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Some Fundamental Considerations of Crystallization and their Implications for the Control of Sugar Boiling.

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Abstract.

Some of the factors influencing crystallization are reviewed and their implications for the conduct and control of vacuum pan crystallization are discussed. This paper considers the requirements of achieving sugar quality at minimum cost. The effect of pan design, operation and control on crystal quality and yield is addressed.

1 Introduction

The objectives of the crystallization stage in a white sugar refinery are to produce sugar at the right quality (size, color and crystal regularity) while at the same time meeting the capacity requirements of the refinery. There is always a trade-off between capacity and quality and pans can generally be boiled more quickly to satisfy capacity requirements but at the expense of sugar quality.

It is also required to meet these requirements at minimum cost. This implies that high sugar yields must be achieved in each boiling to minimize the steam requirement and do so with the lowest labor cost, generally implying full automation of the batch pan process.

2 Sugar quality

The most important quality parameter is usually color. This is determined by the color of the feed to the pan as well as the proportion of color in the mother liquor which ends up in the crystal. This can be significantly affected by the conditions in the pan, in particular the rate of crystallization, the circulation in the pan which affects the degree to which uniform crystallization conditions are experienced, and the degree of control of the process which determines the uniformity of factors affecting crystallization in practice.

The correct crystal size has to be achieved to satisfy the market requirements. This is different in different places but in most cases is fairly easily satisfied. More important is the achievement of average size without excessive fines. In other words a uniform size distribution is required. A wide size distribution makes efficient centrifuging more difficult and excessive fines lead to a pre-disposition to caking. It is probably even more important to avoid the formation of conglomerates, which have an adverse effect on all downstream operations.

2.1 Pan Design

The pan capacity required depends on the rate at which the pans are designed to boil. Pan cycles (including discharging, washing and filling the pan) vary between 80 minutes and 3 h, depending on the design of the pans and the calandria steam pressure.

Many types of batch pan have been used over the years, from the original design introduced by *Howard* in 1813. They have evolved over the years through a mixture of experience, common sense and science to a fairly standard geometry, in the form of a vertical cylindrical vessel, using vertical tube calandrias, with steam or vapor condensing on the outside of the tubes. The most common design incorporates a calandria with horizontal flat plate tube sheets. A single central downtake is surrounded with vertical tubes, an efficient design of simple construction and low maintenance.

Batch vs. continuous pans. Although continuous boiling is now widely applied in the raw sugar mills, it has found little application in refineries. Because of the high liquor purity, boiling and crystallization are much faster. This normally makes continuous pans an inappropriate choice, as the seed still needs to be produced in a batch pan, which cannot be done in a time much shorter than 90 min. One of the problems with batch boiling is the fact that the process is crystallization rate limited in the first part of the boiling and evaporation rate limited in the latter part. There is only one point in a batch boiling when the pan is ideally suited to both simultaneous processes.

Vertical continuous pans are in use in a limited number of new refineries (*Hempelman* 2005). The main advantage is the steady steam load, which makes the achievement of high steam economy easier through process integration and the use of vapor recompression on the steady load. They require a magma to be fed to the pans, which comprises a massecuite produced in batch pans.

Continuous pans are well-suited to the conditions in a recovery house. The better circulation and more homogeneous crystallization conditions enable a better exhaustion to be obtained more readily. *Moodley* et al. (1996) reported on a system of using a single continuous pan for all three recovery house boilings, by cycling the different massecuite grades through the pan sequentially. This concept can lead to a very simple and compact recovery house, as described by *Rein* (1997). With higher exhaustion possible in the continuous pan, only two crop boilings are envisaged. In this arrangement, the pan boils both first and second crop sugars in rotation. This minimizes capital expenditure and offers potential for steam savings.

