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ADAPTIVE MODEL PREDICTIVE CONTROL OF CONSISTENCY

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Abstract: Adaptive model predictive control is applied to the control of consistency. The process model used for the control is updated on the basis of the operation conditions. The process gain is calculated by the mass balance equations or by using the preceding consistency measurement. The process dynamics is determined on the basis of the scheduling variable, i.e. the flow rate. The performance of the proposed consistency control strategy is studied by simulation and the simulation results are compared with the results of the present control strategy.

Keywords: Process control, predictive control, adaptive control, performance analysis, simulation.

1. INTRODUCTION

Consistency is one of the key factors for achieving uniform paper quality and, simultaneously, it has significant effect on the other controlled process variables in papermaking. Therefore, the requirements for the performance of the consistency control are stringent and will become stringer in the future as a result of increased paper quality requirements, process simplifications and decreased process volumes, see (Pekkarinen and Kaunonen, 1999).

The consistency is defined by dividing the dry weight of the sample, that is, fibers and additives by the total weight of the sample and it is controlled by bringing dilution water into the stock to be diluted on the basis of the measured consistency (Ostroot, 1993). Until recently, the consistency has been controlled with a fixed parameter PI-controller despite process non-linearity and a dominant process dead time. As a result, the control performance is far from optimal since the controller tuning must be performed at the operating point in which the process gain is the highest and the time delay longest in order to maintain stability in all operation conditions.

There are several approaches for compensating the changes in process dynamics. Ostroot gives (1993) an extensive introduction on how to compensate the changes in process dynamics. According to Ostroot, one of the alternatives is to stabilize the control loop with process modifications, e.g. pass-by lines, but these methods are rarely used due to additional costs caused by process modifications and energy wasted in stock re-circulating. In contrast, Dumdie (1997) took the changes in the process into account by introducing an adaptive PI-controller where the tuning parameters were changed on the basis of the scheduling variable, i.e. the flow rate. Because the proposed control strategy lacked feedforward compensation from the production changes, a ratio consistency control strategy was also introduced in the same report. However, this consistency control strategy requires additional instrumentation, a dilution flow sensor, and therefore it is still seldom applied to existing plants even though the price of flow sensors is comparatively low. Therefore, one of the objectives of this study is to indicate that a small investment can improve the control performance substantially and is worth investing in. Furthermore, there was one thing in common in both studies. They both used a PI-controller for regulating the consistency even though the PI-control algorithm does not necessarily give satisfactory results for the dead time dominant processes (Åström and Hägglund, 1995). Therefore, one of the objectives of this study is to investigate if the control performance can be improved in terms of disturbance rejection if the adaptive model predictive control is used for consistency control instead of a PI-controller.

Model predictive control (MPC) applied to consistency control is not a new idea. Allison and Ball (1998) demonstrated that control performance of blow tank consistency could be improved with model predictive control. Also Jansson (1999) indicated that the dead time of the control loop can be compensated with a Smith predictor. However, this approach only improves the control performance in the setpoint changes, which rarely occur when consistency is considered (Bialkowski, 1996). In addition, neither of these studies took into account the non-linear dilution process in the control design. Briefly, neither an adaptation of the process models nor adaptations of the controller parameters were performed despite changes in process dynamics. The objective of this study is to overcome illustrated drawbacks by introducing an adaptive model predictive control strategy for controlling consistency. The control algorithm is based on generalized predictive control (Clarke *et al.*, 1987a; 1987b; 1989). The process model used in the control is updated depending on the operation conditions. The process gain is calculated by mass balance equations or by using the preceding consistency measurement located in the process upstream. The time constant of the process and the process dead time are based on the physical model of the dilution process and these model parameters are updated on the basis of the scheduled variable, i.e. the flow rate.

This paper is organized as follows. In Section 2 a dilution process is introduced and an analysis for the process is performed. In Section 3 the proposed consistency control strategy is presented and Section 4 gives the simulations results for the proposed control strategy. In addition, the simulation results are compared with the results of the present control scheme. Conclusions are drawn in Section 5.

2. PROCESS DESCRIPTION AND ANALYSIS

Fig. 1 illustrates a typical dilution process where the consistency of the stock C_s is lowered to C_o by bringing dilution water into the pump suction. The consistency is measured after the pump and based on the measured consistency, dilution flow F_d is controlled by the dilution valve (V1). The pump (P1) is used for mixing dilution water and the stock to a homogenous mixture and for feeding the diluted stock to a downstream of the process. The flow rate F_o of the process is controlled and typically it is based on the production or the level of the chest located in the process downstream.

