tendency of the FC crystallizer to normally produce a smaller CSD compared with the DTB and Oslo crystallizers.

The heat exchanger has a 1- or 2-pass configuration, often with 1.5-in., 12-gage tubes. It is essential to limit the differential temperature between the steam and slurry that passes through, in order to avoid boiling in the tubes, which can lead to scaling, plugging and nucleation. Moreover, there must be adequate liquid head at the point of entry to the crystallizer body (i.e., submergence of the slurry exiting the heat exchanger) to avoid flashing at the liquid-vapor surface.

Similarly, in an adiabatic evaporative-cooling FC unit, in which the heating element is omitted, the feed is admitted to the crystallizer body at a point sufficiently below the liquid-vapor surface to avoid flashing during the mixing process. In both configurations, supersaturation is minimized by maintaining high circulation rates, typically 7 ft/s, and by limiting the temperature drop at the liquid-vapor surface to 3–8°F, with some inorganic salt slurries being limited to a 1–5°F decrease.

Vortexing in the main body of the crystallizer can cause the supersaturated slurry to bypass the boiling liquid surface, resulting in higher levels of supersaturation throughout the body, variable nucleation rates and, consequently, cycling of the CSD between fine and coarse particles as the crystal surface area varies. A number of design features may be used to minimize this problem. These include:

• installation of a vortex breaker at the bottom of the cone
• a slurry inlet that is nearly tangential with the main body (the angle of installation will depend on the diameter of the crystallizer vessel)
• adequate submergence of the feed as it enters the crystallizer body to prevent vortexing and flashing
• conforming the slurry inlet velocity through the recirculation line to a maximum Froude number, defined as:

$$N_{Fr} = \frac{V}{\sqrt{gD}} \tag{1}$$

where $V$ is slurry velocity in ft/s; $D$ is the vessel diameter in ft; and $g$ is the gravitational constant equal to 32.2 ft/s$^2$. For example, a light organic slurry with a viscosity of 1–3 cP and a specific gravity of 1.2–1.3, would call for $N_{Fr} < 0.14$.

FC units typically range from $D = 2$–20 ft, with some units approaching $D = 40$ ft. They are especially useful for high evaporative loads. For example, a unit used for evaporating water at 380 mmHg can typically be designed to handle 250–300 lb/(h)(ft$^2$). The CSD is wide with a theoretical coefficient of variation (CV) of 50%. CV is a parameter, expressed as a percent, that measures the spread of a CSD. The higher the value, the broader the size distribution (Figure 2). CV is derived from a plot on arithmetic-probability paper of the particle diameter ($d$) on the x-axis and the cumulative undersized weight percent (wt.%) of particles on the y-axis. From the Figure 2, one obtains the particle sizes corresponding to $y = 16\%$ (i.e., $x = d_{16\%}$) and $y = 84\%$ (i.e., $x = d_{84\%}$) of the distribution, and dividing the difference in these particle sizes by two times the average particle size ($x$ value at $y = 50\%$, or $x = d_{50\%}$):

$$CV = \frac{100(d_{84\%} - d_{16\%})}{2d_{50\%}} \tag{2}$$

To obtain meaningful results using this method, the plot should be a straight line from $y = 10$ to 90% cumulative retained. Together, with the mean particle diameter, CV provides a satisfactory description of the CSD. Using Eq. 2 and Figure 2, $CV = (1,270 \mu m - 440 \mu m)/(2 \times 870 \mu m) = 48\%$.

Other than allowing one to adjust the RT or slurry density, the FC crystallizer affords little opportunity to change the CSD. Alteration of operating conditions, such as the mean RT of the MSMPR, has a minimal effect on the mean particle size and CV. Greater control of the CSD is possible by adjusting the RT of the crystals in a
certain size range, so that it differs from the RT of the liquid phase. This can be achieved in two ways — fines destruction and classified product removal. Fines destruction has been attempted, but it is not very efficient with the FC unit, due to turbulence in the vessel. Optimizing fluid distribution (i.e., eliminating vortexing in the crystallizer) to effect a change in the CSD can be done by allowing the feed to enter the body through a cone within the body, thereby altering the feed distribution and crystal nucleation patterns.

**Draft-tube baffle crystallizer**

Mechanical circulation has a significant impact on the level of nucleation in the crystallizer. For this reason, low-speed impellers in draft tubes are sometimes incorporated into the body to reduce the shear forces seen by the circulating pump. The draft tube efficiently suspends solids with lower agitator power input. One example of this technology is the DTB crystallizer (Figure 3). In this unit, a slow-moving (60–125 rev/min) impeller installed in a draft tube pumps the circulating slurry upward to the boiling surface, where the supersaturation created by concentration and cooling is relieved as crystal nucleation and growth.