Pan shape and dimensions. Some designers adopted a conical enlargement of the body above the calandria, to increase the capacity of the vessel without increasing the strike height and also to give a lower ratio of graining volume to final strike volume. However, these flared pans were largely discredited because of the negative effect on circulation (*van der Poel* et al. 1998:783-784). This has been confirmed by computer modeling (*Rein* et al. 2004). *Tippens* (1972) showed that straight-sided pans perform better than flared pans while still being able to achieve an acceptable graining volume/massecuite final volume of less than 0.33.

The shape of the pan bottom should promote an even distribution of massecuite to the underside of the calandria, without restricting circulation or providing stagnant areas, and allow the discharge of massecuite within an acceptable time. Pans with a "W" bottom shape have become more popular, particularly with large pans. This keeps the graining volume small while still allowing good circulation and acceptable striking times.

Good massecuite circulation helps to achieve uniform conditions in the pan, a vital requirement for good crystallization. The achievement of all three objectives, namely good circulation, high heat transfer rates and uniform conditions in the massecuite, can be realized through good design and appropriate choice of pan geometry. As with all design, some compromise is required to achieve satisfactory performance without excessive equipment cost. In general terms, good circulation is favored by pan designs that incorporate shorter tubes, of larger diameter, with a low strike height and large downtake diameter.

Some basic issues which affect massecuite circulation need consideration:

- Massecuite flow paths should be clean and unobstructed. Any unnecessary steel inside the pan represents a resistance to circulation. Louvers, other steel devices installed in the pan, and any unnecessary ironware in the massecuite should be removed.
- Incondensable gases should be vented through the outside of the calandria, and not up through the massecuite.
- Likewise, condensate outlets should be positioned at the periphery of the pan, and should not run from the bottom tube plate down through the massecuite.
- The syrup or molasses feed system should also constitute a minimum obstruction to the circulating massecuite.
- If the feed is conditioned and is at a higher temperature than the boiling massecuite, the feed must be directed under the calandria so that the flash will aid circulation.
- The conical enlargement or flaring above the calandria has a negative effect on pan circulation.
- Circulation can be assisted by installing a stirrer in the downtake or in raw sugar pans using circulation steam (jigger) to promote circulation.

Tube length and diameter. Tube diameters < 100 mm are considered to affect circulation adversely (*Tippens* 1972). Tubes above this size exhibit less resistance and promote circulation, but the lower area/volume ratio is unfavorable and the graining volume consequently increases (*Rouillard* 1985). Short tubes (<1 m) give the best heat transfer coefficients. Because of the lower viscosity and higher heat transfer rate, tube lengths may be a little longer than in low grade boilings, without any penalty.

Downtake diameter. The downtake diameter is generally not less than 0.4 times the pan diameter, unless a stirrer is fitted. A smaller diameter has been shown to restrict circulation (*Tippens* 1972). The ratio of the cross sectional area of the tubes to the area of the downtake (circulation ratio) should be less than 2.5 to obtain a pan with good circulation. Nonetheless a number of pans that have circulation ratio values up to 2.8 have given reasonable results.

Strike height. As the massecuite head above the top tube plate increases, the hydrostatic pressure on the boiling massecuite increases, raising the boiling temperature. As a result, the available temperature difference between vapor in the calandria and massecuite becomes smaller, leading to a reduction in evaporation rate and massecuite circulation. The situation is particularly critical at the end of the strike, when the highest level coincides with the maximum density and viscosity of the massecuite, all factors that impair circulation.

Tippens (1972) reported that the maximum evaporation rate in a stirred refined sugar pan occurs at a massecuite level of 1.4 m above the calandria. Taking into account the fall-off in evaporation rate above this level, he calculated that the maximum production rate is achieved with a strike height between 1.5 and 2 m above the calandria, and is little affected within this range. Tests performed by *Austmeyer* (1986) in stirred beet sugar pans suggested that a maximum heat transfer coefficient in low grade boilings is reached when the massecuite level is 0.8 m above the top tube plate, after which it starts reducing progressively as the massecuite level increases. However he showed that for white boilings, the maximum heat transfer coefficient is achieved at a level of 0.15 to 0.65 m above the tube plate. A final height or strike height around 1.2 – 1.6 m usually provides the best balance between quality, performance and capacity in raw sugar mills; in refineries a 2 m strike height is not uncommon.

