DESIGN, SIMULATION AND IMPLEMENTATION OF DECENTRALIZED PI CONTROLLERS FOR A MULTIVARIABLE FEED BLENDING SYSTEM AT THE FALCONBRIDGE SUDBURY SMELTER

B. Boulet
McGill University
Department of Electrical and Computer Engineering, Centre for Intelligent Machines
3480 University Street, Montréal, Québec H3A 2A7
Telephone: (514) 398-1478
Fax: (514) 398-4470
E-mail: benoit@ece.mcgill.ca

L. Ryan
Falconbridge Limited
Falconbridge Technology Centre, Falconbridge, Ontario, Canada P0M 1S0
Telephone: (705) 693-2761
Fax: (705) 699-3976
E-mail: lionel_ryan@sudbury.falconbridge.com

C. Graves, B. MacKay
Hatch
128 Pine Street, Sudbury, Ontario, Canada P3C 1X3
Telephone: (705) 688-0250
Fax: (705) 688-0244
E-mails: cgraves@hatch.ca, bmackay@hatch.ca

ABSTRACT

Falconbridge's new Raglan mine started to ship nickel ore concentrate to Sudbury for smelting in the spring of 1998. The smelter's new continuous feed blending process allows for the blending of concentrate slurry from the Strathcona mill with a dry Raglan concentrate before it is roasted and fed into the electric smelting furnace. Accurate control of the concentrate ratio and density of the blended slurry was deemed critical to achieve efficient roaster and furnace operation. This paper reports on the nonlinear multivariable process modeling, decentralized controller design, and simulation studies that led to a successful commissioning of the feed blending control system.

INTRODUCTION

Falconbridge Limited's new Raglan mine in Northern Quebec started to ship its nickel ore concentrate to Sudbury for smelting in the spring of 1998. The new feed blending process at the Sudbury smelter allows for the blending of concentrate slurry coming from the Strathcona mill with a dry Raglan concentrate before it is roasted and fed into the electric smelting furnace. Accurate control of the ore concentrate ratio and density of the blended slurry was deemed critical to achieve efficient roaster and furnace operation. A process modeling and simulation study was undertaken to assess the feasibility of using decentralized proportional-integral (PI) controllers for this coupled nonlinear multivariable process. The PI controllers were implemented and successfully commissioned in May of 1998. This paper discusses the dynamic modeling, simulation, control design and implementation aspects of the project.

The basic idea behind the process is to mix dry Raglan concentrate with Strathcona slurry and water in a repulper tank to maintain both a concentrate dry mass ratio setpoint, and a density setpoint at the output of the repulper tank. The most important controlled variable is the pulp density, followed by the Raglan to Strathcona dry mass ratio (R:S in short). The repulper tank level must also be regulated. A simplified flowsheet of the open-loop process is given in the next section. The overall process actually consists of two repulper tanks and two roaster feed tanks. However, since it is symmetrical, only one repulper/roaster feed tank pair was analyzed in this work. Note that the roaster feed tanks downstream of the repulpers are operated in open loop. These tanks are large, so their dynamics are quite slow. An operator keeps an eye on their levels, and simply changes the roaster feed rate when the levels get too close to their limits.

FEED BLENDING PROCESS

Description of the Feed Blending Process

The model of the feed blending process is essentially composed of three interconnected subsystems (Figure 1):

