



## **The effect of process design on refiner pulp quality control performance**

Downloaded from: <https://research.chalmers.se>, 2026-04-04 13:32 UTC

Citation for the original published paper (version of record):

Sund, J., Sandberg, C., Karlström, A. et al (2021). The effect of process design on refiner pulp quality control performance. *Nordic Pulp and Paper Research Journal*, 36(4): 594-607.

<http://dx.doi.org/10.1515/npprj-2021-0011>

N.B. When citing this work, cite the original published paper.

## Mechanical pulping

Johan Sund\*, Christer Sandberg, Anders Karlström, Göran Thungström and Per Engstrand

# The effect of process design on refiner pulp quality control performance

<https://doi.org/10.1515/npprj-2021-0011>

Received February 11, 2021; accepted May 28, 2021

**Abstract:** In this study, the effect of process- and online analyser configuration on pulp quality control is explored. The following parameters were included: analyser sampling interval, time delay, measurement error magnitude, and latency chest residence time. Using different values of parameters in a process model, a range of configurations were constructed. For each configuration, the achievable control performance was evaluated using an optimization approach. PI controller settings were chosen based on minimization of the integrated absolute error (*IAE*) in pulp quality after an input step disturbance. The results show that reducing the sampling interval improves performance also when the interval is smaller than the chest residence time or the analyser delay. Moreover, reducing the chest residence time can reduce the *IAE* by up to 40 %. However, reducing the residence time to lower than 1/3 of the sampling interval does not improve performance. Further improvement is possible if the analyser delay is reduced. The compromise between reducing the *IAE* and avoiding creating variation by acting on measurement error has a strong influence on the results. In conclusion, pulp quality control performance can be improved significantly by making changes to the studied configuration parameters.

**Keywords:** latency chest; measurement error; mechanical pulp; optimization; sampling rate.

\*Corresponding author: Johan Sund, Mid Sweden University, SE-85170 Sundsvall, Sweden; and Holmen Paper, SE-60188 Norrköping, Sweden, e-mail: [johan.sund@holmen.com](mailto:johan.sund@holmen.com)

Christer Sandberg, Holmen Paper, SE-60188 Norrköping, Sweden; and Mid Sweden University, SE-85170 Sundsvall, Sweden, e-mail: [christer.sandberg@holmen.com](mailto:christer.sandberg@holmen.com)

Anders Karlström, Department of Electrical Engineering, Chalmers University of Technology, SE-41296 Göteborg, Sweden, e-mail: [anderska@chalmers.se](mailto:anderska@chalmers.se)

Göran Thungström, Department of Electronics Design, Mid Sweden University, SE-85170 Sundsvall, Sweden, e-mail: [goran.thungstrom@miun.se](mailto:goran.thungstrom@miun.se)

Per Engstrand, FSCN, Department of Chemical Engineering, Mid Sweden University, SE-85170 Sundsvall, Sweden, e-mail: [per.engstrand@miun.se](mailto:per.engstrand@miun.se)

## Introduction

Changing the design of a process and its instrumentation may enable significant improvement of control performance (e.g. Sharifzadeh 2013). The extent to which re-design is possible will of course differ between applications. In mechanical pulp production, control performance is affected by the choice of e.g. on-line analyser sampling rate, and the size of the preceding pulp chest (Hill et al. 1979, Blanchard and Fontebasso 1993). The combined effects of such parameters have not been studied systematically. Following an introduction to refiner pulp production this study explores the effect of design choices on pulp quality control performance.

## Refiner mechanical pulp production

Although there is a large variety of processes for refiner-based mechanical pulp production (Sandberg et al. 2020), the challenge of pulp quality control is similar. As an example, a single stage pressurized refiner mechanical pulping process is described in Figure 1. Wood chips are fed into a pressurised pre-heater, with a metering screw in the bottom which sets the throughput. Chips are forced to the entry-zone between a rotating and a stationary disc, where a coarse pattern of bars create a coarse pulp. Fibres and fibre bundles are further refined as they move through the gap between the discs (Atack 1980). After refining, the high-consistency pulp is transported by the steam flow via a blow-line to a steam-separator. At its discharge, hot water is added, and the pulp falls into a large chest where it is mixed to a consistency of 3–6 %. After this chest, pulp is sampled by an online pulp quality analyser. The results are delayed, infrequent, and include significant random errors. This can limit their use for control purposes.

The main reason for including a large agitated pulp chest is that refiner pulp requires disintegration at high temperature. When fibres exit the refiner, they are gathered in aggregates (Karnis et al. 1986), and individual fibres and fibre fragments are twisted, bent, or distorted in other ways (Mohlin 1980, Karnis 1993, Gao et al. 2013). Hot disintegration breaks up aggregates and straightens

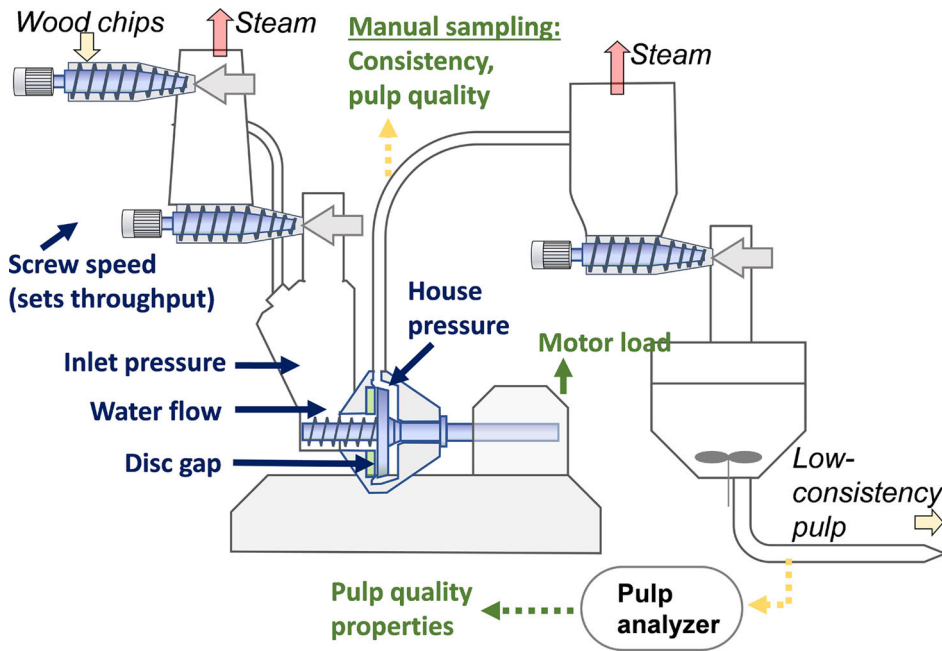


Figure 1: Schematic illustration of a single stage refining process, showing some of the process variables.

fibres. This is called latency treatment, since it releases the “latent strength properties” of the pulp (Beath et al. 1966). Most of the early refiner lines incorporated a latency chest, often with a residence time around 20 minutes (Evans 1978). There has however been a large variation in chest designs in terms of size, agitation, etc. Later, smaller chests have been more common (e. g. Tien-rii et al. 1999, Mokvist et al. 2005). Latency removal proceeds faster with higher temperature, impeller power input, and consistency (Gao 2014). This suggests that a relatively small chest can achieve latency removal, perhaps at lower energy use (Dawson et al. 1978). Some mills seem to have implemented such designs (e. g. Tamminen et al. 1987, Sikter et al. 2007).

The main objective when operating a refining process is to produce pulp with low variation around the target quality, preferably with as low energy use as possible. The basic instrumentation and control set-up includes control of steam pressures, water flows, and disc gap (or hydraulic pressure), as well as protection systems. Many processes also include strategies to control refining consistency and/or motor load. The choice of control strategies for the use of continuous process measurements has a clear impact on the variation of pulp quality (e. g. Roche et al. 1996, Kortelainen et al. 1997, Eriksson and Karlström 2009, Karlström et al. 2020). However, the topic of this paper is feedback control of pulp quality, and how it is influ-

enced by design choices regarding the pulp tester and the pulp chest.

### Design choices for pulp chest and online analyser

In Figure 2, the most frequently used chest designs are shown: A) a large latency chest, B) a standpipe (no mixing), C) a small well-mixed chest. The small chest may be operated at higher consistency, and higher impeller power input, especially if it is a “pulper” design with the impeller in the bottom. In contrast, there is essentially no mixing in a standpipe (B). Several mills have added a small “transfer chest” before an existing latency chest, to enable better use of online pulp testing (Blanchard and Fontebasso 1993). Some latency chests may be described as a series connection of a mixed volume and a plug-flow volume (Tessier et al. 1997, Ein-Mozaffari et al. 2004).

