A MATHEMATICAL MODEL OF VACUUM PAN CRYSTALLISATION

P. G. Wright
Sugar Research Institute, Mackay, Queensland.

and E. T. White
Chemical Engineering Dept, University of Queensland, Brisbane

ABSTRACT

This paper outlines the results of a systematic study of the vacuum pan crystallisation process, directed towards the formulation of a mathematical model which can be used to predict pan performance over a range of conditions.

The paper deals with the selection of the model equations, and the setting up of a model which would predict the mass and size distribution of product crystals from data on the initial conditions and the progressive feed quantities and operating conditions. The model describes the crystal size distribution by its moments, and incorporates a size dispersion function to predict accurately the spread of sizes resulting from crystallisation. The important parameters of the model were evaluated from data logged from an experimental pan at a Queensland raw sugar factory. A brief outline of the uses of such a model in simulation of the batch and continuous boiling of sugar strikes is given.

INTRODUCTION

Over the last 15 years in the chemical industries there has been a gradual trend towards a more quantitative approach to problems in the design and operation of processes. The trend has been made possible by the increasing availability of powerful electronic computers for the solution of complex systems of mathematical equations. This computerised analytical approach offers a wider scope than the older methods in investigating alternative designs and methods of operation of commercial batch and continuous processes.

In the raw sugar factory there are a number of processes to which the new approach may be applied to advantage. Some of the processes such as the clarification operation may not be so amenable due to the difficulty in prediction of the basic rate steps. The crystallisation step in the vacuum pans is however, one operation whose complexity has in the past discouraged attempts at quantitative analysis, but which now can yield to the computerised analysis procedures. An analysis which leads to a mathematical model (a set of equations which can be used to predict pan behaviour) is considered here.

This system study, which is a science of formulating the equations describing the system, is necessarily a specialist task, particularly if the necessary data is lacking and thus further experimentation is required. Once a model is obtained, however, it can be applied by others to a wide range of problem applications to give useful predictions of performance. As further comparisons of the real process with the predictions are made, the original model can be modified and improved to increase the degree of confidence in its use.
This paper deals with the derivation of the model equations and the evaluation of the important unknown parameters in the model when applied to the raw sugar pan process. Other papers will develop some of the applications of the model in two areas which are now receiving close attention i.e. computer control of pans and continuous systems of pan boiling.

The Vacuum Pan Process

Much basic work on crystallisation systems has been restricted to idealised laboratory experimentation, which is often quite unrelated to the performance of the industrial crystallisation process. However, following the work of Bransom,2 Saeman,3 and others, an upsurge of interest in the prediction of the crystal size distribution, production rate and recovery from industrial plants has taken place. The earlier treatments covered the mathematics of crystal size prediction from simple continuous stirred tank crystallisers. However, more complex situations are now being considered.

The aim of vacuum pan operation is to produce a mass of slurry containing crystals with an acceptable size and size distribution and crystal content, given an initial base footing of liquor and seed crystals, and a given quantity of feed syrup of a given purity.

The control and operating conditions now used are the end result of a considerable body of experience in pan design and operation rather than that of the direct application of knowledge of the process kinetics. As more sophisticated control hardware becomes available, however, there will be a greater need for an analytical approach to the fundamental processes in crystallisation.

In the sugar crystallisation process we are concerned with production rate, yield and crystal size distribution. In addition to the consideration of the basic principles usual in chemical engineering, i.e. material balances, energy balances, equilibrium data and rate mechanisms, we require for crystallisation two extra types of relationships. These are the size-property relationships for the crystals, and the crystal population balances. If the latter can be formulated then it is possible to model the crystalliser for the purposes of design, control or optimisation of the process.

Some attempts have been made at modelling the pan process (Doucet and Gidde, Withers and Bass, Evans, Treachis and Jones), but these have been simplified models aimed at the control requirements of the strike. A preliminary model in greater depth has been presented by the authors (Wright and White). In this work the need for the model relationships to be investigated in more detail was discussed, and, in particular, an investigation of the factors governing crystal size distribution of the product was thought desirable. This latter involves a study of the aspects of size dispersion in growth, and of size dependent growth.