- Repulper tank
- Transfer pump
- Roaster feed tank

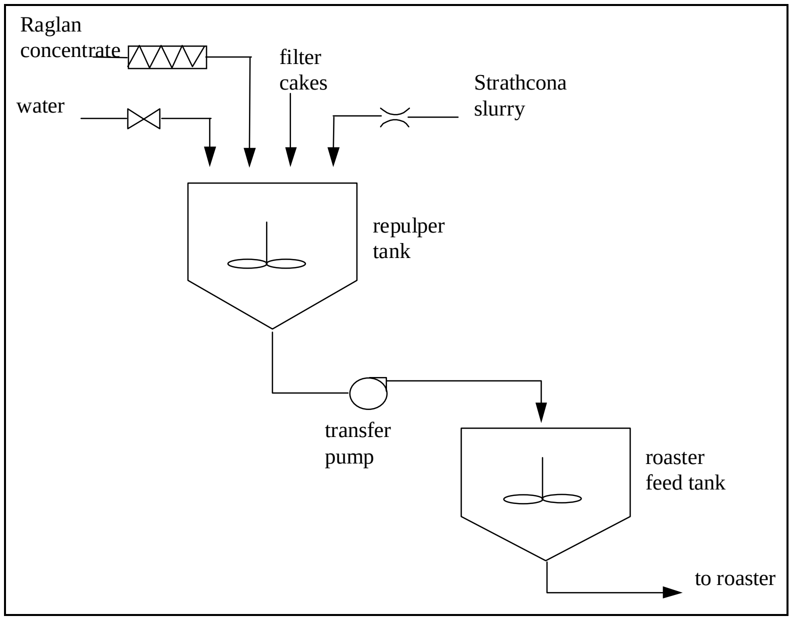


Figure 1: Feed blending system

There are at least four significant disturbances in this process: the Strathcona filter cakes that are dropped in the repulper tank when one or two filters are switched on, the specific gravity of the incoming slurry, its density, and its pressure. Only the filter cakes are modeled as an external disturbance. In the past, these filters were used to increase the density of the pulp going into the roasters. This is still required, although usually the addition of dry Raglan concentrate is sufficient to increase the density to a point where finer density control using process water becomes possible. The effects of the slurry disturbances are modeled as perturbations to the linearized process dynamics rather than exogenous disturbance inputs. A nonlinear open-loop model of the feed blending process was developed and programmed as a simulator in Matlab Simulink™. In order to use a frequency-domain loopshaping method for the PI controllers, the full nonlinear model was linearized around the desired operating point (see, e.g., [Kuo] on linearization).

The R:S ratio r is not measurable, and the dry mass flow rate of Strathcona $\dot{m}_S$ is not measured. However, the latter can be indirectly measured by subtracting the known mass flow rate of Raglan concentrate $\dot{m}_R$ into the repulper from the measured total dry mass flow rate coming out of the repulper. With the tank level well regulated and having fast closed-loop dynamics, the total dry mass flow rate at the output rapidly follows a change of concentrate mass flow rate at the input, and makes the indirect measurement of dry mass flow rate of Strathcona accurate enough for our purposes. Moreover, the dynamics of R:S become almost independent of the repulper tank level dynamics. This is the reason why we chose to design the tank level PI controller first, with fast closed-loop dynamics.

Given the desired total mass flow rate in the process and the desired R:S ratio, the Raglan mass flow rate setpoint is computed for the loss-in-weight controller feeding the repulper tank. The R:S ratio is indirectly controlled by computing the corresponding setpoint for dry mass flow rate of Strathcona concentrate, and by forming an error signal using the indirect measurement of Strathcona for the ratio PI controller. The density loop, which uses a nuclear density meter, was designed last. Simulated open-loop and closed-loop responses to process disturbances were obtained, together with actual closed-loop responses. It was found that the adopted decentralized PI control strategy performed well in spite of the nonlinear couplings between variables. This was achieved through careful analysis and design procedures, and by running several simulation experiments to test the robustness of the closed-loop system under various process disturbances and model perturbations.

Modeling Assumptions

The following assumptions were made to obtain the models and design the controllers.

- Perfect mixing in the repulper tank
- Linear transfer pump characteristic
- Linear slurry valve characteristic

The last assumption may be the least valid as valves are inherently nonlinear, but closed-loop robustness against variations in the slurry valve gain was checked in the simulations. This robustness analysis indicated that the actual control system should perform as expected. The Raglan concentrate mass flow rate into the repulper is controlled via a fast and accurate loss-in-weight control system. We chose to ignore its dynamics in the model. The repulper tank subsystem model is the most complex of the three, because the evolution of three different types of materials inside the tank (Raglan, Strathcona and water) have to be simulated individually. Our approach to model this is to keep track of the total mass of each material in the repulper tank by using separate integrators.

The transfer pump is assumed to have a volumetric flow varying linearly with pump motor speed. The roaster feed tank is modeled using integrators for dry mix, water, and mix slurry. The volume of mix varies according to the mass of slurry coming in from the repulper via the transfer pump, and going out to the roaster. Both the nonlinear and the linearized models are described in some details in the next section.