On-line analysers for pulp drainage rate gained widespread use for mechanical pulp production in the 1980ies (Brewster and Rogers 1985). While the drainage test method differs between models, the results are usually calibrated to match laboratory results of Canadian Standard Freeness (CSF) (Roche et al. 1997). Today, the most popular analysers measure several pulp properties. The basic tests are drainage rate, and optical analysis of fibre and shive size distributions.

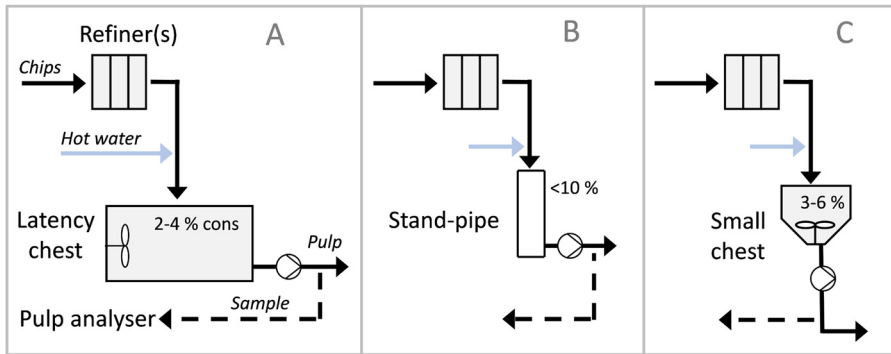


Figure 2: Different choices of pulp chest.

The time required for a pulp test is around 1–10 minutes, depending on the time used for the analysis step, and for sample preparation (Dilution, disintegration, sub-sampling). The transport from the sampling position also adds a short delay.

When an analyser serves multiple positions, the sampling interval for each position is increased. Connecting e. g. 4 positions usually gives an interval in the range 20–35 minutes. While such configurations are common, shorter intervals are possible (e. g. Tamminen et al. 1987, Roche et al. 1996, Tessier et al. 1997, Johansson et al. 2018).

Each test result includes a random error due to variations in sample extraction, preparation and analysis. Moreover, there is a random “misrepresentation” error due to infrequent sampling and fast variations. This can be described as “sampling noise” (e. g. Alsip 1981), and it can be explained by the aliasing effect (e. g. Åström and Wittenmark 1997). Variations which are fast, in relation to the sampling interval, give a contribution to each sample which is not representable. The presence of fast variations in pulp quality is known from manual blow-line sampling and indicated by variations in blow-line consistency as well as motor load (Strand 1996, Ferritsius et al. 2017). Since a well-mixed pulp chest smoothens out fast variations, it can be used to reduce sampling noise.

Automatic control of pulp quality has in many applications been limited to a single property, often pulp freeness. Mean fibre length has also been controlled (Karlström et al. 2015). Pulp quality can be controlled by manipulating one of the basic process variables, e. g. second stage disc gap, or by manipulating e. g. refiner motor load or consistency (e. g. Roche et al. 1996). Some control strategies use the pulp tester in a model-based approach, where the test results are used to correct a pulp quality model. Different forms of model predictive control (MPC) have been studied and applied, in some cases together with a pulp quality model (Strand et al. 1999, Karlström and Isaksson 2009,

Harinath et al. 2013). In the context of this study, it is not the under-lying control structure which is of interest.

Although few studies have investigated the impact of refining process design choices on pulp quality control performance, the literature contains many short statements that the latency chest is too large or that the sampling rate is too low (e. g. Honkasalo et al. 1989, Blanchard and Fontebasso 1993). The importance of the sampling interval was investigated by Hill et al. (1979), based on logged quality variation from an online analyser. Methods are available for evaluating control performance of different processes at varying sampling interval (e. g. Lennartson 1990, Åström and Wittenmark 1997, Horch and Isaksson 2001). To better understand the importance of refining process design parameters, they should be studied with respect to suitable indices of control performance, and for the large range of possible configurations.

## Objectives and limitations

The objective of this study is to explore how control performance is affected by the size and design of the latency chest, as well as the configuration of the online analyser. Direct feedback control of a single quality property is studied. Latency treatment is not studied, it is assumed that it can be achieved with a relatively small chest, see Gao (2014) and Dawson et al. (1978).

## Materials and methods

When comparing different process designs (or control strategies) it is important that relevant aspects of control performance and robustness are included. In the case of pulp quality control, the most relevant aspects are: the response to disturbances (such as raw material variation),

the response to random measurement error, and that the behaviour of the refiner will change over time.

The effect of process- and instrumentation alternatives on control performance can be evaluated by removing the influence of the choice of controller. This can be done by designing a controller for each configuration using a strict specification. Several aspects of performance can be included by setting the controller to optimize a main performance index, with constraints on robustness and/or sensitivity to measurement noise (e. g. Kristiansson and Lennartson 2006).

In this paper, the effect on control performance will be investigated for the following configuration parameters:

- **Mixed chest time constant ( $T_R$ )**, which is determined by the residence time of the part of the pulp chest volume which can be considered well-mixed.
- **Time delay ( $T_D$ )**, due to sample preparation and analysis, transport in piping to the analyser, and in some cases due to a plug-flow region in the pulp chest.
- **Sampling interval ( $T_S$ )**, of the pulp quality analyser.
- **Random error**, i. e. deviations of analyser results due to variations in sample extraction, preparation and analysis. Sampling noise may also be included, since it has a similar effect.

The chosen approach is to design a controller for a wide range of “process designs”, defined by values of  $T_R$ ,  $T_D$  and  $T_S$ . For each design, PI controller parameters were determined by optimization (see Garpinger and Hägglund 2015, Soltész et al. 2017). The main performance index is the integrated absolute error (*IAE*) after a step disturbance. Values of controller parameters were chosen to minimize the *IAE*, under constraints for two complementary indices; 1) The peak amplification of disturbances must be less than a given level, 2) The sensitivity to measurement error must be less than a given level.

The effect of  $T_R$ ,  $T_D$ , and  $T_S$  on control performance was compared based on the resulting *IAE*. The importance of random error was examined by varying the level of the second constraint. In the following, the method is explained in detail. An additional aspect is that a mixed chest filters out fast variations, reducing “sampling noise”. This is treated separately.

## Process model and control algorithm

A refining process is modelled as shown in Figure 3. The manipulated variable,  $u$ , is denoted disc gap. The input disturbance represents variations in e. g. chip quality or throughput. The pulp chest is modelled as “perfectly

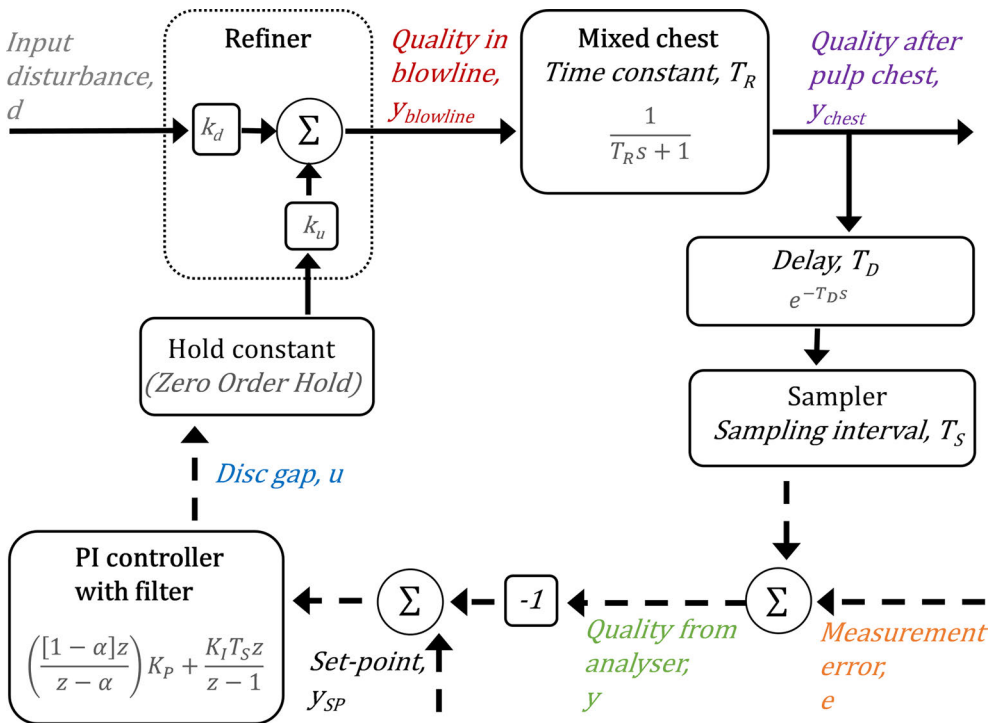


Figure 3: Process model for simulation of pulp quality control. Colours are used in simulation results.