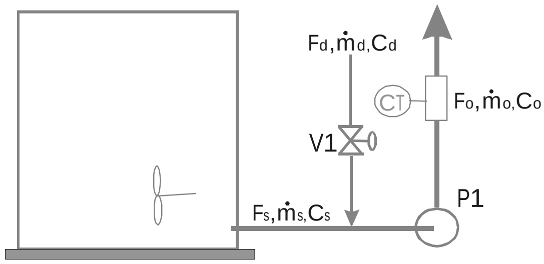

Fig. 1. Typical dilution process.

The following equations (1-3) can be derived for the dilution process shown in Fig. 1 on the basis of the mass balance equations:

$$F_o = F_d + F_s \tag{1}$$

$$\dot{m}_o = \dot{m}_d + \dot{m}_s \tag{2}$$

$$\dot{m} = \frac{C}{100} * F \tag{3}$$

where F is the volume flow [l/s] and $\dot{m}$ is the dry substance flow, i.e. fibers and additives [kg/s], and C is the consistency [%].

The process gain for the dilution process can be determined by rearranging the equations 1-3 as follows:

$$C_o = (m_d + m_s) / F_o \tag{4}$$

$$C_o = C_d F_d / F_o + C_s (F_o - F_d) / F \tag{5}$$

$$C_o = C_s + (C_d - C_s) F_d / F_o \tag{6}$$

When the equation 6 is differentiated with respect to the dilution flow F_d, the process gain for the dilution process can be finally given by:

$$\frac{dC_o}{dF_d} = \frac{C_d - C_s}{F_o} \tag{7}$$

The derived process gain equation (7) shows that the dilution process is strongly non-linear and that the process gain is inversely proportional to the flow rate F_0 and directly proportional to the consistency of the stock to be diluted C_s. It follows that, if the flow rate is halved, the process gain is doubled. For this reason, the tuning of the controller is carried out at low flow rates in order to maintain stability in all operation conditions. However, the control performance becomes very sluggish at high flow rates due to the change in the process gain. In the derivation above it was assumed that the dilution valve behaves linearly. However, if the non-linear control valve is used, it can be used for compensating changes in the process gain. The process gain for this case can be derived as follows:

$$F_d = f(u) \tag{8}$$

$$\frac{dC_o}{du} = \frac{(C_d - C_s) f^{'}(u)}{F_o} \quad (9)$$

where *f(u)* describes the relation between the dilution flow via the dilution valve at different valve openings at constant pressure. When the equations 8 and 9 are examined, it can be seen that if the constant process gain for the dilution process is desired, the derivate of the valve gain should decrease proportional to the flow rate. However, it is better to compensate the changes in the process gain and to improve the control performance by changing the control strategy to ratio cascade control where the consistency controller regulates the flow ratio between the dilution flow and the total flow. Dilution flow is calculated by multiplying together the flow ratio and the total flow. See Fig. 2.

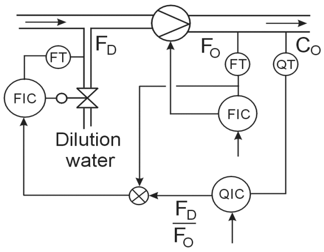

Fig. 2. Ratio cascade consistency configuration.

When the equations 10 and 11 are examined, it can be seen that there are two advantages in using the flow ratio as a control signal. Firstly, non-linearity of the control loop is removed, i.e. the process gain is not dependent on the flow rate anymore. Secondly, the configuration provides a feedforward compensation for production rate changes. As a result, the dilution flow is changed simultaneously when the flow rate changes. Finally, the dilution flow controller in the cascade control strategy eliminates effectively dilution header pressure variations and compensates non-linearity of the dilution valve.

$$F_d = uF_o \quad (10)$$

$$\frac{dC_o}{du} = C_d - C_s \quad (11)$$

where u is the control value of the consistency controller.

The process gain is not the only difficulty in the dilution process. Also the process dead time as well as the process dynamics are dependent on the flow rate. The time delay can vary on quite an immense scale due to large flow rate changes caused by sheet breaks, etc. The dead time of the process is mainly governed by the location of the consistency sensor and the flow velocity in the pipe. The measurement method determines the location of the consistency sensor. If the blade type of consistency sensors is used, the process dead time is fairly long since the sensor cannot measure the consistency from the turbulent stock flow and therefore a long settling length is needed to get a stable measurement. With sensors that use different measuring technologies, e.g. microwave, the sensor can be installed closer to the pump outlet and as a result the process dead time is considerably shorter and better control performance can be facilitated (Jansson, 1999). However, the varying time delay still remains.

