Control of a batch pulp digester using a simplified mechanistic model to predict degree of polymerisation

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Abstract

Effective control of the sulphite pulp digestion process in the production of dissolved pulp in a batch digester is limited by three important restrictions, namely:

• the inability to measure the degree of polymerisation (DP) of the cellulose in the wood pulp, which is the controlled variable,
• the fact that flowrate of steam to an external heat exchanger, through which the cooking liquor circulates, is the only manipulated variable available and that,
• due to scheduling requirements, cook time per digester is fixed.

The availability of substantial computational capability in industrial environments has rendered the traditional S-factor prediction of cook time to control DP inadequate, compared to what can be achieved with more sophisticated predictive models. A fundamental model, in simplified form, with adjustable parameters, was developed and its accuracy to predict DP based on given operating conditions was tested using available plant data. This model was built into a control structure for implementation on an operational batch digester.

Because the parameters form part of fundamental relationships in the model, realistic bounds can be placed on them during the optimisation process, enabling a better understanding of the behaviour of the model.

Overall control of the batch digestion process takes place in three interdependent, but distinct time frames, namely continuous control of digester temperature based on a calculated setpoint, using a model-based approach to predict DP on a cook-to-cook basis and lastly while the final DP must be kept on target through a number of cooks (N cooks), taking into account long-term changes in the process. This long-term optimisation takes place in the third time frame.

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1. Introduction

In the production of dissolved pulp, special emphasis is placed on the quality of the pulp, especially with respect to fibre length. Unlike many other processes, where back-blending of product can ensure adherence to product quality requirements, this is not possible in the case of dissolved pulp production. Although modern pulp manufacturing plants make use of continuous pulping digesters, several large plants are still operational, where batch digesters are used for production of dissolved pulp.

To enable production targets, the control challenges on such a plant require innovative approaches to ensure sustained trouble-free operation.

The control challenges to be addressed are:

• Each digester operates on a batch cycle over an operating range.
• The controlled variable for quality, the degree of degradation of the pulp, is not continuously measurable.
• Only one manipulated variable, steam supply rate, is available.
• In order to ensure continuous supply of pulp to the downstream processes, strict requirements exist for each digester.
with respect to available time within which to complete cooks.

- The controller has to operate within specific constraints, which include limits on available steam and continuously varying heat transfer area in addition to unmeasured disturbances associated with liquor strength, wood chip composition, moisture content, etc.

The conventional approach towards control of pulp viscosity is to estimate cook duration based on a prediction of required energy which is in turn based on a simplified reaction rate for delignification of the wood chips. This methodology is commonly known as the S-factor approach, in the case of dissolved pulp production.

In order to achieve tighter control of pulp viscosity under the stated specifications, a mechanistic model, with adjustable parameters was used to achieve improved control of pulp viscosity. In this paper it is shown that substantially improved control is possible by making use of a more sophisticated mechanistic model, and applying optimisation techniques to adjust model parameters, ensuring tighter control.

2. Traditional control of pulp viscosity

The SAPPi SAICCOR plant in Umkomaas, KwaZulu-Natal in South Africa produces dissolving pulp for the production of a wide range of products. Acid sulphite pulping in 23 batch digesters is used to produce the pulp from mainly Eucalyptus wood. Since dissolving pulps are produced, the aim of the process is to produce pulp with a specified final degree of polymerisation.

Temperature in the digester is used to control the cellulose degradation, while the steam flow to the external heat exchangers is used to control digester temperature in a cascaded arrangement as shown in Fig. 1.

During the manufacture of dissolving pulps, the controlled variable is the degree of polymerisation (DP) of the cellulose, which gives an indication of the average length of the polymer chains, but which cannot be measured directly. DP is strongly affected by temperature and pH.

pH is rarely measured on a batch plant, because of the cost of installing and maintaining pH probes on the large number of individual digesters. The DP of pulp is related to the intrinsic viscosity and can be determined by measuring the viscosity of a pulp sample. This method is simple and relatively quick and is most commonly used for the estimation of the DP value of the cellulose. The correlation between the DP value and the measured viscosity was determined by Watson (1992) and was found to be:

\[
\text{DP} = 285.4\mu^{0.345}
\]  

This correlation is valid over a range of 20 cP < \(\mu\) < 150 cP.