MODEL RELATIONSHIPS

In the derivation of a satisfactory model, the behaviour of the batch crystalliser has first to be expressed in terms of the basic relationships or laws governing the various subsystems of the process. In the case of the sugar vacuum pan these can be tabulated as follows:—
1) Material Balances
   a) Water Balance
   b) Sucrose Balance
   c) Impurity Balance
   d) Total Material Balance
2) Energy Balance
3) Equilibrium Phase Relationships
   a) Sucrose Solubility Relationships
   b) Elevation of Boiling Point
4) Kinetic Expressions
   a) Nucleation Kinetics — Heterogeneous and Homogeneous
   b) Crystal Growth Rate
   c) Crystal Dissolution Rate
   d) Crystal Size Distribution Changes
5) Crystal Inventory and Crystal Material Balances
   a) Distribution Moments related to number, area, volume and mass of crystals
   b) Crystal Population Balance
6) Heat Transfer Relationships
   a) Temperature Variations in the mass of the pan
7) Fluid Flow Relationships
   a) Concentration variations

The above are the basic relationships which could be of importance in sugar pan crystallisation studies. Certain of these relationships (such as the material and energy balances) are generally true, i.e. they are not affected by the application of the model to syrups of varying purity, seed crystals of different mean size, etc. Other relationships, including growth, nucleation, solubility and size dispersion, are quite dependent on these conditions.

A list of symbols is appended. The Model here developed is specific to the following conditions:
   a) A well mixed batch pan crystalliser is assumed;
   b) A controlled temperature situation is assumed, rather than a system where the temperature is a function of heat transfer, circulation and absolute pressure variables;
   c) Feed syrup composition and concentration is assumed constant, except for one or more step changes;
   d) A condition of no nucleation is assumed, i.e. the concentration of the massecuite is kept below the predicted "critical" value for nucleation;
   e) The model operates on the crystal size distribution by predicting changes in its moments, and assumes that the size distribution can be obtained from the first three moments. This is quite valid for the normal distribution. A definition and a discussion on the moments of the size distribution are given in other papers by (Wright);
   f) No inflow or outflow of massecuite during the batch is allowed.

It is relatively easy to extend the basic model to allow inflow or exit of massecuite and hence a model of a continuously operating unit but, for simplicity, this will not be covered here.
Material Balances

The mass balance expressions (equations 2 to 5) are easily derived for the batch process where the material inflows arise only from syrup, water or molasses feed, and outflows only from the evaporation of water. The total sucrose in the pan is divided into two components, i.e. the soluble sucrose $S$ and the crystalline sucrose $C$.

\[
\frac{dC}{dt} = \rho_c \frac{d\mu_3}{dt} \quad (1)
\]

\[
\frac{dS}{dt} = F \cdot C_t \cdot P_t = \frac{dC}{dt} \quad (2)
\]

\[
\frac{dI}{dt} = F \cdot C_t \cdot (1 - P_t) \quad (3)
\]

\[
\frac{dW}{dt} = W_t + F \cdot (1 - C_t) - E \quad (4)
\]

\[
\frac{dM}{dt} = \frac{dW}{dt} + \frac{dI}{dt} + \frac{dS}{dt} + \frac{dC}{dt} \quad (5)
\]

The mass of crystal $C$ is derived in equation 1 from the number ($N$) and density ($\rho$) of the crystals and the third moment, $\mu_3$, of the crystal size distribution. The third moment is related to the average volume of a crystal as the measure of size used in the moments is the “volume equivalent diameter” size as found by the method of direct crystal weighing proposed in an earlier paper (Wright and White).

The moments are designated as $\mu_j$ where

\[
\mu_j = \int f(1)^j \, dl \quad \text{for } j = 0, 1, 2, 3 \text{ etc.}
\]

and have direct physical significance. As $N$ is the total number of crystals we have that:

- $N \mu_0 =$ the total number of crystals;
- $N \mu_1 =$ total size of crystals;
- $N \mu_2 =$ total surface area of crystals;
- $\pi / 6 \cdot N \mu_3 =$ total volume of crystals.

The expressions for the rate of change of moments are listed below. The parameter $p$ relates to the size spreading effect during growth and is discussed later.
Moment Expressions

$$\frac{dN}{dt} = 0 = \frac{d\mu^3}{dt}$$  \hspace{1cm} (6)

$$\frac{d\mu_1}{dt} = \frac{d\bar{I}}{dt}$$  \hspace{1cm} (7)

$$\frac{d\mu_2}{dt} = 2\mu_1 \frac{d\bar{I}}{dt} + p\frac{d\bar{I}}{dt}$$  \hspace{1cm} (8)

$$\frac{d\mu_3}{dt} = 3\mu_2 \frac{d\bar{I}}{dt} + p\mu_1 \frac{d\bar{I}}{dt}$$  \hspace{1cm} (9)

Energy Balance

An energy balance to relate steam consumption to feed quantities and temperatures and evaporation rate can be drawn up, but, for a model mainly aimed at prediction of crystal growth it is unnecessary, and is not included here.

