FINAL REPORT

Project Title: Model-Based Approach to Soft-Sensing and Diagnosis for Control of a Continuous Digester

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Project Objective: The goal of the work was to develop and demonstrate computing based modeling and control methodologies that will facilitate integrated operations on the continuous pulp digester. The technical work required to achieve these goals included development of efficient methods for soft-sensing using fundamental models, integration of fault monitoring and control algorithms, and the development of computationally feasible formulations of model predictive control for profile management in the digester. The developed operational methodology for digester grade transition control was benchmarked against an industrial design in cooperation with our collaborators at Weyerhaeuser and Westvaco.

Background: The pulp and paper industry, a large and important sector of the CPI, is the most capital intensive among the sectors. While other components of the chemical industry have rapidly exploited model-based methods for improved operation, the PPI has lagged behind. Furthermore, as the US paper industry is moving from high volume single grade production to multi-grade continuous production, which offers the possibility for faster response to customer needs and production of higher margin specialty products, there arises the technical challenge of controlling the plant under the transient conditions which tend to propagate through the entire

mill, and hence influence other processes. This can lead to production losses, increased variability in product quality, and increased environmental loads.

In this project, we focused on a key bottleneck unit in the pulp mill flow sheet - the pulp digester. Digesters are very capital intensive (\$50-\$100 million), and their performance is of paramount importance to maximize the produced pulp quality and yield, reduce the overall operating costs and minimize the adverse environmental impacts of pulp mills. With more pulp and paper companies replacing their pulping processes with modern fiber lines using continuous digesters to meet increasing competitiveness in the global market and tighter environmental regulations, there is an increasing need for improved control of continuous digesters. One of the key technical challenges to operation of this unit is the management of production rate changes and grade swings between hardwood and softwood feedstocks.

Status:

I. FUNDAMENTAL MODELS OF CONTINUOUS DIGESTER

I.1 Digester Model - Summary

Continuous pulp digesters are large moving bed reactors with significant residence time. They play a critical role in shaping the pulp and paper qualities achieved at the product-end of a mill and significantly influence energy balance and environmental burden of the operations. There is renewed interest and much progress towards developing fundamental dynamic models to be used for better process understanding and especially in the development of model based controllers.

One of the challenging areas in fundamental model development is the incorporation of grade transition with dynamic compaction (in packed bed of chips) and the transport/conversion effects due to chip size distribution. The challenge arises not from the theoretical understanding of how to include these factors into the fundamental model, but from the considerable size of the resulting equations that can easily exceed 100,000 ODEs representing the dynamic behavior.

Technical aspects of this work are focused on the numerical methods for reducing the model order to a tractable size so that process dynamics can be captured with a fast computational approach without any loss of information. Overall utility of this high fidelity stochastic dynamic model is in its ability to capture concealed process variability that could not be quantified with a distributed parameter approach constructed with hidden

lumped parameters, as it is typical in sequential stirred tank approximations of plug-flow reactors.

Although the fundamental model has many parameters associated with reaction rate and transport correlations only two adjustable parameters are needed to “tune” the model to represent an operational digester. Model tuning is done using data representing typical steady state behavior, yet validation studies show that the dynamic predictions of the model are also on target and reliable. The simulation code developed with the model is a practical tool for both research purposes and for practical needs of mill engineers in improving process performance.

I.2 Digester Model – Process Description

The particular digester design chosen for this project is the dual vessel EMCC (extended modified continuous cooking) arrangement as shown in Figure 1.

Wet chips are steamed to remove air in the pores and fed into the impregnation vessel (IV) together with white liquor. In the impregnation vessel, white liquor penetrates into the chips and equilibrates with initial moisture for about 30 minutes depending on the production rate. In the IV, both chips and liquor move in the co-current downward direction.

From the IV, the chips are carried into the top section of the digester with hot liquor that brings the mixture to the desired reaction temperature. The top section of the digester, referred to as the cook zone, is a co-current section where the main reactions take place. Chips react from inside out owing to the significant internal pore volume and associated surface area. Therefore, overall reaction rates depend on the concentration levels of entrapped liquor and the diffusion rates from free liquor that replenishes the active ingredient holdup in the pores. Spent liquor saturated with dissolved solids at the end of the cook zone is extracted for chemical recovery elsewhere in the mill. Chips follow into the MCC (modified continuous cooking) and the EMCC zones, now counter-current to fresh dilute white liquor that simultaneously continue mild delignification reactions and extract valuable dissolved solids from the pores of chips.

Figure 1. Dual Vessel EMCC continuous digester.

