

Managing steam and concentration disturbances in multi-effect evaporators via nonlinear modelling and control

G.J. Adams* B.J. Burke** G.C. Goodwin* J.T. Gravdahl***
R.D. Peirce** A.J. Rojas*

* *ARC Centre of Excellence for Complex Dynamic Systems and Control, School of Electrical Engineering and Computer Science, The University of Newcastle, NSW 2308, Australia.*

** *Technical Systems, CSR Limited Victoria Mill, Ingham, QLD 4850, Australia.*

*** *Department of Engineering Cybernetics, Norwegian University of Science and Technology, N-7491 Trondheim, Norway.*

25 September 2007

Abstract: Evaporators are core units in many industrial processes including sugar mills. The dynamics of these systems are complex and hence the systems have been frequently used to demonstrate unusual systems and control behaviour. In this paper we explore a particular control architecture commonly employed in industry. We show that the architecture can lead to poor performance due to steam and concentration disturbances. An alternative architecture is then proposed which overcomes the difficulties.

1. INTRODUCTION

Evaporators are core units operating in many industrial processes. For example they are used as a key step in the extraction of sugar crystals from sugar syrup. Other uses are found in pulp and paper, desalination and others.

CSR Limited, owner and operator of seven raw sugar mills in Queensland, Australia, wish to enhance the operation of the Pioneer Sugar Mill (near Townsville, North Queensland) in order to reliably export a greater amount of power from a recently installed co-generation plant.

The operation of the multiple-effect evaporators (MEEs) has been identified as a critical station at which enhancements will deliver many benefits. Thus the present paper focuses on analysis of the evaporators, especially the last stages of the evaporator chain.

1.1 Background

The basic evaporator unit process involves heating a liquid (called the juice end product) containing a volatile and a non-volatile component. The goal is to remove the volatile component as vapour. In a sugar mill, the volatile component is water, with the remainder being sugar. Steam is used as a heating medium, and the boiling takes place in an enclosed unit at reduced pressure (to lower the boiling point of the water).

As only a certain amount of water removal takes place in each unit, evaporators are placed in chains to create an MEE process. The juice is passed through each evaporator in the chain, progressively gaining a higher content of sugar (termed the *brix* level, which is the percentage of soluble

product). At Pioneer Mill, the brix content starts at 15%, and exits the last evaporator at 67%.

1.2 Modeling

Based on first-principles modeling Newell and Fisher (1972) developed a tenth order nonlinear model for a double-effect evaporator. The model was reduced, using various assumptions, to fifth order and further to third and second order. The second order model has the concentrations of the first and second effect as states. In Newell and Lee (1989) a third order model for a single-effect evaporator with separator level, output concentration and operating pressure as states was reported.

The model of Newell and Fisher (1972) has inspired a lot of publications on control of double-effect or MEE.

The second order model of Newell and Fisher (1972) was further elaborated upon by Montano and Silva (1991). The goal of this work was modeling for control. This simplified model, which is a 2-state Newell-model, tight level control is also assumed, so that the evaporator level is no longer a state in the model. The two states then represent concentrations in two evaporators in series. In this model the steam flow to the evaporator is considered the control variable. This paper in turn motivated examples in papers such as Sira-Ramírez and Llanes-Santiago (1994), Silva-Navarro and Alvarez-Gallegos (1997) and Sira-Ramírez and Silva-Navarro (1999). In Silva-Navarro and Alvarez-Gallegos (1997) the model of Newell and Fisher (1972) and Montano and Silva (1991) is analyzed from a nonlinear and positive-systems point of view. In Nielsen et al. (1996) a first-principles model of a MEE was also derived.

In Elhaq et al. (1999) the authors augment the Newell model with first order equations for the flow and pressure variables in a five effect evaporator. The purpose of this is to capture the significant dynamics and response times otherwise not accounted for by the simplified Newell-model. It is also argued that steam demand from the crystallization process introduces a significant disturbance to the system, and that this explains some deviations between simulations and measurements from the five-effect plant.

There has also been an interest in modeling multi-effect evaporators with economic optimisation issues in mind, see Kaya and Sarac (2007) for an example of this approach.

1.3 Control and analysis

In Sira-Ramírez and Silva-Navarro (1999) the model of Newell and Fisher (1972) and Montano and Silva (1991) is controlled using passivity tools and trajectory planning. The paper is a fine example of the employment of tools mostly used for mechanical Euler-Lagrange systems to a process control problem. In Sira-Ramírez and Llanes-Santiago (1994) it is demonstrated how dynamic discontinuous feedback strategies, that is different types of modulation, can be applied to a model of a double-effect evaporator. Simulations in this paper illustrate inverse response in the states of the closed loop system. In Nielsen et al. (1996) a combination of dynamic decoupling, gain scheduling and PI-controllers is used, while Pitteea et al. (2004) propose a fuzzy-logic controller for a five-effect evaporator plant. The model of Newell and Fisher (1972) is employed, but is tuned to real data using genetic algorithms.