Equilibrium Phase Relationships

a) Sucrose Solubility Relationships

The pure sucrose solubility (equation (10)) was determined from the expression of Charles,9 and an expression for the effect of impurity concentration on the solubility coefficient of the solution (equation (12)) was derived from the work of Kelly10 on molasses with a reasonably high reducing sugar to ash ratio. The calculation of the supersaturation in impure solutions (SS) is given in equation (11). It is based on the ICUMSA definition of supersaturation for solutions crystallizing at constant Impurity to Water (I/W) ratio.

The solubility coefficient (equation (12)) assumes no influence of temperature on the solubility coefficient. This was established for the actual material used, but would have to be re-examined for each material for which the model is required.

Solubility Data

$$\text{SAT} = 64,407 + 0.07251T + 0.002057T^2 - 9.035 \times 10^{-6}T^3$$  \hspace{1cm} (10)

$$\text{SS} = \frac{(100 - \text{SAT}) \times S \times 1}{\text{SAT} \times W \times SC}$$  \hspace{1cm} (11)

$$\text{SC} = 1.0 - P_4 \times \frac{I}{W}$$  \hspace{1cm} (12)

$$\text{OS} = \text{SS} - 1.0$$  \hspace{1cm} (13)

b) Elevation of Boiling Point

As for the energy balances the equilibrium relationships for elevation of boiling point in sugar massecuites are not strictly required for many simulations
i.e. those assuming that the vacuum is manipulated to give a designated massecuite temperature. Recently new correlations of boiling point rise for cane syrups have been published (Batterham, Frew and Wright), together with a correlation for the superheat effect above the equilibrium boiling point for a large raw sugar pan. The superheat effect is essential for consideration when predicting massecuite temperature from the concentration and pan absolute pressure. The details of the relationships will not be given here.

**Kinetic Expressions**

a) **Nucleation kinetics**

The treatment of nucleation kinetics here supposes that the rate of formation of new nuclei in the pan contents in the presence of large (> 0,05 mm) crystals is very low at supersaturation levels less than a fairly reproducible critical value (SCRIT). This supposition has been challenged by certain workers e.g. Nicol, but in our experience in raw sugar work does hold to all practical purposes. Thus we first require an expression to predict the critical values.

The critical supersaturation expression first chosen was derived from work done at Sugar Research Institute some years ago (Penkils and Wright, Hesp and Dunton). A new expression is now available (Broadfoot and Wright) which lies close to this at 60 C but does not have a temperature term. This change does effect optimal control considerations markedly. It is anticipated that more work will soon be forthcoming to assist a decision as to the best form of this expression. The Broadfoot and Wright expression is given in equation 19. These authors also give an indication of the rate of nucleation per unit volume of solution for supersaturation values in excess of the critical value. However this was derived only for pure sucrose solutions. It might be sufficient at this stage to accept that, as expressed in equations 17 and 18, the nucleation rate per ml (B_n) is zero for supersaturation values below the critical and some very high value above it. It is possible to operate with the present model only if the critical supersaturation is not exceeded or, if once exceeded, the syrup is brought quickly into a state of undersaturation sufficient to remove all the fine crystals so formed. If a known amount of nuclei do appear it is possible that these can be handled by considering that a burst of nuclei can be followed as a distribution separate from that of the original seeds and its moments evaluated with time. The total weight of crystals can be obtained by summing the third moment of each distribution before applying the weight-size factor.

Nucleation Expression

\[
\frac{dN}{dt} = B_n = 0 \text{ for } SS < \text{SCRIT} \\
= 10^{4.6} \text{ for } SS > \text{SCRIT}
\]

\[
\text{SCRIT} = 1,1289 - 0,284 \frac{(I)}{(S+I)} + (2,333 - 0,0709 (T - 60)) \times \frac{(I)^2}{(S+I)}
\]
<table>
<thead>
<tr>
<th>MANUFACTURING — PROCESSING</th>
</tr>
</thead>
</table>

\[
\begin{align*}
\left( \frac{1}{1.09} \times \left( \frac{1}{1.09} + \frac{1}{1.09} \right) \right) X & \text{ as above, cf. for condition} \\
\frac{1}{1.09} - \frac{1}{1.09} - \frac{1}{1.09} & = \frac{1}{1.09} - \frac{1}{1.09} \\
\text{Overall effect} & \geq 1.09 \% \\
\text{Arrows indicate relative crystal size} & \geq 0 \\
\text{Where } p & = \text{ fraction} \\
\text{Crysal size} & \geq 0 \\
\text{Weil-Gobert industrial pant} & \geq 0 \\
\text{Aspiration conditions} & \geq 0 \\
\text{Impurity (l/l W ratio)} & \geq 0 \\
\text{Temperature (l/°C)} & \geq 0 \\
\end{align*}
\]