Pan feed system. The pan feed system should be designed to mix the feed into the bulk of the massecuite as soon as possible. Good rapid mixing of feed with massecuite is important to ensure that no areas of under-saturated conditions endure, which can lead to partial crystal dissolution. The feed should be into the downtake if a stirrer is fitted. Otherwise it should be directed uniformly under the calandria, particularly if the feed is hot and flashes on entry; this assists circulation.

Heating steam / vapor. The heating steam / vapor should be at a high enough pressure to achieve an appropriate temperature difference for heat transfer. Generally pans require a temperature difference between condensing steam and massecuite of 35 °C for a stirred pan or 45 to 50 °C for an unstirred pan (Ziegler 1978). Since the massecuite temperature is around 65 °C on average, this means that the calandria steam should be at a temperature of at least 110 °C (144 kPa) or 100 °C (102 kPa) for unstirred and stirred pans respectively.

Steam distribution. The steam system should be designed to ensure uniform distribution to all parts of the calandria. In addition, a positive purging of incondensables to appropriately placed outlets is required.

Stirrers. Pan stirrers if correctly designed can significantly improve the performance of a batch pan. The assisted circulation improves heat transfer and shortens the duration of the batch boiling, thus improving capacity. It has also been shown that stirrers improve the quality of high-grade sugar produced (*Rein* 1990). This is a consequence of the better circulation leading to more homogeneous crystallization conditions within the pan. However a stirrer can never totally compensate for a poorly designed pan.

There is complete consensus that well-controlled pans with mechanical circulation make a vast contribution to proper conditioning by producing uniform crystals and reducing conglomerate counts. *Rodgers* and *Lewis* (1963) reported conglomerate counts as low as 15 %, and attributed this largely to mechanical circulation. *Chapman* (1970) noted that mechanical circulation reduced syrup inclusions (in conglomeration) by 33 – 50 %. *Stachenko* et al. (1966) found a drop in conditioning time from 4 or 5 days to 2 days after installing pan stirrers, and asserted that increased capacity on the pan floor (to allow slower boiling) pays for itself by allowing a smaller conditioning plant.

The first stage of a strike is characterized by a high evaporation rate. This normally corresponds to the period over which graining is done and the crystal established. The mechanical circulator is particularly useful at this time, enabling the steam to the calandria to be cut back without sacrificing circulation. The alternative is to add water, which is wasteful of steam. During the last stages of the strike, the evaporation rate is reduced, resulting in lower vapor generation. The effect of the forced circulation becomes important at this stage, and is reflected in higher heat transfer coefficients compared to natural circulation vacuum pans (*Austmeyer* 1986).

Its installation gives all the advantages associated with good circulation mentioned previously. In contrast, high capital costs, air leakage and high power consumption, particularly at the end of the strike, work against the installation of stirrers.

Mechanical agitators provide the option of achieving an acceptable heat transfer with a lower temperature difference. The use of lower pressure vapors becomes possible, enabling reductions in the factory steam requirements.

A stirrer makes it feasible to have a smaller diameter downtake, thus enabling a larger heating area to be installed for a given pan diameter. The stirrer itself is located in the downtake if it is an axial flow impeller (pitched blade turbine or marine propeller), or else just below the downtake in the case of a radial flow impeller. Experiences with Kaplan (mixed flow) and a helical axial flow impeller are described by *Purdham* and *Cox* (1990).

The tip speed has to be kept below a designated maximum speed, or else false grain will form. *van der Poel* et al. (1998:786) recommend a maximum tip speed of 5.8 m/s and *Kiuijvenhoven* (1983) a maximum of 10 m/s. In practice, tip speeds below 7 m/s have given good results.

