Improved Batch Pan Monitoring and Control – a soft sensor approach

by

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1. Summary

Cane sugar is manufactured in process vessels called vacuum pans, which are closely supervised by operating personnel, known as pan boilers. Traditionally, the pan boiler's job has been made easier through the use of a control system based on electrical conductivity. However, conductivity-based vacuum pan control is far from perfect, since it requires continual attention from the pan boiler to ensure that the vacuum pan is running efficiently.

Existing and emerging technologies are now available which should allow for more advanced monitoring and control schemes to be developed for vacuum pans. The ultimate goal for these technologies is to develop a control system that can run the pan closer to its optimal operating point, with less operator intervention.

The main aim of this project is to develop optimal operating policies for an industrial vacuum pan and implement them on a production scale. Rather than use conductivity as the control variable, it is proposed that more process relevant variables should be considered. These variables are the molasses sucrose oversaturation, or OS, and the massecuite crystal content, or CC.

Once these policies have been developed, they need to be implemented. Before that can happen, the OS and CC variables need to be available to the controller. The problem is that the variables cannot be measured directly – there are no devices that can measure either of these variables. However, a technique called soft sensing can be used to estimate the values of OS and CC in the vacuum pan. A soft sensor is actually a computer model of the vacuum pan that uses other available process measurements to predict the values for the OS and CC in the pan. This project's second aim is to develop a soft sensor that can be implemented on a factory vacuum pan, so that the OS and CC variables are available for process control purposes.

The third aim of the project is to evaluate the effectiveness of controlling the vacuum pan by using OS and CC, in place of electrical conductivity.

The final aim of this project is to control the full-scale vacuum pan in such a way that it follows the predicted optimal operating policies.

This final aim is very important to the Australian sugar industry, since it can help to make each vacuum pan operate faster and therefore be more productive. An extension of this goal is that the vacuum pan might also run with less operator intervention, giving them more time to see to other, more pressing tasks in the factory.

The testing phase of this project was carried out in partnership with Macknade Mill, who have availed a 120 tonne batch vacuum pan. This pan is unique, since it has a number of high quality process sensors attached to it. This makes it an ideal testing ground for the evaluation of alternative process control techniques.
The main outcomes from this work are that while advanced monitoring and control are possible on the 100 tonne vacuum pan, it is currently not feasible to operate the vacuum pan to the predicted optimal operating policies.

However, it has been clearly demonstrated that controlling the pan using a massecuite Brix sensor is efficacious. The advantage of using this for process control, instead of conductivity, is that this device is much less sensitive to changes in cane purity. As a result, it can be argued that this method of control requires much less operator intervention, which could lead to cost savings for the industry.

If the findings of this study are applied to the Australian sugar industry, it is likely that productivity will improve through batch time reductions and potential reductions in plant operating personnel.

In summary, this project has succeeded in evaluating the applicability of advanced process control to the operation of batch vacuum pans. A great deal of knowledge has been gained in terms of what level of advanced control is practicable. Furthermore, it proposes a new method of pan control, based on massecuite Brix, that could be used in place of conductivity control.
2. Introduction
This final report completes the requirements for SRDC project JCU018. The report presents background information, which leads naturally into the objectives of the project. Research methodologies are presented in order to clarify how the research was executed. Results of the work undertaken then follow. Project outputs and expected outcomes are delineated in the next two sections. Recommendations for further work are presented in the next section. Finally, copies of papers published as a result of this research appear in the last section of the report.

3. Background
For the last several decades the Australian Sugar Industry has clung to a simple, conductivity-based pan control scheme, given its low installed cost, simplicity and high reliability. The main hurdle to overcome in this type of control scheme was in determining the setpoint for the conductivity controller; since this measurement is not fundamentally process relevant. Initially, this did not present a problem, since skilled operators were on hand to make that determination.

Paradoxically, the problem of pan control really isn’t the problem at all. In fact the control problem is actually a measurement problem. If the molasses oversaturation (OS) and massecuite crystal content (CC) in the pan can be measured (or estimated) then it is a trivial exercise to control them. The main difficulty lies in knowing the setpoint trajectories (or ramps) to choose for these process variables throughout the duration of the strike.

The overriding goals of this project were as follows:

- to develop a reliable scheme that would estimate, in real time, the OS and CC process variables.
- to determine the best operating policy for a batch vacuum pan through a computer optimisation study

In order to pursue the above goals, a series of objectives were established and are presented in the next section.

4. Objectives
The objectives of this investigation, as stated in the original proposal, and a statement of the extent to which the project has achieved them is summarised below.

*To develop practical batch pan monitoring and control strategies based on a soft sensor approach*

A soft sensor was developed in order to predict two key process variables on-line and in real time

- molasses sucrose oversaturation (OS)
- massecuite crystal content (CC)

on a 120-tonne industrial batch vacuum pan.

This sensor proved to be useful in terms of process monitoring, since it gave operating personnel an alternative way of evaluating the process. However, it must be said that the uptake of information was only slight. This is explained by the fact that pan boilers lack a deep understanding of the process fundamentals involved.
Trying to control the pan using the above two derived process variables was not as successful. This is due in part to errors in the two predicted process variables. This was further complicated by the fact that the operating policies predicted by computer simulation did not stand-up to industrial implementation. In other words, what worked well in computer simulation, did not work well on the actual process.

**To evaluate these schemes in simulation and on a 120 tonne production vacuum pan**

In simulation, the proposed batch pan monitoring and control scheme worked extremely well. By incorporating newly available process measurements (such as the microwave Brix device), it was straightforward to develop a reliable soft sensor for monitoring and control purposes.

In terms of actual process monitoring, a soft sensor set-up to run in parallel with a 120 tonne batch vacuum pan, located at CSR’s Macknade Mill. This soft sensor was able to predict reasonable values for the molasses oversaturation and massecuite crystal content, which gave operating personnel a completely new view of the process. It is of interest to note that operating personnel were hesitant to use this information, most likely since they lacked sufficient training to fully utilise this information.

It was demonstrated that controlling the molasses oversaturation and massecuite crystal content led to the control the full-scale batch vacuum pan. This represents an important departure from traditional control schemes that rely on process outputs, such as conductivity.

**To revisit, and consequently redefine, the optimal control problem first presented by Frew, but in a more practical and direct manner.**

The Postdoctoral Fellow attached to this project, Dr Linh Vu, revisited the optimal control problem of the batch vacuum pan. The objective of the optimisation method was to reduce the time required to grow the seed crystals to their target size.

Dr Vu’s work in this area was significant, since she was able to incorporate various real-world operating features into her optimisation studies, including:

- **A-molasses boilback** - The optimum switching time for the commencement of A-molasses boilback was determined by this approach, determined by setting a target molasses purity for the end of the strike.

- **Process constraints** – Constraint handling was incorporated into the optimisation scheme, including limits on sucrose oversaturation, massecuite crystal content and the flow of steam and feed to the pan. Constraint handling on the state variables (like oversaturation and crystal content) has not been achieved to date.

- **Uncertainties** – The effect of plant-model mismatch and uncertain operating conditions (footing conditions, feed composition) could also be handled by the approach developed by Dr Vu, which makes the optimisation process more robust.
To implement these optimal control policies on the 120 tonne batch vacuum pan

The operating policies predicted by the computer-based optimisation studies were implemented on the same 120 tonne batch vacuum pan located at CSR's Macknade Mill.

It became evident very early in these trials that the predicted optimum operating policies were not superior to current pan operation. This was a rather discouraging result, to say the least. Many attempts were made to overcome this issue, but to no avail.

The reasons for this implementation failure are most likely related to some combination of the following effects:

- **a poor heat transfer model** -- modelling heat transfer across the calandria is a key requirement to solving the optimal control problem posed by Dr. Vu (and others before her).
- **no model of pan circulation** -- the model employed in Dr. Vu's optimal control study assumed that the mixing behaviour of the pan is perfect. It is possible that the sensors being used in the soft sensor were not indicative of average conditions in the pan. In this case, the control policies being implemented would be compromised.
- **a poor thermodynamic model** -- the model of the batch pan employs rather coarse relations describing the sucrose solubility in the mother molasses. The relations employed were chiefly developed for lower purity molasses, since no relations exist for high purity molasses.
- **inaccurate crystal growth model** -- the model description of the crystal growth rate kinetics is far from perfect. However, the sensors selected for measurement feedback to the soft sensor should afford some protection against this problem.
- **calibration errors** -- the sensors used for measurement feedback may have been inaccurately calibrated. The net effect of this is that the estimated OS and CC values would be biased away from their true values.

Based on the mediocre performance of the plant-based soft sensor experiments, additional project funds were requested, and subsequently granted, in order to pursue one final objective.

To evaluate alternative control schemes based on direct measurement of the process

Given that the optimal control of the pan by the soft sensor approach did not live up to expectations, it was decided to attempt control of the vacuum pan using novel process sensors.

A simulation study was carried out in which the molasses and massecuite Brix were controlled, by manipulating the feed and steam flows to the pan. This approach was successful in computer simulation.

During the 2002 crushing season, a series of trials were carried out at Macknade Mill in which the output from a microwave Brix device was controlled for almost the entire period of the strike.

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1 pro-M-tec Theisen GmbH, Pforzheimer Str. 162, D – 76275 Ettlingen
Operator acceptance of this scheme was high. During the 2002 crush, Macknade Mill took on factory make-up water with brackish water (due to tidal effects on the creek adjacent to the plant). At these times automatic pan operation was effectively impossible, since the conductivity probes were driven off-scale, due to the high conductivity of the incoming liquor streams, concentrated in salt. At these times the pan boilers requested that the microwave-based control system be switched on, freeing them from the need to monitor this pan closely. Since the microwave device is not influenced by the presence of salts, ongoing adjustment of setpoint trajectories would no longer be necessary.

By the end of the 2002 crush, the pan was being controlled using the microwave device at least as well as it was under traditional conductivity control. Unfortunately, due to time restrictions and issues related to the batch automation structure, it was not possible to optimise the pan for microwave Brix control. However, it is anticipated that batch optimisation based on microwave Brix control would be straightforward.

In summary, the project objectives were largely met. It is apparent that the more complex pan control system, based on soft sensing, is not ready for factory implementation. However, it is clear that improved output base control is feasible and should lead to more uniform and effective batch pan control.

5. Methodology
The methodologies for this research project were diverse and fell into three distinct categories. They are described in turn.