As packed reactors, digesters are very unique in that the packing (main ingredient of the process) is continuously in motion, non-uniform in size and undergo through variable compaction both with respect to conversion and differential head pressure. Extent of reaction, defined through the blow-line (exit) Kappa number, is the major performance measurement. Other important factors are the yield of the process and the fiber properties of the final product. Although various operating conditions may yield the same Kappa number, important fiber properties like strength are reaction path dependent.

I.3 Digester Model – Features of the New Generation Model

1. Solid phase moves in distinct “plugs” downward in the digester satisfying the experimental evidence of plug-flow behavior. Plug size can be as small as desired for numerical accuracy. For example, about 1000 solid plugs in a 6hr residence time digester capture 0.36 min (~22 seconds) flow intervals.

2. Chip size distribution can be represented by discrete size mass fraction allocations. Usually 5 to 7 size cut categories are sufficient to represent practical differences between chip thickness variations. Each

plug has its own representative mass fractions for predefined size cut dimensions.

3. Solid, entrapped liquor and free liquor are the three distinct phases in transport and reaction calculations. Liquor-solid reactions occur between solid segments of each chip size and the entrapped liquor contained within. Diffusion between entrapped liquor segments and free liquor determine liquor density (composition) gradients and replenishment via free liquor. Solid volume segments are represented in equivalent spherical shapes and multiple discrete "shells" are used to capture diffusion and reaction interactions within large solid particles (see Figure 2).

4. Each plug entering the digester system carries its own distinct set of physical, chemical and transport properties as model parameters. Thus, there may be as many simultaneous wood species in the digester at any time as there are distinct plugs.

5. Solid compaction is a computed dynamic state for each plug that depends on chemical and hydrodynamic properties of the digester.

Figure 2. Chip size distribution and the approximation of diffusion effects on delignification rates by discrete treatment of multi-volume compartments for "larger" chips using spherical geometry with equal total chip volume.

I.4 Digester Model – Summary of Model Equations

Governing material and energy balance equations of the new generation model are similar to those reported by Wisniewski [5]. Details will be kept to a minimum here due to space limitations. It is worth pointing out that, to track chip size and diffusion rate limitations, inter-particle solid and liquor density gradients are calculated in a manner similar to Gustafson [2] as compared to the lumped-parameter approach of [5]. Furthermore, dynamic compaction calculations are done with similar assumptions and correlations as in Michelsen [6], which are originally proposed by Harkonen [8].

Reaction rates for solid densities are specified as

$$R_{Si} = \frac{1}{\tau_{Si}} \left[k_{Ai} \left(\frac{C_{Li}}{C_{Li}^0} \right)^{1/2} + k_{Bi} \left(\frac{C_{Li}}{C_{Li}^0} \right)^{1/2} \right] (C_{Si} - C_{Si}^0)$$

$i = 1, \dots, 5$

with reaction rate constants as

$$k_{Ai} = k_{Aoi} \exp(-E_{Ai}/RT) \text{ and } k_{Bi} = k_{Boi} \exp(-E_{Bi}/RT).$$

Liquor component rates are related through stoichiometric relationships as

$$\begin{aligned} R_{L1} &= \eta_{EAL} R_{LG} + \eta_{EAC} R_C + \eta_{L1} \frac{\eta_{L1} \eta_{L2} \eta_{L3}}{\eta_{L1} \eta_{L2} \eta_{L3}} \\ R_{L2} &= \eta_{HSL} R_{LG} + \eta_{L2} \frac{\eta_{L1} \eta_{L2} \eta_{L3}}{\eta_{L1} \eta_{L2} \eta_{L3}} \\ R_{L3} &= \eta_{LS} R_S + \eta_{L3} \frac{\eta_{L1} \eta_{L2} \eta_{L3}}{\eta_{L1} \eta_{L2} \eta_{L3}} \\ R_{L4} &= \eta_{LG} R_{LG} + \eta_{L4} \frac{\eta_{L1} \eta_{L2} \eta_{L3}}{\eta_{L1} \eta_{L2} \eta_{L3}} \end{aligned}$$

Where $R_{LG} = R_{S1} + R_{S2}$, $R_C = R_{S3} + R_{S4} + R_{S5}$ and $R_S = R_{LG} + R_C$

Entrapped liquor density diffusion coefficient is

$$D = 8 \times 10^{-8} T^{1/2} \exp(-2450/T)$$

Chip compaction under pressure is given by the Harkonen [8] correlations as

$$\begin{aligned} \eta &= 0.644 + [0.831 + 0.139 \ln(Kappa \#)](P/10)^{0.59} \\ \eta \frac{P}{L} &= 4.6 \left(\frac{1}{\eta} \right)^2 v + 3900 \left(\frac{1}{\eta} \right) v^2 \end{aligned}$$