Control of temperature is accurate, indicating that the control problem is associated with the difficulty in determining the temperature setpoint profile for the temperature controller. A modified S-factor model, which is currently in use at SAIC-COR relates the degree of polymerisation to temperature in the digester in order to be able to calculate the temperature setpoints that will ensure obtaining the desired DP in the available cook time.

The S-factor model was originally developed by Yorston and Liebergott (1965) and was based on the assumption that a correlation exists between the lignin content of the solid phase in the reaction and the pulp viscosity. The widely accepted delignification rate equation, in the form also later reported by Hagberg and Schöön (1973), was used to model delignification and relate the pulp viscosity to that. The Arrhenius temperature dependency of the reaction had already been confirmed and this was used in the following equation:

\[
\text{SF} = -\int_{L_0}^{L_f} \frac{1}{A_0[L]} d[L] = \int_{t_0}^{t_f} e^{\left\{-\frac{B}{RT}\right\}} dt
\]  

Integration of the left-hand side of this equation yields:

\[
[L_f] = k_1 e^{\left\{-A_0\text{ SF}\right\}}
\]  

From the initial assumption, the residual lignin concentration was taken to be proportional to the cuprammonium viscosity of the pulp, leading to the following:

\[
\mu = k_2 e^{\left\{-A_0\text{ SF}\right\}}
\]  

The parameters \(k_1\) and \(k_2\) are constant. Eq. (4) formed the basis of the “S-factor” model. Individual mills adapted the S-factor model to their own data and observations in order to improve the control of their respective plants, for example using the actual SO\(_2\) concentration of the liquor instead of the partial pressure of the SO\(_2\) (Marr & Bondy, 1986).
The weakness of the S-factor model lies in the assumption that the residual lignin content is proportional to the degree of polymerisation of the cellulose.

The basis for any new control strategy would be a model that can relate the degree of polymerisation to the manipulated variable or variables. This will be necessary, since the degree of polymerisation cannot be measured continuously. The digester conditions as well as the degree of polymerisation must therefore be simulated and the control done based on the predictions of this simulation. It was shown (Kilian & de Vaal, 2000) that the best variable to use for the control of the final degree of polymerisation will be the temperature profile. That will also be the manipulated variable used in this study. A model is therefore necessary to relate the temperature to the degree of polymerisation.

3. Improved model-based approach

3.1. The mechanistic model

A lot of detail is necessary in a fundamental model to define the system completely, which can make the task of fundamental modelling very complicated. Empirical modelling can be a good alternative to fundamental modelling if the necessary equipment is available. It can often be performed more quickly and requires only minimum insight into the process. As both modelling approaches have their obvious drawbacks, they are frequently combined to a certain extent, in order to obtain meaningful models of limited complexity with minimum modelling effort, while still covering the operating regime of interest for the optimisation.

The approach taken during this study for the modelling of the digester was to combine the empirical and fundamental modelling techniques. A general fundamental model was constructed from the knowledge of the acid sulphite process and the reactions taking place during the pulping process. This model structure contained the basic reaction mechanisms, as well as the relevant equilibrium conditions. However, one of the biggest problems during fundamental modelling is to obtain accurate parameters for the various reaction rate equations and the corresponding equilibrium conditions. The model was therefore created with the ability to easily adjust the parameters. The model was then fitted to experimental and actual plant data and the parameters computed to best fit the real observations. Computer software was written to assist in the initial process of adjusting the parameters in order to obtain an accurate model.

The software programme facilitates the process of constructing an accurate process model over a batch pulp digester. However, the basic model structure has to be in place already before the package can be used to fit the model to actual data. This implies that the programme doesn’t apply process identification techniques or artificial intelligence to fit a black-box model to the process. It is used as a tool to fine-tune the parameters of an already developed model, as well as for the verification of the model and is generally applicable to batch pulping processes, and by changing the process model used, could also be used for other batch processes as well.