Soft Sensor Development

The first area of research investigated the design of a feasible soft sensor for factory implementation. The ultimate goal of the soft sensor is to run in real-time, in parallel with the process that it is monitoring. Its development was achieved by first implementing the soft sensor on a computer simulated batch pan. Next, the soft sensor was set up to process data logged from the full-scale factory pan. The third stage was to implement the soft sensor in real-time, in step with the factory vacuum pan. An advanced control prototyping and development package, called UNAC2, was employed for all soft sensor development.

The soft sensor was formulated using a modified version of Wright’s model, which assumed a point crystal size distribution. The model employed flow and temperature data to in order to make predictions of the OS and CC in the pan. Measurement feedback from the process was based on massecuite Brix (microwave probe), molasses Brix (process refractometer) and massecuite level (?P cell). This choice of measurement feedback affords excellent feedback to the soft sensor, since it is receiving information that is highly process relevant. These soft sensor input and feedback measurements are highlighted in Table 1 below.

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2 Matrikon Inc., PO Box 516, Mayfield NSW 2304
Table 1: Soft sensor input and feedback measurements

<table>
<thead>
<tr>
<th>Soft Sensor Inputs</th>
<th>Soft Sensor Feedback</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Variable</strong></td>
<td><strong>Device</strong></td>
</tr>
<tr>
<td>feed</td>
<td>mag flow</td>
</tr>
<tr>
<td>steam</td>
<td>ΔPcell/orifice plate</td>
</tr>
<tr>
<td>movement water</td>
<td>mag flow</td>
</tr>
<tr>
<td>temperature</td>
<td>100ohm RTD</td>
</tr>
</tbody>
</table>

**Vacuum Pan Optimisation**

The second area of research involved the development of the batch pan optimisation codes. All optimisation work was carried out using GAMS\(^3\), which is specifically designed for modelling linear, nonlinear and mixed integer optimisation problems. To avoid solving the ODEs required for each iteration step, which is computationally expensive, the state variables are approximated by the Lagrange interpolation polynomial. These were then differentiated and back substituted into the state equations, converting them to a set of algebraic equations. The approximating problem now becomes a normal nonlinear programming problem and can be solved by existing optimisation techniques. The approximating method of Orthogonal Collocation on Finite Elements is especially advantageous for this type of optimal control problem, since it is solved for not only the minimum operating time but also the duration of each stage of the whole batch pan process. Any number of finite elements can be specified for each stage; it is the lengths of these elements that are decision variables in the optimisation problem. The advantage of this approach was that events, such as A-molasses boilback, could be determined based on a specified target purity for the mother molasses at the end of the strike.

**Factory Implementation**

The final area of this research involved the implementation of the proposed control schemes on the 120 tonne batch vacuum pan at Macknade Mill. Again UNAC was employed for this task, since the soft sensor was commissioned using UNAC. The first step in this process was to merge the soft sensor with the existing pan control scheme. Macknade Mill employs a reasonably complex set of batch sequencing to effect very good control of their pans. These batch sequences are complex and difficult to account for when attempting to monitor the pan with the soft sensor, since all of the possible events need to be built into the soft sensor code. The complexity of this task should not be underestimated, since it plays a key role in the success of a robust the soft sensor.

The next stage in the factory implementation involved the development of the control system that employed either the soft sensor outputs (OS and CC) or the microwave Brix device. Below is a screen shot of a portion of the UNAC-based control system,

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\(^3\) GAMS Development Corporation, 1217 Potomac Street, NW, Washington, DC 20007
showing details of the microwave controller. It should be noted that a significant amount of this configuration is devoted to bullet-proofing, so that the controller could be implemented with confidence. This bullet-proofing was also required on the existing factory control system, in case UNAC froze-up for some reason.

**Figure 1: UNAC implementation of a PID feedback controller**

Feedback control was affected using simple PID formulations. In fact, only PI control was required for the OS and CC control, while P-only control was required for the microwave Brix control. In all cases, some kind of setpoint trajectory was required. The computer optimisation worked performed by Dr Vu was used to generate these trajectories.

The table below shows the different configurations considered in this study for the control of the batch vacuum pan.

**Table 2: Controller configurations considered for this study**

<table>
<thead>
<tr>
<th>Option</th>
<th>Controlled variable(s)</th>
<th>Manipulated variable(s)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td>Feed</td>
</tr>
<tr>
<td>1</td>
<td>OS</td>
<td></td>
</tr>
<tr>
<td></td>
<td>CC</td>
<td>✓</td>
</tr>
<tr>
<td>2</td>
<td>CC</td>
<td>✓</td>
</tr>
<tr>
<td>3</td>
<td>OS</td>
<td>✓</td>
</tr>
<tr>
<td>4</td>
<td>microwave Brix</td>
<td>✓</td>
</tr>
<tr>
<td></td>
<td>refractometer Brix</td>
<td>✓</td>
</tr>
<tr>
<td>5</td>
<td>microwave Brix</td>
<td>✓</td>
</tr>
</tbody>
</table>
6. Results

As stated above, it was possible to estimate the OS and CC in the 120-tonne factory batch pan. A variety of factory experimental trials were run in order to determine the efficacy of the soft sensor. The estimated OS and CC were repeatable and reasonable.

Effective process control these variables was also achieved. The first three options shown in the table above were evaluated at the factory. While it was possible to control the pan using these options, it became clear that the predicted optimum profiles for the OS and CC did not produce optimal factory results. In fact, many of the trials had to be aborted, due to concerns expressed by the operator (and shared by the author!). The reasons for this have been explained in the Objectives section above. While this is a disappointing result, it is nonetheless valuable, since it is the first time that such a study has been carried out and a great deal of insight into this problem has been gained.

The last two options (4,5) were considered in simulation. The detailed results of this study have been published (Schneider, 2003) and appear in the Appendix. Factory trials were carried out evaluating only option 5 in the table above, whereby the microwave Brix signal was controlled by manipulating the feed flowrate to the pan. These results are shown below in Figure 2. In this case the microwave signal was controlled to a constant profile of 86.5 Brix. Clearly the pan is under tight control.

![Figure 2: Control of microwave Brix on 120 tonne batch vacuum pan](image)

The departures from setpoint in the early and late stages of the strike are not due to the alternative control scheme. At the start of the strike (i.e. the first 40 sample times), the pan was being controlled by the factory’s existing batch sequencing code. Late in the strike (i.e. between 360 and 400 sample time) the pan was cutting massecuite out to a receiving pan. The microwave signal changes during this time, since temperature compensation was not being used. Therefore, when the pan drops vacuum (to motivate flow to the receiving pan) the pan temperature rises, biasing the signal.
Figure 3 shows the flowrate of feed to the pan during the same period. It is of interest to note that very good control of the massecuite Brix could be achieved, despite the fact that a simple proportional mode controller was employed. Clearly this result is encouraging, since it means that a more process relevant and potentially more robust measurement is available for the control of batch vacuum pans.

![Figure 3: Feed flow to the 120 tonne vacuum pan under microwave Brix control](image)

### 7. Outputs

There are a variety of important outputs resulting from this research.

- The batch pan optimisation studies, carried out by Dr Vu, demonstrate that it is possible to optimise the operating policy of a pan, while honouring various state constraints, such limits on OS and CC. This is important, since it delivers a more realistic picture of what can be achieved on the real process. This optimisation work also includes the determination of when the pan should commence A-molasses boilback, which is a significant step forward in pan stage optimisation.

- It is possible to run a soft sensor on a 120 tonne batch vacuum pan without interfering with its routine operation. Derived process variables, like OS, CC and others, can be reported to operating personnel in real time as a pan control advisory tool. The soft sensor developed in this project produces valid estimates of these process variables.

- It was also demonstrated that the outputs from the soft sensor running on the full-scale vacuum pan can be controlled, replacing traditional conductivity-based control schemes.

- The predicted optimal setpoint trajectories derived from the computer optimisation study lead to relatively poor control of the pan, compared with traditional conductivity control. This failing is most likely due to errors in the model employed within the soft sensor and/or errors in calibration of the sensors used for measurement feedback to the soft sensor.
• A simulation study indicates that Brix-based control is an alternative to conductivity control.

• Plant studies carried out on a 120 tonne batch pan demonstrate that massecuite Brix control (manipulating feed flowrate) is a highly effective way in which to control a pan. The controller structure is simple and very effective. It could be argued that Brix-based control is superior to conductivity control, since it is not sensitive to inorganic salts, which tend to obscure conductivity measurements.

8. Expected Outcomes
There are a variety of potential outcomes from this research. Key outcomes are highlighted below.

• The work of Dr Vu indicates that it is now possible to more rigorously optimise the operation of a batch vacuum pan. The obvious outcome of this study is that factories could benchmark their pan operations against the predicted optimum. In this way, each pan on the stage could be assessed in terms of its overall productivity. There is clearly an economic incentive to carry out such benchmarking programs, although it is difficult to estimate the net dollar value benefit to the Australian industry.

• Another key outcome is that this project has clearly demonstrated that massecuite Brix control of the pan is feasible. In fact, it could be argued that this approach is superior to the conductivity paradigm. Based on its reliability, it might be possible to run batch vacuum pans with much less supervision. This would have a significant economic impact on the Australian and world sugar industry.

The main goal of this project was to evaluate the applicability of advanced process control theory to the operation of an industrial vacuum pan. Further research work should be carried out in the following areas.

• to recast the batch pan optimisation work in terms of molasses and massecuite Brix. It is clear, from the preceding discussion, that optimal control of the OS and CC are not feasible at this time. However, the development and implementation of optimal Brix profiles should have a much higher chance for success, since the soft sensor would not be used. This type of project is ideally suited to a PhD student.

• assuming that optimal control of an individual pan is achievable, the next step is to carry out an optimisation study on the entire pan stage. While optimising a single pan is important, it is critical to ensure that the pan stage is operated such that it capitalises fully on the optimal operation of each pan. Due to the advanced nature of this work, it would be best left to a Postdoctoral Research Fellow.

• it is also necessary to better understand heat transfer in batch vacuum pans. If this problem can be fully addressed, improved optimal control policies can be developed. Presently research is being carried out by Mr Darryn Rackemann in the area of pan circulation at the Masters level. It is felt that another postgraduate student, at Masters level, would be required to evaluate and model heat transfer in a mill-based study.
finally, it is highly recommended that a study commence, evaluating the utility of advanced process control applied to continuous vacuum pans. Given the significance of these units in the Australian sugar industry, this research would be well justified.

10. Recommendations
It is recommended that the following steps be taken to maximise potential benefits stemming from this research.