CONTROL STRATEGY

The objectives of the feed blending control system are to:

1. Maintain the pulp density around 70% (or other subsequent setpoints), to within 1%, at the output of the repulper
2. Maintain the R:S dry mass ratio close to its setpoint
3. Maintain the repulper tank level close to its setpoint
4. Reject disturbances that may affect the density and the ratio
5. Provide robustness to variations in process dynamics, including slurry valve characteristics

Three single-input, single-output decentralized PI controllers, described in the PI Controller Design section, were designed to achieve the above objectives. See the closed-loop system block diagram of Figure 5 for more details on the architecture. The use of decentralized PI controllers was favored as a first design for ease of implementation on the smelter's Foxboro DCS, tunability, fault isolation, and also because plant personnel are familiar with this technology.

Sensors and Measured Variables

The following sensors are used:

- density meter at the output of the repulper, downstream of the transfer pump,
- magnetic flowmeter at the output of the repulper, downstream of the transfer pump,
- ultrasonic level transmitter for repulper tank level,
- magnetic flowmeter for process water going into repulper,
- ultrasonic level transmitter for roaster feed tank.

In addition to the direct measurements provided by these sensors, they are used to compute the dry mass flow rate of mix (or pulp) at the output of the repulper. A measurement of the dry Raglan concentrate feed rate from the loss-in-weight controller is used for R:S ratio control. The measurements directly used for feedback control are:

- pulp density at the output of the repulper
- repulper tank level
- dry mass flow rate of mix at the output of the repulper
- mass flow rate of Raglan

Actuators and Control Variables

The following actuators are used:

- control valve for Strathcona slurry input to repulper
- AC motor with variable-frequency drive for transfer pump
- control valve on the water line going into the repulper

These actuators are used to modulate the following variables for feedback control:

- volumetric flow rate of Strathcona slurry into the repulper
- volumetric flow rate of mix at the output of the repulper
- water flow rate into the repulper

A low-level water flow rate controller not discussed in this paper allows the density controller to output a water flow rate setpoint to the flow rate controller. For the purpose of this study, the roaster feed rate was considered as an exogenous input, i.e., it is not used for feedback control and its variations are seen as disturbances.

NONLINEAR OPEN-LOOP MODEL

Nonlinear Open-Loop Evaluation Model of the Repulper Tank

Due to space limitations, we only give the equations for the repulper tank. The total masses of Strathcona m_{SR}, Raglan m_{RR}, and water m_{wR} in the repulper tank vary according to the following first-order nonlinear differential equations. The notation used can be found in Appendix A. The dot above a symbol indicates time derivative.

Mass of Strathcona in repulper:

$$\dot{m}_{SR} = \dot{m}_S + \dot{m}_f - \frac{m_{SR}}{m_{SR} + m_{RR}} \dot{m}_m \qquad (1)$$

Mass of Raglan in repulper:

$$\dot{m}_{RR} = \dot{m}_R - \frac{m_{RR}}{m_{SR} + m_{RR}} \dot{m}_m \qquad (2)$$

Mass of water in repulper:

$$\dot{m}_{wR} = \frac{(1 - f_s)}{f_s} \dot{m}_S + \dot{m}_w + \frac{(1 - f_f)}{f_f} \dot{m}_f - \frac{m_{wR}}{m_{SR} + m_{RR}} \dot{m}_m \qquad (3)$$

The three mass equations above allow us to compute the mix density at the output of the repulper tank f_m, and the R:S ratio r.

Density at the output of the repulper tank:

$$f_m = \frac{m_{SR} + m_{RR}}{m_{SR} + m_{RR} + m_{wR}} \tag{4}$$

Raglan to Strathcona ratio:

$$r = \frac{m_{RR}}{m_{SR}} \tag{5}$$

Repulper tank level:

$$h_{\text{Rep}} = \left(\frac{4}{\pi D_{\text{Rep}}^2}\right)\left(\frac{m_{RR}}{SG_R} + \frac{m_{SR}}{SG_S} + m_{wR}\right) \tag{6}$$

PI CONTROLLER DESIGN

Three single-input, single-output proportional-integral (PI) controllers, described in the next subsections, were designed to achieve the objectives listed in the Control Strategy section. The PI controller designs were carried out in the frequency-domain using a model linearized around the operating point. The classical technique of loopshaping was used to ensure sufficient gain and phase margins for each loop. See the block diagram of the closed-loop system for more details on the architecture.

Since the feed blending system is multivariable and coupled, we proceeded sequentially as follows to design the overall control system, starting from the linearized model.