Development and verification of the digester model was divided into three parts, namely:

- the kinetics of the reactions involving the wood constituents,
- the liquor composition, and
- the pressure and the temperature inside the digester.

For verification of the model, the first two parts were considered together and the third part separately. The reaction kinetics and the liquor composition were verified using data obtained from a pilot scale facility (Watson, 1992), since intermediate measurements of the wood and liquor conditions were possible using the laboratory equipment. Modelling of the actual conditions inside the digester were verified with real plant data. It was reasoned that if the verification of the kinetics, using known conditions for the modelling of the reactions, shows that the kinetic model is correct and the verification of the modelling of the conditions inside the digester give good results, then the entire model can be assumed to be correct.

The software programme was used to do the verification of the model, as well as to simplify the adjustment of the parameters in the model to obtain a model that represents the real-world situation.

The approach taken in this study was similar to the one mentioned by Terwiesch and Agarwal (1995). This approach requires that the basic model structure be chosen from first-principles considerations, while the model parameters are typically computed so that the model predictions best fit the data. The software therefore allows for the parameters in the model to be adjusted so that the modelling results best fit the data.

The most important variable in the model for control purposes, is the degree of polymerisation of the cellulose. The primary aim is therefore the accuracy of the simulation of the degradation increase. However, it will also be necessary to verify the correctness of the other parameters. This enables one to locate the source of the largest deviation from the real values. Therefore, if large errors in the degree of polymerisation occur, the variable that is mainly responsible for the errors can be located. The possibility to adapt the model for the variables that give the worst results is then created.

The correctness of the following variables were verified:

(i) Liquor samples:
- bisulphite ion concentration;
- pH of the liquor.

(ii) Pulp samples:
- lignin content;
- hemicellulose content;
- degree of polymerisation of the cellulose.

The simulation was done using the software package that was created. After the simulation had been completed, the measurements taken on the mini digesters were compared with the calculated values from the simulation at the corresponding times. A correlation coefficient was then calculated for each variable.
The derivation of a simplified mechanistic model (Kilian & de Vaal, 2000), used by the controller can be summarised by the following assumptions:

- The reactions in the digester can be accounted for by modelling the rate of cellulose degradation, lignin dissolution and strong acid equilibria in the digester. The hemicellulose degradation rate is implicitly calculated in these equations.
- The initial composition of the reaction mixture is known from a test of the liquor.
- The temperature and pressure in the digester are known.

To model the reactions, the following reaction rates are used:

\[
- \frac{d[L]}{dt} = k_L[L]^\alpha[\text{HSO}_3^-]^\beta[H^+]^\gamma \quad (5)
\]

\[
\frac{d[\text{SA}^-]}{dt} = \left\{ \frac{g}{v} + \frac{2h}{v} \left( [L]_0 - [L] \right) \right\} r_L + \frac{k_{SA}(T)}{v} \left( [L]_0 - [L] \right)[\text{HSO}_3^-]^\beta[H^+]^\gamma \quad (6)
\]

\[
- \frac{d[C]}{dt} = k_C[H^+]^\delta \quad (7)
\]

The temperature dependence of all the k factors above, can be described by the Arrhenius relationship, \( k = k_0 e^{-E/T} \), with \( E \) and \( k_0 \) known for each of these reactions.

The hydrogen and bisulphite ion concentrations are obtained via electroneutrality arguments combined with the assumption that equilibrium is attained between the SO2 in the vapour and liquid phases, as shown below:

- Chemical equilibrium:
  \[
  K_{\text{SO2}} = \frac{[H^+][\text{HSO}_3^-]}{[\text{SO}_2]} \quad (8)
  \]

- Vapour–liquid equilibrium:
  \[
  K_p = \frac{[H^+][\text{HSO}_3^-]}{P_{\text{SO2}}} \quad (9)
  \]

- Electron neutrality:
  \[
  [M^+] + [H^+] = [\text{HSO}_3^-] + [\text{SA}^-] \quad (10)
  \]
  \[
  [M^+] = 2[\text{combined SO}_2] \quad (11)
  \]