• extension work is recommended in order to more fully develop microwave Brix control of industrial batch vacuum pans. A participating mill is required to see host this work. Macknade Mill would make an ideal candidate for this work.

• carry out an industry-wide survey in order to determine the level of interest in adopting the use of massecuite Brix control.

11. Publications
The following papers resulted from this research project. Full copies are found in the Appendix.


Appendix
Evaluation of an alternative vacuum pan control scheme

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Keywords
vacuum pan, process control, Brix

Abstract
Conductivity-based vacuum pan controllers dominate the Australian sugar industry. A new scheme, based on Brix control of the massecuite and mother molasses is proposed. Multi-loop control of the process is demonstrated, whereby the massecuite Brix is controlled by manipulating the liquor/A molasses feed rate to the pan and the mother molasses Brix is controlled by manipulating the steam flow rate to the pan. A single loop controller is also proposed, which controls massecuite Brix by manipulating the feed flow to the pan. The steam flow rate is set manually (i.e. mother molasses Brix is uncontrolled). A gPROMS simulation of a pivot pan under typical operating conditions demonstrates the efficacy of this approach.

Introduction
To date the Australian raw sugar industry has clung to a simple, conductivity-based control scheme for vacuum pan operation. In this scheme, liquor is fed to the pan such that a predefined conductivity trajectory is maintained, so that a suitably ‘heavy’ massecuite results. The pan boiler chooses an appropriate set point for the steam flow rate controller, in order to set the overall process rate. While this system is inexpensive and reliable, it is not always easy to relate the measured conductivity output with process-relevant variables, since variations in conductivity may result from factors that are not linked to the crystallisation process. Examples of this would include changes in cane variety or maturity, changes in geographical region of the cane supply, dirt loading in the cane and temperature excursions during vacuum pan operation.

This paper reports on an alternative pan control scheme, which is based on process measurements that are arguably more process relevant, compared with conductivity. The first is a process refractometer, which gives a measure of mother molasses Brix, and the second is a microwave moisture device, which delivers an estimate of the massecuite Brix. Two multi-loop controllers are proposed in this scheme. One
controls the process refractometer output, by manipulating the steam flow rate to the calandria. The other controls the microwave output, by manipulating the feed flow rate to the pan. Results from a series of computer simulations are presented, which demonstrate the feasibility of the proposed method.

**Traditional vacuum pan control**

Ideally, massecuite crystal content should be controlled as high as possible, but not so high that pan circulation suffers. This ensures that the amount of exposed crystal surface per unit massecuite mass is maintained at the highest possible level, leading to enhanced net deposition rates of sucrose. Additionally, the mother molasses oversaturation should be controlled to a point just below the secondary nucleation limit, in order to sustain the highest possible crystal growth rates. Since it is currently not possible to measure directly either crystal content or oversaturation, they must be controlled inferentially.

In Australian sugar factories, pan control (whether batch or continuous) typically employs electrical conductivity probes to determine the condition of the massecuite within the pan. Adjusting the flow rate of fresh liquor/molasses to the pan controls the conductivity to its desired set point. The set point for massecuite conductivity is usually scheduled against the tonnage of massecuite in the pan. This resulting set-point trajectory attempts to anticipate (i.e. model) changes in the process, including the build-up of impurities in the mother molasses and the boil back of A molasses, which is a higher Brix/lower purity feed. This scheme is pictured below in Figure 1. Steam flow to the pan is normally controlled by feedback, using an operator-selected set point and in that sense is manually controlled. The conductivity controller maintains the massecuite at an appropriate “weight” over the duration of the strike, while the steam flow controller affects the rate at which the process proceeds.

![Fig. 1: Schematic of typical batch vacuum pan conductivity control scheme](image-url)
The advantages of this scheme are manifold. Firstly, it is relatively inexpensive in terms of capital, installation and maintenance. Secondly, the system is highly reliable, requiring little, if any, attention. Thirdly, it is effective in terms of controlling the massecuite condition, as long as skilled operators can divine suitable set point schedules. The optimality of the chosen set point trajectory can be evaluated in terms of how closely the massecuite crystal content is controlled to its optimal level. It will be assumed for the purposes of this paper, that the optimal crystal content is a constant value\(^1\).

There are disadvantages when using conductivity for pan control. Massecuite conductivity is a complex function of a range of different effects (Wright 1984), namely;

- mother molasses temperature,
- mother molasses saturation temperature,
- crystal volume fraction,
- mother molasses purity and
- mother molasses Brix.

Massecuite conductivity also depends on the type of impurities present. This effect usually manifests itself when cane of changing condition enters the mill. A skilled pan boiler will attribute the observed conductivity variations to their root cause and alter the scheduled set point, delivering consistent massecuite between rounds. Another less frequent, but much more sinister effect, occurs when factory make-up water contains significant quantities of sea salt, attributed to very low creek levels around some factories. The net result for the pan stage was that conductivity readings often went off scale and many, if not all, pans boiling liquor had to be manually controlled. This situation was, at least in the opinion of the operators, catastrophic to the operation of the pan stage.

The effect of sudden process temperature changes on conductivity also presents challenges for effective vacuum pan control. Residual heat, retained in the conductivity probe after pan steam out or during cutover of massecuite to a receiving pan, can artificially bias the conductivity signal, which is often falsely attributed to massecuite condition changes.

\(^1\) This issue remains unresolved, especially under factory, as opposed to simulated, conditions!
While conductivity-based control does work, it suffers one key weakness; the mother molasses oversaturation and crystal content variables cannot be decoupled from one another, and therefore can never be independently controlled. It is clear that better control of this process would result if these variables could be independently monitored.

It has been demonstrated (Schneider and Vigh 1997, Schneider 1998) that it is technically possible to employ state estimation schemes (i.e. Kalman filtering) to predict, and subsequently control, the oversaturation and crystal content in an industrial scale vacuum pan. However, these methods are highly complex and costly to implement and maintain. More significantly, in the case of the vacuum pan control problem, they are not robust enough to cope with the rigours of an industrial setting and show little promise of being embraced by the industry.

**Proposed vacuum pan control scheme**

An alternative vacuum pan control scheme is proposed, in which the massecuite Brix, and optionally the mother molasses Brix, in the pan are controlled by manipulating the feed and steam flow rates to the vacuum pan. While it is possible – even tempting - to present a complex, multivariable controller, simple multi-loop controllers will suffice. The proposed scheme is shown in Figure 2.

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**Simulation studies**

In order to evaluate the proposed control scheme, a series of simulations were carried out using gPROMS², which is a general process simulator that can optimise and execute parameter estimation for both steady state and dynamic processes. The simulations were based on a variation of Wright’s dynamic model (Wright 1971) of a batch vacuum pan, which assumes a point distribution for the crystal population

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² Process Systems Enterprise Ltd., Bridge Studios, 107a Hammersmith Bridge Road, London W6 9DA UK.
and an updated correlation for the solubility coefficient developed at the Sugar Research Institute (Broadfoot per. comm. 1993).

A gPROMS project was created in which an event-driven process, using a dynamic model for the pan, mimics the operation of a pivot pan. This pivot pan runs a 2\textsuperscript{nd} seed strike followed by an instantaneous cut out, which drops the pan back to 50\% of its full volume. An A strike follows, feeding on liquor to 90\% of the nominal full mass, and then boils back A molasses up to the 100\% level. Heavy-up is not presently modelled, since this is essentially an open loop (manually controlled) operation, and is independent of the type of boil-on controllers employed. During the boil-on phases, both the feed flow and steam flow to the pan are constrained to realistic values. No movement water is used. Heat transfer limitations are not included in the model used in the process simulation. Table 1 highlights the conditions used. It is important to note that, apart from the initial footing level, all initial conditions in the A strike are defined by the terminal conditions of the 2\textsuperscript{nd} Seed strike.

**Table 1: Conditions for gPROMS simulation study**

<table>
<thead>
<tr>
<th>Initial Conditions</th>
<th>2\textsuperscript{nd} Seed</th>
<th>A strike</th>
</tr>
</thead>
<tbody>
<tr>
<td>Footing, % full mass</td>
<td>50</td>
<td>50</td>
</tr>
<tr>
<td>Massecuite Brix</td>
<td>86</td>
<td>defined</td>
</tr>
<tr>
<td>Massecuite Purity</td>
<td>88</td>
<td>defined</td>
</tr>
<tr>
<td>Crystal Content, % mass</td>
<td>38</td>
<td>defined</td>
</tr>
<tr>
<td>Crystal size, mm</td>
<td>0.45</td>
<td></td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Runup Conditions</th>
<th>2\textsuperscript{nd} Seed</th>
<th>A strike</th>
</tr>
</thead>
<tbody>
<tr>
<td>Temperature, °C</td>
<td>65</td>
<td>65</td>
</tr>
<tr>
<td>Liquor Brix</td>
<td>69</td>
<td>69</td>
</tr>
<tr>
<td>Liquor Purity</td>
<td>90</td>
<td>90</td>
</tr>
<tr>
<td>Molasses Brix</td>
<td>-</td>
<td>73</td>
</tr>
<tr>
<td>Molasses Purity</td>
<td>-</td>
<td>72</td>
</tr>
<tr>
<td>Feed limits, % full mass/h</td>
<td>0-100</td>
<td>0-100</td>
</tr>
<tr>
<td>Steam limits, % full mass/h</td>
<td>0-20</td>
<td>0-20</td>
</tr>
<tr>
<td>Boilback point, % full mass</td>
<td>-</td>
<td>90</td>
</tr>
<tr>
<td>Massecuite Brix set point</td>
<td>86.5</td>
<td>86.5</td>
</tr>
<tr>
<td>Mother molasses Brix set point</td>
<td>79</td>
<td>79</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Heavy-Up Conditions</th>
<th>2\textsuperscript{nd} Seed</th>
<th>A strike</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heavy-Up point, % full mass</td>
<td>-</td>
<td>90</td>
</tr>
<tr>
<td>Steam Flow, % full mass/h</td>
<td>-</td>
<td>7.5</td>
</tr>
<tr>
<td>Massecuite Brix</td>
<td>-</td>
<td>91.5</td>
</tr>
</tbody>
</table>

Figure 3 shows output of the gPROMS simulation for the above conditions. In this case the massecuite mass, feed and evaporation rates, are shown. These profiles are a direct result of the PI controllers that maintain the massecuite Brix and mother molasses Brix to their prescribed set points.
The controlled Brix profiles in the pan are shown on Figure 4. Both Brix variables are tightly controlled to their respective set points throughout the feeding on period of the pan’s operation.