In Elhaq et al. (1997), analysis and control of the Newell and Fisher (1972) model is carried out. Both inflow and steam flow are considered control inputs. Variations in steam flow is mentioned as a disturbance to the process. In Elhaq et al. (1999), the same authors design controllers for sugar evaporation using multivariable generalized predictive control. In all of the above papers, the steam flow is considered a control variable.

In the current paper, we will treat the inflow of steam to the process as a disturbance, and use the inflow and out-flow of juice as control variables to control both evaporator level and output brix.

1.4 Our Focus

The current paper arose from a two day workshop dedicated to the study of control issues at Pioneer Mill near Townsville, Australia¹. One of the issues arising from the workshop was that of process disturbances, both in level and brix. Observations from real data obtained at Pioneer Mill (see Figure 1) show the existence of three different oscillatory behaviours: a long term sustained oscillation with a period of approximately two hours, a second sporadic oscillation with a period of approximately ten minutes, and a faster and smaller oscillation with a period of approximately a minute (particularly in juice level).

¹ There were six participants at the workshop and this explains the larger than usual number of authors of this paper.

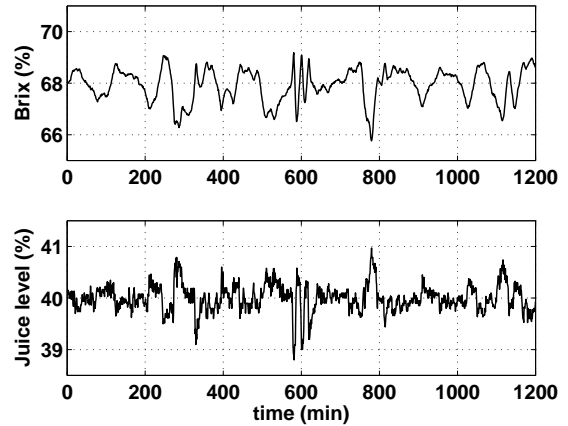


Fig. 1. Brix and juice level responses from the last stage evaporator at Pioneer Mill.



Fig. 2. Sugar process schematic overview .

The slowest oscillatory perturbation is linked to the evaporator stage being coupled with the pan's batch process (where the liquor is crystallised) through the steam flow requirement (see Figure 2). This was also discussed in Elhaq et al. (1999). The second sporadic oscillatory episode is linked to interruptions in the sugar cane's crushing stage and can be observed in Figure 1 around 600 minutes. The juice from the crushing mill is stored in the effet supply juice (ESJ) storage tank to which water is added when a stop occurs. The water addition maintains the MEE stage in operation, until crushing can restart. However, this addition dilutes the sugar concentration, producing a disturbance in the incoming brix content for the first MEE stage. The brix disturbance is then transmitted through all the other stages. Finally, it is believed that the fastest of the three oscillations (observable in the level trace in Figure 1) may be due to excessive control action for the juice level, the presence of sticky valves or a small-scale level fluctuation.

The above observations lead the authors to develop a detailed model for a double-effect evaporator. A first stage model describes the lumped chain of MEEs from the first to the second from last, and a second model describes the last evaporator producing the final brix content.

Having established a likely cause for the oscillatory behaviour, an alternative control system architecture is proposed to solve the low frequency oscillation in brix due to steam flow disturbances from the pans. The sporadic disturbance episodes (linked to water addition from sudden stops in the crushing mill) is the subject of on-going research.

The proposed alternative architecture is shown in the sequel to attenuate the effect of disturbances in the steam flow. Simulation results confirm the superior nature of the proposed architecture.

In summary, the key contributions of the present paper are: the modelling of a particular co-generation sugar mill plant, the identification of a set of disturbances relevant to a specific installation and finally a proposed new control architecture aimed at mitigating the effect of those disturbances.

2. PHYSICAL MODELLING

An important aspect of the operation of the particular evaporators of interest here is that maximal evaporation efficiency occurs when the juice level is at some specific operating point. Indeed, the evaporator efficiency versus juice level (denoted $f(L(t))$) is as in Figure 3 (see Hugot (1986) for details). Obviously, there is a strong incentive to maintain the level at the maximum of the efficiency curve in Figure 3. This is typically achieved by some form of level control.