The high flowrates generated by the agitator and mixed with the feed stream, limit surface cooling to 1–2°F during flashing. This, in turn, limits the degree of supersaturation. Cooled slurry is returned to the bottom of the vessel through the space between the outside of the draft tube and an annular baffle, and is recirculated through the impeller, where it mixes with the heated slurry returning from the heat exchanger, or with adiabatic feed.

The active volume of a DTB (including areas inside and outside the draft tube, and excluding areas behind the baffle), which typically contains a solids loading equal to 25–50% of the apparent settled volume (i.e., volume of slurry occupied by settled crystals and retained ML that is trapped in settled crystals) is turned over 1–4 times/min, thereby limiting nucleation and scaling on the vessel walls, and promoting crystal growth on the existing crystal surfaces. Crystal RTs are typically 4–6 h and crystal production rates are normally 60–85 kg/m³h. As with all continuous crystallizers, the period between washouts usually increases with increased crystal RT and slurry density as a result of decreasing supersaturation and resultant scaling.

The annular baffle area functions as a settling zone through which a stream of ML and fines are separated from the slurry in the active volume and removed via gravitational settling. The result is a RT for the fines that is less than the RT for the product. The baffle area can be segmented such that one or more settling sections are utilized (e.g., to separate fine crystalline material from coarse crystals). The CSD in the body of the crystallizer is controlled by adjusting the vertical velocity of the slurry in the baffle area and the maximum crystal size that will be removed and dissolved. This is accomplished either by changing the flowrate in the active baffle area or the amount of area selected for the baffles. The maximum size of the crystalline material (\( L_{F} \)) that is theoretically removed by the baffle, may be used to change the CSD of the product. \( L_{F} \) has a hindered settling velocity equal to the free vertical velocity (FVV). If \( L_{F} \) is too large, the increased supersaturation created by recrystallizing the dissolved fines will cause nucleation to occur homogeneously throughout the unit, resulting
in periodic upsets in the system and cycling of the CSD. Increasing the solids content of the slurry within the crystallizer body to a concentration above the natural make of the slurry (a result of equilibrium cooling of the incoming feed solution to its final temperature), is sometimes done by withdrawing a stream of ML from the baffle zone, which thickens the slurry in the body. High slurry densities tend to reduce the efficiency of baffle performance. To improve performance, one may add a lamella plate in the main crystallizer body to direct the flow vertically at the baffle entrance, or install alternating donut baffles in the settling zone, behind the regular baffles. These donuts dissipate large liquid eddies that can trap and carry out undesirable larger crystals.

A fines-destruction system, generally comprising the settling zone surrounding the crystallizer body, a circulating pump and a heat exchanger, may be employed to produce a larger CSD with a narrower size distribution. Ideally, the heating element should supply enough heat to meet the crystallizer’s evaporation requirements and raise the temperature of the solution removed from the settler enough to destroy unwanted fines. For applications where the only heat removed is that required for adiabatic cooling of the incoming feed solution, fines destruction is achieved by selectively removing a slurry stream containing fines and redissolving them via dilution prior to their return to the crystallizer, typically in the region below the impeller. This technique increases supersaturation in the crystallizer body, which enhances both crystal growth rates and nucleation rates, because the excess fines are destroyed by heat or dilution. Further improvements in the CSD are possible by utilizing an elutriation device. This technique should be coupled with a fines- destruction system to avoid upsetting the dynamics in the unit, which leads to cycling of the CSD of the product.

When destruction of fines is not desired, baffles are omitted, and the internal circulation rate is adjusted to minimize nucleation in the suspension (or active volume).

**Fluidized suspension/Oslo crystallizer**

In the fluidized suspension/Oslo crystallizer (Figure 4), supersaturation can be achieved via evaporative cooling (i.e., adiabatically), as shown in Figure 4, or by the evaporation and cooling of a heated stream exiting a heat exchanger within the recirculation loop, and then relieved by passing the supersaturated liquor through a fluidized bed of crystals. Like the FC crystallizer and DTB units, the Oslo crystallizer requires the feed inlet to be submerged enough in the suspension to prevent flashing at the feed point, as well as across the heat exchanger.

For the adiabatic system, hot concentrated feed is mixed with a circulating stream of ML, and the mixed liquor is pumped via an axial-flow pump to the vaporizer (or flash chamber), where solvent is evaporated and the solution is cooled to create a supersaturated solution. The decrease in temperature due to flashing is typically less than 6°F for mixtures with normal solubilities, and less than 4°F for those with inverted solubility curves. Measured at the boiling surface, the increase in solute concentration in the circulating liquid due to evaporation and cooling is about 2–3 g/L of solvent.

