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

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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 set.

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 sets to create an MEE process. The juice is passed through each evaporator in the set, 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 Modelling

Based on first-principles modelling 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 modelling for control. This simplified model, which is a two-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.

[1] The authors gratefully acknowledge the support of CSR Ltd, and the Australian Research Council

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 modelling 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 an 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 [1].

1.4 Our Focus

The current paper arose from a two day workshop and subsequent research effort dedicated to the study of control issues at Pioneer Mill near Townsville, Australia [2]. 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).

The slowest oscillatory perturbation is linked to the evaporator stage being coupled with the pan's batch process

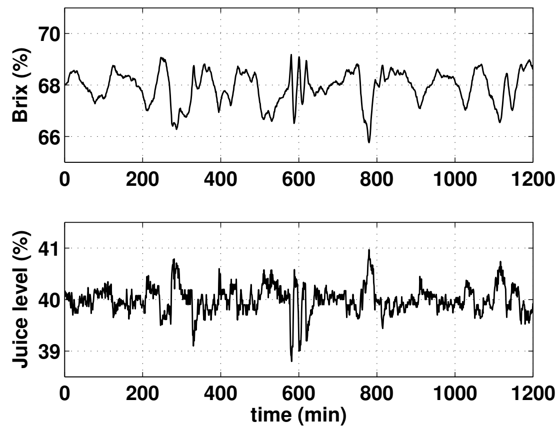

Fig. 1. Brix and juice level responses from the last stage evaporator at Pioneer Mill.

Fig. 2. Sugar process schematic overview.

(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. 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 set 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 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 generic sugar mill plant, the identification of a set of disturbances relevant to a specific installation and finally a new control architecture aimed at mitigating the effect of those disturbances.

[1] In the current paper, we will treat the inflow of steam to the process as a disturbance, and use the inflow and outflow of juice as control variables to control both evaporator level and output brix.
[2] There were six participants at the workshop and this explains the larger than usual number of authors of this paper.

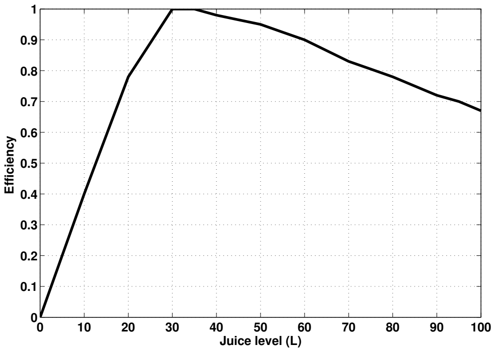

Fig. 3. Evaporator efficiency versus juice level $f(L(t))$ (as percentage of total height).

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). 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 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

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

where[3] 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:

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

Rearranging leads to

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

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

[3] 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.

These considerations, and a simple mass balance on juice, lead to the following model[4] linking inflow $F_{in}(t)$, outflow $F_{out}(t)$ and juice level $L(t)$ (as a percentage of total evaporator chamber height).

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

where E_{vol} denotes the evaporator volume and

$$O(t) = V(t)f\left(L(t)\right) \tag{5}$$

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

$$\begin{aligned} \frac{dL}{dt} = \frac{1}{E_{vol}}\Big\{ & \Big(1 - f(L(t))\frac{h_{in}}{h_v}\Big)F_{in}(t) - \\ & \Big(1 - f(L(t))\frac{h}{h_v}\Big)F_{out}(t) - f(L(t))\frac{(h_s - h_c)}{h_v}S(t)\Big\}. \end{aligned} \tag{6}$$

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 $B(t)$, we first obtain an expression for sugar volume, $C(t)$, in the evaporator:

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

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

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

Rearranging this, and using (4), gives

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

So

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

Substituting for $O(t)$ gives

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

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

The MEE at Pioneer Mill consists of two sets of six effects each. 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 set). Thus the model is

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

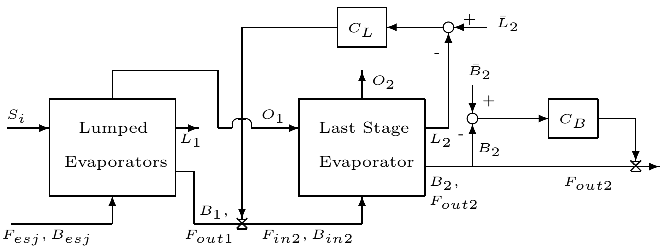

Fig. 4. Control architecture at Pioneer Mill; $\bar{L}_2$ and $\bar{B}_2$ are level and brix setpoints for the final evaporator.

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

where the coupling takes place by means of O_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 used at Pioneer Mill, and shown in Figure 4. 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 set, as well as the flows between each unit. Unfortunately, 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. This figure shows level and brix 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 replicate the small-scale fluctuations in level.)