1. Design the repulper tank level controller,
2. Close the tank level control loop and recompute the transfer function matrix,
3. Design the ratio controller based on the transfer function matrix of step 2,
4. Close the ratio control loop and recompute the transfer function matrix,
5. Design the density controller based on the transfer function matrix of step 4.

Repulper Tank Level PI Controller

The repulper tank level is controlled by a PI controller computing speed commands for the transfer pump motor. The two controller inputs are:

1. the pulp level h_{Rep} as measured with an ultrasonic transducer, and
2. the level setpoint h_{Rep0}.

The output of the controller is a correction $\Delta\omega_{pm}$ in pump motor speed, given by:

$$\Delta\omega_{pm}(t) = K_{ph}[h_{\text{Rep0}}(t) - h_{\text{Rep}}(t)] + K_{ih}\int_0^t [h_{\text{Rep0}}(\tau) - h_{\text{Rep}}(\tau)]d\tau \qquad (7)$$

which is then subtracted from the nominal speed ω_{pm0} to form the speed command. Figure 2 below shows a block diagram of the repulper tank level PI controller, where the Laplace operator "1/s" represents an integrator.

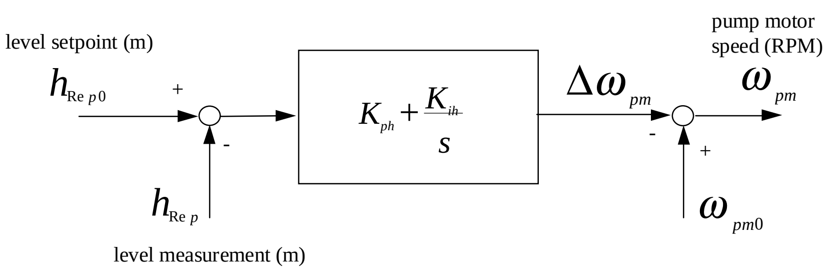


Figure 2: Repulper tank level PI controller

Raglan to Strathcona Ratio PI Controller

The R:S ratio is controlled by a PI controller computing slurry valve opening commands (in %) for the Strathcona slurry line, from a computed error in dry mass flow rate of Strathcona. The three controller inputs are:

1. the dry mass flow rate of the mix at the output of the repulper as computed from magnetic flowmeter and density meter measurements,
2. the mass flow rate of Raglan going into the repulper,
3. the R:S ratio setpoint.

The output of the controller is a correction Δx_v in valve position, given by

$$\Delta x_v = K_{pr}e_S + K_{ir}\int_0^t e_S(\tau)d\tau \qquad (8)$$

where $e_S := \dfrac{\dot{m}_m}{1+r_{SP}} - (\dot{m}_m - \dot{m}_R)$ is the error in Strathcona feed rate, and to which the nominal valve opening x_{v0} is added to form the valve command. Note that the first term on the right-hand side of the last equation defining e_S is the Strathcona dry mass flow rate setpoint. Figure 3 shows a block diagram of the ratio PI controller. The block with the symbol "X" simply multiplies its two inputs.

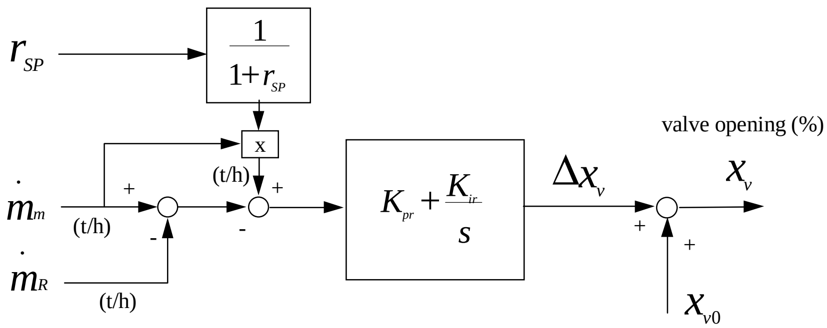


Figure 3: Ratio PI controller

Density PI Controller

Density at the output of the repulper is controlled by a PI controller acting on the process water flowrate setpoint, based on the density error. An internal water flow control loop, not included in this study, ensures that the actual flow is close to the setpoint. The two controller inputs are:

1. density measurement f_m from density meter at the output of the repulper,
2. density setpoint f_{m0}.