A further relationship required is for the total pressure of the system:

\[
P = p_{\text{SO}_2} + p_{\text{H}_2\text{O}} \quad (12)
\]

The two temperature-dependent equilibrium constants can be calculated from parameters in the shape of the Antoine equation, as follows:

\[
\log_{10} K_{\text{SO2}} = B_{\text{SO2}} + C_{\text{SO2}} T \quad (13)
\]

\[
\log_{10} K_p = B_p + C_p T \quad (14)
\]

In the 10 independent Eqs. (5)–(14), there are 15 variables, this leaves five variables to be specified.

The temperature and pressure are read from the environment during a cook and specified by approximations during the simulation. At the start of a cook, the liquor-to-wood ratio is known and the free and combined SO2 concentrations are read during the 1-h test. Specifying these five variables completely specifies the problem.

The variables defining the digester model are given in Table 1 and the associated parameters in the simplified mechanistic model in Table 2 below.

<table>
<thead>
<tr>
<th>Variable</th>
<th>Symbol</th>
<th>Units</th>
<th>Typical range</th>
</tr>
</thead>
<tbody>
<tr>
<td>Residual lignin concentration</td>
<td>([L])</td>
<td>mass%</td>
<td>0</td>
</tr>
<tr>
<td>Cellulose concentration</td>
<td>([C])</td>
<td>kmol/m³</td>
<td>300</td>
</tr>
<tr>
<td>Sulfite concentration</td>
<td>([\text{HSO}_3^-])</td>
<td>kmol/m³</td>
<td></td>
</tr>
<tr>
<td>Hydrogen ion concentration</td>
<td>([H^+])</td>
<td>kmol/m³</td>
<td></td>
</tr>
<tr>
<td>Strong acid concentration</td>
<td>([\text{SA}^-])</td>
<td>kmol/m³</td>
<td></td>
</tr>
<tr>
<td>Temperature</td>
<td>(T)</td>
<td>K</td>
<td>288</td>
</tr>
<tr>
<td>Pressure</td>
<td>(P)</td>
<td>Pa</td>
<td>103</td>
</tr>
<tr>
<td>Partial pressure of SO2</td>
<td>(p_{\text{SO2}})</td>
<td>Pa</td>
<td>700</td>
</tr>
<tr>
<td>Partial pressure of water</td>
<td>(p_{\text{H}_2\text{O}})</td>
<td>Pa</td>
<td>0</td>
</tr>
<tr>
<td>Equilibrium constant of SO2</td>
<td>(k_{\text{SO2}})</td>
<td>kmol/m³</td>
<td></td>
</tr>
<tr>
<td>Vapour–liquid equilibrium constant</td>
<td>(K_p)</td>
<td>(kmol/m³)²/Pa</td>
<td></td>
</tr>
<tr>
<td>Combined SO2</td>
<td>Combined SO2</td>
<td>kmol/m³</td>
<td></td>
</tr>
<tr>
<td>Free SO2</td>
<td>Free SO2</td>
<td>kmol/m³</td>
<td></td>
</tr>
<tr>
<td>Total positive charge</td>
<td>([M^+])</td>
<td>kmol/m³</td>
<td></td>
</tr>
<tr>
<td>Liquor:wood ratio</td>
<td>(v)</td>
<td>l/kg</td>
<td>3:1</td>
</tr>
</tbody>
</table>

Table 1: Variables defining the digester model
Table 2
Parameters representing the simplified mechanistic model