Each stirrer should be individually designed for the particular pan and its duty. There is no universal stirrer for all occasions. It is seldom possible to transfer a stirrer from one pan to another without redesigning the arrangement, if optimum results are to be obtained.

2.2 White pan operation

The choice of pan absolute pressure, and hence massecuite boiling temperature, is determined by a number of factors. Higher temperatures lead to an increased rate of crystallization, promoting pan capacity and reduced boiling times. However, high temperatures also promote the formation of color and increase sucrose losses. In some cases factors such as available steam pressure, pan design and throughput requirements allow little choice in choosing the optimum conditions.

Conglomerates. Conglomerate formation is a phenomenon that can occur soon after graining and involves the growing together of several crystals. In the raw house graining is carried out at a lower purity, at which level conglomerates practically never form. At high purity however conglomerates may form if the supersaturation is not closely controlled. To avoid conglomerate formation, circulation in the pan must be good, supersaturation must be kept below 1.2 and the crystallization rate should not be too high. Conglomerates inevitably trap some mother liquor inside the crystal and the consequences of conglomerates may be listed as follows:

- Nonsucrose trapped in the conglomerate reduces sugar purity
- The efficiency of separation in the centrifugals is impaired
- The bulk density of the sugar is increased.
- Conglomerates are more prone to crystal breakage, leading to dust formation
- The sugar becomes more difficult to condition.

Color. Some color generation occurs during boiling, with massecuite color increases between 8 and 16 % being reported. This is minimized by good pan design and the use of stirrers in pans (*Rein* 1990). In general lower massecuite temperatures lead to less color formation, as most color-forming reactions have an exponential relationship with temperature.

Color removal in refinery white boilings is about 94 % and may vary between 90 and 95 %, although ash removal is about 99 %. The color removal as a percentage is higher with higher feed colors and may be estimated from the following relationship for carbonatation refineries (*Thompson* et al. 2005):

$$\text{Color elimination factor} = 15.5 + 0.007 \cdot \text{feed color} \quad (1)$$

The color elimination factor is the ratio of massecuite color to sugar color. This is affected substantially by the conditions in the pan, but gives estimates of color removal slightly higher than values given by *Rein* (1990) above i.e. elimination factor averaging 16.7 with a range of 10 to 20.

3 White sugar yields

The crystal yield obtained on each boiling is very important, particularly from a steam economy point of view. Yield is defined as kg crystalline sucrose /100 kg massecuite dry solids. High yields lead to significantly reduced massecuite quantities, which also reduce steam consumption. This requires boiling a clean conglomerate-free grain to a high crystal content in the pan. Centrifuging must also be carefully controlled to ensure that the minimum amount of water is used to achieve the desired quality, and that the basket is ploughed clean without leaving crystal in it to get redissolved in the basket wash between cycles.

3.1 Target sugar yields

The limit to crystal content can be calculated, assuming that the crystals are in contact with each other and only the void space is filled with mother liquor. In this case, the porosity or void fraction ε can be calculated from the bulk and crystal densities:

$$\varepsilon = (1 - \rho_b / \rho_{Cr}) \qquad (2)$$

Assuming the voids are filled with mother liquor with a dissolved solids content $w_{DS,ML}$ and density ρ_{ML}, the crystal content is given by:

$$w_{Cr,Ma} = 100 \cdot \frac{(1-\varepsilon)\cdot \rho_{Cr}}{(1-\varepsilon)\cdot \rho_{Cr} + \varepsilon \cdot \rho_{ML} \cdot w_{DS,ML}/100} \qquad (3)$$

$$w_{Cr,DS} = 100 \cdot \frac{(1-\varepsilon)\cdot \rho_{Cr}}{(1-\varepsilon)\cdot \rho_{Cr} + \varepsilon \cdot \rho_{ML}} \qquad (4)$$

Assuming that the mother liquor has a supersaturation of 1.05, the dissolved solids content may be found in *Bubnik* et al. (1995). Calculations for two massecuite purities (corresponding to white and recovery boilings) and two temperatures are shown in Table 1. The sucrose crystal density ρ_{Cr} is 1587 kg/m^3 (*Bubnik* et al. 1995) and the mother liquor density is obtained from tables or correlations.