Fig. 3: Output from gPROMS dynamic simulation of a fed batch vacuum pan.

Fig. 4: Massecuite Brix and mother molasses Brix profiles under proposed control scheme

Figure 5 shows the corresponding trajectories for the massecuite crystal content and the mother molasses oversaturation and nucleation limit. Clearly these unmeasured variables are being well controlled to reasonable levels. The mother molasses oversaturation is far from the nucleation limit, so this vacuum pan would not be expected to throw a crop of false grain.
Fig. 5: Massecuite crystal content and mother molasses oversaturation and critical oversaturation in simulated Brix control scheme.

The efficacy of the proposed Brix control scheme is demonstrated by comparing the actual growth rate against the growth rate necessary to maintain constant massecuite crystal content, which according to Wright (1984), is defined as

\[
G_{\text{Required}} = \frac{1}{3} L \frac{\text{Evaporation}}{\text{Massecuite}} \frac{\text{Brix}_{\text{Feed}}}{(\text{Brix}_{\text{Massecuite}} - \text{Brix}_{\text{Feed}})}
\]

where

- \(G_{\text{Required}}\) linear growthrate of sugar (mm/h)
- \(L\) average volume equivalent diameter of crystals (mm).
- \(\text{Evaporation}\) evaporation rate (t/h)
- \(\text{Massecuite}\) massecuite mass (t)
- \(\text{Brix}_{\text{Feed}}\) feed Brix
- \(\text{Brix}_{\text{Massecuite}}\) massecuite Brix in the pan

Figure 6 shows that, apart from transient effects, due to undersaturated conditions at the beginning of the strike, and instantaneous process changes (in pan mass during cut out and changes in feed Brix, during boil back) the actual crystal growth rate superimposes that of the growth rate necessary to maintain constant massecuite crystal content in the pan. Since this criterion is being satisfied, the crystal content in the pan should be essentially constant. Re-evaluation of Figure 5 shows this is indeed the case, which indicates that accurate inferential control of crystal content is within reach, using this combination of Brix variables.
Fig. 6: Simulated and required crystal growth rate trajectories, during batch pan simulation of Brix pan control scheme.

It is clear that the proposed Brix pan control scheme should deliver improved control of fed batch vacuum pans, since the crystal content and mother molasses oversaturation are maintained at productive levels throughout the course of the batch. It is also noteworthy that the controller gain and reset times could be set to very high values, without making the closed loop process unstable. If this situation translates to industrial scale, then tuning tightly controlled massecuite Brix controllers should be straightforward.

Another issue of interest is to evaluate the case when only the massecuite Brix is controlled and the mother molasses Brix control loop is left open. In this case, the evaporation rate from the pan is set manually, by giving the steam flow controller some arbitrary set point, while the feed rate is determined by a massecuite Brix control loop, as before. Figure 7 shows the corresponding Brix profiles and Figure 8 shows the massecuite crystal content and oversaturation and critical oversaturation, when the steam flow rate is set to 15% full mass/h. Clearly a constant steam flow rate to the pan does not lead to significantly degraded performance, compared with the case where steam flow is the manipulated variable for the mother molasses Brix. In fact, in this case the boiling-on time has been reduced from 2.125 hours to 1.75 hours. This batch time reduction results from a higher mother molasses oversaturation, leading to a higher crystal growth rate, which is ultimately attributed to the higher boiling rate.
Fig. 7: Brix profiles in single loop massecuite Brix pan control simulation. Steam flow set to 15% full mass/h.

Fig. 8: Crystal content, oversaturation and critical oversaturation profiles in single loop massecuite Brix pan control simulation. Steam flow set to 15% full mass/h.

It is also of interest to note that when the mother molasses Brix is controlled to a constant value, the growth rate declines during the boilback of A molasses, due to a reduction in mother molasses oversaturation. In this case, the molasses Brix set point should be ramped upwards to account for impurities that have built-up in the pan, thereby ensuring adequate mother molasses oversaturation.

Conclusions and outlook

An alternative vacuum pan control scheme, based on the control of massecuite and mother molasses Brix, is proposed. The key benefit of this scheme is that crystal content and mother molasses oversaturation can be effectively decoupled from one another and therefore independently controlled for high purity strikes. By controlling
massecuite and mother molasses Brix to constant values, crystal content in the pan is maintained at a constant value, giving improved process control. It is also feasible to control only the massecuite Brix (using feed flow rate) and manually set the steam rate to the pan.

**References**


Improving the Control of an Industrial Sugar Crystalliser: a Dynamic Optimisation Approach

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Abstract
The controllability of an industrial sugar crystalliser is investigated. Using a simultaneous integration and optimisation approach, the optimal control problem is formulated in GAMS, based on a dynamic model of the vacuum pan validated against plant data. MINOS 2.50 solves this open-loop model, yielding: the minimum batch time, set-point trajectories and the optimal switching from high to low purity feed. These results are implemented within double-loop PI controllers. Alternative control variables are proposed, replacing traditional process outputs, which are not fundamentally process-relevant. Dynamic simulation results show that the proposed control schemes are satisfactory in the face of errors in growth rate correlations and variations in feed properties and initial batch conditions. Preliminary full-scale investigations have shown promise and will be extended into next year’s sugar cane crushing season.

1. Introduction
The number 6 fed-batch evaporative crystalliser, also known as a vacuum pan, located at CSR’s Macknade Sugar Mill (North Queensland, Australia) operates through two distinct phases, composed of many sub-steps. The existing conductivity-based control scheme, developed several decades ago, dictates the feed policy to this unit operation. Increasing production capacity, while maintaining process flexibility and quality, is of paramount importance to CSR and so they made this unit available for testing purposes. In the first phase of the batch Vu and Schneider (2001) propose two control schemes, using state-based control variables to dictate the feed and steam rates. Our investigation into the second phase of the batch is presented in this paper. The second phase of the batch is composed of three sub-steps and is more complex than the first phase, since this phase includes a switch from one feed material to another.

Previous authors (Frew, 1973 and Chew and Goh, 1989) concentrated on solving the optimal control problem, using an open-loop model. Their solutions could not handle uncertainties in initial conditions, variations in feed properties and plant-model mismatch. Their proposed optimal control solutions were never implemented.

This paper first briefly describes the process at hand. A brief overview of the solution technique is then given. The keynote is the comparison between two different control schemes to determine the feed and evaporation rate policies of the fed-batch process. Finally closed-loop responses within vacuum pan dynamic simulations and preliminary implementation data from the factory are presented for discussion.
2. Process Description

A schematic of a vacuum pan is presented below in Figure 1, showing the steam and various feed inlets. The starting and full levels of the vessel are indicated.

![Schematic of a Vacuum Pan](image)

**Figure 1: Schematic of a Vacuum Pan**

The second phase of the crystallisation process includes three main sub-steps.

1. **Foundation**: steam heating is resumed with approximately 50 tonnes of footing material, known as massecuite ($M$), which is composed of sugar crystals ($x_4$) suspended in mother molasses ($Mol$). Molasses ($Mol$) is a solution comprising dissolved sucrose ($x_1$), dissolved non-sucrose impurities ($x_2$) and water ($x_3$). Liquor, a high purity feed, is simultaneously introduced to enrich the mother molasses, depleted due to sugar crystallisation.

2. **Boilback**: liquor feed is switched to A-molasses, a lower purity feed material, until the pan reaches its full volume.

3. **Heavy-up**: the pan is full. Feeding is stopped, but steam continues, leading to exhaustion of the mother molasses. After this step the massecuite is centrifuged, in order to recover the product sugar crystal.

Rigorously determining the schedules for feed and steam rates, as well as the feed switching point, requires a more sophisticated solution method than is presently available to industrialists. The essential steps leading to batch vacuum pan optimisation are described next.

3. Methods

The following sections present the problem formulation, including discussion on the constraints acting upon this system, followed by a brief description of the implementation.
3.1 Problem formulation

The process dynamic model of the vacuum pan can be fully described by mass, energy and population balances. However, the dynamics of the crystal size distribution are simplified, while the energy balance is written in the simplest form possible. In other words the evaporation rate \( E \) is assumed to be proportional to the steam rate to the vessel. The mass balances of water, impurities and sucrose in the pan are straightforward. Readers should consult Vu and Schneider (2000, 2001) for more details.

Three main constraints involve the crystal content \( CC \), the solution oversaturation \( OS \) and the target purity of the mother liquor at the end of Heavy-up.

The crystal content is defined as the mass fraction of crystal in the vessel, according to

\[
CC = \frac{x_4}{M} \leq 0.55
\]  

(1)

The solution sucrose oversaturation \( OS \) is the driving force for crystallisation. The higher the \( OS \), the faster the crystal population will grow. However, the \( OS \) has an upper limit, termed the critical oversaturation or \( OS_{crit} \). Beyond this point, nucleation of new unwanted crystals occurs, resulting in downstream processing inefficiencies. This critical level has been previously defined by (Broadfoot and Wright, 1972). At all times the fractional oversaturation, known as \( FOS \), should be kept below unity.

\[
FOS = \frac{OS}{OS_{crit}} \leq 1
\]  

(2)

The last constraint determines the switching point from fresh liquor to A-molasses feed in order to achieve the desired final molasses purity, or \( Pty \left( = \frac{x_1}{x_1 + x_3} \right) \), defined as mass ratio of sucrose to total dissolved solids in the molasses.

\[
Pty \leq 0.75 \quad \text{at Heavy-up}
\]  

(3)

The upper limits of massecuite, crystal content and purity also serve as termination conditions for the second phase. In other words, when \( M \) reaches 100 tonnes and \( Pty \) drops to 0.75, feeding is stopped, but steam continues until the crystal content reaches 0.55.

Challenges facing the batch pan problem include uncertainties in feed properties and initial conditions, such as purity. \( Brix = \frac{x_1 + x_2 + x_4}{M} \), \( CC \), footing seed diameter, etc. Due to upstream process fluctuations, the ranges of these variations can be large at the beginning of the batch. However, during the second phase, these variations become less important. After Boilback, the crystal content increases to a high level, reducing the natural circulation in the vessel and, consequently, the rate of heat transfer. This loss of circulation and the subsequent reduction in heat transfer are not presently modelled, which must be taken into account for factory implementations.