<table>
<thead>
<tr>
<th>Expression</th>
<th>Range</th>
<th>Effect</th>
</tr>
</thead>
<tbody>
<tr>
<td>G = \frac{1}{1.09} \times \left( \frac{1}{1.09} + \frac{1}{1.09} \right)</td>
<td>\geq 0</td>
<td>\text{as above, cf. for condition}</td>
</tr>
<tr>
<td>G = \frac{1}{1.09} - \frac{1}{1.09} - \frac{1}{1.09}</td>
<td>\geq 0</td>
<td>\text{Overall effect}</td>
</tr>
</tbody>
</table>

| TABLE I (crowd and dissolution expression for the industrial crystallization model) |
b) Crystal Growth and Dissolution Rate Expression

The growth-dissolution expression chosen, which is detailed in Table 2, is plotted in Fig. 1 to illustrate its form. This is a simplification of the growth expression derived by Silin\textsuperscript{16} and which was favourably reviewed by Smythe.\textsuperscript{17} The dissolution process is assumed to be five times as fast as is the crystallisation process at the same deviation from the saturated condition, following the work of Mantovanii\textsuperscript{18} and others. More recently Nicol and Parker\textsuperscript{19} have shown that this ratio might be temperature dependent.

\begin{figure}[h]
\centering
\includegraphics[width=\textwidth]{fig1.png}
\caption{The effect of oversaturation on the growth of sugar crystals.}
\end{figure}

In the growth rate expressions the activation energy selected derives from laboratory tests on cane sugar massecuites. This value does control the variation of growth rate with temperature. As there is a great deal of variability of estimates of activation energy in the literature, care is needed when operation at temperatures much different from those values (about 60 to 70°C) at which the growth parameters were evaluated.

Van Hook\textsuperscript{20} had criticised an earlier simplification by the authors that the activation energy for growth decreased linearly with increasing temperature. He points out that a linear approximation would only hold over a small temperature range say 45 to 75°C, as the activation energy—temperature slope tends to be relatively flatter at lower temperatures and again is small at temperatures above 80°C. The value for activation energy at constant supersaturation for pure sucrose is given by him as 46 kJ at 60°C. A value for cane syrups of I/W rate of about 0.5 was found by the authors to be 62.8 kJ at 60°C. Equation (16) is an approximate estimate of the effect of impurity and temperature on the activation energy based on these considerations.
Growth Rate Expression

As for Table 2, e.g. for OS > 1.5 \times P_0

\[ G = \frac{dl}{dt} = P_1 (OS - P_0) \exp \left( \frac{P_3 I}{W} \right) \]

\[ \exp \left( \frac{-E_{\text{act}}}{R \times 2303} \times \left( \frac{1}{273.16 + T} \right) - \frac{1}{333.16} \right) \quad (14) \]

where

\[ R = 8.9872 \times 10^{-4} \]

\[ E_{\text{act}} = 11.0 \times 8.0 \times I/W - 0.84 \times (T - 60) \text{ kJ} \]

for T of 60 \pm 15 \text{ C} \quad (15)

The parameters \( P_0, P_1, \) and \( P_3 \) have to be determined for the particular material type and pan being considered. These have been obtained for factory conditions by methods described later.

Crystal Size Distribution Changes

The use of the moments of the crystal size distribution has been made in equations 6-9. In essence the distribution is described by the first two moments, \( \mu_1 \) and \( \mu_2 \), and the third and higher moments derived from this, with the assumption that the distribution is approximately normal (Gaussian) in shape. This assumption of normal shape is made following experimental evidence, and greatly simplifies the description of size distribution in the model. Absence of nucleation is necessary to this approach.