Another aspect which affects evaporator operation is the throughput of juice. Thus the per-unit volume evaporation rate V depends on this flow and on the steam flow. In reality, the vapour flow rate also depends on evaporator pressure, temperature, and probably other effects (viscosity and density of the liquid etc). Note that total throughput depends on the smaller of inflow and outflow, and for non-zero evaporation the outflow will always be the smaller of the two.

A total energy balance for the system leads to

$$\frac{d}{dt}(M_{total}H_{total}) = (h_s - h_c)S + h_{in}F_{in} - h_vV - hF_{out}, \quad (1)$$

where² M_{total} is the mass of both solute and steam, H_{total} is the total enthalpy content in solute and steam, S is the inflow of steam, h_s is the enthalpy of the steam, h_{in} is the enthalpy of the inflowing juice, V is the vapour flow in the stage, h_v is the enthalpy of the vapour, h is the enthalpy of the outflowing juice and h_c is the energy content in the condensated steam. We consider a static balance, that is $d(M_{total}H_{total})/dt = 0$. This is justified given the relatively slower dynamics related to the heating of liquid. This leads to the following simplified energy balance:

$$0 = (h_s - h_c)S + h_{in}F_{in} - h_vV - hF_{out}. \quad (2)$$

Rearranging leads to

$$\begin{aligned} V &= \frac{1}{h_v} \{h_{in}F_{in} - hF_{out} + (h_s - h_c)S\} \\ &= \frac{h_{in}}{h_v}F_{in} - \frac{h}{h_v}F_{out} + \frac{(h_s - h_c)}{h_v}S. \end{aligned} \quad (3)$$

We see that V is linearly dependent on the inflow rate $F_{in}(t)$, the outflow rate $F_{out}(t)$ and the steam flow $S(t)$.

These considerations, and a simple mass balance on juice, lead to the following model³ linking inflow $F_{in}(t)$, outflow $F_{out}(t)$ and juice level $L(t)$ (as a percentage of total evaporator chamber height).

² The mass and flow quantities in (1) are dependent on time, but the (t) dependence is not shown in (1) for clarity; enthalpies are assumed constant.

³ Time-dependent variables will now have the (t) qualifier, including V ; enthalpies are still constant.

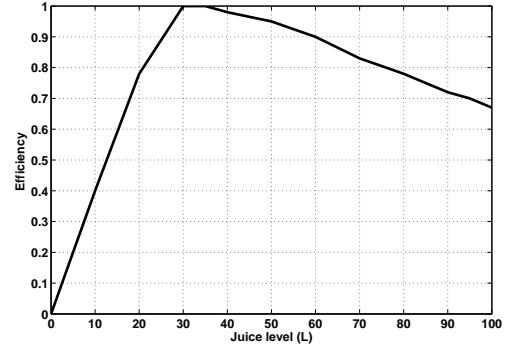


Fig. 3. Evaporator efficiency versus juice level $f(L(t))$ (as percentage of total height).

$$\frac{dL}{dt} = \frac{1}{E_{vol}} \{F_{in}(t) - F_{out}(t) - O(t)\} \quad (4)$$

where E_{vol} denotes the evaporator volume and

$$O(t) = V(t)f(L(t)) \quad (5)$$

is the actual vapour flow. Substituting for the vapour flow (3) in (4) gives

$$\frac{dL}{dt} = \frac{1}{E_{vol}} \left\{ \left(1 - f(L(t))\frac{h_{in}}{h_v}\right)F_{in}(t) - \left(1 - f(L(t))\frac{h}{h_v}\right)F_{out}(t) - f(L(t))\frac{(h_s - h_c)}{h_v}S(t) \right\}. \quad (6)$$

The second variable of interest is brix in the evaporator $B(t)$ (which, assuming perfect mixing, is the same as brix in the outflow). To develop an expression for $B(t)$, an expression for sugar volume in the evaporator will first be obtained. The sugar volume $C(t)$ changes according to

$$\frac{dC}{dt} = B_{in}(t)F_{in}(t) - B(t)F_{out}(t), \quad (7)$$

where $B_{in}(t)$ is the brix level in the inflow. Next we note that the brix level is related to the sugar volume according to $C(t) = L(t)E_{vol}B(t)$. So relating (7) to $C(t)$ gives

$$\begin{aligned} \frac{dC}{dt} &= \frac{d}{dt}(L(t)E_{vol}B(t)) = E_{vol} \left(L(t)\frac{dB}{dt} + B(t)\frac{dL}{dt} \right) \\ &= B_{in}(t)F_{in}(t) - B(t)F_{out}(t). \end{aligned}$$