Massecuite Purity	100	100	85	85
Temperature (°C)	70	60	70	60
Mother liquor DS $w_{DS,ML}$ (g/100 g)	77.3	75.2	79.7	77.6
Mother liquor density (kg/m^3)	1369	1363	1385	1379
Sugar bulk density (kg/m^3)	900	900	850	850
Void fraction	0.433	0.433	0.464	0.464
Massecuite solids (g/100g massecuite)	91.0	90.2	91.3	90.4
Crystal content $w_{Cr.Ma}$ (g/100g massecuite)	60.3	60.4	56.9	57.0
Crystal content $w_{Cr.Ma}$ (g/100g dry matter)	66.3	67.0	62.4	63.1

Table 1. Calculated maximum massecuite crystal content. Mother liquor concentration and bulk densities estimated from *Bubnik* et al. (1995).

In practice it is not possible to achieve these pan yields and a value of 95 % of this is a practical maximum. With a high yield boiling, it is sometimes necessary to flush out the pan on striking with some of the jet (runoff) from that boiling.

Yields in the pan house are calculated using the runoff from the centrifugals. Because of the high purities involved, yields cannot be determined by purities and ash contents are used instead. The calculation for yield uses the following relationship:

$$Yield = 100 \cdot \frac{w_{A,ML} - w_{A,Ma}}{w_{A,ML} - w_{A,S}} \qquad (5)$$

where $w_{A,Ma}$ is the ash content of the massecuite, $w_{A,ML}$ the ash content of the mother liquor and $w_{A,S}$ the ash content of the sugar.

Some loss in centrifugal washing is also inevitable. Again the best case probably assumes that 5 % of the sugar gets dissolved, suggesting a practical maximum yield of 59 g crystal/100 g DS in white pans. Yields are highest with first boilings and drop progressively in subsequent boilings. Centrifugal yields range between 80 and 95 % in practice and typical attainable yield values over the pan and centrifugal, expressed as kg crystal/100 kg massecuite DS are:

1st boiling	56
2nd boiling	53
3rd boiling	50
4th boiling	46

Using these yields and assuming a massecuite of 90 % DS and density of 1470 kg/m^3, a material balance on white sugar crystallization can be undertaken and the results are shown in Table 2.

Boiling	**1st**	**2nd**	**3rd**	**4th**	**Total**
DS in massecuite (t/h)	100	44.0	20.7	10.4	
Massecuite volume (m^3/h)	75.6	33.3	15.6	7.8	132.3
DS in runoff (t/h)	44.0	20.7	10.4	5.6	
DS in sugar product (t/h)	56.0	23.3	10.4	4.8	94.5
Proportion of total product sugar	0.59	0.25	0.11	0.05	1.0

Table 2. Typical material balance on white sugar crystallization in a four-boiling scheme refinery with good yields.

3.2 Effect on pan capacity

In practice, average yield values over the pan and centrifugal may be somewhat lower, such as the following:

1st boiling	52
2nd boiling	48
3rd boiling	44
4th boiling	40

Using these yields and assuming a massecuite of 90 % DS and density of 1470 kg/m^3, the material balance looks somewhat different and the results are shown in Table 3.