1.5 Digester Model –Fitting with Mill Data

Previous applications of the digester model have shown that only two adjustable parameters are sufficient to fit the model to a specific mill data. These two parameters are the effectiveness factors that act as the scaling parameters associated with the overall reaction rate terms and the diffusion rates as depicted below. The pre-multiplying factors η_1 and η_2 are the two adjustable tuning parameters. Experience has also shown that the former, the reaction rate effectiveness factor, is more influential than the latter. Therefore, in any model-fitting situation one fits η_1 first with η_2 set at 1, and then fine-tunes η_2 as necessary. In most applications the two adjustable parameters are usually between 0.6 and 1.0.

$$\begin{aligned} R_{Si} &= \eta_1 [k_{Ai} \eta_{L1} + k_{Bi} \eta_{L1}^{1/2} \eta_{L2}^{1/2}] (\eta_{Si} \eta_{Si}^{\eta}) \\ i &= 1, \dots, 5 \end{aligned}$$

$$D = 8 \times 10^{-28} T^{1/2} \exp(2450/T)$$

Figure 3. Dynamic response match and model validation.

1.6 Digester Model – Practical Applications of the Simulation Code

The digester model can be used for a variety of applications including controller design and development, operator training, and process improvement. We want to briefly focus here on the potential applications of the model first for process improvement studies and then for transition swing decisions.

Considering the needs of a process engineer we will demonstrate that the model and specifically the GUI (Graphical User Interface) can be formulated to create a practical and easy tool for process simulations with the purpose of performance improvements. We will take two different approaches that will complement each other. One approach will be to provide a GUI option so that the simulation model can be run by deciding process variables in a similar fashion to the choices of a process operator. The other approach will provide an alternative GUI option where the simulation can be run by deciding on performance results directly.

There are basically 12 decision variables that determine the performance of the simulation model. These are flowrates (chip, sluice, total extract, blow, wash, white liquor, sluice trim or make-up white liquor, MCC trim, EMCC trim) and temperatures (cook, MCC, EMCC). From a production point of view only two performance variables are of paramount importance, production rate and blow line Kappa number. For a non-

linear process like the continuous digester there are numerous combinations of the operating conditions that will give the desired final Kappa number at the target production rate. Among these multiple choices some are obviously better for product quality than the others. We can raise two questions: (1) how does one find feasible operating conditions, and (2) how does one choose better operating conditions among the many that are feasible? The two alternative GUI options mentioned above for running the simulation model are designed to answer these two questions respectively.

Transition management for continuous digesters is another one of the many challenges faced by the operators and process engineers alike. One way to approach the problem is to design a simple policy and then follow it through the transition (swing) period. An example is provided here for Pine (softwood) to hardwood transition. Initially we need to know both the starting and final nominal operating (decision variables) conditions. Obviously, the initial nominal conditions are known by default. Expected final nominal conditions should also be known to a certain degree of accuracy from past experience and confirmed by simulation studies. Once the two end points are determined, then the remaining choices are the paths to follow for all of the variables between their corresponding two states. A simple policy is to ramp the decision variables from one state to the other after a period of appropriate delay following the initiation of specie swing. Delay is needed to approximately synchronize a change with the moving front of the chips and the ramp is needed to smoothly transition from one state to the other. Thus, from an optimization point of view every decision variable has two degrees of freedom, delay and ramp rate. A simulation code is very helpful in finding optimum transition decisions, which can be computed either through a rigorous optimization algorithm or through a trial-and-error search where some heuristic decisions can be made to decide between the merits of similar conditions.

I.7 Digester Model - Conclusions

A new generation continuous digester model is developed to provide high fidelity simulation capabilities specifically suitable for specie and rate transition dynamics. Undesirable "filtering" effects of CSTR approximation and numerical difficulties associated stiff ODEs are avoided by using moving solid plugs with a cinematic algorithm for numerical solutions. Dynamic compaction and chip size distribution effects are easily incorporated in the model. On a P4 2.8GHz PC the model operates with a detailed animation GUI at a speed about 350 times faster than real process time; i.e. 60 minutes of digester behavior under full dynamic conditions and at 5 minute sampling intervals can be simulated and its

animation displayed in less than 10 seconds. Mill data are used to tune and validate the model where it is shown that only two simple scaling factors, associated with reaction and diffusion rates, are needed for model tuning.

Simulation code is designed to operate in a variety of conditions that are useful for both the theoretical needs of research students and the practical needs of mill engineers. One of the novel GUI options of the simulation code is the choice of performance target variables as model inputs that in turn determine required process conditions to operate the digester. It is demonstrated that an effective specie transition policy can be formulated through smooth ramping of operating variables between two nominal states in approximate synchronization with the transition front in chip column.