If further evaporation is required to produce the driving force for crystallization, a heat exchanger can be installed before the vaporization chamber to supply adequate heat for the required rate of vaporization. Typically, the supersaturated liquid leaves the flash chamber and flows through the downcomer at a velocity of approximately 3 ft/s into the suspension chamber, where it fluidizes the crystal bed, and relieves its supersaturation on the growing crystals. Coarse product crystals are continuously removed by pumping slurry from the lower part of the chamber. An elutriation leg can be employed to achieve a narrower cut.

The suspension chamber has a semi-hemispherical head and a tapered body that progressively increases in cross-sectional area (CSA) from the bottom to the top of the chamber. This shape serves to classify both the CSD and percent solids in the bed based on the particles’ hindered settling velocities versus the circulating fluid velocity.

Crystals separate from the ML in the upper portion of the chamber, so that the ML overflowing the baffle and entering the circulation loop to the vaporizer has a low quantity of fines. The baffle can be used to adjust the percent solids in this loop. An adjustable pipe may be installed to remove the fines and destroy them by dissolution or heat input. Care must be taken to ensure that the crystallizer is designed and operated at ML velocities that both fluidize and classify the bed. During classification, the larger particles settle out, causing the slurry crystal size and percent solids to decrease. Meanwhile, the liquid velocity decreases, due to increasing cross sectional area in the vertical direction; the lower velocities aid the classification process. Improper flowrates and poor overall design can lift the crystal bed and cause the unit to function as a FC crystallizer. The active volume for the crystallizer must be at least 3 m deep in order for the larger crystals to settle out from the ML.

As with the other evaporative configurations, Oslo crystallizers are designed to operate within the metastable zone for the desired level of supersaturation. (In the metastable zone, the supersaturation range enables growth and secondary nucleation in the absence of primary nucleation.) Many applications of Oslo crystallizers have concentration changes of 1–3 g/L (5). The presence of small quantities of crystals in the recirculation line entering the vapor body can cause nucleation and scaling, presenting an onstream cycling problem and a broadening of the CSD.

The fluidized-bed configuration, like the DTB, enables the production of a larger CSD with a narrow CV. The CSD is dictated by the RT in the suspension chamber and the flow rate of fines to dilution or heating, for destruction purposes. The RT is determined by the crystal production
Given forces depend on the settling velocity of the crystals. For low crystal production rates, large CSA is required, which means large crystallizers and particles is reduced. Crystals with slow settling velocities require large CSAs, which means large crystallizers and slow crystal production rates.

The Oslo crystallizer is best suited for use with compounds that have high settling rates — i.e., greater than 20–40 mm/s. If the crystals have high settling velocities, the larger particles will settle out more quickly. Thus, the CSA required to induce settling/separation of the larger particles will settle out more quickly. Thus, the CSA required to induce settling/separation of the larger particles is reduced. Crystals with slow settling velocities require large CSAs, which mean large crystallizers and low crystal production rates.

The suggested productivities for concentration driving forces depend on the settling velocity of the crystals. For a given ΔC, the higher the the settling velocity, the higher the allowable crystallizer productivity. For example, for a DC of 2 g/L, the recommended productivity increases from 125 to 250 kg/h/m² as the settling rate increases from 20–30 mm/s for any particle size (5).

### Challenging phenomena

Crystallizers are complex units that encounter a number of process challenges:

- **Entrainment** — The entrainment of significant quantities of liquid droplets in the vapor body above the boiling surface of a crystallizer causes severe scaling and plugging problems for the areas and equipment above the boiling surface. There are several rules of thumb that can be heeded to reduce entrainment. One design criterion is the entrainment factor:

  \[ F = v \rho_v^{0.5} \]  

  (3)

where \( v \) is the vapor velocity in ft/s and \( \rho_v \) is the vapor density in lb/ft³. Values of \( F < 0.3 \) are usually indicative of a conservative design. The installation of a demister with spray nozzles at the top of a vapor body reduces contamination and plugging of condensers and vacuum equipment.

- **Operation in the metastable zone** — For ideal crystallization, systems must be operated within the metastable zone. To this end, it is essential to have adequate crystal surface area available at the point of supersaturation, at the boiling surface for the FC and DTB units, and at the bottom portion of the fluidized bed for the Oslo crystallizer. Failure to meet these design criteria will result in the creation of fines, scaling/plugging and reduced onstream time.

- **Impurities** — In all crystallization systems, impurities have a dramatic influence on kinetics and the resultant CSD and crystal morphology. For an optimal design, laboratory and pilot investigations should be performed with a feed composition that is representative of the one that will be used in the plant.

Continuous crystallizers may experience product degradation if impurities are brought into the system. For example, the introduction of filter aid into the feed may decrease the CSD, due to heterogeneous seeding with the small filter-aid particles. Inadequate purging of the ML from the active system may result in a slow or sudden buildup of impurities within the crystallization loop, which can reduce the product purity, while modifying the CSD and crystal morphology. Product purity can be increased by washing or re-slurrying the crystals in fresh solvent.