Rearranging this, and using (4), gives

$$\begin{aligned} E_{vol}L(t)\frac{dB}{dt} &= B_{in}(t)F_{in}(t) - B(t)F_{out}(t) - E_{vol}B(t)\frac{dL}{dt} \\ &= B_{in}(t)F_{in}(t) - B(t)F_{out}(t) \\ &\quad - B(t)(F_{in}(t) - F_{out}(t) - O(t)) \\ &= F_{in}(t)(B_{in}(t) - B(t)) + B(t)O(t). \end{aligned}$$

So

$$\frac{dB}{dt} = \frac{1}{L(t)E_{vol}} \{F_{in}(t)(B_{in}(t) - B(t)) + B(t)O(t)\}. \quad (8)$$

Substituting for $O(t)$ gives

$$\begin{aligned} \frac{dB}{dt} &= \frac{1}{L(t)E_{vol}} \left\{ \left(B_{in}(t) - B(t) \left(1 - f(L(t))\frac{h_{in}}{h_v}\right) \right) F_{in}(t) \right. \\ &\quad \left. - B(t)f(L(t))\frac{h}{h_v}F_{out}(t) + B(t)f(L(t))\frac{(h_s - h_c)}{h_v}S(t) \right\}. \end{aligned} \quad (9)$$

This equation shows that rate of change of brix is not affected by outflow, except through the vapour flow term. This is valid because brix only changes when water leaves as vapour, or when the inflow brix is different to the evaporator brix and the inflow changes. So if $B_{in}(t)$ is less

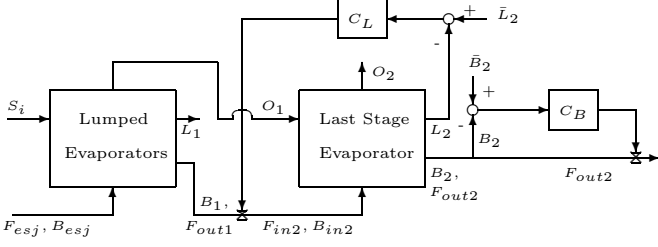


Fig. 4. Control architecture at Pioneer Mill, with \bar{L}_2 and \bar{B}_2 the level and brix setpoints for the last stage evaporator.

than evaporator brix, then an increase in inflow will result in a drop in brix $B(t)$.

The model represented by equation (6) and equation (9) (and also (3)) describes only one of the effect stages involved in an MEE. In particular the MEE at Pioneer Mill consists of two chains of six effects each. Thus we would require six such models coupled together for each chain. Nonetheless, in a trade-off between precision and simplicity of the model, in this paper we only explicitly consider the last stage evaporator, whilst lumping together the five previous evaporators in one fictitious effect (and only consider one chain). In summary the model under consideration is given by

$$\begin{aligned} \frac{dL_1}{dt} &= \frac{1}{E_{vol1}} \left\{ \left(1 - f_1(L_1(t)) \frac{h_{in1}}{h_{v1}}\right) F_{esj}(t) - \right. \\ &\quad \left. \left(1 - f_1(L_1(t)) \frac{h_1}{h_{v1}}\right) F_{out1}(t) - f(L_1(t)) \frac{(h_{s1} - h_{c1})}{h_{v1}} S_i(t) \right\} \\ \frac{dB_1}{dt} &= \frac{1}{L_1(t) E_{vol1}} \left\{ \left(B_{esj}(t) - B_1(t) \left(1 - f_1(L_1(t)) \frac{h_{in1}}{h_{v1}}\right) \right) F_{esj}(t) \right. \\ &\quad \left. - B_1(t) f_1(L_1(t)) \frac{h_1}{h_{v1}} F_{out1}(t) + B_1(t) f_1(L_1(t)) \frac{(h_{s1} - h_{c1})}{h_{v1}} S_i(t) \right\} \\ \frac{dL_2}{dt} &= \frac{1}{E_{vol2}} \left\{ \left(1 - f_2(L_2(t)) \frac{h_{in2}}{h_{v2}}\right) F_{in2}(t) - \right. \\ &\quad \left. \left(1 - f_2(L_2(t)) \frac{h_2}{h_{v2}}\right) F_{out2}(t) - f_2(L_2(t)) \frac{(h_{s2} - h_{c2})}{h_{v2}} O_1(t) \right\} \\ \frac{dB_2}{dt} &= \frac{1}{L_2(t) E_{vol2}} \left\{ \left(B_{in2}(t) - B_2(t) \left(1 - f_2(L_2(t)) \frac{h_{in2}}{h_{v2}}\right) \right) F_{in2}(t) \right. \\ &\quad \left. - B_2(t) f_2(L_2(t)) \frac{h_2}{h_{v2}} F_{out2}(t) + B_2(t) f_2(L_2(t)) \frac{(h_{s2} - h_{c2})}{h_{v2}} O_1(t) \right\} \\ O_1(t) &= \left(\frac{h_{in1}}{h_{v1}} F_{esj}(t) - \frac{h_1}{h_{v1}} F_{out1}(t) + \frac{(h_{s1} - h_{c1})}{h_{v1}} S_i(t) \right) f_1(L_1(t)) \\ O_2(t) &= \left(\frac{h_{in2}}{h_{v2}} F_{in2}(t) - \frac{h_2}{h_{v2}} F_{out2}(t) + \frac{(h_{s2} - h_{c2})}{h_{v2}} O_1(t) \right) f_2(L_2(t)), \end{aligned} \quad (10)$$