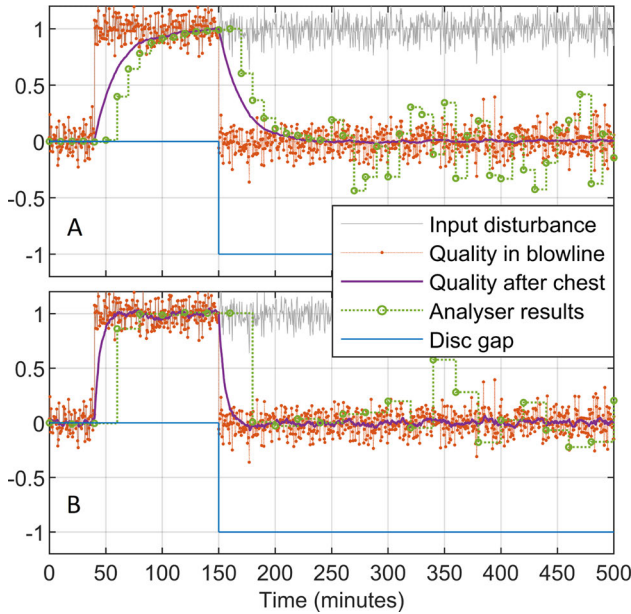
mixed”, i. e. a low-pass filter, with time constant  $T_R$ . Analyser results are available at discrete instants separated by uniform intervals set by  $T_S$ . Results are delayed by  $T_D$  minutes. Note that delay is often mainly in the analyser, but it is placed before the sampler to simplify calculations. This has no effect on results (see Appendix). The model can be used to simulate the pulp quality response to input disturbances and measurement error, generated from a normal distribution with standard deviation  $\sigma_e$ .

Since the residence time of a refiner is relatively short, it is modelled without dynamics, according to

$$y_{blowline}(t) = k_d d(t) + k_u u(t), \quad (1)$$

where  $k_d$  and  $k_u$  are the gains from input disturbance and disc gap. The relative scale of variables is not important in this study. Therefore,  $k_d$  and  $k_u$  are set to 1 unless stated otherwise.

The character of two different process designs is illustrated in Figure 4 (without controller), showing the response to input disturbance (grey) and disc gap (blue).



**Figure 4:** Response of two different process designs to a step in input disturbance and disc gap. From  $t = 250$ , measurement error is added. A:  $T_R = 20$ ,  $T_D = 10$ ,  $T_S = 10$ . B:  $T_R = 5$ ,  $T_D = 10$ ,  $T_S = 20$  minutes.

The blow-line quality (red) is directly affected by the input disturbance. After the pulp chest, the pulp quality has been filtered (purple). The analyser measurements (green) are separated by intervals of length  $T_S$  and are also delayed by  $T_D = 10$  minutes. After the disc gap step in 4B,

nearly the full effect is visible already in the next sample, while in 4A some of the gradual change is captured. The response in measured quality in 4A is that of a discrete “first order with time delay process”. The process in 4B is simpler since  $T_S$  is large compared to  $T_D$  and  $T_R$ .

Proceeding to the controller, a PI algorithm with filter was chosen for this study. The controller acts on infrequent measurement results (samples), given by  $y$  in Figure 3. Each time a new result is presented, the control algorithm decides on a new disc gap,  $u$ , based on the deviation between the measured value and the set-point,

$$y_{dev}(k) = y_{SP}(k) - y(k). \quad (2)$$

A simple control algorithm is to change the disc gap in proportion to the deviation, i. e.

$$\Delta u(k) = K_I T_S y_{dev}(k), \quad (3)$$

$$u(k) = u(k-1) + \Delta u(k). \quad (4)$$

This is referred to as a discrete time integral controller, where  $K_I$  is the integral gain. The sample index  $k$  is used to identify the sample at the time  $t = k * T_S$ . In-between sampling instants  $u(t)$  equals the value set by the last control action.

A proportional controller is tuned by the proportional gain,  $K_P$ , and sets gap in relation to the current deviation,

$$u(k) = K_P y_{dev}(k). \quad (5)$$

A discrete PI controller is given by the sum of the integral part and the proportional part (in incremental form),

$$\Delta u(k) = K_I T_S y_{dev}(k) + K_P [y_{dev}(k) - y_{dev}(k-1)]. \quad (6)$$

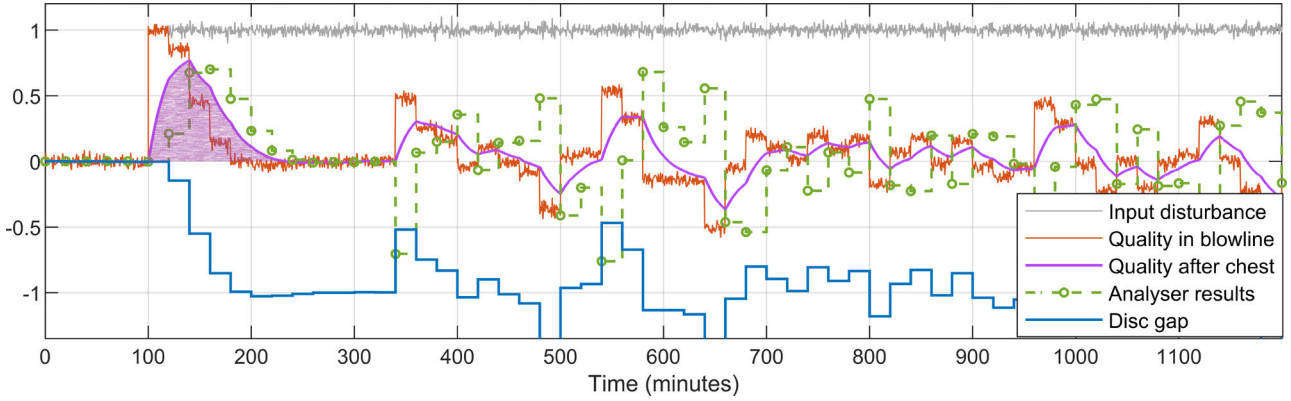
The output of the first order low-pass filter,

$$y_{filt}(k) = \alpha y_{filt}(k-1) + (1 - \alpha) y_{dev}(k), \quad (7)$$

is a weighted average of past results, where  $\alpha$  sets the strength of the filter. In Figure 3, a filter on the P-part is included in the controller (in transfer function form). Filtering the I-part is less motivated, since it is already a form of filter.

## Control performance and process design comparison

For many refining processes, the response to set-point change is of secondary interest. The main objective is to reduce the variation in pulp quality due to changing raw material and other disturbances. An additional objective is to



**Figure 5:** Response to an input disturbance step at  $t = 100$ , and to random measurement error after  $t = 340$ .  $T_R = 20$ ,  $T_D = 15$ ,  $T_S = 20$  minutes. The IAE is given by the shaded purple area. PI controller with  $K_p = 0.26$ ,  $K_I = 0.44$ . Noise is added to the input disturbance for graphical reasons.

avoid acting on measurement error, since this creates quality variation. These objectives require a compromise. To illustrate this, a simulation is presented in Figure 5, where a given process is controlled by a PI controller. Moreover, the process behaviour can change over time, which also motivates defensive controller settings to achieve robustness to process change.

A popular index for performance is given by the shaded purple area in Figure 5, between the pulp quality curve (purple) and its set-point (which is 0 in this case). This is the integrated absolute error, and is in this case given by

$$IAE = \int_0^{\infty} |y_{SP}(t) - y_{chest}(t)| dt, \quad (8)$$

where  $y_{chest}$  is the pulp quality after the chest (Figure 3). When there is no over-shoot in the step response, the IAE equals the integrated error (IE).

The chosen method for controller design is based on minimization of the IAE after an input step disturbance. For many processes, the lowest possible IAE is achieved at the cost of other performance objectives. For this reason, constraints are added to limit the sensitivity to measurement noise, and to ensure robustness (Garpinger and Hägglund 2015, Soltesz et al. 2017). This involves performance indices which are usually defined and calculated using transfer functions. These indices are presented briefly below, and the calculations are presented in the Appendix.