The optimal control problem formulated for the second phase is based on the nominal conditions. The objective function, \( OF \), presented in (4) contains two different targets: minimum batch operating time and optimum trajectories of \( M, CC \) or \( FOS \) set points and, importantly, the switching point from the Foundation to Boilback sub-steps.
This type of dynamic optimisation problem can be solved using a simultaneous integration and optimisation technique, which approximates the state variables by interpolation polynomials (Vu and Schneider, 2000, 2001). These are differentiated and then back-substituted into the state equations, converting them to a set of algebraic equations. The scaling factors \( w_m, w_c \) or \( w_f \) in the above equation force \( M, CC \) or \( FOS \) to follow the desired optimum trajectories of \( M_{sp}, CC_{sp} \) or \( FOS_{sp} \). The \( time \) variable in (4) is actually replaced by the summation of the element lengths. It is important to note that the number of finite elements \( nfe \) during Foundation, Boilback and Heavy-up are specified, but their lengths are decision variables in the optimisation problem. Thus the optimal switching points from one sub-step to another can be determined.

### 3.2 Implementation

Instead of using feed and steam flow profiles, set-point trajectories will be implemented within two PI controllers. Two different control schemes are proposed. Both schemes select the mass-controller as a primary loop because mass can be directly and accurately measured and it is one of the ending conditions stated above. Another advantage is that the mass set-point trajectory will not be affected by the presence of errors in the growth rate expressions. This dominant loop controls the mass in the pan by adjusting the feed rate.

**First Scheme: M-F/CC-Steam** - The second loop controls \( CC \) by adjusting steam. \( CC \) cannot directly be measured, but can be estimated using either a state estimation scheme or by combining other process outputs. Since \( CC \) is the second ending condition, this loop is simpler to be implemented. The greatest disadvantage of the first scheme is that the \( FOS \) remains uncontrolled. This might result in constraint (2) violation due to variations in feed properties and initial conditions. The gains and reset times of these two controllers therefore must be tuned against a worst-case scenario, requiring an iterative dual-level optimisation, as discussed in Vu and Schneider (2001).

**Second Scheme: M-F/FOS-Steam** - To avoid these tuning problems, the second loop could control \( FOS \) by adjusting the steam flow rate. The gains should be set at the highest values, without leading to oscillations in the controlled or manipulated variables. Therefore they must be tuned against the fastest situation and tested with other cases. At high values of \( CC, FOS \) control is safer and more robust but this loop pairing also has some disadvantages. First, the secondary loop in this case must include a logical operation to stop the batch, using the ending condition of \( CC \). Second, \( FOS \) is derived from many other correlations, which might bring in a large error in \( FOS \) determinations.

Dynamic simulations obtained from two control schemes and preliminary implementation results are compared in the following section.

\[
OF = \min \left\{ w_i \text{time} + w_m \sum_{i=1}^{nfe} \left[ M(i) - M_{sp}(i) \right]^2 + w_c \sum_{i=1}^{nfe} \left[ CC(i) - CC_{sp}(i) \right]^2 + w_f \sum_{i=1}^{nfe} \left[ FOS(i) - FOS_{sp}(i) \right]^2 \right\}
\]
4. Results

The scheme $M$-Feed/$FOS$-Steam was selected as the option to test under factory conditions during the first phase of operation. Figure 2 shows $FOS$, $CC$ and $M$ profiles in the pan. Solid curves representing factory implementation are plotted against broken lines for simulation. It should be noted that the mass variable, $M$, was indirectly controlled, but it nonetheless follows the path prescribed by the optimisation results. This factory trial was necessarily run very conservatively, so that plant personnel would not abandon the experiment. As such, the set point for $FOS$ was kept quite low, given the anticipated plant-model mismatch. Future experiments at the factory will determine how much higher the $FOS$ set point can be, before the system suffers.

Dynamic simulation results using both control scheme options applied to the second phase of operation are shown in Figure 3: $CC$-Steam in the left and $FOS$-Steam on the right, showing the controlled ($M$, $CC$, $FOS$) and manipulated variables (Feed and Steam). In terms of batch time reductions, the $FOS$-Steam loop should be used over that of the $CC$-Steam, since it maintains a higher level of sucrose oversaturation throughout the batch, and therefore sustains crystal growth at a higher rate.

5. Conclusions

The controllability of an industrial sugar crystalliser has been investigated through the second phase of operation in a similar fashion used in the first phase investigation. New controlled variables: $M$, $CC$ and $FOS$ can replace the traditional output-based conductivity control. A double-loop control strategy is also proposed to replace the existing single loop controller. Based on the preliminary implementation, results from the factory indicate that the loop pairing of $M$-Feed/$FOS$-Steam functions well in the second batch phase. Future work will evaluate alternative implementations of both proposed control schemes across both phases of this fed-batch crystallisation.

6. References


7. Acknowledgements

Both authors would like to acknowledge CSR Ltd for their strong and enthusiastic support of this project. Dr Vu would like to acknowledge the SRDC for funding under their CP2002 project program.
Figure 2: FOS, CC and M Profiles During Phase One of the Batch (M-F/FOS-Steam)
(Solid line: implementation; broken line: simulation)

Figure 3: Dynamic Responses Using Different Control Schemes in Phase Two of the Batch.
INTRODUCTION
Over the past few years, the number 6 evaporative vacuum pan, at CSR’s Macknade Sugar Mill (North Queensland) has been upgraded with advanced process control software and sensors. Within this vessel raw sugar is crystallised in a fed-batch mode. At present a simple conductivity-based control scheme, developed several decades ago, controls the vacuum pan process. The output conductivity, which is not fundamentally process relevant, dictates the feed policy through a set point that is scheduled against the pan tonnage. Experienced operators set the parameters that describe the set-point profile. Steam flow to the calandria heating section is manually supervised and controlled. Recently pan stage efficiency and process reliability have been reduced, given the loss of skilled process operators. Production has been unreliable because the process cannot be exactly monitored, predicted and controlled.

This paper focuses on the vacuum pan automation as a means to increase production volume, improve product qualities and maintain production flexibility. The proposed control schemes, which manipulate the feed and steam rates, should produce minimal operating times, while satisfying path constraints, regardless of variations in feed and initial conditions in the batch. The selection of loop-pairings is vital to eliminate interactions, but more importantly the control scheme must be reliable and simple enough to be implemented industrially. The tuning parameters and set-point trajectories therefore should be obtained within an optimisation framework. The mathematical dynamic model of the vacuum pan was recently redefined in the previous paper of Vu and Schneider [1]. Some model equations and definitions related to the discussed topic and the technique used to solve the optimal control problem for optimal batch operation time will be reviewed. The implementation of the result found by means of two proposed control strategies will be discussed in detail afterwards.

PROBLEM FORMULATION
A glossary of some common terms used in the sugar industry and a schematic of a 100 tonne vacuum pan are presented first.

- **Brix** The mass fraction of sucrose, impurities and sugar crystals in massecuite.
- **Crystal content** The mass fraction of sugar crystals in massecuite.
- **Dry substance** The mass fraction of sucrose and impurities in molasses.
- **Massecuite** The mixture of sugar crystals, impurities and water in the vacuum pan.
- **Molasses** The mother liquor obtained from the massecuite by centrifugation.
- **Purity** The mass ratio of sugar crystals over impurities and sugar crystals.
Pan 6 at Macknade Mill normally operates through nine batch phases.

1. **Cut-in**
   A solution of sucrose, impurities, water and accompanying seed crystals are drawn into the pan to approximately 50 tonnes.

2. **Run-up**
   The mixture is heated by steam while liquor, a high purity feed is added to the vacuum pan to approximately 100 tonnes, and then heating is paused.

3. **Cut-out**
   Half of the contents are sent to another, waiting pan.

4. **Foundation**
   Heating is resumed with approximately 50 tonnes of massecuite in the pan. Liquor is added to enrich the depleted mother molasses.

5. **Boilback**
   A molasses recycle is fed to the pan until it reaches its full mass of 100 tonnes.

6. **Heavy-up**
   The pan is full. Steam is still admitted and limited crystal growth occurs.

7. **Break-vacuum**
   The pressure in the calandria is brought up to ambient pressure.

8. **Discharge-pan**
   All contents are dropped into receivers below the pan.

9. **Raise-vacuum**
   Vacuum is raised in the pan to start a new batch.

Phases from Foundation to Heavy-up (four to six) will be included in the Foundation problem. Run-up is treated as an independent problem in this study. Other phases are not considered. Hold water usage through these phases is optional and will be discussed later within each problem. The same dynamic model fully described by the energy, mass and population balances is applied for both Run-up and Foundation problems but their initial conditions and the variations of these conditions are different. Table 1 shows the variations in feed and initial conditions of the Run-up problem.

**Table 1: Variations in Feed and Initial Conditions of The Run-up Problem**

<table>
<thead>
<tr>
<th>Initial Conditions</th>
<th>Lower Limit</th>
<th>Nominal Value</th>
<th>Upper Limit</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Massecuite, Mo (tonne)</strong></td>
<td>40.0</td>
<td>50.0</td>
<td>60.0</td>
</tr>
<tr>
<td><strong>Purity, Ptyo</strong></td>
<td>0.800</td>
<td>0.835</td>
<td>0.870</td>
</tr>
<tr>
<td><strong>Crystal Content, CCo</strong></td>
<td>0.230</td>
<td>0.280</td>
<td>0.330</td>
</tr>
<tr>
<td><strong>Brix, Brixo</strong></td>
<td>0.825</td>
<td>0.845</td>
<td>0.865</td>
</tr>
<tr>
<td><strong>Seed Diameter, dimo (mm)</strong></td>
<td>0.400</td>
<td>0.500</td>
<td>0.600</td>
</tr>
<tr>
<td><strong>Dry Substance, Dryf</strong></td>
<td>0.658</td>
<td>0.680</td>
<td>0.702</td>
</tr>
<tr>
<td><strong>Purity, Pryf</strong></td>
<td>0.899</td>
<td>0.915</td>
<td>0.916</td>
</tr>
<tr>
<td><strong>Steam Factor, stcof</strong></td>
<td>0.900</td>
<td>0.960</td>
<td>1.000</td>
</tr>
</tbody>
</table>

The following equation shows the relationship of Brix, dry substance and crystal content.
\[ D_{\text{Dry}}(1 - CC) = Brix - CC \]  
(1)

The optimal control problem formulated for the Run-up phase is based on nominal values shown in Table 1. The feed purity is assumed constant since its variation is expected to be small. If the crystal content is not high, the evaporation rate is assumed to be proportional to the steam rate. The variation of this ratio (stcof) is found in the above table.

The mass balances and constraints related to the discussion are reviewed.