The phenomenon of size dispersion, or the tendency of a crystal distribution to spread in size during growth, was explored rather fully by Wright. This is a concept widely ignored in the past, but it is now obvious that it must be considered if accurate prediction of crystal size distribution is required. This is so in the investigation of size distribution from continuous crystallisation schemes. The expression for the size dispersion parameter here is an empirical relationship with variables including the temperature, purity and growth rate associated with the operating factory conditions, as well as the variance and mean size of the crystals. The amount of increase of size spread (variance) with each increment of growth is then proportional to the size dispersion parameter, \( p \), which is defined as the ratio of the increase in variance to the increase in mean size of the crystals size distribution after a small increment of growth. The value of \( p \) is influenced by the two main mechanisms of dispersion or spreading of sizes. One possible mechanism implies that the larger crystals grow faster than small ones in the same distribution due to diffusion and settling effects. This results in a spread in which \( p \) is itself proportional to the existing ratio of variance to mean of the crystals.

Another, which implies that crystals grow in random spurts due to surface effects which cause increase in size spread, results in a \( p \) value independent of the existing spread. Evidence is that both mechanisms are important and the estimation of \( p \) is thus quite complex. The form of equations given in equation 20 and 21 for the prediction of \( p \) is a model drawn from the first mechanism,
but which should apply reasonably well to raw and refined sugar boiling where
the normal size spread of crystals exist i.e. above 0.10 coefficient of variation on
a number basis. It would not apply for narrowly sized seed crystals.

Size Dispersion parameter \( p \) (microns)

\[
p = 2 \times n^* \cdot \left( \frac{\mu_2 - \mu_1^b}{\mu_1} \right)
\]

(20)

where

\[
n^* = 8,105 - 2,203 \log_{10} \left( \frac{d_l}{n} \right) - 0,369 \frac{I}{T} - 0,0396 T
\]

(21)

In the model most of the parameters have been fitted from data in the
literature, leaving four for further determination on the factory pan. These are:

a) the proportionality constant in the linear growth expression \((P_1)\);

b) the null oversaturation value above which the linear growth rate-
oversaturation relationship appears \((P_2)\);

c) the parameter determining the retardation effect of impurities on the
growth rate \((P_3)\);

d) the parameter in the solubility coefficient expression \((P_4)\).

DETERMINATION OF MODEL PARAMETERS

As the prime purpose of the model is the simulation of the industrial
crystallisation process it was considered highly desirable that some of the im-
portant parameters be evaluated directly from industrial process data. Such
process data are, normally, rather difficult to obtain directly from factory
units which are tied to a tight production schedule. In addition, the measure-
ment of variables and incremental feed rates would have to be done at frequent
intervals in view of the rapid variations associated with a loosely controlled
batch process. This would involve a formidable problem in data collection.
However, the solution to these problems was available. In the first instance an
experimental pan installation at Marian factory was available for “off produc-
tion” experimental boilings, and secondly the instrumentation of the pan could
be directly connected to a mobile computer for logging of data signals directly
onto coded paper tape. Thus data from various types of raw sugar strikes could
be collected, and used to evaluate some of the model parameters. At this stage
only those parameters associated with the growth model have been considered
in this way, but possibly nucleation parameters could be obtained from the
factory by this method also.

The data from the factory pan tests was reduced to sets of:
(i) initial condition data;
(ii) feed condition data and;
(iii) analytic data on the condition of the massecuite at several times
during the run.

This was all stored on a disc file. The mathematical model was then used
for the estimation of the massecuite condition at any time from the initial
condition and feed data, using a given set of parameters in the model. The
calculated condition was compared with the real value (from sample analysis) at all sample points and the squares of the differences between these were summed. The parameters were then varied in a systematic manner by a pattern search algorithm so that the set that gave the least square of the condition deviations was obtained.

**Results of Parameter Estimations**

The parameters which minimise the error between calculated and actually measured supersaturation values in the test runs are tabulated below. Figure 2 shows how the model prediction using these parameters fitted the experimental points for some strikes.

**FIGURE 2.** Model prediction and data points for several test runs.

An estimate of the accuracy of the parameter estimates is also given. The parameter estimates would apply only to the conditions of operation applicable during the data collection runs and care should be taken in wider application. The result can be summarised as follows:

**Conditions**

- **Material:** Raw Sugar, 85, 75 and 65 purity strikes;
- **Crystalliser:** Natural circulation calandria pan;
- **Operating temperature:** 60 to 70 °C;
- **Impurity:** Cane syrups, reducing sugar to ash ratio in impurities of approximately 1:3:1;