where the coupling takes place by means of S_1 , the fact that $F_{out1} = F_{in2}$ and that $B_1 = B_{in2}$. The subindex 1 is used for the lumped stage model, whilst the subindex 2 is used for the final stage evaporator; see Figure 4.

3. TYPICAL CONTROL ARCHITECTURE

Several architectures are in common use to regulate evaporators. A typical architecture is the one shown in Figure 4. Indeed, this is the one used at Pioneer Mill. As explained earlier, optimal heating efficiency is attained at a certain juice level inside each evaporator, requiring good level control. At Pioneer Mill, controllers are set up to manipulate both the inflow and outflow to the final evaporator unit in each chain, as well as the flows between each unit. The rationale behind the architecture is as follows:

- Juice level inside the evaporator is determined by integrating the difference between the inflow and out-

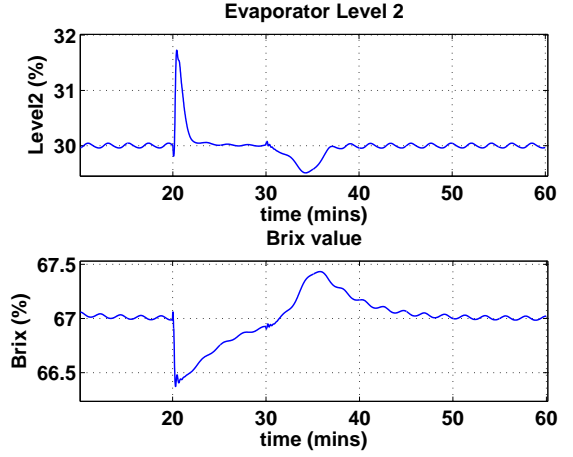


Fig. 5. Poor performance in outputs of stage 2, due to a drop in inlet brix (simulated data).

flow; inflow was chosen as the manipulated variable for level.

- Brix level is controlled by the total flow of juice through the evaporator. Total flow is governed by the smaller of inflow or outflow, which will be the outflow for non-zero evaporation. Also note that inflow is already used for level control.

However, the performance of this control system is often unsatisfactory.

Pioneer Mill is a co-generation plant, which supplies power to the grid under contract. The ability to deliver the contracted power is compromised somewhat by the unsatisfactory operation of the evaporators, due to the fundamentally important role the evaporation process plays in sugar milling.

We have already seen an example of unsatisfactory behaviour of Pioneer Mill brix control in Figure 1. Note in particular the sporadic behaviour associated with abrupt lowering of brix in the inlet flow (at around 600 minutes). Simulations using the model in (10) replicate this problem, as shown in Figure 5 and Figure 6. These figures show responses to an abrupt drop in inlet brix (due to water being added to the ESJ tank) when under feedback control of the type described above. Observe the reactive behaviour from brix control changes causing level changes, thereby causing level control corrections and so on. (N.B. Higher frequency oscillations are intended to replicate the small-scale fluctuations in level.)

4. IMPROVED CONTROL ARCHITECTURE (STEP 1)

Examination of the phenomenon described above suggests that the key issue is to uncouple changes in level from changes in brix. This can be achieved by adding a feedforward element d as shown in Figure 7. This then has the effect of introducing a matrix

$$D = \begin{bmatrix} 1 & d \\ 0 & 1 \end{bmatrix} \quad (11)$$

at the input to the last stage evaporator, so that

$$\begin{bmatrix} F_{in2}(t) \\ F_{out2}(t) \end{bmatrix} = \begin{bmatrix} 1 & d \\ 0 & 1 \end{bmatrix} \begin{bmatrix} u_l(t) \\ u_b(t) \end{bmatrix}, \quad (12)$$

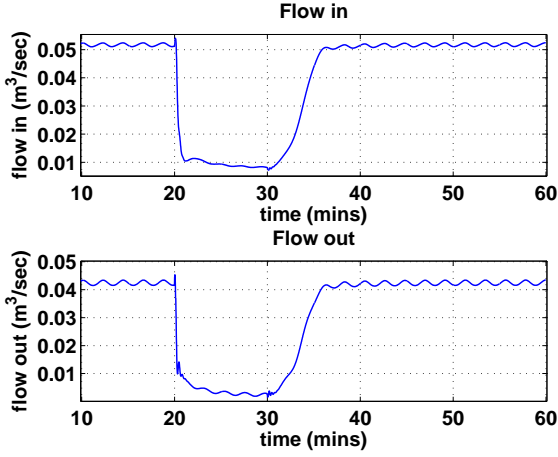


Fig. 6. Inlet and outlet flow for the stage 2 evaporator, in response to a drop in inlet brix (simulated data).