II. PLANTSCAPE & DIGESTER MODEL OPEN SYSTEM IMPLEMENTATION AT MEADWESTVACO MILL PROJECT SITE

II.1 Open System Implementation – Model Simulation

At the mill site Honeywell DCS system resides on a dedicated local area network. For purposes of communication between the external model application, located on a local area network, and the mill DCS, an application node was installed. The model communicates with Honeywell's Plantscape server through an application task that communicates with MATLAB research code. The application node communicates with the DCS and the Plantscape server using OPC protocols. A schematic of the arrangement is shown in Figure 4.

Figure 4. Schematic depicting communication between the mill DCS and the MATLAB model via an application node.

The model task on Plantscape was developed by using the Mathworks' C/C++ compiler to convert the Matlab code to: (a) open C++ master file with (b) standalone executable model code called from the master C++ file, such that the proprietary aspects of the code are preserved. For the Plantscape application, a single DLL project was generated from the combined two parts (master m-file calling the executable code) using Mathworks' C/C++ compiler. All necessary input/output points were created in the Plantscape server. Communication between the server and the DLL project was established using a GUI. Further, the modular nature of the research code allowed the user to perform on-line tuning of the model without recompiling the code. Figure 5 shows the user interface to the model in Plantscape. The user is allowed to change model parameters from the screen and monitor model outputs.

Figure 5. User interface

III. ADVANCED CONTROL OF A DUAL VESSEL KAMYR PULP DIGESTER USING MODEL PREDICTIVE CONTROL

III.1 MPC Design – Honeywell RMPCT Controller

Honeywell's Profit® Controller (RMPCT) program is a state of the art multivariable predictive controls package that deals rigorously with interacting systems subject to time varying constraints. Profit Controller (RMPCT) has been designed to provide both control and optimization in a robust fashion. Honeywell's Profit Controller (RMPCT) program controls and optimizes processes that have significant interaction between variables. The Profit Controller (RMPCT) incorporates a model of the process dynamics. From this model, Profit Controller predicts future behavior of the process and determines how to adjust the controller's outputs to bring all process variables to setpoints or within constraints. Then if there are any degrees of freedom remaining, the controller adjusts the process to optimize operations, for example by maximizing total product value.

The approach taken in this work involved development and preliminary validation of a rigorous fundamental model of the pulp digester followed by testing of model based strategies for control, fault diagnosis, and controller performance monitoring. Closed-loop control using Honeywell's commercial RMPC technology is demonstrated by developing a prototype. Here, the fundamental model played the role of a virtual digester, while the controller was based on linear models identified by step tests.

The Kappa number constitutes a key quality variable of the resulting pulp, which measures the residual lignin content. A typical control objective of digester operation is minimization of variation in Kappa number from a prescribed value. Challenges in operation and control can be partly attributed to large dead-times (4 to 8 hours) between the quality variables and manipulated inputs such as temperature as well as the large variability in the naturally occurring feedstock. The advanced control problem of the pulp digester using RMPCT is characterized as Kappa profile control and EA control. The control problem of interest has 5 controlled variables (CVs), 7 manipulated variables (MVs) and two measured disturbance variables (DVs). These are displayed in Table 1 below. The numerical values in parentheses refer to nominal operation. Values corresponding to controlled variables can be taken as desired setpoints.

RMPCT responses for setpoint changes, measured disturbance rejection and unmeasured disturbance rejection were tested for the pulp digester.

Setpoint Change: Setpoint changes are often encountered at the mill while attempting to optimize fiber yield. In this example RMPCT was tested for setpoint changes in the Kappa profile (characterized the three Kappa measurements. Simulation results for the CV and MV responses are shown in Figure 6. RMPCT exhibits a short rise time (19 hours) for

the key quality variable of EMCC Kappa. The response has an undershoot with the settling time of approximately 30 hours. Figure 6 reveals that RMPC makes relatively smooth manipulations as seen in the MV responses.

Controlled Variables (CVs) (Nominal/Setpoint)	Manipulated Variables (MVs) (Nominal Values)	Measured Disturbance Variables (DVs) (Nominal Values)	Unmeasured Disturbance (Nominal Values)
CV1 - Cook Zone Kappa (173.0754)	MV1 - Cook zone temperature (315.28 °F)	DV1 - Production rate change (900.3726 GPM)	White Liquor EA, g/L as Na ₂ (91.6 g/L)
CV2 - MCC Kappa (57.2384)	MV2 - MCC temperature (328.099 °F)	DV2 - Grade Transition (softwood to hardwood) 0 – SW; 1 – SW–>HW	
CV3 - EMCC Kappa (23.7686)	MV3 - EMCC temperature (326.899 °F)		
CV4 - Upper Extract EA (9.8114 g/L as Na ₂ O)	MV4 - White liquor to IV (355.45356 GPM)		
CV5 - Lower Extract EA (4.2845 g/L as Na ₂ O)	MV5 - Cook Trim flow (64.9972 GPM)		
	MV6 - MCC Trim flow (127.1576 GPM)		
	MV7 - EMCC Trim flow (63.4656 GPM)		