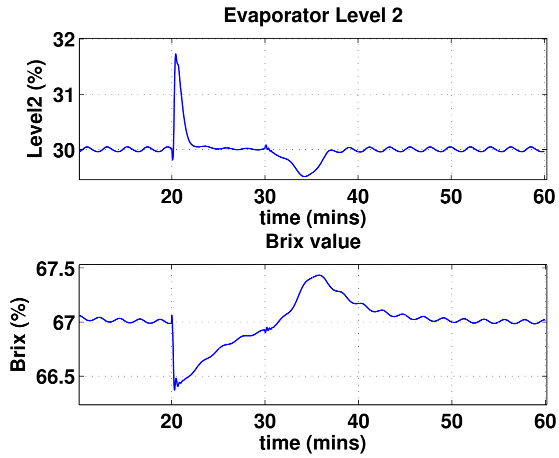

Fig. 5. Poor performance in outputs of stage 2, due to a drop in inlet brix (simulated data).

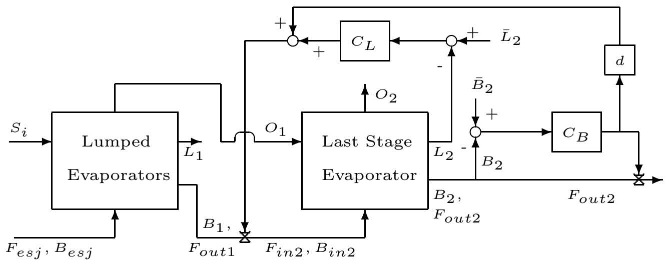

Fig. 6. Improved control architecture.

4. IMPROVED CONTROL ARCHITECTURE

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 6. This then has the effect of introducing a matrix

$$
D = \begin{bmatrix} 1 & d \\ 0 & 1 \end{bmatrix} \tag{11}
$$

at the input to the last stage evaporator, so that

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

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) gives

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

Then d is constructed as follows:

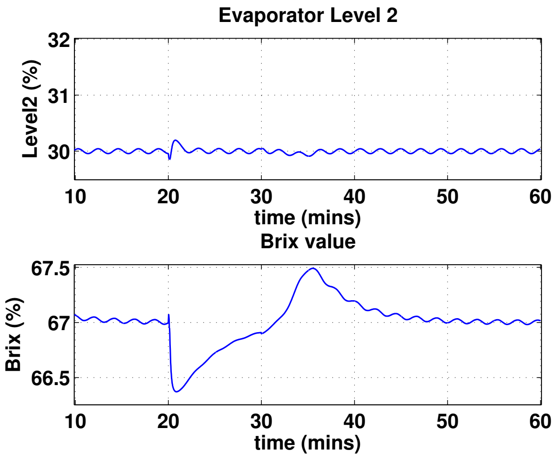

Fig. 7. Improved performance in output responses due to a drop in inlet brix (simulated data).

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

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

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

and so using (14) gives

$$\frac{dL_2}{dt} = -\frac{f_2(L_2(t))}{E_{vol2}}\frac{(h_{s2}-h_{c2})}{h_{v2}}O_1(t) + \left[\frac{1}{E_{vol2}}\left\{1 - f_2(L_2(t))\frac{h_{in2}}{h_{v2}}\right\} \quad 0\right] \begin{bmatrix} u_l(t) \\ u_b(t) \end{bmatrix}. \tag{16}$$

This change in control architecture then causes the level control to be, in principle, uncoupled from brix variations. Simulation results (for the same brix setpoint change as in Figure 5) are shown in Figure 7. It can be seen by comparing Figure 7 with Figure 5 that the performance is significantly better for level control using the modified architecture. This improved performance has been confirmed by plant trials as shown in Figures 8 and 9, where the improved architecture of this section was used in the B set evaporator (darker lines), and original controllers remained on the A set (lighter lines).

5. IMPROVED CONTROL ARCHITECTURE WITH STEAM FEED-FORWARD

Next we focus on steam flow perturbations arising from interactions with the pans. 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 11, which shows brix and level control for the original control architecture.

To reduce the effect of steam disturbances of the type experienced at Pioneer Mill, a feedforward scheme was developed in which the measurement of steam flow to the evaporators, $S_i(t)$, is filtered by an appropriately-tuned bandpass-filter to form a steam fluctuation value

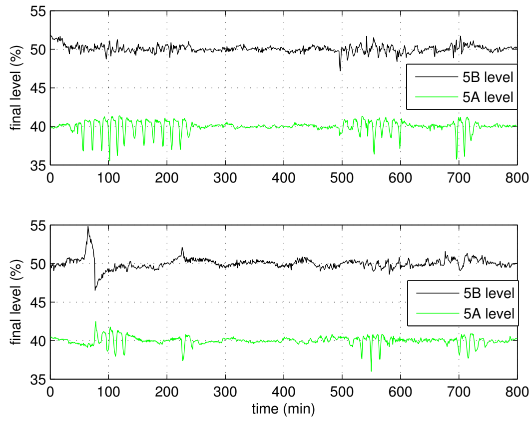

Fig. 8. Improved performance in level control (real data).