Boiling	**1st**	**2nd**	**3rd**	**4th**	**Total**
DS in massecuite (t/h)	100	48.0	25.0	14.0	
Massecuite volume (m^3/h)	75.6	36.3	18.9	11.3	142.1
DS in runoff (t/h)	48.0	25.0	14.0	8.4	
DS in sugar product (t/h)	52.0	23.1	11.0	5.6	91.7
Proportion of total product sugar	0.57	0.25	0.12	0.06	1.0

Table 3. Typical material balance on white sugar crystallization in a four-boiling scheme refinery with average yields.

By comparing the results in the two Tables, the large effect of yields on massecuite quantities can be seen. In addition the greater proportion of sugar produced in first boilings relative to third and fourth boilings influences the average sugar color significantly.

The effect on pan capacity can be illustrated by an example. If the batch pan cycle time is 2.5 h , the massecuite quantities in Table 3 indicate a pan capacity requirement of 142.1 x 2.5 = 355 m^3 or 3.55 m^3/t DS in fine liquor. In the particular situation depicted in Table 3, the pan requirement may be satisfied by having 4 x 90 m^3 pans, two on first boilings, one on second boilings and one for thirds and fourths. In the higher yield case shown in Table 2 the 4 pans could be smaller at around 84 m^3 each.

3.3 Steam usage

The amount of steam required is directly proportional to the amount of massecuite to be boiled. It becomes clear therefore that crystal yields are also important in limiting the steam requirement of a refinery.

4. Automatic pan control

4.1 Measurement transducers

Conventional methods of measurement of pressure, flow and level provide no particular problems. The absolute pressure control requires an absolute pressure transmitter, rather than a vacuum transmitter (a pressure transmitter measuring pressure relative to atmospheric pressure). This ensures that the pan boiling temperature is unaffected by changes in atmospheric pressure. Measurements used in pan control are sensitive to temperature variations and a controlled temperature is therefore important (*Saska* and *Rein* 2001).

The major challenge in pan control is the control of the dissolved solids concentration, the crystal content and the consistency of the massecuite. There are no commercial instruments available that can do this directly, and secondary measurements have to be used to infer the quantities of interest. At the start of a boiling, and particularly when graining, the supersaturation of the liquor needs to be controlled. As the boiling progresses, control of crystal content becomes more important, and is the overriding factor dictating the consistency at the end of a batch.

Boiling point elevation and online refractometers can be used to measure mother liquor concentration (supersaturation). They are useful at the start of a boiling, but only measure the state of the mother liquor and cannot be used to control the whole cycle. Boiling point measurements are also affected by changes in massecuite level.

Some transducers are used which are sensitive to both the concentration at the start of the boiling and throughout the cycle:

- Electrical conductivity is a useful method but only suitable in raw sugar factories (or in the recovery house). It actually measures the activity of the ionic species, but depends on both mother liquor concentration and massecuite consistency as affected by increasing crystal content. Conductivity is non-linear with respect to dissolved solids concentration.
- Viscosity transducers behave in a manner similar to conductivity, but can be used on all grades of boiling. A *Ziegler* consistency meter was commonly used in the past, but it and other viscosity transducers are nowadays less in evidence.
- Radio frequency conductivity measurements measure electrical properties at radio frequencies (*Radford* et al. 1988; *Reichard* et al. 1992). The measurements are affected by both conductivity and dielectric constant. They have some advantages over conventional conductivity, in that they can operate in refined boilings, and it is possible to separate outputs representative of the resistive and capacitive characteristics of the massecuite.
- Nuclear density meters are sensitive to crystal content but are relatively insensitive to changes in mother liquor concentration. However they can still be used for graining control (*Donovan* 1988). They are nowadays less acceptable for environmental reasons.
- Microwave absorption measurements are largely responsive to water content and so measure total solids accurately. They are similar to nuclear density measurements since they are more responsive to crystal content than mother liquor concentration. A big advantage is the fact that the measurement is not affected by changes in purity (*Saska* and *Rein* 2001).

It is important to give careful thought to the placement of the measuring devices. They should be far enough away from the feed and from vapor bubbles so that these do not affect the measurement. The placement becomes most important when the pan is not circulating fast, particularly near the end of a boiling.