The output of the controller is a correction $\Delta \dot{m}_w$ in water flow rate, given by:

$$\Delta \dot{m}_w(t) = K_{pd}[f_{\mathrm{m}}(t) - f_{\mathrm{m0}}(t)] + K_{id}\int_0^t [f_{\mathrm{m}}(\tau) - f_{\mathrm{m0}}(\tau)]d\tau \tag{9}$$

to which the nominal water flow rate $\dot{m}_{w0}$ is added to form the setpoint for the internal water flow control loop. Figure 4 shows a block diagram of the density PI controller.

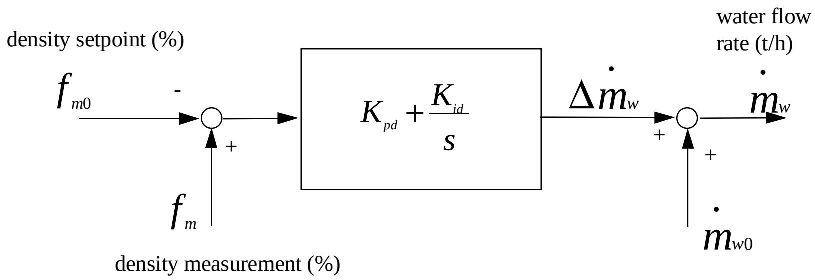
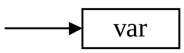


Figure 4: Density PI controller

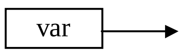
NONLINEAR CLOSED-LOOP SIMULATION

The nonlinear model of the feed blending system was implemented as a simulator using the software package MATLAB SIMULINK^{TM} [THE MATHWORKS, 1998]. The main simulator block diagram is depicted in **Figure 5**. It combines the nonlinear feed blending system model described above with the level, ratio, and density PI controllers (gray blocks). The simulator was built from low-level continuous-time dynamic function blocks to form an interconnection of the three main components of the system, namely: the repulper tank, the transfer pump, and the roaster feed tank. Each simulation run represented 12 hours of real-time operation, and the time unit used was the hour. A step size of 10/3600 hour (10 seconds) in the fifth-order Runge-Kutta integration routine was used to solve the nonlinear differential equations.

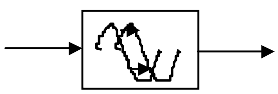
Legend for Simulink Symbols



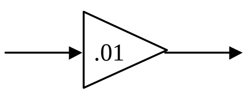
output to variable “var” in Matlab workspace



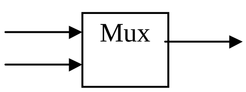
input from variable “var” in Matlab workspace



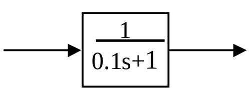
transport delay



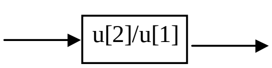
gain multiplying the input



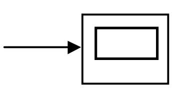
multiplexer block combines its inputs to form a vector output