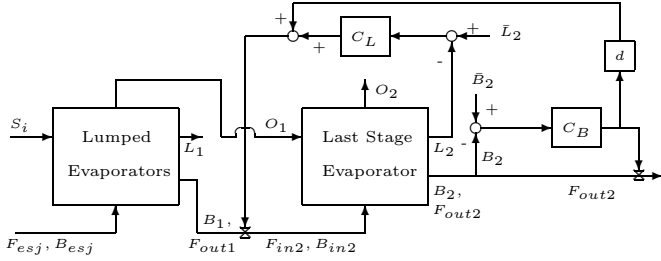


Fig. 7. Improved control architecture.

where $u_l(t)$ and $u_b(t)$ are the outputs of the level and brix controllers, respectively. The feedforward term d is usually constructed using input/output relationships for the variable in question. In our case this is the evaporator level due to the inlet and outlet flow rates. Rearranging the expression for dL_2/dt in (10) will give us

$$\frac{dL_2}{dt} = -\frac{f_2(L_2(t))}{E_{vol2}} \frac{(h_{s2} - h_{c2})}{h_{v2}} O_1(t) + \left[\frac{1}{E_{vol2}} \left(1 - f_2(L_2(t)) \frac{h_{in2}}{h_{v2}}\right) - \frac{1}{E_{vol2}} \left(1 - f_2(L_2(t)) \frac{h_2}{h_{v2}}\right) \right] \begin{bmatrix} F_{in2}(t) \\ F_{out2}(t) \end{bmatrix}. \quad (13)$$

Then d is constructed as follows

$$d = \frac{\left(1 - f_2(L_2(t)) \frac{h_2}{h_{v2}}\right)}{E_{vol2}} / \frac{\left(1 - f_2(L_2(t)) \frac{h_{in2}}{h_{v2}}\right)}{E_{vol2}} = \frac{h_{v2} - f_2(L_2(t)) h_2}{h_{v2} - f_2(L_2(t)) h_{in2}}, \quad (14)$$

where it is assumed that the valve characteristics for F_{in2} and F_{out2} are identical. From (12) and (13) we have

$$\frac{dL_2}{dt} = -\frac{f_2(L_2(t))}{E_{vol2}} \frac{(h_{s2} - h_{c2})}{h_{v2}} O_1(t) + \left[\frac{1}{E_{vol2}} \left\{1 - f_2(L_2(t)) \frac{h_{in2}}{h_{v2}}\right\} - \frac{1}{E_{vol2}} \left\{1 - f_2(L_2(t)) \frac{h_2}{h_{v2}}\right\} \right] \begin{bmatrix} 1 & d \\ 0 & 1 \end{bmatrix} \begin{bmatrix} u_l(t) \\ u_b(t) \end{bmatrix} \quad (15)$$

and so using (14) gives

$$\frac{dL_2}{dt} = -\frac{f_2(L_2(t))}{E_{vol2}} \frac{(h_{s2} - h_{c2})}{h_{v2}} O_1(t) + \left[\frac{1}{E_{vol2}} \left\{1 - f_2(L_2(t)) \frac{h_{in2}}{h_{v2}}\right\} \right] \begin{bmatrix} u_l(t) \\ u_b(t) \end{bmatrix}. \quad (16)$$

This change in control architecture then causes the level control to become uncoupled from brix variations. Sim-

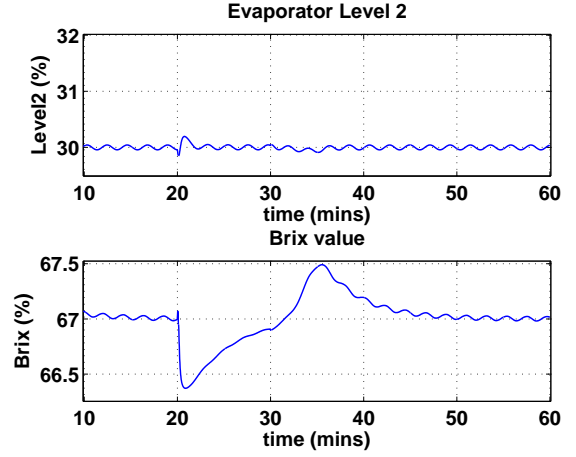


Fig. 8. Improved performance in output responses due to a drop in inlet brix (simulated data).