<table>
<thead>
<tr>
<th>$\lfloor L \rfloor &gt; 12.4$</th>
<th>$\lfloor L \rfloor &lt; 12.4$</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Lignin reaction kinetics</strong> (Hagberg &amp; Schön, 1973)</td>
<td></td>
</tr>
<tr>
<td>$k^0_L$ (%$g\text{m}^{-1}$)</td>
<td>$1.346 \times 10^{12}$</td>
</tr>
<tr>
<td>$E_L$ (K)</td>
<td>12,500</td>
</tr>
<tr>
<td>$a$</td>
<td>0.646</td>
</tr>
<tr>
<td>$a$</td>
<td>0.819</td>
</tr>
<tr>
<td>$\beta$</td>
<td>0.705</td>
</tr>
<tr>
<td><strong>Cellulose reaction kinetics</strong> (Kilian &amp; de Vaal, 2000)</td>
<td></td>
</tr>
<tr>
<td>$k^0_C$ (K)</td>
<td>7.68 $\times 10^{12}$</td>
</tr>
<tr>
<td>$E_C$ (K)</td>
<td>17,100</td>
</tr>
<tr>
<td>$\delta$</td>
<td>1</td>
</tr>
<tr>
<td><strong>Strong acids reaction kinetics</strong> (Kilian &amp; de Vaal, 2000)</td>
<td></td>
</tr>
<tr>
<td>$k^0_{SA}$ (kmol/m$^3$)</td>
<td>$1.887 \times 10^{16}$</td>
</tr>
<tr>
<td>$E_{SA}$ (K)</td>
<td>19,860</td>
</tr>
<tr>
<td>$q$</td>
<td>2.2231</td>
</tr>
<tr>
<td>$b$</td>
<td>1.681</td>
</tr>
<tr>
<td>$c$</td>
<td>0.653</td>
</tr>
<tr>
<td>$g$</td>
<td>$6.87 \times 10^{-3}$</td>
</tr>
<tr>
<td>$h$</td>
<td>$-1.310 \times 10^{-4}$</td>
</tr>
<tr>
<td><strong>Temperature range (°C)</strong></td>
<td>$B_{SO2}$</td>
</tr>
<tr>
<td>Antoine coefficients (Kilian &amp; de Vaal, 2000)</td>
<td></td>
</tr>
<tr>
<td>$K_{SO2}$</td>
<td></td>
</tr>
<tr>
<td>20–80</td>
<td>$-5.077$</td>
</tr>
<tr>
<td>80–120</td>
<td>$-6.916$</td>
</tr>
<tr>
<td>120–150</td>
<td>$-7.971$</td>
</tr>
<tr>
<td><strong>Temperature range (°C)</strong></td>
<td>$B_p$</td>
</tr>
<tr>
<td>Antoine coefficients (Kilian &amp; de Vaal, 2000)</td>
<td></td>
</tr>
<tr>
<td>$K_p$</td>
<td></td>
</tr>
<tr>
<td>20–80</td>
<td>$-9.430$</td>
</tr>
<tr>
<td>80–120</td>
<td>$-10.208$</td>
</tr>
<tr>
<td>120–150</td>
<td>$-10.948$</td>
</tr>
</tbody>
</table>

3.2. The model-based controller

The control objective of any controller can be seen as keeping the effect of disturbances small while allowing the controlled system to produce the specified outputs, as well as allowing the system to move between desired outputs in a satisfactory manner. This leads to a conceptual picture of the control process required for the digester.

The controller is based on a general feedforward structure, as shown in Fig. 2. It should be clear that some form of inverse model is required as well as all feedforward control. This inverse is obtained iteratively by obtaining the value of $T_{top}$ that gives $DP = DP_{set}$. This is due to the fact that a highly non-linear time domain model has already been developed, and finding this inverse is computationally intensive.

3.3. Adaptive model

3.3.1. Control time frames

Control time frames are an important aspect of layered control systems. Control of the digester takes place in three different time frames. The first is the continuous time frame. Control on this level involves keeping the top temperature ($T_{top}$) on a temperature setpoint. Control of the final DP takes place per cook, representing the second time frame, while the final DP must be kept on target through a number of cooks ($N$ cooks), taking into account long-term changes in the process. This long-term optimisation takes place in the third time frame. The second and third time frames both encapsulate discrete control problems with small sampling rates.

In Figs. 2 and 3, increased sampling frequency is indicated by thicker lines connecting units.