The sensitivity to random measurement error is shown in the later part of the time series in Figure 5. Although the step response looks good, the controller creates much quality variation by acting too strongly on random analyser errors. The random errors are uncorrelated (i. e. white

noise) and generated from a normal distribution. The controller acts on measurement error with standard deviation  $\sigma_e = 0.30$ , and thereby creates variation in blowline pulp quality (red) with standard deviation  $\sigma_{y,e} = 0.225$ . An index for the sensitivity to measurement error is the ratio  $\sigma_{y,e}/\sigma_e$ , here 75 % (e. g. Garpinger and Hägglund 2015). This index is here referred to as the “noise transfer ratio”.

Robustness constraints are often set using a frequency domain description of the controlled process. When periodic disturbances act on a process under feedback control, those with low frequency are reduced in amplitude (compared to when the manipulated variable is held constant), but variations with high frequency are amplified. Since pulp quality is controlled using sampled (discrete) results, we use the step-invariant discretization of the process and study the discrete frequency response (e. g. Åström and Wittenmark 1997). Details are given in the Appendix.

The *peak amplification of disturbances*,  $M_S$ , is a complementary index of control performance, and it also indicates robustness to change in process behaviour (e. g. Åström et al. 1998). It is common to include a limit on  $M_S$  as a part of a controller design specification. For comparison of different controllers, the limit has often been set in the range 1.4–1.7 (e. g. Åström et al. 1998, Kristiansson and Lennartson 2006, Soltesz et al. 2017).

For a given process design, the optimization problem to be solved is:

Find values of controller parameters  $K_p$ ,  $K_I$  and  $\alpha$  that minimize IAE, under the constraints  $M_S < c$ , and  $\frac{\sigma_{y,e}}{\sigma_e} \leq d$ .

The effect of the process design is explored by varying  $T_R$ ,  $T_D$ , and  $T_S$ , while the importance of measurement error is studied by varying the level of the constraint on  $\sigma_{y,e}/\sigma_e$  between 30 and 70 %.

The constraint on  $M_S$  was set to 1.4. The choice of controller was either PI ( $\alpha=0$ ) or PI with filter (all parameters optimized). Optimization was performed using a general-purpose interior-point-algorithm, available in the Optimization Toolbox for Matlab. The values of  $M_S$  and  $\sigma_{y,e}/\sigma_e$  were calculated using transfer functions, whereas the  $IAE$  was calculated based on results of simulation using Simulink (details are given in the Appendix).

## Pre-filtering of fast variation before pulp sampling

In the previous section the aspect of sampling noise was not included. A smaller mixed chest time constant (residence time) enables better step response performance ( $IAE$ ), but worse pre-filtering (more sampling noise). The impact of sampling interval and pre-filter time constant on control performance is complex (Lennartson and Middleton 2014). Since there is a lack of knowledge about the character of variations after the refiner, we use a simple method and estimate that the pre-filtering function can be made good enough with a small negative impact on the  $IAE$ . The level of sampling noise depends on the choice of mixed chest time constant and sampling interval. It also depends on the variance and character of the fast quality variations after the refiner. The variance of fast blowline quality variations can be large, but the character of these variations is less known, since few have studied a large number of samples (see Ferritsius et al. 2017).

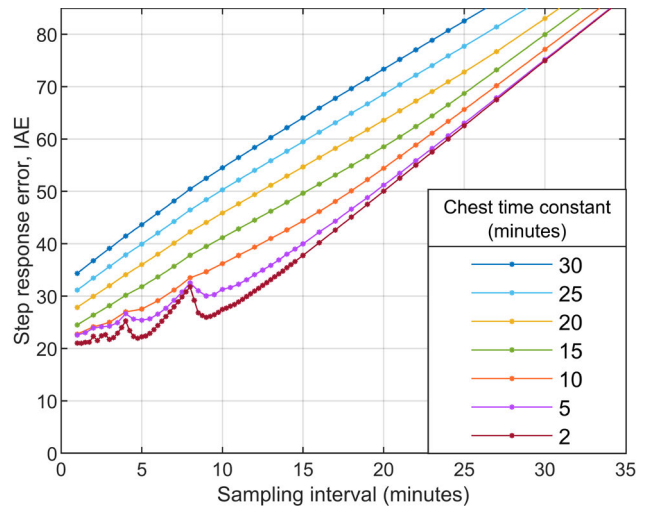
A simple index of pre-filter performance is its dampening at the Nyquist cycle time,  $2 * T_S$ . A well-mixed chest can be modelled as a first order low pass filter with time constant  $T_R$ . At the cycle time  $2 * \pi * T_R$ , this filter reduces the amplitude to 71% (square root of 0.5), and faster frequencies are damped more. To have 71% dampening at  $2 * T_S$  requires that  $T_R$  is around  $1/3$  of  $T_S$ . With limited knowledge of the character of fast quality variations, it seems reasonable to estimate that the choice  $T_R > 1/3 * T_S$  will reduce sampling noise variance to a level where it is small in relation to the other random errors.

## Results

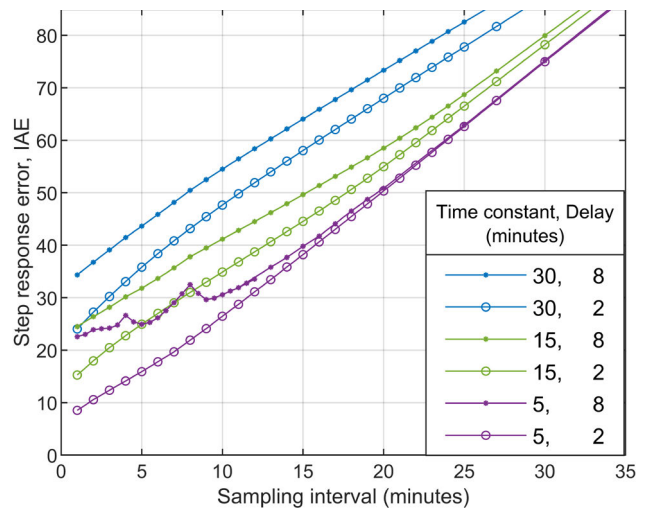
The achievable control performance was determined for a wide range of configurations, with varying choice of mixed chest time constant ( $T_R$ ), analyser time delay ( $T_D$ ), and sampling interval ( $T_S$ ), all expressed in minutes. For a well-mixed latency chest, we can interpret  $T_R$  as its residence

time. Time delay is often mainly due to the analyser but may also be due to plug flow in a latency chest and sample transport. The main performance index is the integrated absolute error ( $IAE$ ) of the response to an input step disturbance. It has the unit minutes, see (8), which may be perceived as a strange measure of performance. These values can be related to e. g. the settling time using the example in Figure 5, and the results can be used to compare the relative effect of changes to an existing process.

As seen in Figure 6 and 7, control performance (step response error,  $IAE$ ) depends strongly on the chosen sam-

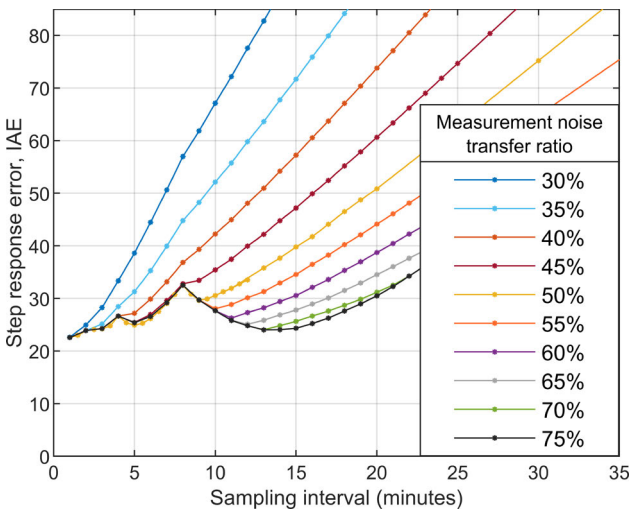


**Figure 6:** Control performance ( $IAE$ ) vs sampling interval for different chest time constants (residence time). Analyser delay = 8 min. PI controller.  $M_S \leq 1.4$ . Noise transfer ratio  $\leq 50\%$ .