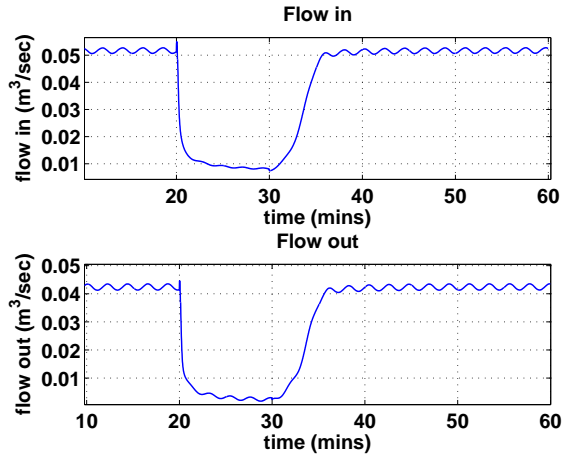


Fig. 9. Inlet and outlet flows for the improved architecture in response to a drop in brix (simulated data).

ulation results (for the same brix setpoint change as in Figure 5) are shown in Figure 8 and Figure 9. It can be seen by comparing Figures 8 and 9 with Figures 5 and 6 that the performance is significantly better for level control using the modified architecture. It is interesting that the inflow and outflow shown in Figures 6 and 9 are only slightly different. However, the difference is sufficient to prevent the level fluctuations clearly evident in Figure 5 but barely visible in Figure 8. This improved performance has been confirmed by plant trials as shown in Figure 10, where the improved architecture of this section was applied from the 160th minute until the 250th minute. We observe fast variability in the outlet flow $F_{in2}(t)$ (dotted line), nonetheless the slow variability in level output $L_2(t)$ is noticeably reduced in amplitude (solid line), producing better tracking of the setpoint \bar{L}_2 for the period in which the modified control architecture was in place.

5. IMPROVED CONTROL ARCHITECTURE WITH STEAM FEED-FORWARD (STEP 2)

Next we focus on steam flow perturbations arising from interactions with the pans. In practice these perturbations have a major effect on performance.

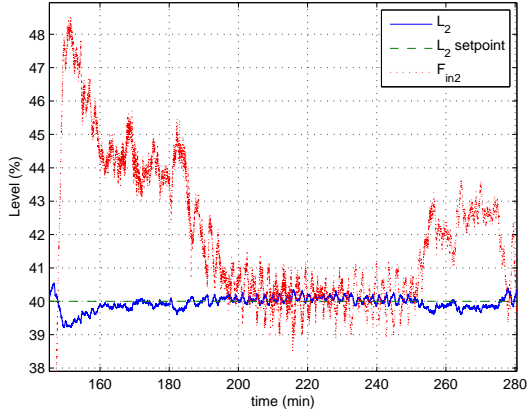


Fig. 10. Improved performance (real data).

As described in Subsection 1.4, slow periodic steam disturbances (due to steam demand from other mill operations) have a significant effect on brix. This can be observed in Figure 12, which shows brix and level control for the original control architecture (without feedforward from brix control to level).

The feedforward described in Section 4 only helps level control, as seen in the simulations shown in Figure 13 where we observe that brix is still affected.

To reduce the effect of steam disturbances of the type experienced at Pioneer Mill, the measurement of steam flow to the evaporators, $S_i(t)$, is averaged (over a long time, say 24[hrs]) and this mean value, \bar{S}_i , is removed from the $S_i(t)$ measurement to leave $\tilde{S}_i(t) = S_i(t) - \bar{S}_i$. The brix output $B_2(t)$ has its setpoint $\bar{B}_2(t)$ removed to leave $\tilde{B}_2(t) = B_2(t) - \bar{B}_2(t)$.

A simple correlation analysis is then performed to determine the relative time lag T_s and gain K_s between $\tilde{S}_i(t)$ and $\tilde{B}_2(t)$. The signal $K_s \tilde{S}_i(t - T_s)$ is then added to the error signal for the brix controller effectively compensating for the steam disturbance $\tilde{S}_i(t)$.

The simplicity of this scheme is of practical relevance due to the variability of the evaporator dynamics, feed material, etc. The feedforward scheme is not intended to remove all of the main periodic disturbance effects, only an economically significant amount.

As observed in Figure 14, this simple scheme has removed much of the brix variability, with the brix oscillations reduced from a range of $\{66.6\% \rightarrow 67.9\% \}$ to $\{66.95\% \rightarrow 67.05\% \}$. Future plant trials are planned to evaluate this scheme on the real MEEs at Pioneer Mill.