Table 1 – Control problem specification

Production Rate Transition: Production rate changes are implemented by ramping up or slowing down the wood chip feeder. Pulping mills frequently need to change the production rate to cope with the downstream paper machine section demands and equipment outages. Simulation results for a production rate change of 100 oven dried tons per day are shown in Figure 7. RMPC displays a very short transition time of 11 hrs. Also the controller moves implemented by RMPC are smooth as

seen in Figure 7. PI controllers were used for chip level control in the impregnation and digester vessels.

Unmeasured disturbance EA change:

Figure 8 shows RMPCT responses to a Step Change In White Liquor EA from 91.59 to 100.75 G/L as Na₂O. It is observed that RMPCT rejects the disturbance effectively with some undershoots in the Kappa profile. For the EMCC kappa it takes approximately 40 hours to adjust to the EA change. In the MV responses (Figure 8) it is observed that RMPCT smoothly cuts the white liquor flow (MV4) to the impregnation vessel and adjusts the trim flows and temperatures to counter the increase in EA.

Figure 6: Controlled and Manipulated Variable responses to setpoint changes.

Figure 7: Controlled and Manipulated Variable responses to production rate transition (Measured Disturbance Rejection).

Figure 8: Controlled and Manipulated Variable responses to a step change in In White Liquor EA From 91.59 To 100.75 G/L As Na_2O (Unmeasured Disturbance Rejection).

III.2 MPC Design – Grade Transition as a Profile Control Problem

In this algorithmic component, we assume availability of real-time kappa measurements at four different locations along the digester, viz. upper and lower heaters, extraction screens, and blowline, as indicators of the kappa profile. Although, real-time measurements are not generally feasible, schemes for inferring kappa numbers using readily available measurements have been reported in literature and can be employed to obtain estimates. A summary of the the control problem is provided below;

?? *Controlled Variables:* Kappa # at upper heater, Kappa # at lower heater, Kappa # at extraction screens, Kappa # at blow line and the residual effective alkali at the extraction screen

?? *Manipulated Variables:* Upper heater temperature, lower heater temperature, makeup white liquor to upper heater, makeup white liquor to lower heater and white liquor flow at top of digester

?? *Measured Disturbance:* Feedstock grade transition treated as a binary switch [9]

?? *Unmeasured Disturbance:* Step change in white liquor EA, which affects all three white liquor flow variables in the manipulated input.

Two control strategies were explored for control of kappa number profile, viz. a decentralized PI scheme consisting of 5 SISO loops and a centralized, linear constrained model predictive controller using step response models with output feedback. Both schemes use a 10 minute sampling interval. Linear transfer function models (first order plus time delay (FOPTD), FOPTD with single zero, second order plus time delay (SOPTD), SOPTD with single zero) were identified from step tests data. Figure 9 shows comparison of residual EA prediction by a SOPTD with

single zero model with the fundamental model when the lower heater temperature is decreased by 5 K.

For the decentralized scheme, output/manipulated input pairings were determined with the help of RGA, steady-state gain matrix, and process analysis. The pairings are detailed in the Control 2002 conference paper resulting from this work. "Improved PI" IMC tuning rules were employed. During simulations, it was observed that large values of the IMC tuning parameter, λ , was necessary to obtain acceptable response. With lower values of λ (however, within recommended range) the controller was found to be very aggressive often leading to sustained oscillations over long periods of time.

Figure 9: Linear model identification with second order plus time delay model with a single zero (solid: fundamental model, dashed: linear model) using step test data generated by decreasing upper heater temperature.

The 5 input x 5 output linear MPC controller used the following parameters during control simulations,

- ?? prediction horizon = 60 samples (10 hours)
- ?? control horizon = 10 samples (100 minutes)
- ?? output weights = [0.03 0.03 0.08 0.09 0.03]
- ?? input weights = [20 20 20 20 20]

The step response models were scaled appropriately allowing meaningful weight matrices.