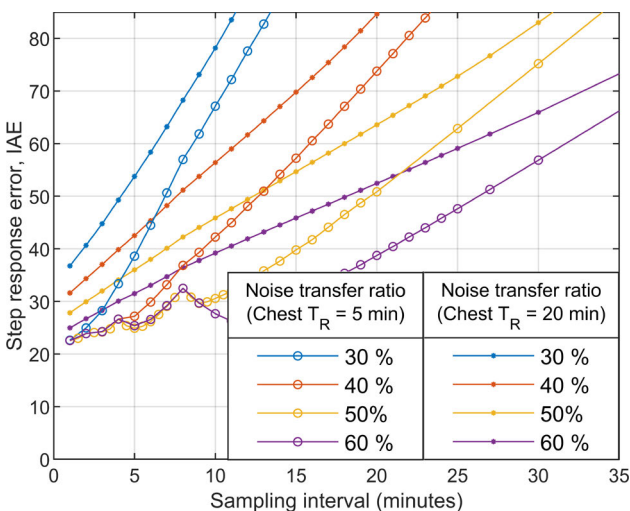


**Figure 7:** Control performance ( $IAE$ ) vs sampling interval for different analyser delay and chest time constants (residence time). PI controller.  $M_S \leq 1.4$ . Noise transfer ratio  $\leq 50\%$ .

pling interval and chest time constant. In these figures, the constraint on the noise transfer ratio is set to 50 %. The relations change depending on this level, as shown in Figures 8 and 9. The PI controller without filter is used in Figures 6–9. Adding a filter had only a small effect on the results (See Appendix), and therefore we present mainly results without filter. In the largest part of the studied range, there is no single parameter which sets a limit for performance.



**Figure 8:** Control performance (IAE) for different levels of constraint on the noise transfer ratio. Chest time constant = 5 min (residence time), analyser delay = 8 min.  $M_S < 1.4$ , except for the bottom line where  $M_S = 1.4$ .



**Figure 9:** Control performance (IAE) for different level of constraint on the noise transfer ratio. Compares two chest time constants ( $T_R$ ). Analyser delay 8 min.  $M_S \leq 1.4$ .

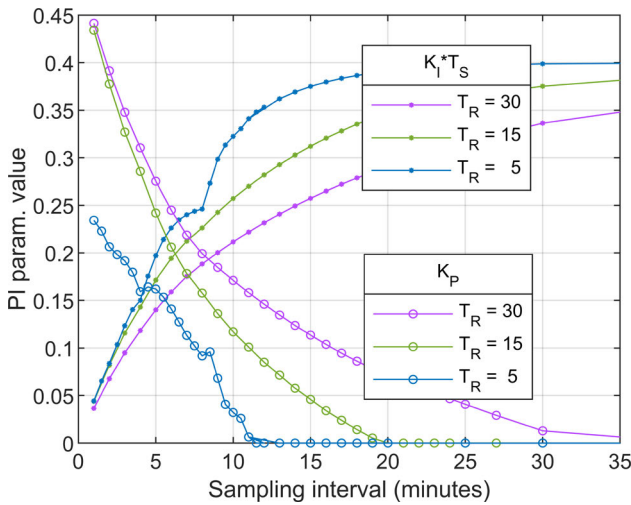
Reducing the sampling interval improves performance (reduces the *IAE*) following an almost linear relationship (Figures 6 and 7). The exception is the curves for 5- and 2-minute chests, which flatten out after small peaks related to the 8-minute analyser delay, in relation to  $T_S$  (See Appendix).

Reducing the chest time constant also improves performance significantly, but not when it is small in relation to the sampling interval. In Figure 6, the *IAE* improvement when changing from  $T_R = 20$  (and  $T_S = 20$ ) to  $T_R = 5$  is the same as that of changing to  $T_S = 13$ . However, the improvement from further reducing the chest size from  $T_R = 5$  (at  $T_S = 20$ ) is small. A change of  $T_R$  from 30 to 5 minutes can reduce the *IAE* by up to 40 %. Reducing the time constant to less than around 1/3 of the sampling interval does not improve performance significantly. Moreover, reducing the time constant to less than the delay does not improve performance.

The effect of reducing the delay is generally stronger than that of reducing the time constant by the same number of minutes. Reducing  $T_D$  is especially effective when  $T_S$  is less than  $T_D$  (Figure 7). In practise, this can occur when there is a latency chest with a plug-flow region, or a very long sample transport pipe. More commonly, the sampling interval is larger than the time delay, and the effect of reducing  $T_D$  is then less pronounced.

The effect of measurement error standard deviation ( $\sigma_e$ ) can be retrieved indirectly from Figure 8, which has two important interpretations. The direct interpretation is that the *IAE* can be reduced by choosing a less strict limit on the noise transfer ratio ( $\sigma_{ye}/\sigma_e$ ). Indirectly, if  $\sigma_e$  can be decreased, the *IAE* can be reduced. For example, reducing  $\sigma_e$  by 30 % ( $\sigma_e^2$  by 50 %) without changing the controller settings results in a reduction of the induced quality error ( $\sigma_{ye}$ ) by 30 %. To stay at the same level of  $\sigma_{ye}$ , the controller can be set with higher  $\sigma_{ye}/\sigma_e$ . The consequence of this can be seen by comparing the curves of 35 % and 50 % ( $35\%/0.70 = 0.50$ ), or those of 45 % and 65 % ( $45\%/0.70 = 0.64$ ). From this comparison, the effect of reducing standard deviation by 30 % is approximately equal to that of reducing the sampling interval by 50 %.

With a stricter limit on the noise transfer ratio, the importance of the sampling interval increases (Figure 9). This is illustrated by comparing performance for designs with  $T_R = 5$  and 20 minutes (see also Figure 4). As an example, the process in Figure 4A is better than that in 4B when the noise transfer ratio is less than 50 %. Since the random error of pulp quality tests is often significant in relation to the real variation in pulp quality (e.g. Hill 1993), the controller design should not allow transfer of more than



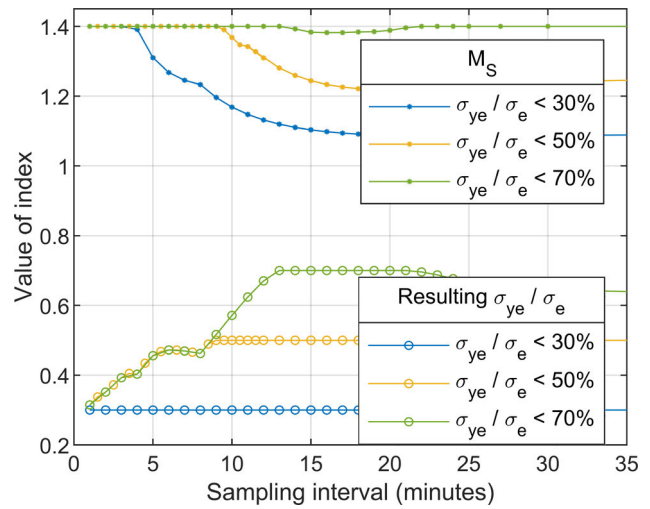
**Figure 10:** PI Controller parameter values ( $K_I$ ,  $K_P$ ) vs sampling interval ( $T_S$ ) for some of the curves from Figure 6, with different chest time constants,  $T_R$ .

20–50%. This is a difficult compromise, since the  $IAE$  increases strongly when the noise transfer ratio is decreased.

The almost linear relations between performance and sampling interval in Figures 6–9 are related to the constraint on the noise transfer ratio. This nature of the results is characteristic for the combination of this type of constraint and the  $IAE$  as the main performance index. The “robustness constraint” ( $M_S \leq 1.4$ ) is inactive for most of the points in Figures 6–9, as shown in Figure 11. In the range where this constraint is active, it generates the peaks in the curves, and bends the otherwise relatively straight curve.

Figure 11 shows the peak amplification and the noise transfer ratio corresponding to the results in Figure 8. The constraint on  $\sigma_{y,e}/\sigma_e$  has a greater influence on the results. Even if a quite strict constraint with  $M_S \leq 1.2$  is used, the solution for  $\sigma_{y,e}/\sigma_e$  is only changed for  $T_S < 10$  minutes, since  $M_S$  is already less than 1.2 for larger  $T_S$ . Moreover, the analyser random error may be of significant magnitude in relation to the actual quality variations. Consequently, for the major part of the studied range of configurations the controller settings can be considered limited by the noise transfer ratio. When a controller is set for a reasonable compromise between reducing the effect of input disturbances and avoiding creation of quality variation by acting on random measurement noise, the robustness properties (here described by  $M_S$ ) will also be good.

The PI controller settings for the curves in Figure 6 are shown in Figure 10. The optimal PI controller turns into an integral-only controller when the sampling interval is relatively long. Also, for the three evaluated chest time con-



**Figure 11:** Peak amplification ( $M_S$ ) and noise transfer ratio ( $\sigma_{y,e}/\sigma_e$ ) for the curves in Figure 8, resulting from varying constraint:  $\sigma_{y,e}/\sigma_e \leq 30, 50$  or  $70\%$ .

stants, as the sampling interval becomes relatively long, the product  $K_I * T_S$  converges to the same value.