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>$P_1$</td>
<td>7420 micron/(hr °SS) ±250</td>
</tr>
<tr>
<td>$P_2$</td>
<td>0.005 °SS ±.006</td>
</tr>
<tr>
<td>$P_3$</td>
<td>1.75 ±.03</td>
</tr>
<tr>
<td>$P_4$</td>
<td>0.088 ±.005</td>
</tr>
</tbody>
</table>

**USES OF THE MODEL**

The uses of the model are manifold but some applications that have already been made are as follows:

1) the examination of the effect of changing initial and control conditions on the batch times for simulated strikes, so that the best control procedures can be found. For example Figures 3 and 4 show how the batch strike time would be affected by choice of initial seed size or of the operating level of oversaturation used;
it has been possible to find an optimal policy or strategy for a given well defined batch process from the use of the mode, and knowledge of the physical constraints which operate in the given particular process. In particular Rees, Batterham and Frew have shown that the optimal policy for a batch lies often along the constraint boundaries, and can be rationalised along some simple principles. These can be stated:

a) as long as the mother liquor purity is above about 50 per cent, the optimum operating policy regarding massecuite temperature is to use the maximum allowable value (here suggested at 70°C for cane syrups);

**FIGURE 3.** The effect of seed crystal size on the batch time for simulated raw sugar strikes (constant seed mass).

**FIGURE 4.** The effect of the oversaturation control point on the batch time for simulated raw sugar strikes.
\[ W \quad \text{mass of water in the pan} \]
\[ M \quad \text{total mass in the pan} \]
\[ t \quad \text{time} \]
\[ \rho \quad \text{specific gravity of sucrose crystals} \]
\[ l \quad \text{volume mean diameter of the crystal (diameter of a sphere of the same volume as a crystal)} \]
\[ N \quad \text{number of crystals in the pan} \]
\[ \mu_0 \quad \text{zero'th moment of the crystal size distribution (CSD)} \]
\[ \mu_1 \quad \text{first moment of the CSD} \]
\[ \mu_2 \quad \text{second moment of the CSD} \]
\[ \mu_3 \quad \text{third moment of the CSD} \]
\[ F \quad \text{feed rate of syrups to the pan} \]
\[ E \quad \text{evaporation rate of water from the pan} \]
\[ \text{W}_f \quad \text{feed rate of water direct to the pan} \]
\[ \text{C}_f \quad \text{concentration of feed syrups (fraction)} \]
\[ P_r \quad \text{purity of feed syrups (fraction)} \]
\[ p \quad \text{size dispersion parameter} \]
\[ \text{SAT} \quad \text{solubility of pure sucrose (by weight)} \]
\[ T \quad \text{temperature} \]
\[ \text{SS} \quad \text{supersaturation coefficient of the solution} \]
\[ \text{OS} \quad \text{oversaturation coefficient of the solution} \]
\[ \text{SC} \quad \text{solubility coefficient of the solution} \]
\[ G \quad \text{crystal growth rate, size basis} \]
\[ E_{\text{act}} \quad \text{energy of activation of the crystallization reaction} \]
\[ R \quad \text{gas constant} \]
\[ B \quad \text{rate of nucleation per unit (ml) of solution} \]
\[ \text{SCRIT} \quad \text{critical value of supersaturation for breeding nucleation} \]
\[ \mu_1 - \mu_2 \quad \text{parameters in the growth and solubility expressions} \]
\[ n^* \quad \text{a size dispersion term from which 'p' the size dispersion parameter is derived} \]
\[ f(t) \quad \text{frequency density function of crystal size} \]

REFERENCES


MODELO MATEMÁTICO PARA LA CRISTALIZACIÓN EN TACHOS

P. G. Wright, y E. T. White

RESUMEN

Este trabajo resume los resultados de un estudio sistemático del proceso de cristalización en tachos, dirigido hacia la formulación de un modelo matemático que puede ser usado para predecir la operación del tacho sobre un rango de condiciones.

Este trabajo discute la selección de las ecuaciones modelo y el diseño de un modelo capaz de predecir la distribución de masa y de tamaño de los cristales del producto partiendo de datos sobre las condiciones iniciales, las cantidades crecientes de alimentación y las condiciones de operación. El modelo describe la distribución del tamaño de los cristales por sus momentos, e incorpora una función de dispersión de tamaño para predecir con exactitud el rango de tamaños resultantes de la cristalización. Los parámetros más importantes del modelo fueron evaluados utilizando información experimental de un tacho que trabaja en un ingenio de Queensland. Se hace un esbozo del uso del modelo ya mencionado en la simulación de templas y cristalización continua de azúcar.