As a first example, the two controllers were implemented for a setpoint change in the Kappa profile (characterized by the four measurements)

and residual EA for the same feedstock grade. This situation is often encountered while attempting to increase fiber yield. Results are shown in Figs. 10 and 11. The MPC controller (solid line) quickly responds to the setpoint change and steers all four measurements to the vicinity of the setpoints. The PI controllers for Kappa # control at upper and lower heaters were tuned aggressively, while the other three PI controllers were sufficiently detuned to avoid a ringing response. Transition times for the linear MPC and PI controllers are seen to be 6 hrs and 12 hrs, respectively. The sluggish response of the PI controller is a necessary consequence of large dead-times in the process. Use of time-delay compensation may provide one possible remedy. Fig. 11 also suggests that there exists an apparent multiplicity of solutions. This fact is reflected in the poor conditioning of the steady-state gain matrix. For example, the Kappa numbers at the extraction screen and blow lines have similar gains and time constants but differ widely in the transport delay.

Figure 10 - Kappa profile setpoint change results using linear MPC (solid) and PI (dot-dashed) controllers.

Figure 11 - Manipulated variables corresponding to Figure 10

The second example deals with rejection of unmeasured disturbance in form of a 2% increase in white liquor EA (see Figure 12). This disturbance affects both the make up white liquor streams as well as that entering at the top. The long time needed (>10 hours) by the linear MPC controller to completely reject the deterministic disturbance is partially attributable to the fact that output feedback provides a poor estimate of unmeasured disturbances entering in the input as in the current case.

Figure 12 - Disturbance rejection: 2% step increase in effective alkali concentration in white liquor

As a final example, the case of accomplishing grade transition via Kappa # profile control was explored. Grade transition is considered as a feedforward disturbance triggered by a binary switch. Details are provided in the accompanying report from the Control Systems 2002 meeting.

IV. PROCESS MONITORING OF CLOSED-LOOP DIGESTER

IV.1 Process Monitoring – Toolbox Summary

The objective of this final part of the project was to develop a tool that allowed industrial practitioners to analyze and assess the performance of the control loops on an industrial plant from routine operating data. To this end, a control performance monitoring (CPM) toolbox including several metrics was developed. Matlab-based Graphical User Interfaces (GUI) were designed and implemented to guide the user.

The various metrics developed in the literature are useful in various contexts (*e.g.* oscillations, control valve stiction *etc.*). Therefore, having methods that allow the identification of the context significantly improves one's confidence in a particular measure. The toolbox was developed for the off-line analysis of the control loops. It offers a variety of tools to detect the presence of oscillations and includes performance metrics for a first-level study, which do not require the process model knowledge. It also allows a quantification of the performance of Proportional-Integral (PI) controllers from closed-loop operating data. In addition, with an objective of supervising hundreds of control loops; the procedures used in the toolbox were chosen of modest complexity.

The main GUI of the CPM toolbox depicted in Figure 13 allows the user to perform the following tasks:

1. Report file creation
2. Select between using mill data or simulation data
3. Oscillation detection
4. Dead-time estimation
5. FCOR performance index calculation
6. Harris and modified Harris performance indices calculation
7. Index for oscillating loops calculation
8. Normalized settling time performance index calculation

The task *Report file creation* allows the user to create a text file, which will contain the diagnosis and result information. The file can be saved as a Word document. A dialog box is used to enter the file name. The file is stored in the current directory. On generating the report during the performance assessment studies, the figures are saved in the enhanced metafiles (*.emf) format. This choice was motivated by its ease to include in a Word document. Experimental or simulation data can be used for analysis. The selection is done via a list box.

On selecting *Mill data* a separate list box offers the possibility to choose the data file. Once the data file is selected and the OK press-button is hit, the GUI *Mill Data Selection*, presented in Figure 13 appears. The user can choose between the following control loops:

- Loop 1: TC_019H: MCC Temperature
- Loop 2: TC_020H: EMCC Temperature
- Loop 3: HC_019: MCC Flow
- Loop 4: HC_020: EMCC Flow
- Loop 5: LC_010: Digester level control loop

On selecting *Simulation* a GUI *Simulation Example* presented in Figure 13 appears. The user can choose the different transfer functions of the control scheme and generate the data or can use SIMULINK to generate the data.

Figure 13 - Main GUI of the CPM toolbox, Mill Data Selection GUI, and Simulation Example GUI

The other tasks (from 3 to 8) can be classified in four studies as follows:

1. Oscillation detection and diagnosis
2. Minimum variance benchmark performance index estimation
3. User-defined benchmark performance index estimation
4. Quantification of controller performance

In a separate technical report, application of the toolbox is detailed for both a simulation case study and for an industrial data record.