A better step response (lower  $IAE$ ) implies a higher bandwidth, i. e. that the effect of faster disturbances can be reduced. This is important, since raw material variations often have a stochastic character, with variation in a wide band of frequencies. The bandwidth for the response to input disturbances can be calculated from the results (See Appendix). For example, consider the 20-minute chest in Figure 6 at a sampling interval of 15 minutes. An input disturbance with cycle time of 7.7 hours is reduced in amplitude to 70% using the optimal PI controller. Faster variations are either only slightly damped or amplified up to a maximum of 130% ( $M_S$ ). This emphasizes that the peak amplification is also a performance index, and that it should not be chosen too high. A fast step response is also important when disturbances appear as short-term deviations from a relatively constant level.

### Pre-filtering of fast variation before pulp sampling

In the last part of the methods section it is estimated that the level of sampling noise can be made relatively low using a mixed chest with time constant (residence time) of  $1/3$  of the sampling interval. Together with the main part of the results, this indicates that pre-filtering can be achieved without significant reduction of step response performance. However, sampling noise can be significant if the pulp chest is not mixed, as in the case of a standpipe.

## Discussion

A wide range of latency chest and analyser configurations are used in mechanical pulping mills, and the range of possible configurations is even wider. Most mills today use multi-property analysers which serve several positions, resulting in long time delay and long sampling intervals. Shorter intervals are possible by using one analyser for each position, or by using alternative analysers (e. g. Tessier et al. 1997, Johansson et al. 2018).

A clear result of this work is that control performance can be improved by reducing the sampling interval. An exception occurs when the measurement noise transfer ratio is high, and the chest is small. Then, performance is limited mainly by the time delay. However, this exception is less relevant. In practise, the noise transfer ratio should not be set higher than 20–50 %. Then, there is an almost linear dependence of  $IAE$  on  $T_S$  (Figure 7). The slope depends mainly on the allowed transfer of measurement error (Figures 8, 9).

A small well-mixed pulp chest after refiners offers a short time constant and thereby enables improved step response performance, but it should also provide latency treatment and pre-filtering to reduce sampling noise. Latency treatment can be achieved in a short time using high intensity agitation, but there is a lack of reported mill results (Dawson et al. 1978, Tamminen et al. 1987, Gao 2014). If the need for latency treatment does not limit the chest size, control performance can be significantly improved by using a smaller chest, especially together with a short sampling interval. There are also other benefits of having a small pulp chest. With a relatively small chest and a short sampling interval, it is possible to detect short-term variations (Karlström and Hill 2018). These are of interest when trying to understand the process, and when evaluating control strategies. Moreover, manual control actions are easier to evaluate, since the final level of the response is approached faster.

To further improve the use of pulp property measurements in different control concepts, the random error of measurement results should be reduced. Sampling noise can be reduced by using a well-mixed chest. Composite sampling can reduce sample extraction error and to some extent also sampling noise (see Ferritsius et al. 2017). For multi-property testers, the random analysis error of each property may be reduced by reconciling the results (Strand et al. 1989). The random analysis error depends on the choice of pulp property and analysis method. The strong trade-off between reducing the effect of disturbances and avoiding action on measurement noise emphasizes the importance of reducing measurement error.

This work shows how the use of pulp quality feedback control can be improved by changes to the latency chest and to the configuration of the online analyser. Variation in pulp quality naturally also depends on the performance of the continuous control strategies, such as control of refining consistency. The importance of pulp quality feedback control in a specific mill depends on the character of the disturbances which cause quality variations. Further assessment of pulp quality control can be made using logged values of variation in quality (Hill et al. 1979, Toivonen and Tamminen 1990, Horch and Isaksson 2001). Two additional aspects on control performance is how fast the controller can react after a set-point change, and how fast it can find the right pulp quality after the controller is activated, e. g. after start-up of the refiner. These functions are related to the input disturbance step response but can be modified separately.

## Conclusions

In this paper, analysis of a range of simulated process designs has shown that:

- Reducing the pulp chest time constant improves control performance, down to a limit set by the sampling interval and the analyser delay.
- Reducing the time delay improves performance, also when it is only a fraction of the sampling interval.
- Using a pulp chest without mixing results in sampling noise due to fast variations. The study indicates that sampling noise is not significant for a small well-mixed chest.
- Reducing the sampling interval improves performance, also when the interval is smaller than the chest time constant or the analyser delay.
- When low sensitivity to measurement error is important, the achievable performance depends strongly on the sampling interval length.

**Acknowledgments:** We gratefully acknowledge FORIC (Forest as a Resource Industrial Research College) at Mid Sweden University. Many thanks to Torsten Wik at Chalmers University of Technology, and Magnus Rydstrand at Holmen Paper for valuable advice.

**Funding:** This work was supported by Holmen Paper AB and the Knowledge Foundation, grant number: 20170305.

**Conflict of interest:** The authors declare no conflicts of interest.

## Appendix

In the following, some additional comments about the results are provided. But first, a detailed description of the control performance evaluation used in the study are given.

The study of a continuous process under feedback control from discrete samples offer some challenges. Advanced methods are available for study of such systems, but a common approach is to form a discrete representation of the system and study only the behaviour at the discrete sampling instants (e. g. Åström & Wittenmark 1997). This approach is used in this paper, but it is combined with simulation to evaluate the continuous-time response to an input step disturbance.

The discrete process model is formed by zero-order-hold (ZOH) sampling of a continuous process model (Figure 3),

$$P(s) = \frac{1}{T_R s + 1} e^{-T_D s}. \quad (9)$$

Note that the analyser delay is included as a part of this continuous process model, even though it in practise occurs after sampling. This makes it easier to form a discretized transfer function to relate the input disturbance to the analyser output.

Since the studied range contains configurations where  $T_D$  is a fraction of  $T_S$ , the step-invariant-transformation is slightly complicated (see Åström and Wittenmark 1997). Functions for step-invariant sampling are available in e. g. Matlab, through the function `c2d`. This function also provides the correct step-invariant discretization in the case of fractional delay.

Step-invariant discretization of  $P(s)$  at the sampling interval  $T_S$  yields the discrete pulse transfer function

$$P(z) = \frac{(1-l)z + (l-k)}{z-k} z^{-m}, \quad (10)$$

$$\text{where } k = e^{-\frac{T_S}{T_R}}, l = k^{m - \frac{T_D}{T_S}},$$

and  $m$  is the integer such that  $(m-1)T_S < T_D < mT_S$ .

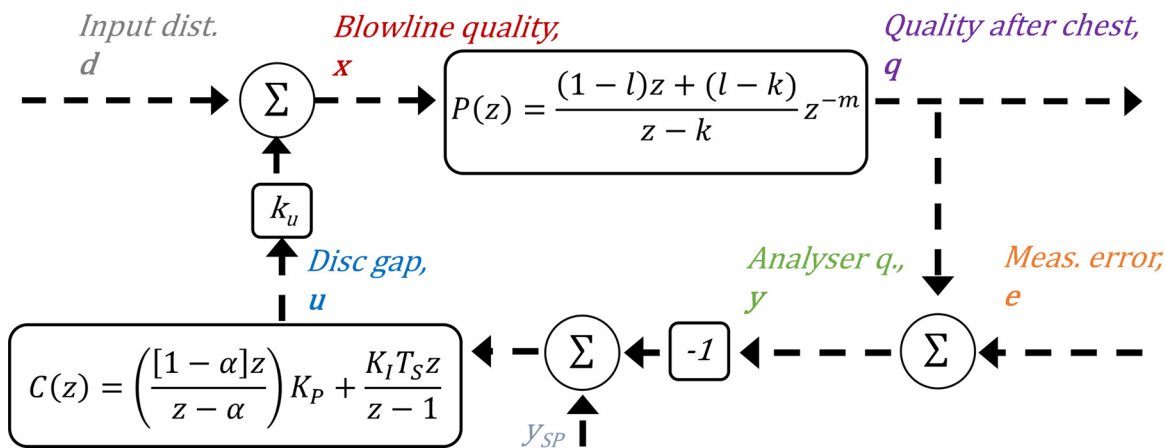
After the discrete pulse transfer function  $P(z)$  has been found, analysis of the controlled process is performed in the same way as for a continuous system. The transfer functions of interest are formed by combining  $P(z)$  with the controller transfer function  $C(z)$ , and the refiner gain,  $k_u$ , see Figure 12. The gain from disturbance to refiner output,  $k_d$ , has been dropped from the process description.