3.3.2. Continuous temperature control

This control configuration corresponds to the normal feedback controller setup. The temperature in the digester is controlled by manipulating the steam flow rate to the external digester. This first level of control is already handled satisfactorily by the feedback controllers on each digester.

The data read continuously are used directly in the process model. They are not stored during the cook process, as this was found to be too slow to allow real time processing. During the control simulation, feedback is given to the personnel using the software by updating graphics on the user interface showing the current temperatures, side circulation flow rate and pressure read from the plant, as well as the values of the calculated pH and estimated DP.

The different phases of the control problem require separate action. The initial phase, until the 1 h test becomes available, simply stores data during this phase. Once the data becomes available, it is used to deduce the starting conditions of the digester.

3.3.3. Per cook DP control

The temperature setpoint for the low-level temperature control system is supplied by a profile generator which ensures that a predefined temperature profile is followed. The profile generator requires the value of $T_{top}$; the constant top temperate for the cook, in order for it to generate a complete profile.

The final DP of the cook is determined by the temperature profile that the digester is taken through. Even though the profile generator determines this profile, the value of $T_{top}$ is specified externally. It should be clear that the value of $T_{top}$ therefore determines the value of the final DP to a great extent.

3.3.4. N cook time frame

The results of a number of cooks are collected using a historical system. These stored results can be analysed to check the performance of the system over time. New model parameters can also be passed to the DP controller so that long-term changes in plant behaviour can be accommodated. The diagram shown in Fig. 2 shows the inputs and outputs of the $N$ cook time frame.

In this time frame, the controller responsible for outputting the correct value of $T_{top}$ to the profile generator is shown. This is the UP controller. The model parameters used in the controllers are changed every $N$ cooks, placing this update in the $N$ cooks time frame. It should be borne in mind that the controller reads process variables from the plant with other sampling frequencies as well. Fig. 2 is intended to show the position of the controller...
in the time frames, but this is by no means a unique representation.

Ultimately all control action is implemented by manipulation of the steam flow rate, but the low level control systems in place to control the temperature profiles are working well enough so that the controller just has to concern itself with determining the optimal top temperature that will result in the correct DP being obtained.

After $N$ cooks have passed (at the moment $N = 10$), the parameters have to be optimised to ensure the best fit between the
Fig. 4. Flow diagram for the control strategy.
model predictions and the real results of the previous $N$ cooks. This is done using Matlab’s built in optimisation routines which are based on the Simplex method.

Upon examination of Fig. 2, it is clear that the controller does not have any specific measurement of control error. When the conditions on the plant change, causing plant-model mismatch, control can be expected to degrade. In order to keep the model parameters in line with the measured results from the plant, it is necessary to provide a feedback mechanism. This is done using a periodic optimisation of the model parameters. At a predetermined frequency, the model parameters are adapted by optimising the fit of the model to the measured DP values for the cooks since the last optimisation. This ensures that the model is constantly in close accord with the conditions on the plant. This is illustrated in Fig. 3.

3.3.4.1. Plant interaction. Interaction with the plant was achieved using direct queries to the plant database to obtain all the data required for operation. Both continuous and historical data can be obtained in this manner, so that operation is entirely automated.

Algorithm. The controller described above is implemented using the following algorithm:

1. Initial model parameters are loaded from a file. The model parameters stored in file are listed in tables.
2. The program reads status information from the plant, waiting for a cook to start.
3. The database system stores a state tag which is read to determine whether a cook has started.
4. As soon as the start of a cook has been detected, the controller reads plant data and stores the information internally. Note that no simulation is done in this phase, as the initial conditions for the integration of the model equations are only available after roughly an hour, once the analysis of the liquor sample as been done.
5. As the program reads data from the plant, it checks whether the results of the 1-h test have become available. This test supplies the liquor composition at the time of the test.
6. When the 1-h test results are available, the controller program simulates the cook up to the current point in time using the stored process variables and initial values calculated from the test results. The simulation is done by integrating the differential equations given in Eqs. (4)–(6) and solving the remaining equations to yield all the variables constituting the pulp degradation model.
7. The program then simulates reactions inside the digester in real time using the same equations, incorporating real time values of the process variables (temperature, pressure and the flow through the heat exchanger) that are available from the plant database system.
8. When the temperature reaches a predetermined temperature, $T_C$, the program starts an optimisation routine that simulates the remainder of the cook using the current states as the initial conditions for each trial cook. The optimisation finds a value for $T_{\text{top}}$ that will minimise the viscosity error at the end of the cook.
9. This temperature is displayed on the screen for entry by an operator into the plant control system. The plant control system uses this value to ensure that the temperature is ramped up to $T_{\text{top}}$ in the required time.
10. The previous two steps are repeated 5 min after the first prediction to verify the correct operation of the predictor.
11. The model equations are integrated up to the end of the cook.
12. In intervals of 10 cooks, the data for the last 10 cooks are read from the plant database, and a separate optimisation process is initiated to reduce the error between the measured DP and the DP predicted by the model for each of these cooks.

This control strategy is shown in a flow diagram in Fig. 4.

4. In situ control results

4.1. Results

The UP controller was used to control 13 cooks during the period from May 2003 to June 2003. Fig. 5 shows the performance of the S-factor controller during the same period. The horizontal dashed line shows an average deviation from target slightly above 10. Note that the DP cycles dramatically between high and low extremes during this time period. This is typical of the overall operation of this controller.

Several reasons for this cyclic behaviour can be suggested. The most likely explanation is aggressive modification of the controller parameters by plant personnel, without waiting for the effects of these modifications to become apparent.

Fig. 6 shows the performance of the UP controller during the same time period. The vertical line shows the point at which new model parameters were obtained using the automatic tuning facility. When comparing the two controllers, it should be noted that the S-factor controller parameters are changed for almost every cook. This makes it difficult to identify contiguous regions

![Fig. 5. Deviation from target for S-factor controller, May to June 2003.](image-url)
where the same parameters are used, as was done in the figure showing UP controller performance.

The figures show clear evidence of the improved control performance of the UP controller. After parameter tuning, the deviation from target was significantly lowered, and the variance is consistently lower than that of the S-factor controller during the same time period.

4.2. Variance reduction

The variance of the S-factor controller during the test was measured as 21. The UP controller variance was 18 before parameter tuning and at 12 after parameter tuning. Results derived from a simulated model of the process (Sandrock & de Vaal, 2003) suggest that this figure can be expected to drop further when the parameters are tuned again. Even if the variance does not improve, the UP controller shows a 43% lower variance than the current controller.

4.3. Error reduction

The S-factor controller showed an average error of 10 units. The high average error can be explained by the fact that personnel are loath to obtain low viscosities. Given that the S-factor controller exhibits a high variance, average viscosities are consistently high to avoid low viscosity results. The UP controller showed an average error of only one unit after parameter tuning. Automatic tuning of parameters will keep the error low.

5. Conclusions

A new model-based control algorithm that interacts with an industrial batch digester was developed and has been implemented on one of the batch digesters in operation. The control predictions made by the new controller led to improved control compared with the current controller which is based on a modified S-factor model.

A number of issues were identified during the course of this investigation:

- The model used by the UP controller does not account for all the variability observed on the plant. If these sources of variability can be identified, they might provide valuable clues as to further ways of reducing the output variance.
- The possibility of adding additional measurements is also an interesting field for further study. One additional measurement such as near-infrared spectrum analysis or moisture content measurements of wood entering the digester could improve control by allowing the model to use correct initial values for the model rate equations.
- If the controller can obtain optimised parameters for different species of wood, these parameters could be stored for recall when switching wood species.
- Control can be improved considerably by removing the constraint that the cook time needs to be constant. The control algorithm can be easily adapted to supply an optimal time for a constant temperature profile instead of an optimal temperature for a constant time cook.
- In order for variable time cooks to be possible, scheduling on the plant has to be reworked. The current constant time constraint is due to the difficulty of manually scheduling the digesters taking into account shared equipment and utilities. If the scheduling could be automated, the constant time constraint could be relaxed.

References