6. CONCLUSION

In the present paper we have developed an evaporator model that includes steam by means of the evaporation efficiency curve and as a process disturbance. We used the resultant model to represent a double-effect evaporator situation. The first effect described a lumped version of the first five effects at Pioneer Mill, whilst the second model effect described the last stage effect from the real plant. The double effect approximation was able to reproduce

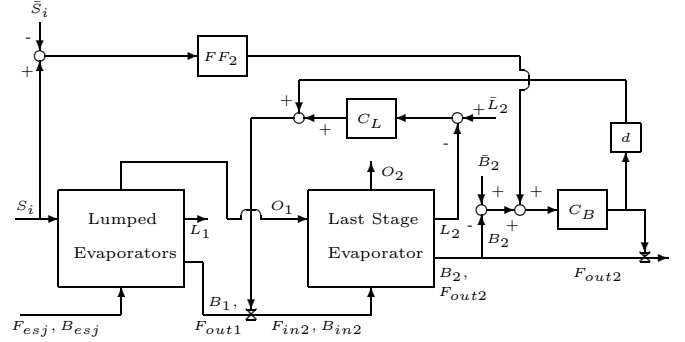


Fig. 11. Improved control architecture with steam flow feed-forward, where \bar{S}_i is the average steam disturbance trend.

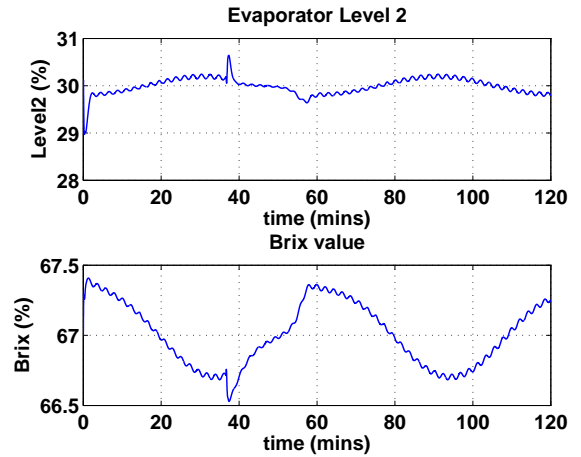


Fig. 12. Original control with sinusoidal steam disturbance, as well as an inlet brix disturbance at time 37 minutes.

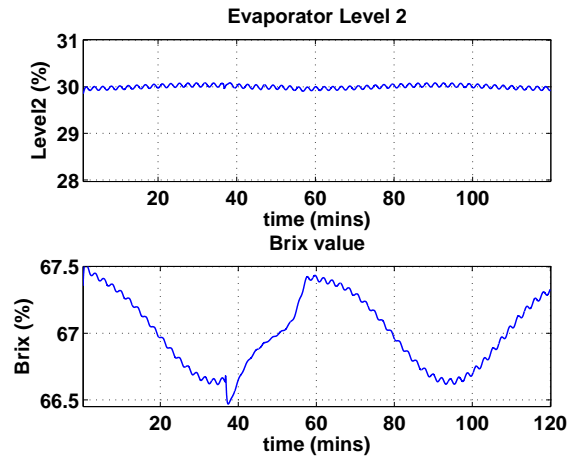


Fig. 13. Feed-forward control from brix controller.

reasonably well the main disturbances observed in the real sugar mill process, that is: a long term sustained oscillation due to steam disturbances, a sporadic oscillation due to concentration disturbances and a faster and smaller oscillation (particularly in juice level) possibly due to small-scale level fluctuation. We then used the double-effect model to simulate the control architecture at Pioneer Mill. Finally we proposed two modification in the existing control architecture to include brix and steam feedforward

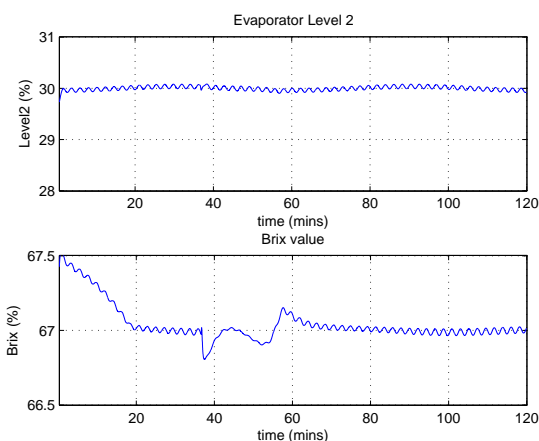


Fig. 14. Feed-forward on steam fluctuations.

which showed a significant improvement in performance via simulation studies. Future work will include the testing of the full scheme on the real plant. However, the simulation results give confidence that this should be successful.

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