In the refining process in Figure 3, pulp quality is defined after the refiner (in the blowline) and after the mixed chest. We will consider quality in the blowline to be the output of the process. Then, the mixed chest and the delay can be regarded as a part of the sensor. With this perspective, the disturbance enters at the output of the process. The transfer from disturbance,  $d$ , to blow-line quality,  $x$ , is described by

$$x(z) = G_{xd}(z) d(z) = \frac{1}{1 + P(z)k_u C(z)} d(z), \quad (11)$$

Note that  $G_{xd}$  is also the transfer function from a disturbance entering at the chest output, to quality after the chest. This is recognized as the sensitivity function,  $S$ . The peak of  $S$ ,  $M_S$ , describes the peak amplification of input disturbances to blow-line quality, and is given by

$$M_S = \max_{\omega} S(e^{i\omega T_S}) = \|S\|_{\infty} = \left\| \frac{1}{1 + PC} \right\|_{\infty}, \quad (12)$$



**Figure 12:** Discrete representation of the controlled refining process. Blowline quality and quality after chest are here denoted  $x$  and  $q$  respectively, for clarity when denoting transfer functions.

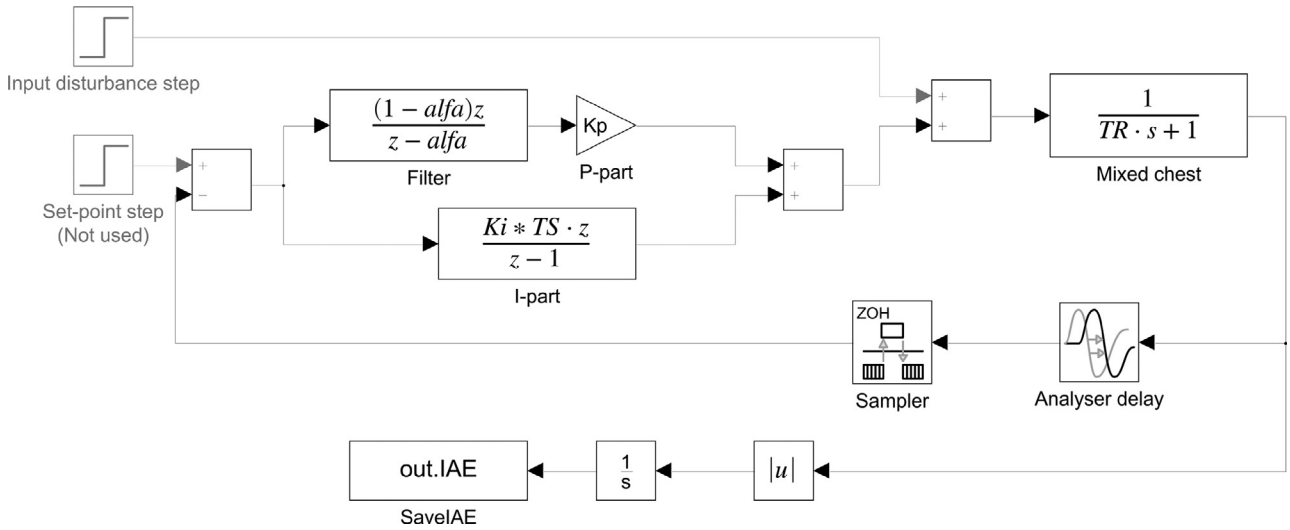


Figure 13: Simulink model used for determining step response error.

where  $\omega$  is the angular frequency, i. e. by the maximum value of the discrete frequency response function.

From the discrete frequency response of  $S$ , the bandwidth is here defined as the highest frequency which is damped to an amplitude of 0.70 (the 3 dB bandwidth). The bandwidth can be expressed in terms of cycle time instead of frequency.

The transfer from measurement error,  $e$ , to blowline pulp quality,  $x$ , is given by

$$x(z) = G_{xe}(z) e(z) = \frac{k_u C(z)}{1 + P(z) k_u C(z)} e(z). \quad (13)$$

When  $e$  is a white noise signal with variance  $\sigma_e^2$ , the standard deviation in blowline quality due to  $e$  is given by

$$\sigma_{xe} = \left\| \frac{k_u C}{1 + P k_u C} \right\|_2 \sigma_e. \quad (14)$$

The ratio

$$\frac{\sigma_{xe}}{\sigma_e} = \|G_{xe}\|_2 \quad (15)$$

is a measure of the sensitivity to random measurement error. This index is here referred to as the noise transfer ratio.

Since measurement error enters as a discrete signal,  $G_{xe}$  is an accurate representation of the response to measurement error (at the sampling instants). In contrast, the discrete frequency response of e. g.  $S$  is only an approximation of the response to a continuous periodic sine wave. The accurate response can be given as a sum of the response to the input (fundamental) frequency and its aliases. This may cause the calculated  $M_S$  to be inaccurate

(too low), especially when the chest (pre-filter) time constant is small. Other approaches to robustness are available (e. g. Toivonen and Tamminen 1990, Lennartson and Middleton 2014). However, the approximation used may be sufficient for most of the studied range of configurations and levels of constraints. It was found that the constraint on noise transfer ratio was more important, and the  $IAE$  was calculated for the continuous-time step response.

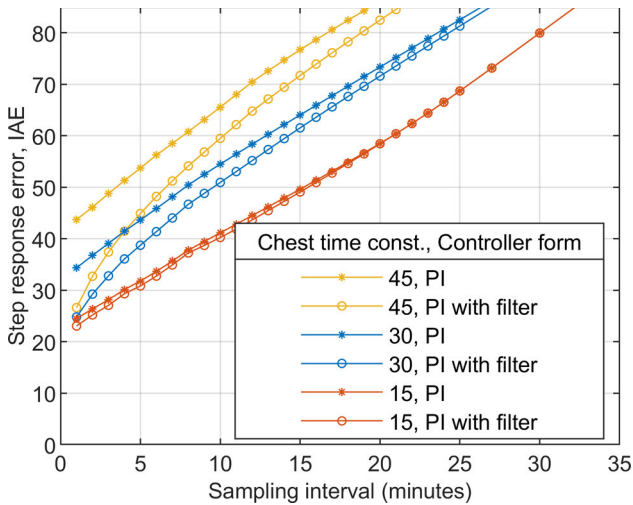
Figure 13 shows the implementation of the model in Simulink, used to evaluate response to input step disturbance. The transfer of measurement error was calculated using the Matlab Control Toolbox function *norm*, while  $M_S$  was calculated using the function *getPeakGain*.

The simulations were performed using a numerical solver (*ODE45*), with maximum step size set to 2 minutes and relative tolerance set to 0.001. The timing of the sampling instants is such that the first sample is at time 0, where also the step disturbance occurs.

## Additional results and comments

Although the PI controller was complemented with a filter, it was decided to leave this out from the results section, since the improvement was small. Optimization with the filter parameter included was also less robust and required some restarts with changed initial guess. A few results are shown in Figure 14.

The addition of a filter on the P-part of the PI controller enables reduced  $IAE$  when the chest time constant is large, but not otherwise. To minimize the  $IAE$  requires high values of  $K_I$ . A step disturbance does not motivate the use of



**Figure 14:** Control performance (IAE) vs sampling interval for different chest time constants, with or without filter in the PI controller. Analyser delay = 8 min. PI controller.  $M_S \leq 1.4$ . Noise transfer ratio  $\leq 50\%$ .

a filter, but a filter is motivated by the mixed chest when there is a large time constant. However, such large chests in relation to the sampling interval should not be used, since performance can be improved by choosing a smaller chest. Consequently, the PI controller without filter was used, which also leads to simpler optimization.

The response to an input step was chosen for the main measure of performance because it is simple. However, it comes with some difficulties. The timing of the input step in relation to the equally distanced sampling instants influences the results. This is especially evident where  $T_D$  is around the same size as  $T_S$ . In the results presented, the step occurs at time 0, together with a sampling instant. If the timing of the input step is set so that it occurs between two sampling instants, the resulting minimized IAE will be different. This is apparent for the result curves which show peaks. The positions of the peaks change depending on the input step timing. The difference is however small for the result curves without peaks, and these can be considered more relevant. The peaks appear also when the constraint on  $M_S$  is excluded, but to a lesser extent.

## References

- Alsip, W. P. (1981) Digital control of a TMP operation. *Pulp Pap. Can.* 82(3):T76–T79.
- Åström, K. J., Wittenmark, B. *Computer controlled systems*. 3rd. ed. Prentice Hall, Upper Saddle River, N.J., USA, 1997.