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Milestone Status Table:

ID Number	Task / Milestone Description	Planned Completion	Actual Completion	Comments
Task 1.	Digester Transition. (U. Delaware)			
1.1.	Formulate digester simulation case.			
1.1.1.	First round mill visits	4/30/00	4/19/00	Completed
1.1.2.	Second round mill visits/discussions	6/30/00 7/18/00	7/18/00	Completed
1.1.3.	Case study defined	4/30/01 6/30/02	5/1/2002	Completed
1.2.	Develop benchmark simulation.			
1.2.1.	Formulate mill-specific strategies	8/31/00	8/31/00	MeadWestvaco mill
1.2.2.	Code into UD simulation package	7/30/01 12/31/01	12/31/01	Completed
1.2.3.	Existing transition strategies reproduced	12/31/01 4/30/02	6/30/02	UD model has been validated with rate transition data. A hypothetical grade transition case has also been demonstrated.
1.3.	Develop fundamental multiple-grade digester model.			
1.3.1.	Formulate mixing rule model	4/30/00	3/31/00	Completed
1.3.2.	Efficient numerical formulation	4/30/01 8/31/01	7/31/01	Completed
1.3.3.	Final UD model developed	4/30/01	4/30/01	Completed
1.4.	Develop framework for model-based control.			
1.4.1.	Formulate multi-rate, inferential MPC	4/30/01	6/30/01	Completed

ID Number	Task / Milestone Description	Planned Completion	Actual Completion	Comments
1.4.2.	Refine profile control	6/30/02 8/31/02 10/31/02	10/31/02	Design of a nonlinear MPC controller for profile control using the validated UD model as the virtual plant is in progress. Honeywell will complete a linear MPC using RMPCT design. Due to computation constraints in on-line linearization, a linear MPC strategy is being adopted.
1.4.3.	Final control package developed	6/30/02 8/31/01 10/31/02	10/31/02	See note above.
1.5.	Validate multiple-grade digester model against industrial data.			
1.5.1.	Literature data validation	6/30/00	4/30/00	Completed
1.5.2.	Project partner (mill) validation	4/30/01 9/30/01 2/28/02	6/30/02	Comparisons underway, expected to coordinate with IETek model
1.5.3.	Final (validated) models developed	12/31/01 4/30/02	6/30/02	See 1.2.3 regarding validation
1.6.	Refine transition control methodologies.			
1.6.1.	Feed system control for rate transitions	4/30/02 8/31/02 10/31/02	10/31/02	This task will be completed simultaneously with task no. 1.4.2 and 1.6.2
1.6.2.	Feedstock swing control strategy	8/31/02 10/31/02	10/31/02	See note above (1.4.2)
1.6.3.	Profile management control strategy	12/31/02	12/31/02	
1.7.	Integrate control methodologies into commercial control packages.			
1.7.1.	Control strategy interface protocol	12/31/01	12/1/01	Completed
1.7.2.	First round beta-test on hardware platform	6/30/02 11/30/02	12/15/02	Testing on RMPCT with UD model in progress
1.7.3.	Final beta-test on hardware platform	10/31/02	1/15/03	

ID Number	Task / Milestone Description	Planned Completion	Actual Completion	Comments
Task 2.	Fault Detection and Handling in Digester. (U. Delaware)			
2.1.	Definition of fault problem.			
2.1.1.	First round mill visits/interviews	4/30/00	4/19/00	Completed
2.1.2.	Second round mill visits/interviews	8/31/00	7/18/00	MeadWestvaco mill
2.1.3.	Fault variables defined	12/31/00 01/31/01	1/31/01	Completed
2.2.	Develop qualitative framework for fault detection.			
2.2.1.	Initial qualitative tool development	10/31/00	11/10/00	Completed
2.2.2.	Framework developed	12/31/00 02/15/01	2/15/01	Completed
2.3.	Develop quantitative framework for fault detection.			
2.3.1.	Build fault variables into model	4/30/01	4/30/01	Completed
2.3.2.	Develop residual methods for fault prediction	8/31/01	12/1/01	Completed
2.3.3.	Integrate with simulation for soft-sensing	12/31/01	12/1/01	No longer relevant (see item 2.3.2)
2.4.	Integrate fault diagnosis with control.			
2.4.1.	Controllability/Observability analysis	4/30/01	4/30/01	Completed
2.4.2.	Incorporate soft-sensing as state estimation	10/31/01	8/31/01	Completed
2.4.3.	Integrated tool developed	2/28/02 5/31/02 8/31/02	8/31/02	Completed (state estimation)
2.5.	Optimization toolkit developed.			
2.5.1.	Mixed-integer formulation defined	4/30/02	4/20/01	Completed
2.5.2.	Preliminary toolkit developed	8/31/02	10/30/01	Completed
2.5.3.	Final toolkit developed after workshop review	12/31/02	2/1/03	
Task 3.	Development of Prototype Software Tools. (IETek)			
3.1.	Fundamental model development.			
3.1.1.	Mill visits to MeadWestvaco & Weyerhaeuser sites	9/30/00	7/18/00	Completed
3.1.2.	Base case specification	7/31/00	6/30/00	Completed