4.2 Control valve sizing

Correct control valve sizing is essential if good, reliable control is to be obtained. The first requirement is to establish the average, maximum and minimum flows to be expected. Then it is necessary to calculate values of C_V for these conditions, to establish from valve supplier's specifications the size of control valve required.

Flashing and cavitation can have a significant effect on the performance of the valves. As the liquid passes through the valve, the velocity increases and the pressure drops according to *Bernoulli's* equation. If the pressure drops sufficiently, the liquid will flash, forming vapor bubbles. This has two potential adverse effects; firstly the ability of the valve to pass the required amount of liquid is severely affected, and secondly if the pressure subsequently rises, the vapor bubbles collapse, which over time can lead to serious damage of the valve. The potential for this to occur is real for the cases of injection water entering the condenser and conditioned pan feed as a result of the reduced downstream pressure and the raised temperature. In the case of injection water, the valve should be located at a low elevation where the static pressure is high enough to prevent cavitation, or right at the entrance to the condenser where the consequences of cavitation are less severe.

In the cavitation regime, the actual pressure difference to be used in the sizing equations is less than the actual pressure drop available. The maximum allowable pressure drop for sizing purposes is given by the critical value at which cavitation occurs for high recovery valves (butterfly and ball valves, the most common choice for pan control), dependent on the water inlet pressure and vapor pressure.

4.3 Batch pan control

Automatic process control of pans leads to consistent results, and when set up optimally leads to good exhaustion, good sugar quality and maximum pan capacity. The fundamental control systems are almost the same on all pans. The control actions involved are:

Absolute pressure control: This is controlled by regulating the water flow to the condenser. The set point is usually kept constant throughout a strike, although it may be varied during a boiling, particularly on refinery boilings. The control system may incorporate a tailpipe temperature override. This limits the over-use of water, which may be caused by an inadequate or faulty vacuum system; if a vacuum pump is not performing, the concentration of incondensable gases rises, which is reflected as an increase in absolute pressure and the control system responds by adding more water.

Level control: This is operative during filling and while controlling the level during the concentration period prior to graining, by regulating the feed valve. The level transmitter also detects the pan full condition and initiates the final concentration phase.

Feed control: This controls the concentration before, during and after graining, by regulating water addition. Once the grain is established, the build-up starts with feed of molasses or syrup. The set point for the control is usually changed as the level rises to achieve a steady increase in crystal content. Once the pan is full, the set point is ramped to the strike value, and the controller will admit water via the water control valve if necessary during this period.

Stirrer motor power: The power absorbed by the stirrer may be used to control feed during the build-up period. The stirrer power is particularly useful as a reliable and reproducible measure of the final strike consistency.

Seed crystal addition: Once the correct concentration for graining is reached, the pan is controlled at this concentration for a time to stabilize conditions. Then the control system may automatically open the slurry valve, admitting the charge of slurry that the operator has put in the feed funnel. Some operators prefer this to be done manually by the pan boiler, who is alerted by a "ready" alarm.

Steam flow rate: The flow of steam flow rate to a pan may be limited to a maximum set value and may be ramped up and down on start-up and shutdown, to minimize the swings caused on the overall steam demand.

To automate pan operations fully, all valves need to be actuated and controlled by a sequence controller. This requires remote actuation of valves on the calandria steam supply, vacuum breaker, steaming out, steamings diversion and ejector steam, as well as automatic valves on the massecuite strike and cut-over systems. Timers set the time for operations such as steaming out, and alarms are necessary to alert the operator to out of control situations.

It is now possible to use the measurements available to compute by mass balance the contents of the pan, particularly crystal content (*Martins* et al. 2005). The availability of cheap, easy-to-use and readily available hardware and software makes online monitoring and optimization of pan boiling a reality.