Mass of water in the pan
\[ \frac{dx_1}{dt} = F(1 - D_{\text{Dry}_{f}}) + W_{h}(1 - D_{\text{Dry}_{h}}) - E \]  
(2)

Mass of impurities
\[ \frac{dx_2}{dt} = F \cdot D_{\text{Dry}_{f}}(1 - P_{\text{ty}_{f}}) + W_{h} \cdot D_{\text{Dry}_{h}}(1 - P_{\text{ty}_{h}}) \]  
(3)

Mass of sucrose
\[ \frac{dx_3}{dt} = F \cdot D_{\text{Dry}_{f}} \cdot P_{\text{ty}_{f}} + W_{h} \cdot D_{\text{Dry}_{h}} \cdot P_{\text{ty}_{h}} - \frac{dx_1}{dt} \]  
(4)

Mass of crystals
\[ \frac{dx_4}{dt} = \frac{? pNGx_2^2}{2} \]  
(5)

Average diameter of crystals
\[ \frac{dx_5}{dt} = G \]  
(6)

G is the linear crystal growth rate. Details of this empirically derived expression can be found in Vu and Schneider [1].

The mass of massecuite is always within limits. The upper limit is one ending condition of the batch.
\[ 40 \leq M \leq 100 \text{(tonne)} \]  
(7)

The crystal content has an upper limit set at 0.37, which is another ending condition of the batch.
\[ CC = \frac{x_4}{M} \leq 0.37 \]  
(8)

The main driving force for crystallisation is the concentration of sucrose in excess of saturation, defined as the solution sucrose oversaturation (OS). The higher the OS, the faster the crystal population will grow. The OS has an upper limit, termed the critical oversaturation (OS_{crit}). Beyond this limit, nucleation of new, unwanted crystals occurs, resulting in downstream processing inefficiencies. Therefore, the operating policy requires that the OS be kept at its highest possible value, without exceeding the OS_{crit}. The oversaturation fraction (FOS) is defined as
\[ FOS = \frac{OS}{OS_{crit}} \leq 1 \]  
(9)

The constraints showing the physical limitations of the pan include steam and liquor feed supply rates.
\[ \text{steam} \leq 40 \text{(ton.hr\textsuperscript{-1})} \]  
(10)

\[ F \leq 100 \text{(ton.hr\textsuperscript{-1})} \]  
(11)

**SOLUTION METHOD**

The first target of the Run-up problem is the minimum batch operating time, described by the following objective function.
\[ OF_1 = \min \left\{ w_{,\text{time}} \sum_{i=1}^{i=\text{inlet}} M(i) - M_{\text{sp}}(i) + w_{,\text{cc}} \sum_{i=1}^{i=\text{inlet}} [CC(i) - CC_{sp}(i)]^2 + w_{,\text{fos}} \sum_{i=1}^{i=\text{inlet}} [FOS(i) - FOS_{sp}(i)]^2 \right\} \]  
(12)

As discussed in our previous paper [1] the state variables are approximated by the Lagrange interpolation polynomials. These are differentiated and back substituted into the state equations, converting them to a set of algebraic equations. The optimal control problem now becomes a normal nonlinear programming
problem and can be solved by existing optimisation techniques. The approximation approach used is Orthogonal Collocation on Finite Elements, developed by Biegler and Cutrell [2]. The term \( t \) in equation (12) is actually replaced by the summation of element lengths, which are decision variables in the optimisation problem.

The scaling factors, \( w_t, w_c, w_f \), play an important role in the optimisation problem. If all scaling factors, except \( w_t \), are set at zero, solving the optimal control problem based on the open-loop model gives the least possible batch operating time \( (t_{\text{min}}) \). However, the corresponding steam and feed profiles cannot be implemented with the existing control system at Macknade Mill, due to equipment constraints.

Some set-point trajectories are required for the automation of the pan, since it is a batch process. Equation (12) shows three set-point variables, \( M_{\text{sp}}, CC_{\text{sp}} \) and \( FOS_{\text{sp}} \), which can be approximated by one or more \( n^{th} \) order polynomials. The coefficients of these polynomials are also decision variables in the same optimisation problem using equation (12). The scaling factor values in this case must be different from zero and small enough so that \( t \) remains close to \( t_{\text{min}} \).

Instead of using feed and steam flow profiles, the set-point trajectories will be implemented to the pan within two PI control loops. The first loop controls the mass in the pan by adjusting the liquor feed.

\[
F = F_o + gm \left( erm + \frac{erm}{rstm} \right)
\]

In equation (13) \( gm \) and \( rstm \) are respectively gain and reset time of the PI controller. The term \( erm \) is the deviation of the mass in the pan from its set point.

\[
erm = M - M_{\text{sp}}
\]

The second controller will control either the \( CC \) or \( FOS \) by adjusting steam.

\[
\text{steam} = \text{steam}_o + gc \left( erc + \frac{erc}{rstc} \right)
\]

or

\[
\text{steam} = \text{steam}_o + gf \left( erfos + \frac{erfos}{rstf} \right)
\]

Similarly, in equation (15) or equation (16), \( gc \) or \( gf \) and \( rstc \) or \( rstf \) are respectively gain and reset time of the PI controller. The terms \( erc \) and \( erfos \) are given by

\[
erc = CC - CC_{\text{sp}}
\]

\[
erfos = FOS - FOS_{\text{sp}}
\]

The above control problem is a mixture of servo and regulator problems, in which disturbances and set points undergo change. If considering only the nominal case, controller gains can be set at arbitrarily high values, until process oscillations set in. However, these values may cause constraint violations due to variations in feed and initial conditions shown in Table 1. The controller tuning parameters must therefore be selected to cope with a worst-case scenario. The sub-level (or inner-level) of an iterative dual-level optimisation developed by Bahri \textit{et al.} [3] is applied to find the worst case, which produces the maximum constraint violation. The objective function of the inner-level optimisation is either

\[
OF_{i1} = \max \left[ \max \left[ \max \left[ FOS(i) - 1 \right] \right] \right]
\]

or

\[
OF_{i2} = \max \left[ \max \left[ \max \left[ CC(i) - 0.37 \right] \right] \right]
\]

depending on which controller is employed. Variables shown in Table 1 are decision variables in either case.

The objective function of the outer-level optimisation problem is the least integral squared deviation of the controlled variables from their set points, as indicated in equation (12). In this case the decision variables are the tuning parameters of the controllers. In general the size of the outer-level problem after the second iteration becomes very large, which can only be handled by the sequential integration and optimisation approach. This technique requires integrating the ordinary differential equations for all iterations. Because of this fact, most of the computational time is spent on the integration of the system equations, which provides gradient information.

The following section shows the simulation responses and comparison of the proposed control schemes.
RESULT AND DISCUSSION

Solving the optimal control problem based on the open-loop model gives the following results.

1. Minimum batch operating time (other scaling factors except \( w_t \) are set at zero), \( t_{\text{min}} = 38 \) mins.

2. The \( M_{sp} \) and \( FOS_{sp} \) trajectories are approximated by a first order polynomial. The \( CC_{sp} \) trajectory is approximated by two first order polynomials. To eliminate the time dependency of \( CC \) or \( FOS \), the \( CC_{sp} \) or \( FOS_{sp} \) is linearly related to the \( M_{sp} \), instead of time. The scaling factors are set at \( w_t = 45 \), \( w_m = 6e^{-6} \), \( w_c = 1 \) and \( w_f = 6e^{-2} \), and the minimum batch time remains at 38 minutes.

3. It is not necessary to use hold water \( (W_h = 0) \) as an additional feed during the Run-up phase. Water must be added at the beginning of the phase before starting the controllers, if \( FOS \) is greater than 1. Similarly, if \( FOS \) is less than 0 at the beginning, steam must be added before starting the feed.

The mass-controller is selected as a primary loop because mass can directly and accurately be measured and is one of the two ending conditions of the batch. Since \( CC \) is the second ending condition, it is more advantageous to select the \( CC \)-controller as the second loop. This variable cannot directly be measured but can be calculated from two measurable variables \( Brix \) and \( Dry \), using equation (1). The secondary loop has set points dependent on set points of the primary loop. That means \( CC \) variation depends on the tonnage of massecuite in the pan instead of time. As \( FOS \) is not controlled, constraint (9) posed on \( FOS \) might be violated due to variations in feed and initial conditions. The inner-level optimisation must be performed to find the worst case and the outer-level optimisation tunes the controller gains against this worst case. The tuning parameters are set at \( g_m = -0.31 \), \( rstm = 1 \), \( gc = -3.6 \), \( rstc = 10 \). Table 2 presents some cases, which might occur. Figure 2 shows the sensitivity of \( FOS \) to gains using \( M-CC \) loops. If the value of \( gc \) is increased from 3.6 to 3.8, \( FOS \) is violated.

### Table 2: Batch Operating Time Using CC-Controller

<table>
<thead>
<tr>
<th>Mo (ton)</th>
<th>Ptyo</th>
<th>CCo</th>
<th>Brixo</th>
<th>diamo (mm)</th>
<th>Dryf</th>
<th>stcof</th>
<th>Time (min)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Nominal Case</strong></td>
<td>50</td>
<td>0.835</td>
<td>0.28</td>
<td>0.845</td>
<td>0.5</td>
<td>0.680</td>
<td>0.96</td>
</tr>
<tr>
<td><strong>Worst Case</strong></td>
<td>40</td>
<td>0.800</td>
<td>0.23</td>
<td>0.825</td>
<td>0.6</td>
<td>0.702</td>
<td>1.00</td>
</tr>
<tr>
<td><strong>Least ISE</strong></td>
<td>60</td>
<td>0.870</td>
<td>0.24</td>
<td>0.855</td>
<td>0.4</td>
<td>0.702</td>
<td>1.00</td>
</tr>
<tr>
<td><strong>Fastest Batch Time</strong></td>
<td>60</td>
<td>0.800</td>
<td>0.23</td>
<td>0.865</td>
<td>0.6</td>
<td>0.702</td>
<td>1.00</td>
</tr>
<tr>
<td><strong>Slowest Batch Time</strong></td>
<td>40</td>
<td>0.870</td>
<td>0.33</td>
<td>0.825</td>
<td>0.4</td>
<td>0.658</td>
<td>0.90</td>
</tr>
</tbody>
</table>

The major advantage in using the \( FOS \)-control loop is that it can handle variations in feed and initial conditions of the batch. The controller gains can be set at the highest values, which do not lead to oscillations in the controlled or manipulated variables. Therefore, the gains are tuned against the fastest case and tested for other cases. Figure 3 shows the sensitivity of the manipulated variable steam to gains using \( M-FOS \) loops. If the value of \( gf \) is increased from 0.8 to 0.9, steam oscillates significantly.