ID Number	Task / Milestone Description	Planned Completion	Actual Completion	Comments
3.1.3.	Structural modification to IETek model	8/31/00 2/28/01	12/12/00	Completed
3.1.4.	Initial model parameter tuning	9/30/00 2/28/01	5/1/01	Completed
3.1.5.	Training of UD staff	11/30/00 3/31/01	3/31/01	Completed
3.1.6.	Final model delivered	12/31/00 3/31/01	5/1/01	Completed
3.2.	Develop computational tools.			
3.2.1.	GUI development for control/simulation	7/31/01 6/30/02	4/02/02	Completed
3.2.2.	Initial Digester model /GUI revisions	5/31/01	5/1/01	Completed
3.2.3.	Second Digester model and GUI revisions	3/1/02 6/30/02	4/02/02	Tuned model refined GUI delivered 4/02/02.
3.2.4.	Final Digester model and GUI revisions	3/31/02 6/30/02	4/02/02	Tuned model refined GUI delivered 4/02/02.
3.3.	Establish open standards for component integration.		9/20/02	Completed
3.3.1.	Workshop material development	11/30/01	6/30/01	Workshop material developed by UD for June 2001 workshop.
3.3.2.	Revisions to software system after workshop	6/30/02	4/02/02	Tuned model refined GUI delivered 4/02/02.
3.3.3.	Workshop on project status & accomplishments	10/31/02	1/20/03	Completed at UCSB
Task 4.	PlantScape Prototype and Evaluation (Honeywell)			
4.1.	Installation of PlantScape system	4/1/01 4/1/01	4/15/01	Completed
4.2.	Develop PlantScape Application for Model Based Sensing.			
4.2.1.	Software Architecture Design	7/1/00	7/1/00	Completed
4.2.2.	Software Development and Test (on test system)	1/1/01	1/1/01	Completed
4.2.3.	Software Development and Test at Mill site	1/1/02	12/1/01	Completed
4.3.	Develop and Implement Multivariable Control Strategy.			
4.3.1.	Control Software Architecture Design	7/1/00	7/1/00	Completed

ID Number	Task / Milestone Description	Planned Completion	Actual Completion	Comments
4.3.2.	UD Digester Model Implementation in test system	6/1/02 8/1/02	9/15/02	Completed
4.3.3.	RMPCT Installation on test system	7/1/02 9/1/02	9/15/02	Completed
4.3.4.	RMPCT Simulation Test using UD model	10/1/02 12/1/02	12/31/02	Completed
4.3.5	Test Report (UD and Honeywell)	12/1/02 12/31/02	12/31/02	Completed
4.4.	Controller Fault Assessment Technology Development.			
4.4.1.	Develop Process Model for Diagnosis from input output data	6/1/02 10/1/02 11/1/02	11/1/02	Completed
4.4.2.	Development of Controller Performance Indicator	10/1/02	9/30/02	Completed
4.4.3.	Software Architecture Design	10/1/02	9/15/02	Completed
4.4.4.	Development and Test of selected loops at mill site	11/1/02	11/1/02	Completed
4.4.4.	Final Report	12/1/02	12/17/02	Completed
Task 5.	Infra-90 Prototype and Evaluation. (ABB)			
5.1.	Mill visit and installation	12/31/00 6/30/01	4/1/02	UD tools employed for Weyerhaeuser mill
5.2.	Interface project tools with ABB hardware UD model customized to Weyerhaeuser mill	12/31/01	6/1/02	Completed
5.3.	Participate in beta test Detailed case study documentation	12/31/02	3/31/03	Final strategic operations study to be completed
Task 6.	Industrial Collaboration (Weyerhaeuser and MeadWestvaco)	Monthly		Ongoing advisement of project activities
Task 7.	Program Management.	Monthly		Ongoing supervision of project partners

Final Budget Data: The approved spending should not change from quarter to quarter. The actual spending should reflect the money actually spent on the project in the corresponding periods.

			Approved Spending Plan			Actual Spent to Date		
Phase / Budget Period			DOE Amount	Cost Share	Total	DOE Amount	Cost Share	Total
	From	To						
Year 1	1/10/00	1/9/01	373,820	223,740	597,560	167,252	120,100	287,352
Year 2	1/10/01	1/9/02	268,843	83,286	352,129	294,913	621,000	915,913
Year 3	1/10/02	1/9/03	253,758	85,084	338,842	434,256	1,800	436,056
Year 4								
Year 5								
Totals			896,421	392,110	1,288,531	896,421	742,900	1,639,321