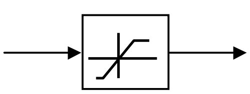
transfer function block



function block mapping vector input into scalar output



scope meter block to view variable trends during simulations



saturation block

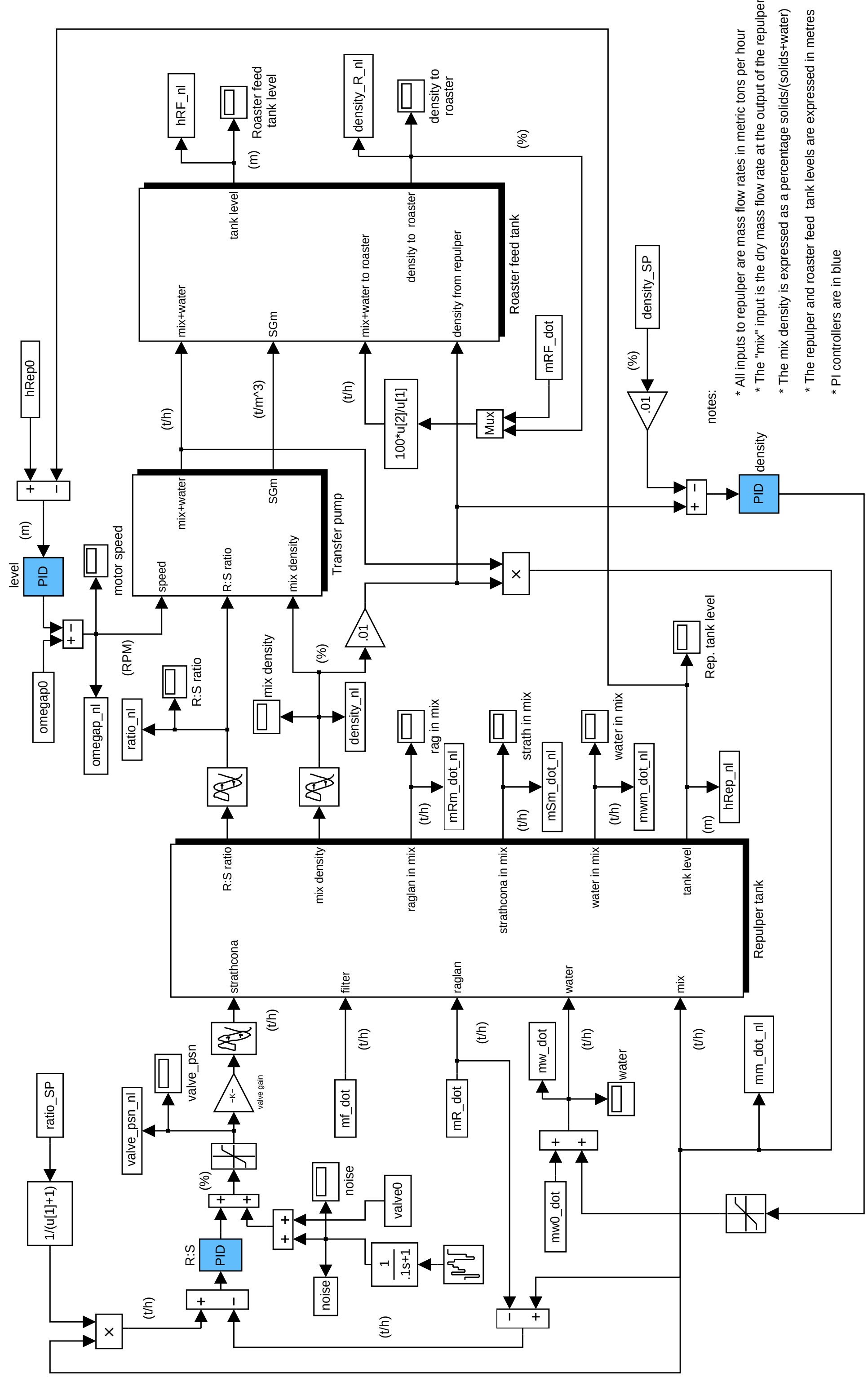


Figure 5: Main block diagram of closed-loop nonlinear simulator

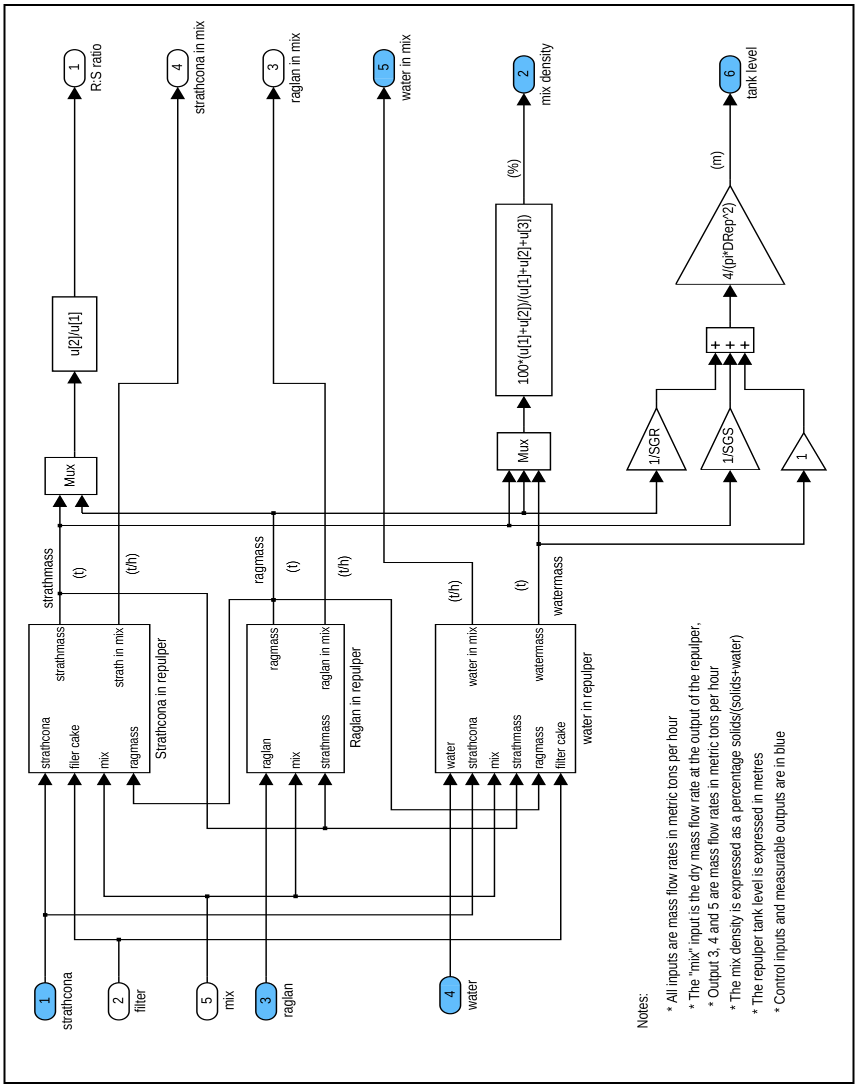


Figure 6: "Repulper tank"

Simulation Results

We ran several simulation experiments to test the following properties:
Robustness and disturbance rejection

- Change in the incoming Strathcona specific gravity
- Change in the incoming Strathcona pulp density
- Change in the incoming Raglan specific gravity
- Change in the roaster feed rate (from the operator)
- Normal lag times and process dynamics typical in a mixing problem
- "Manual" change in the Raglan flow rate being fed to the system

Tracking

- Change in the R:S ratio setpoint (from the operator)
- Change in the density setpoint (from the operator)

Operation with filters

- Addition and removal of filter(s)

Due to space limitations, we present only one simulation case: a step change in the R:S ratio setpoint (from the operator). The step change is from the nominal $r_0 = 0.309$ to $r = 1$.