References

Austmeyer K.E. (1986): Analysis of sugar boiling and its technical consequences: Part II. Heat transfer during sugar boiling. Int. Sugar J. 88, 1046, 23-29.

Bubnik Z.; Kadlec P.; Urban D.; Bruhns M. (1995): Sugar Technologists Manual: Chemical and Physical Data for Sugar Manufactures and Users. Verlag Dr Albert Bartens, Berlin. 8th Ed.

Chapman F.M. (1970): Aspects of white sugar conditioning and storage. Proc. 20th Conf. Brit. Sugar Corp.

Donovan M. (1988): Sensors for computer control of white sugar vacuum pans. Proc. Sugar Ind. Technol. 47, 78-89.
Hempelman R. (2005): Continuous crystallization of refined sugar – present situation and concepts for the future. Proc. Sugar Ind. Technol. 64, 131-143.
Kiuijvenhoven L.J. (1983): Aspects of continuous sucrose crystallization. Rept. WTHD No 156, Delft Univ. Technol.
Lyle O. (1957): Technology for Sugar Refinery Workers. 3rd Ed. Chapman & Hall, London.
Martins, P. M.; Rocha, F. A.; Rein, P. (2005). Modeling sucrose evaporative crystallization. Part 1. Vacuum pan monitoring by mass balance and image analysis methods. Ind. Eng. Chem. Res. 44, (23), 8858-8864.
Moodley M.; Schorn P.M.; Singh I. (1996): Off-crop refining at Malelane. Proc. S.Afr. Sugar Technol. Assoc. 70, 189-194.
Purdham P.R.; Cox M.G.S. (1990): Design and evaluation of a pan stirrer. Int. Sugar J. 92, 1100, 160-167.
Radford D.J.; Tayfield D.J.; Cox, M.G.S. (1988): Further developments in automated white pan boiling using radio frequency control. Proc. Sugar Ind. Technol. 47, 90-107.
Reichard S.R.; Broadfoot R.; Wright P.G. (1992): Development of conductivity and capacitance sensors for pan boiling. Proc. Int. Soc. Sugar Cane Technol. 21, 659-669.
Rein P.W. (1990): The effect of pan design and operation on white sugar quality, in Raw Sugar Quality and White Sugar Quality, Ed. M.A.Clarke. Verlag Dr A. Bartens, Berlin. 109-116.
Rein P.W. (1997): The application of continuous pan boiling in white sugar refineries and the impact on energy consumption. Proc. Sugar Ind. Technol. 56, 308-312.
Rein P.W.; Echeverri L.F.; Acharya S. (2004): Circulation in vacuum pans. J. Am. Soc. Sugar Cane Technol. 24, 1-17.
Rodgers T.; Lewis C.L. (1962-63): The drying of white sugar and its effect on bulk handling. Int. Sugar J. 64, 359-362; 65, 12-16, 43-45, 80-83.
Rouillard E.E.A. (1985). Massecuite boiling. Proc. S. Afr. Sugar Technol. Ass. 59, 43-47.
Saska M.; Rein P.W. (2001): Supersaturation and crystal content control in vacuum pans. Proc. Sugar Ind. Technol. 60, 251-261.
Stachenko S.; Allen W.A.R.; Swan J.A. (1966): Theory and practice of sugar curing at Canada and Dominion Sugar Co. Proc. Sugar Ind. Technol. 25, 75 - 122.
Tippens D. E. (1972): Modern calandria vacuum pan design criteria. Proc. Sugar Ind. Technol. 31, 81-97.
Thompson P.; Fry P.; Hasrajani N. (2005): Thoughts on refinery boiling schemes. Proc. Sugar Ind. Technol. 64, 89-100.
van der Poel P.W.; Schiweck H.; Schwartz T. (1998): Sugar Technology. Verlag Dr Albert Bartens, Berlin.
Ziegler J. (1978): Sugar boiling – some useful strategies. Sugar J. 40, 8, 8-15.