However, the \( FOS \) loop also has some disadvantages, which should be considered. First, \( FOS \) is controlled but is not an ending condition of the batch. The secondary loop in this case must include a logical operation to stop the batch, using the ending condition of \( CC \). This is not a straightforward action to be implemented. Second, \( FOS \) is derived from many correlations, which might bring in a large error in \( FOS \) calculation. Table 3 presents batch operation time of the control loops, using \( g_m = -10 \), \( rstm = 1 \), \( gf = -0.8 \), \( rstf = 10 \).

### Table 3: Batch Operating Time Using FOS-Controller

<table>
<thead>
<tr>
<th>Mo (ton)</th>
<th>Ptyo</th>
<th>CCo</th>
<th>Brixo</th>
<th>diamo (mm)</th>
<th>Dryf</th>
<th>stcof</th>
<th>Time (min)</th>
</tr>
</thead>
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<td>0.658</td>
<td>0.90</td>
</tr>
</tbody>
</table>

Tables 2 and 3 show that due to variations in feed and initial conditions both nominal batch operating times of \( M-FOS \) and \( M-CC \) loops are greater than \( t_{\text{min}} \). However, the time variation from the slowest to the fastest batch, using the first scheme is more than an hour, two-fold increase over the nominal case. This is also considered as a drawback of the \( M-FOS \) loop.
Figure 2: Sensitivity of FOS to Controller Gains Using M-CC Loops

Figure 3: Sensitivity of Steam to Controller Gains Using M-FOS Loops
CONCLUSION
The controllability of an industrial crystalliser has been analysed for the Run-up phase of its operation. Using a simultaneous integration and optimisation technique, the optimal control problem based on the open-loop model is solved for \( t_{\min} \) and profiles of \( M_{sp} \), \( CC_{sp} \) and \( FOS_{sp} \) result. These set-point trajectories are then implemented within a vacuum pan simulation with two control loops, which manipulate the feed and steam policies. Using a sequential technique, the dual-level optimisation problem is solved for the tuning parameters against the worst-case situation. Two control schemes were proposed. The mass controller is selected as the primary loop, to monitor the feed flow. The secondary can be either the CC or FOS controller to monitor the steam flow. The \( M-FOS \) loops can handle variations in feed and initial conditions but some other logic conditions are required to halt the batch. Another drawback is a possibly greater error in \( FOS \) calculation, due to errors in the correlations describing it. Moreover the variation of batch time, from the slowest to the fastest batch is large, twofold greater than using the \( M-CC \) loops. The latter however must be tuned against the worst case to cope with disturbances and model uncertainties. Future work will evaluate a study of the Foundation phase optimisation in a similar fashion. However, this problem becomes even more complicated with the requirement of the Boilback phase, which switching from liquor to molasses feed. That complication will be addressed within the Foundation problem in the next paper.

ACKNOWLEDGEMENTS
The work presented in this paper is the result of a project funded by the Sugar Research and Development Corporation. The authors would like to thank the technical staff at Macknade Mill and Victoria Mill for their support.

NOMENCLATURE
Symbols

- **Brix** (dimensionless) defined as \((x_2 + x_3 + x_4)/(x_1 + x_2 + x_3 + x_4)\)
- **CC** crystal content (dimensionless) defined as \((x_4)/(x_1 + x_2 + x_3 + x_4)\)
- **Dry** dry substance (dimensionless) defined as \((x_2 + x_3)/(x_1 + x_2 + x_3)\)
- **E** evaporation rate (tonne.h\(^{-1}\)).
- **er** error
- **F** feed (tonne.h\(^{-1}\)). Liquor during Foundation and A molasses during Boilback
- **FOS** oversaturation fraction (dimensionless)
- **g** Gain
- **G** growth rate (mm.h\(^{-1}\))
- **ISE** integral of the square of the error
- **M** mass or massecuite (tonne.h\(^{-1}\)) defined as \((x_1 + x_2 + x_3 + x_4)\)
- **OF** objective function
- **OS** solution sucrose oversaturation (°OS)
- **Pty** purity (dimensionless) defined as \((x_3)/(x_2 + x_3)\)
- **rst** reset time
- **stcof** ratio of evaporation rate over steam (dimensionless)
$x_1$ mass of water in the pan (tonne.h$^{-1}$)

$x_2$ mass of impurities in the pan (tonne.h$^{-1}$)

$x_3$ mass of sucrose in the pan (tonne.h$^{-1}$)

$x_4$ mass of crystals in the pan (tonne.h$^{-1}$)

$x_5$, diam volume equivalent average diameter of crystals in the pan (mm)

w scaling factor

W flowrate of water (tonne.h$^{-1}$)

Subscripts

crit critical condition

f feed

frac fractional

h movement water

o initial

sp set point

REFERENCES


Figure 4a: Mass, Feed and Steam Profiles of the M-CC Control Loops

Figure 4b: Crystal Content and Fractional Oversaturation Profiles of the M-CC Control Loops
Figure 5a: Mass, Feed and Steam Profiles of the M-FOS Control Loops

Figure 5b: Crystal Content and Fractional Oversaturation Profiles of the M-FOS Control Loops
Operability Analysis of an Industrial Crystallizer –
An Optimal Control Approach

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ABSTRACT
In the first part of the operability analysis, the process dynamic model of a batch raw sugar vacuum pan is reviewed and redefined in a more practical and direct manner. Using the simultaneous integration and optimization technique, the optimal control scheme is solved by minimizing the time required for the crystals to reach the target average size. This solution technique determines the minimum operating time as well as the duration of each stage of the crystallization process. Due to the uncertain size distribution of the crystals, back-off feed and steam policies are proposed. A correlation for the overall heat transfer coefficient is then incorporated into the model to account for heat transfer limitations on the evaporation rate from the pan. This results in increasing the duration of the heavy-up stage, during which the content of the pan becomes highly viscous and the evaporation rate rapidly reduces.

1. INTRODUCTION
Over the last several decades, the Australian Sugar Industry has clung to a simple, conductivity-based control scheme for the operation of evaporative crystallizers, also known as vacuum pans, which produce raw sugar. Raw sugar quality depends greatly on experienced operators since the output conductivity variable is not fundamentally process relevant. As a result, the pan stage efficiency has gone down with the loss of skilled workers and with increasing throughput requirements. To start the series of crystallizer studies, in this paper the mathematical dynamic model of the vacuum pan is redefined. With the assumption that the evaporation rate is proportional to the steam rate, the optimal control problem is solved for the optimal feed and steam policies and the trajectories of two set points: the impure sucrose oversaturation (OS) and the crystal content (CC) by minimizing the operating time. The solution found assists as a guideline to understand the actions of experienced operators in the plant. However, this first set of results cannot represent the real performance of the vacuum pan, which is mainly heat transfer limited. The process dynamic model should therefore be augmented with some forms of correlations for the overall heat transfer coefficient across the heating section.

Frew [1] have formulated and solved the optimal control problem based on a mathematical model of sugar crystallization developed to represent practice in the Queensland sugar industry. However, the minimum time to operate a batch was not directly solved in that problem and another equivalent problem of maximizing the final crystal size in a fixed terminal time was considered instead. Moreover path constraints were not directly handled but some penalties must be included in the objective functions. Sixteen years later, Chew and Goh [2] addressed the same problem of Ritch [3] as a continuous model. The time control problem was then directly solved by a straightforward method namely Control Vector Parameterization via a FORTRAN program. In that approach the control profiles were approximated by a series of piecewise constant functions and the gradients are computed in a roundabout fashion.

To avoid the requirement of solving the Ordinary Differential Equations (ODEs), which is time consuming, in this paper the simultaneous integration and optimization technique is used to approach a similar problem of Frew. A brief description of the process and the model of the batch are presented in the following section. Readers should consult Frew and Ritch for more details on crystallization stages and terms used in the sugar industry.

1 Author to whom correspondence should be addressed
2. **Problem Formulation**

A normal batch consists of five stages.

1. **Initiation** The batch begins after a solution of sucrose, impurities, water and accompanying seed crystals are drawn into the pan to approximately 45% of its maximum volume. The mixture is heated by steam and water is boiled off until the molasses surrounding the crystals reaches the desired level of supersaturation.

2. **Foundation** Liquor, a high purity feed is added to the vacuum pan to approximately 90% of the full volume of the pan.

3. **Boilback** Molasses recycle is fed, until the pan reaches its full volume.

4. **Heavy up** The pan is full. Steam and, perhaps hold water, are still admitted and limited crystal growth occurs. The main goal of this stage is molasses exhaustion, allowing it to reach its target purity.

5. **Pan drop** At the end of the batch, the contents are dropped into receivers below the pan. Depending on the conditions of the material crystallisation may continue, although this feature was not considered in this study.

During Foundation and Boilback, *movement water*\(^2\) is also fed to the pan. While the compositions of sucrose, impurities and water in fresh liquor, molasses and hold water vary from plant to plant, typical compositions are shown in Table 1.

<table>
<thead>
<tr>
<th>Table 1: Typical Compositions of Feed Materials to the Vacuum Pan</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Liquor</strong></td>
</tr>
<tr>
<td>Purity (Pty)</td>
</tr>
<tr>
<td>Dry Substance (Dry)</td>
</tr>
</tbody>
</table>

The process dynamic model is fully described by the energy, mass and population balances. However the dynamic of the crystal size distribution is neglected and the energy balance will be included later.

Mass of water in the pan

\[
\frac{dx_1}{dt} = F \left( 1 - Dry_f \right) + W_h \left( 1 - Dry_h \right) - E \tag{1}
\]

Mass of impurities

\[
\frac{dx_2}{dt} = F \cdot Dry_f \left( 1 - Pty \right) + W_h \cdot Dry_h \left( 1 - Pty_h \right) \tag{2}
\]

Mass of sucrose

\[
\frac{dx_3}{dt} = F \cdot Dry_f \cdot Pty_f + W_h \cdot Dry_h \cdot Pty_h - \frac{dx_4}{dt} \tag{3}
\]

Mass of crystal

\[
\frac{dx_4}{dt} = \frac{\partial \Omega}{\partial X} \frac{x_5^2}{2} \tag{4}
\]

Average diameter of crystal

\[
\frac{dx_5}{dt} = G \tag{5}
\]

\(^G\) is the linear crystal growth rate, an empirical expression determined by Wright [4].