### Instrumentation and control

To achieve consistent product characteristics, fluctuations in crystallizer operation must be kept to a minimum. To this end, various instrumentation systems may be employed.

- **Liquid level control** — Most vacuum crystallizers operate at a constant liquid level. This parameter should not vary by more than 6 in. With vertical inlets, such as those featured in a DTB, the distance between the inlet and the liquid surface controls the exposure of the solids to the highest levels of supersaturation, which occur at the boiling surface. The distance between the inlet and the liquid surface also affects the head on the circulation system. Since the shut-off head of the internal agitators is relatively low, small changes in the liquid level can adversely effect crystallizer performance the operation due to circulation. Commonly employed liquid-level-control technologies include purged differential-pressure cells (with one cell located in the vapor space and the other in the liquid, at a predetermined distance from the former), continuous nuclear level detectors (in which a narrow ray of radiation is directed...
through the vessel to a detector. As the process level rises, it shields the detector from the radiation. The more radiation the detector “sees,” the lower the process level, and vice versa.

**Absolute pressure control** — The temperature in a vacuum crystallizer should be maintained within 0.5°C of the setpoint. Typically, this is done via an absolute differential-pressure cell mounted on top of the vessel, so that purge drainage is directed back into the vapor space and the control signal is transmitted to a remote recorder. The controller receives its signal from an absolute-pressure recorder-controller, which purges air or nitrogen into the vacuum system, or bleeds off vent gas from a condenser if the vessel is operated above atmospheric pressure.

**Slurry density** — The density of the slurry in the crystallizer body should be kept constant. Control of this parameter is achieved by measuring slurry density to determine the rate at which it is changing, and adjusting the crystallizer discharge rate or ML recycle rate to maintain a constant slurry density in the crystallizer body.

Slurry density can be measured using two purged, submerged differential-pressure cells that are installed a known distance apart, or via a nuclear density meter, or by measuring the amperage draw of a circulating pump or agitator that has been calibrated to percent solids.

**Steam flowrate recorder-controller** — The heat flux in an evaporative crystallizer defines the maximum level of supersaturation. Adequate heat flux can be maintained by controlling either pressure or steam flowrate. Experience indicates that controlling steam flowrate yields better results. For most evaporative units, the steam flowrate is directly proportional to the crystal production rate, the temperature rise in the recirculation loop and the log-mean temperature difference (\( \Delta T_{\text{log}} \)) in the heat exchanger. Both \( \Delta T \) values have an important effect on the scaling rate, especially where materials have inverted solubility curves, or where there are scale-forming components. In systems where heat flux is critical, the \( \Delta T_{\text{log}} \) can be monitored and used to reset the steam flowrate.

**Feed or product flowrate** — This parameter is often measured by a magnetic flowmeter or a micromotion mass flowmeter. A cascade control loop is usually employed to measure the concentration and flowrate of the feed, while modulating the setpoint of the steam flowrate. One can choose to operate this loop in either a “push” or “pull” mode. In the “push” mode, the feed flowrate is set and the discharge rate is varied to achieve the proper liquid level and slurry density in the crystallizer, while in the “pull” mode the discharge flowrate is set and the feed flowrate is varied to achieve the proper liquid level and slurry density in the crystallizer.

**Cost considerations**

It is difficult to provide an accurate estimate of crystallizer cost without consideration of extensive design specifications, such as crystallizer type, capacity, materials of construction, auxiliary equipment and instrumentation. In general, the classified slurry crystallizer is the most expensive system, followed by the DTB, DT, FC and tank crystallizers, respectively. The installed cost of these five units can range from 4:1 to 3:1 when compared with tank crystallizers, depending on capacity.

Based on 2004 prices, DTB and growth (DT) crystallizers producing 25 ton/d of MgSO₄·7H₂O, respectively cost an estimated $470,000 and $625,000. A 20-ton/d FC unit used to crystallize LiCl costs approximately $800,000.

Now, let us consider larger units. A DTB producing 100 ton/d of (NH₄)₂SO₄ costs roughly $900,000, while a Na₂SO₄ triple-effect FC train (i.e., where three units operate in series to maximize steam efficiency) operating at the same capacity costs $1,600,000.

The cost of process equipment tends to vary non-linearly with changes in capacity or size according to:

\[
C/C_b = (Q/Q_a)^n
\]

where \( C \) is the purchase cost of the equipment (including delivery, installation, etc.), \( Q \) is equipment capacity or size, \( a \) and \( b \) represent differently sized units, and \( n \) is the cost scaling exponent. The value of \( n \) can vary from 0.2 to greater than 1, depending on equipment complexity. A value of \( n = 0.6–0.65 \) typically provides an accurate estimate of crystallizer costs at different system capacities.

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**Literature Cited**


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