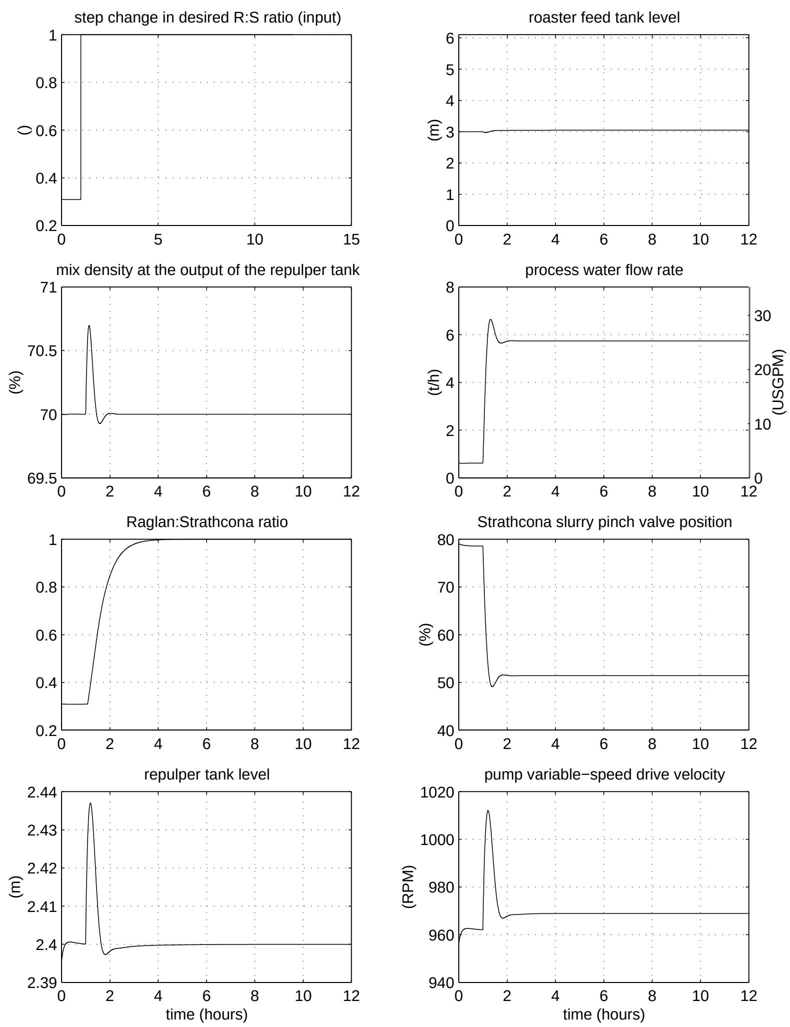


Figure 7: Response to a +0.691 step change in R:S ratio setpoint

Based on the closed-loop simulation results, the following observations were made:

- A decentralized PI control architecture was adequate
- The simulated control system was robust to changes in Strathcona slurry density and specific gravity
- The simulated control system tracked density and ratio setpoints adequately
- The addition or removal of a filter had only a minor effect on density at the output of the repulper
- For very large additive disturbances and perturbations on the slurry valve modeling the effects of changes in Strathcona properties, the output pulp density was still maintained within 1% in most cases. However, the R:S ratio started to vary significantly in some cases.
- Setting the Raglan flow rate manually only had a minor effect on density and ratio at the output of the repulper (unless saturation occurred in the slurry valve).

IMPLEMENTATION AND COMMISSIONING

The density, ratio and level PI controllers were implemented on the smelter's Foxboro distributed control system (DCS) using standard PID blocks. During commissioning, the calculated PI gains were tested, as well as other sets of gains to try to improve the response further. After a few days of testing it was found that the calculated gains offered the best trade-off between performance and robustness to variations in slurry and pulp density, which affect valve dynamics, transfer pump characteristics, etc. However, we decided to add a bit of derivative action to improve the closed-loop damping, leaving the low frequency loop gain basically unchanged.

CONCLUSION

The feed blending control system has been in operation since the spring of 1998, producing a consistent pulp density that ultimately improves the efficiency of Falconbridge's smelting process.

REFERENCES

KUO, B. C., 1985.
Automatic Control Systems. 5th Edition, Prentice-Hall, New-Jersey, 720 p.

THE MATHWORKS Inc., 1998.
Simulink User's Manual.

APPENDIX A: NOTATION

Symbol		**Variable**
H_{Rep}	:	Height of repulper tank (m)
D_{Rep}	:	Diameter of repulper tank (m)
V_{Rep}	:	Active volume of repulper tank (m^3)
$\dot{m}_S$	:	Dry mass flow rate of Strathcona feed (t/h) from valve
f_S	:	Strathcona slurry density in % solids
SG_s	:	Specific gravity of dry Strathcona feed (t/m^3)
x_v	:	Strathcona slurry pinch valve opening (%)
$\dot{m}_f$	:	Dry mass flow rate of Strathcona feed from filters (t/h)
f_f	:	Density of Strathcona slurry from filters (% solids)
$\dot{m}_R$	:	Dry mass flow rate of Raglan concentrate (t/h)
SG_R	:	Specific gravity of Raglan concentrate (t/m^3)
$\dot{m}_w$	:	Mass flow rate of water (t/h)
h_{Rep}	:	Pulp level in repulper tank (m)
r	:	Ratio of dry Raglan/Strathcona mass flow rates in mix
m_{SR}	:	Total mass of dry Strathcona in repulper tank
m_{RR}	:	Total mass of dry Raglan in repulper tank
m_{wR}	:	Total mass of water in repulper tank
$\dot{m}_m$	:	Dry mass flow rate of mix out of repulper and into roaster feed tank (t/h)
f_m	:	Mix slurry density in % solids
h_{RF}	:	Pulp level in roaster feed tank
$\dot{m}_{m+w}$	:	Mass flow rate of mix (including water) out of repulper tank (t/h)
$\dot{m}_{Sm}$	:	Dry mass flow rate of Strathcona in mix out of repulper tank
$\dot{m}_{Rm}$	:	Dry mass flow rate of Raglan in mix out of repulper tank
$\dot{m}_{wm}$	:	Mass flow rate of water in mix out of repulper tank
SG_m	:	Specific gravity of mix (t/m^3)
ω_{pm}	:	Velocity of transfer pump variable speed drive (RPM)