---

\(^2\) so-called, because the diluting effect is said to improve boiling circulation in the vessel. More recently this movement *water* has been replaced by the dilute evaporator supply juice. The term *water* has been retained for the sake of posterity.
Typical values of parameters and constant used in this study can be found in Table 2

Table 2: Typical Values of Constants and Parameters used in the Study

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>$W_h$</td>
<td>0.9 tonne.h$^{-1}$</td>
</tr>
<tr>
<td>$N$</td>
<td>$1.2 \times 10^{11}$</td>
</tr>
<tr>
<td>$P_1$</td>
<td>7.418 mm.(h.OS)$^{-1}$</td>
</tr>
<tr>
<td>$P_2$</td>
<td>0.04 OS</td>
</tr>
<tr>
<td>$P_3$</td>
<td>-1.75</td>
</tr>
<tr>
<td>$P_5$</td>
<td>0.0409</td>
</tr>
<tr>
<td>$P_6$</td>
<td>0.8065</td>
</tr>
<tr>
<td>$E_{ac}$/R</td>
<td>$3.06 \times 10^3$ K</td>
</tr>
<tr>
<td>$P_7$</td>
<td>0.8585</td>
</tr>
</tbody>
</table>

Constraints acting on the system include:

- The mass of massecuite is always within limits.
  
  \[50 \leq x_1 + x_2 + x_3 + x_4 \leq 100 \text{ (tonne)}\]  
  
- The mass fraction of crystal in the massecuite, $CC$, has an upper limit at which circulation will suffer.
  
  \[CC = \frac{x_4}{x_1 + x_2 + x_3 + x_4} \leq 0.55\]  
  
- The main driving force for crystallization is the concentration of sucrose in excess of saturation, defined as the solution sucrose oversaturation ($OS$). The higher the $OS$, the faster the crystal population will grow. The $OS$ also has an upper limit, termed the critical oversaturation ($OS_{crit}$). Beyond this limit, nucleation of new, unwanted crystal occurs, resulting in downstream processing inefficiencies. Therefore, the operating policy requires that the $OS$ be kept at its highest possible value, without exceeding the $OS_{crit}$. The oversaturation fraction ($OS_{frac}$) is defined as

  \[OS_{frac} = \frac{OS}{OS_{crit}} \leq 1\]  

  The correlations of Broadfoot and Wright [5] devised for the oversaturation and the nucleation threshold are used in this work.

- Switching the feed from the fresh liquor during Foundation to A molasses during Boilback will achieve the target purity in the pan after heavy up. This can avoid the problem of purity rise during the centrifuging stage. The following constraint determines the switching point from fresh liquor to A molasses feeding.

  \[P_{ty} \text{ at Heavy up} \leq 0.71\]  

- Other constraints show the physical limitations of the pan, including the feed and steam supply rates.

---

Massecuite is the mixture of sugar crystals in a molasses suspension.
\[ 0 \leq F \leq F_U \] (14)

\[ St_L \leq St \leq St_U \] (15)

The target of the problem is the minimum of batch operating time.

\[
\text{ObjectiveFunction} = \min \sum \left( \text{Initiation} + \text{Foundation} + \text{Boilback} + \text{Heavyup} \right)
\] (16)

Equation (16) leads to the loss of mass balance fidelity because of two levels of approximation: 1/ the parametric uncertainty in the estimated growth rate parameters and 2/ the neglect of size-dependent growth effects. The state equations (1-5) are written for a population with some nominal average crystal size. Since the crystal population entering the process could display a range of average sizes, the time optimal control problem will be formulated and simultaneously solved for two identical vacuum pans, rather than one. These pans have the same operating conditions except that \( N \) crystals of the smallest expected average diameter are seed for the first pan and the same mass of crystals of the largest expected average diameter are seed for the second pan. The feed policies determined are known as back-off policies because they will satisfy constraints (10-15) over the range of average sizes anticipated.

To avoid solving the ODEs required for each iteration, which is computationally expensive, the state variables are approximated by the Lagrange interpolation polynomial. These can be easily differentiated and back substituted into the state equations (1-5), converting them to a set of algebraic equations. The approximating problem now becomes a normal nonlinear programming problem and can be solved by any existing optimization technique. The approximating method of Orthogonal Collocation on Finite Elements developed by Biegler and Cutrell [6] is especially advantageous for this type of optimal control problem since it is solved for not only the minimum operating time but also the duration of each stage of the whole crystallization process. Any number of finite elements can be specified for each stage; it is the lengths of these elements that are decision variables in the optimization problem.

The evaporation rate is assumed to be proportional to the steam feed rate (i.e. no dynamics are considered). However, to describe well the performance of the vacuum pan, a correlation relating the boiling rate to some key variables such as crystal content, sucrose content, purity, etc. must be developed. For two decades many authors have contributed works on boiling heat transfer coefficients for systems like that of the vacuum pan. In general these correlations are too complicated to be incorporated into optimization problems of this type and, given their high standard errors (up to ±40%), the value of doing so is questionable. Thus a simple pool boiling correlation is used instead. Notation and more details of equation (17) can be found in Ozisik [7].

\[
\frac{C_p \Delta T}{\lambda_{fg} Pr^a} = \frac{q}{\mu_{lf}} \left[ \frac{\sigma}{g \left( \rho_f - \rho_s \right)} \right]^{0.33}
\] (17)

Equation (17) shows the relationship between the energy required for boiling and the viscosity of the liquid. The correlations of Broadfoot and Steindl [8] and Awang & White [9] devised for viscosity of the molasses and the massecuite respectively are used in this paper. The following section presents some important results.

3. Results and Discussion

The feed and steam policies are plotted in Figure 1. In searching for the minimum operating time, the solver also determines the stage duration of the crystallization process. During 10-minute Initiation, no feed should be added to the pan. The mixture is boiled off until the content reaches the highest permitted level of supersaturation (\( OS_{fus} = 1 \)). This is shown in Figure 2, plotting the fractional over saturation (OS) and crystal content (CC) profiles. The Foundation stage lasts for about fifty minutes, then the fresh liquor feed is turned off to be replaced by the A molasses feed during 10-minute Boilback. The duration of Boilback is dependent on the purity in the pan required at the final stage. During the above stages, the maximum steam rate is constrained to a fixed maximum value, independent of the state of the system. The crystal content reaches its highest level and the content of the pan becomes very viscous. Meanwhile the OS_{fus} level starts to drop during Foundation but it goes up again during Boilback. Figure 1 shows only the feed back-off profiles. Figure 2, however shows the range profiles of OS_{fus} and CC. The two solid dark curves represent the smallest and the largest sizes of the crystals, the broken curve for the mean size of the crystals. The heavy up stage is hardly seen in both figures since it lasts for only two minutes.

If the mobility of the mixture is considered as one of the key factors in the evaporation, Foundation and Heavy up become the major stages of the whole crystallization process as shown in Figure 3. Both stages last for almost sixty minutes. Since the highly viscous fluid limits the evaporation rate (see Figure 3), the level of CC in
the pan is kept at the lowest level during Foundation (see Figure 4) in order to achieve as high an evaporation rate as possible. Then this level starts to go up during Boilback and Heavy up. Meanwhile the evaporation rate rapidly drops during the last two stages because of the high CC level. The solution throughout the study has represented fairly well the behaviour of the crystallizer although the correlation used was not developed for viscous fluids such as the massecuite. One major disadvantage in using this correlation is that it relates evaporation rate to viscosity. Additional correlations relating viscosity to measurable variables (crystal and sucrose contents, impurities, etc.) are therefore required, which bring in higher standard errors.

4. CONCLUSIONS

This paper considers using the simultaneous integration and optimization technique to solve the optimal control problem of a batch raw sugar vacuum pan. Since a normal batch consists of many stages, the chosen technique is found to be more advantageous than the sequential optimization approach in computational time and path constraint handling. The solution found provides the minimum time for the sugar crystals reaching the desired average size as well as the optimal stage duration of the whole crystallization process. The operating conditions during each stage, namely $OS_{frac}$ and CC levels are also determined. A simple form of the pool boiling correlation is used in the model to take into account the effect of crystal content, sugar content and impurities on the boiling rate. The results show fairly well the operation of the pan in the plant such as no feed added to the pan during Initiation or the switch between two feeds: liquor and A molasses. However, a better correlation for heat transfer in massecuite boiling, which relates the boiling rate to other measurable variables must be developed. This will be our next target in the next crushing season.

5. NOMENCLATURE

SYMBOLS

- $CC$: crystal content (dimensionless) defined as $(x_4)/(x_1 + x_2 + x_3 + x_4)$
- $Dry$: dry substance (dimensionless) defined as $(x_2 + x_3)/(x_1 + x_2 + x_3)$
- $E$: evaporation rate (tonne.h$^{-1}$)
- $F$, feed (tonne.h$^{-1}$). Liquor during Foundation and A molasses during Boilback
- $OS_{frac}$: oversaturation fraction (dimensionless)
- $G$: growth rate (mm.h$^{-1}$)
- $m$: mass (tonne.h$^{-1}$)
- $N$: number of crystal in the pan
- $OS$: solution sucrose oversaturation (°OS)
- $Pty$: purity (dimensionless) defined as $(x_2)/(x_1 + x_2)$
- $x_1$: mass of water in the pan (tonne.h$^{-1}$)
- $x_2$: mass of impurities in the pan (tonne.h$^{-1}$)
- $x_3$: mass of sucrose in the pan (tonne.h$^{-1}$)
- $x_4$: mass of crystals in the pan (tonne.h$^{-1}$)
- $x_5$: volume equivalent average diameter of crystals in the pan (mm)
- $T$: temperature of the pan (°C)
- $W$: flowrate of water (tonne.h$^{-1}$)

GREEK SYMBOLS

- $\pi$: pi (3.14159…)
- $\rho$: density of sucrose (kg.m$^{-3}$)

SUBSCRIPTS

- $f$: feed
- $h$: movement water
- crit: critical condition
- frac: fractional

6. ACKNOWLEDGEMENTS

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7. REFERENCES


Figure 1: Feed and Steam profiles (fixed constraint on maximum steam rate)

Figure 2: Profiles of $OS_{pow}$ and $CC$ (fixed constraint on maximum steam rate)
Figure 3: Profiles of Feed and Steam Rates (heat transfer constraint on maximum steam rate)

Figure 4: Profiles of $OS_{frac}$ and $CC$ (heat transfer constraint on maximum steam rate)