- Åström, K. J., Panagopoulos, H., Hägglund, T. (1998) Design of PI controllers based on non-convex optimization. *Automatica* 34(5):585–601.
- Atack, D. (1980) Towards a theory of refiner mechanical pulping. *Appita* 34(3):223–227.
- Beath, L. R., Neill, M. T., Masse, F. A. (1966) Latency in mechanical wood pulps. *Pulp Pap. Mag. Can.* October:T423–T430.
- Blanchard, P., Fontebasso, J. (1993) Mill implementation of a Freeness – Specific Energy control strategy for high consistency refiners. In: *Int. Mech. Pulping Conf., Oslo*. pp. 215–244.
- Brewster, D. B., Rogers, J. H. (1985) Analysis of on-line pulp drainage testers. *Pulp Pap. Can.* 86(7):T186–T190.
- Dawson, J., MacLellan, G., Shallhorn, P., Karnis, A. (1978) A rapid method for developing latent properties of mechanical pulps. *Pulp Pap. Can.* 79(C):T124–T129.
- Ein-Mozaffari, F., Bennington, C. P. J., Dumont, G. A. (2004) Dynamic mixing in agitated industrial pulp chests. *Pulp Pap. Can.* 105(5):T118–T122.
- Eriksson, K., Karlström, A. (2009) Refining zone temperature control: A good choice for pulp quality control. In: *Int. Mech. Pulping Conf., Sundsvall, Sweden*.
- Evans, J. C. W. (1978) TMP survey. *Pulp Pap.* 52(8):99–110.
- Ferritsius, O., Ferritsius, R., Hill, J., Karlström, A., Ferritsius, J., Eriksson, K. (2017) Process considerations and its demands on TMP property measurements – A study on tensile index. *Nord. Pulp Pap. Res. J.* 32(1):45–53.
- Gao, J., Bennington, C. P., Martinez, D. M., Olson, J. A. (2013) Latency removal of mechanical pulps: Phenomenological observations. *Nord. Pulp Pap. Res. J.* 28(2):198–202.
- Gao, J. *Modelling latency removal in mechanical pulping processes*. PhD Thesis, Dept. of Chemical and Biological Engineering, Univ. of British Columbia, Vancouver, Can., 2014.
- Garpinger, O., Hägglund, T. (2015) Software-based optimal PID design with robustness and noise sensitivity constraints. *J. Process Control* 33:90–101.
- Harinath, E. H., Biegler, L. T., Dumont, G. A. (2013) Predictive optimal control for thermo-mechanical pulping processes with multi-stage low consistency refining. *J. Process Control* 23(7):1001–1011.
- Hill, J., Westin, H., Bergström, R. (1979) Monitoring pulp quality for process control purposes. In: *Int. Mech. Pulping Conf., Toronto*. pp. 111–125.
- Hill, J. (1993) Process understanding profits from sensor and control developments. In: *Int. Mech. Pulping Conf., Oslo, Norway*. pp. 201–207.
- Honkasalo, J. V., Pölkkyinen, E. E., Vainio, J. A. (1989) Development of control systems in mechanical pulping (GW, TMP) at Rauma. In: *Int. Mech. Pulping Conf. Helsinki, Finland*. pp. 376–389.
- Horch, A., Isaksson, A. J. (2001) Assessment of the sampling rate in control systems. *Control Eng. Pract.* 9(5):533–544.
- Johansson, O., Rydstrand, M., Andersson, N., Brundin, H., McGuinness, M., Yadav, R. (2018) Single point morphology instrument yields insight to short term variations in a commercial TMP process. In: *Int. Mech. Pulping Conf., Trondheim, Norway*.
- Karlström, A., Eriksson, K., Hill, J. (2015) Refiner optimization and control part IV: Long term follow up of control performance in TMP processes. *Nord. Pulp Pap. Res. J.* 30(3):426–435.
- Karlström, A., Hill, J. (2018) Refiner optimization and control Part II:

- Test procedures for describing dynamics in TMP refining processes. *Nord. Pulp Pap. Res. J.* 29(4):653–662.
- Karlström, A., Hill, J., Johansson, L. (Submitted 2020) Soft sensors in pulp refining process control. Submitted to *Nord. Pulp Pap. Res. J.*
- Karlström, A., Isaksson, A. J. (2009) Multi-rate optimal control of TMP refining processes. In: *Int. Mech. Pulping Conf.*, Sundsvall, Sweden.
- Karnis, A., Atack, D., Stationwala, M. I. (1986) What happens in refining – Part II. *Pulp Pap. Can.* 87(11):T416–T423.
- Karnis, A. (1993) Latency in mechanical pulp fibres. *Paperi ja puu* 75(7):505–511.
- Kortelainen, J., Nystedt, H., Parta, J. (1997) Optimizing the refining conditions with on-line controls. In: *Int. Mech. Pulping Conf.* Oslo, Norway. pp. 103–110.
- Kristiansson, B., Lennartson, B. (2006) Evaluation and simple tuning of PID controllers with high-frequency robustness. *J. Process Control* 16(2006):91–102.
- Lennartson, B. (1990) On the choice of controller and sampling period for linear stochastic control. *Automatica* 26(3):573–578.
- Lennartson, B., Middleton, R. (2014) Optimal anti-aliasing filter based on multi criteria sampled-data  $H_\infty$  control. In: 53<sup>rd</sup> IEEE Conference on Decision and Control, Los Angeles, CA. pp. 4348–4354.
- Mohlin, U. B. (1980) Latency in thermomechanical pulps. *TAPPI* 63(3):83–86.
- Mokvist, A., Russell, G., Lauritzen, J. (2005) The new simplified concept of cost effective, high quality pulp manufacturing, for newsprint at Stora Enso Port Hawkesbury. In: *Int. Mech. Pulping Conf.*, Oslo, Norway. pp. 209–218.
- Roche, A., Owen, J., Miles, K., Harrison, R. (1996) A practical approach to the control of TMP refiners. In: *Control Systems* 96. pp. 129–135.
- Roche, A. A., Jack, J. S., Savoie, M. (1997) Successful applications of process control and on-line pulp quality sensors – a survey. *Pulp Pap. Can.* 98(12):T440–T444.
- Sandberg, C., Hill, J., Jackson, M. (2020) On the development of the refiner mechanical pulping process – A review. *Nord. Pulp Pap. Res. J.* 35(1):1–17.
- Sharifzadeh, M. (2013) Integration of process design and control: a review. *Chem. Eng. Res. Des.* 91(12):2515–2549.
- Sikter, D., Karlström, A., Engstrand, P., Czmaidalka, J. (2007) Using the refining zone temperature profile for quality control. In: *Int. Mech. Pulping Conf.*, Minneapolis, MN, USA. pp. 894–902.
- Soltész, K., Grimholt, C., Skogestad, S. (2017) Simultaneous design of PID controller and measurement filter by optimisation. *IET Control Theory Appl.* 11(3):341–348.
- Strand, B. C., Mokvist, A., Jackson, M. (1989) Improving the reliability of pulp quality data through factor analysis and data reconciliation. In: *Int. Mech. Pulping Conf.*, Helsinki, Finland.
- Strand, B. C. (1996) Model-based control of high-consistency refining. *Tappi J.* 79(10):140–146.
- Strand, B. C., Reichert, T., Filler, A., Brooks, Z., Nunn, B., Farmin, D., Poquette, P., Machtolf, P., Bardwell, B., Jarrell, M., Pattysen, G. (1999) Control and optimization of mechanical pulping systems. In: *Int. Mech. Pulping Conf.*, Houston, Texas.
- Tamminen, J., Perkola, M., Poutanen, H., Viljakainen, E. (1987) Mill scale trials with three on-line freeness testers in a TMP plant. In: *Int. Mech. Pulping Conf.*, Vancouver, Can. pp. 53–60.
- Tessier, P., Broderick, G., Desrochers, C., Bussiere, S. (1997) Industrial implementation of motor load and freeness control of chemimechanical pulp refiners. *Tappi J.* 80(12):135–142.
- Tienvieri, T., Huusari, E., Sundholm, J., Vuorio, P., Kortelainen, J., Nystedt, H., Artamo, A. (1999) Thermomechanical pulping. In: *Mechanical pulping*, Ed. Sundholm, J., Fapet Oy, Helsinki, Finland.
- Toivonen, H. T., Tamminen, J. (1990) Minimax robust LQ control of a thermomechanical pulping plant. *Automatica* 26(2):347–351.