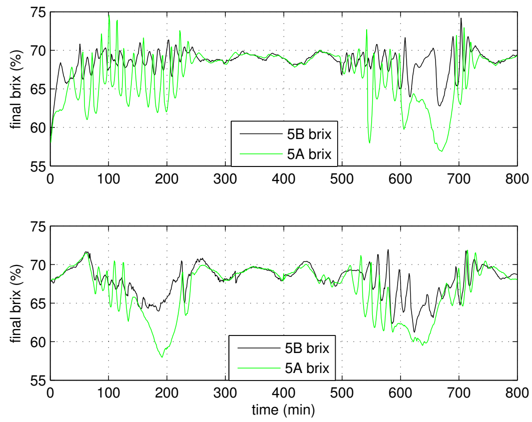

Fig. 9. Improved performance in brix control (real data).

$\tilde{S}_i(t)$. 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 $B_2(t)$. The signal $K_s\tilde{S}_i(t - T_s)$ is then added to the brix measurement being fed back to the brix controller, effectively compensating for the steam disturbance $\tilde{S}_i(t)$ (see Figure 10).

As observed in Figure 12, this simple scheme has removed much of the brix variability in simulations, with the brix oscillations reduced from ±0.3% to very close to zero. Some studies on real data have been promising. For example, Figure 13 demonstrates the similarity between the final brix measurements from the two sets of evaporators at Pioneer Mill, and a scaled and delayed pan steam demand signal. Note the similarity between these measurements. 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. 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

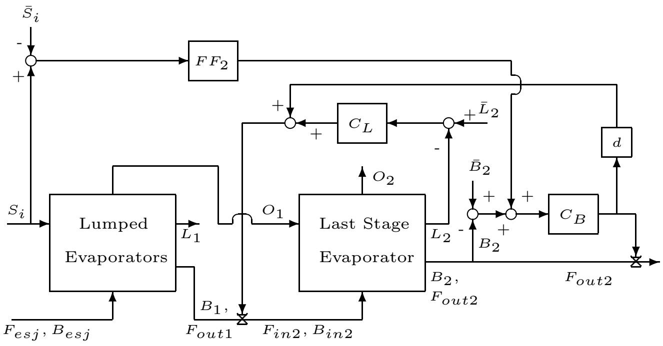

Fig. 10. Improved control architecture with steam flow feed-forward.

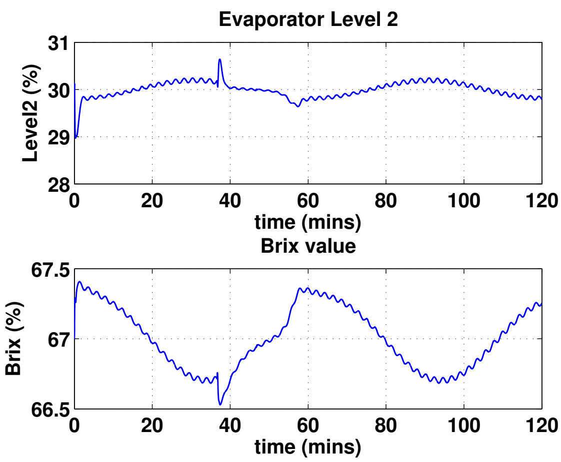

Fig. 11. Original control with sinusoidal steam disturbance, and inlet brix disturbance at 37 mins.

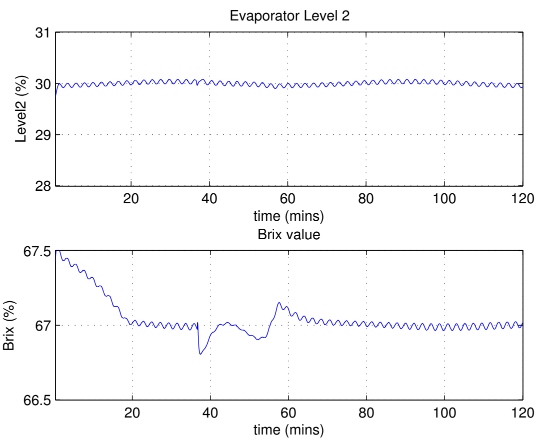

Fig. 12. Feed-forward on steam fluctuations.

double effect approximation was able to reproduce the main disturbances observed in the real sugar mill process. We also used the double-effect model to simulate the control architecture at Pioneer Mill. Finally we proposed two modifications to the existing control architecture to include brix and steam feedforward. This showed a significant improvement in performance via simulation studies. Future work will include the testing of the full scheme on the real plant.

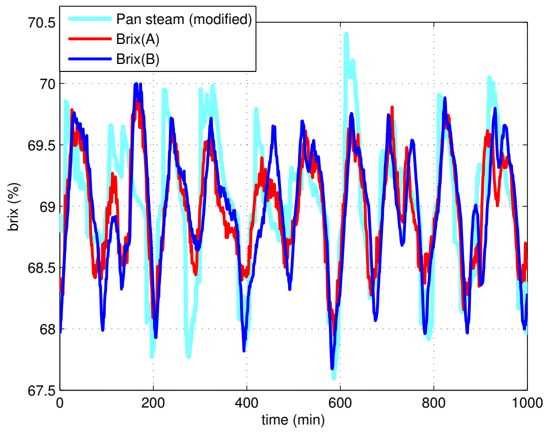

Fig. 13. Comparison of final brix measurements and modified steam demand signal.

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