The dynamics of the dilution process is mainly governed by the dynamics of the dilution flow controller, that is, by tuning of the control loop. The dilution flow dynamics should be as fast as possible in order to attain good disturbance rejection for the disturbances, i.e. pressure fluctuations and feed consistency deviations. In addition to dilution flow control loop dynamics, the dynamics of the consistency sensor affects the dynamics of the controlled process. The time constant of the filter is mainly determined by the process noise, which must be filtered before using the measurement for the control. The long filter time constants should be avoided if possible.

3. PROPOSED CONSISTENCY CONTROL STRATEGY

The proposed consistency control strategy is based on the control configuration presented in Fig. 2. The inner dilution flow controller of the cascade control uses PI-control algorithm for regulating the dilution flow. The outer consistency controller is based on the model predictive control, that is, a generalized predictive control algorithm introduced by (Clarke *et al.*, 1987a; 1987b; 1989). The output predictions of the MPC are based upon using an ARIMAX (auto regressive integrated moving average with exogenous inputs) model. See Eq. 12.

$$A(q^{-1})y(k) = q^{-d}B(q^{-1})u(k-1) + C(q^{-1})e(k)/\Delta \quad (12)$$

where A, B and C are polynomials in the backward shift operator q^{-1}, d is the process time delay in sampling intervals and $\Delta=(1-q^{-1})$. The C (q^{-1}) polynomial can be selected to fit the actual disturbance model or can be treated as a design weighting to provide greater robustness for unmodelled process dynamics (Clarke *et al.*, 1989).

$$A(q^{-1}) = 1 + a_1 q^{-1} + a_2 q^{-2} + \ldots + a_{nA} q^{-nA} \quad (13)$$

$$B(q^{-1}) = b_0 + b_1 q^{-1} + b_2 q^{-2} + \ldots + b_{nB} q^{-nB} \quad (14)$$

$$C(q^{-1}) = 1 + c_1 q^{-1} + c_2 q^{-2} + \ldots + c_{nC} q^{-nC} \quad (15)$$

The control value of the consistency controller, i.e. the flow ratio between the dilution flow and the total flow is calculated by minimizing the quadratic cost function Eq. 16. See Henttonen (1996) for details. The first control value of the calculated control sequence is sent into the process and the procedure is repeated in the next sampling time.

$$J = \{(\sum_{j=N_1}^{N_2} \left[\hat{y}(t+j|t) - r(t+j)\right]^2 + \sum_{j=1}^{N_u} \lambda \Delta u^2 (t+j-1|t)\} \quad (16)$$

where $\hat{y}(t+j|t)$ is the predicted output j steps into the future based upon information available at time t, $\Delta u(t)=(1-q^{-1})u(t)$, r(t+j) is the reference signal j steps into the future, N_1 is the minimum costing horizon, N_2 is the maximum costing horizon, N_u control horizon and λ control weighting factor.

Basic rules for selecting the tuning parameters for GPC have been given by Clarke *et al.* (1987a). The tuning parameters should be selected by setting N_1 equal to the process dead time and N_2 close to the rise time of the plant. The control horizon can be selected between 1 and N_2. A value of N_u of 1 gives generally slower control and increase in N_u gives a faster control performance. The role of λ is to penalize excessive incremental control action. The greater value of λ, more sluggish the control will become (Clarke *et al.*, 1987a).

The proposed consistency control strategy uses a physical model of the dilution process in the control. If there is a consistency measurement available from the preceding dilution stage, it can be used for calculating the process gain. However, the dynamics of the pipe and the chest between the dilution stages must be taken into account in this case. If there is no consistency measurement available or the preceding consistency measurement is not reliable enough, the process gain can be calculated by using the available process measurements. First, the consistency of the stock to be diluted is calculated by equation 17. Then, the calculated consistency is filtered by the first-order filter and, finally, the filtered consistency is used for calculating the process gain by Eq. 19.

$$C_{Cal}(k) = \frac{C_o(k)F_o(k) - C_d F_d(k)}{F_o(k) - F_d(k)} \quad (17)$$

$$\hat{C}_s(k) = \alpha \hat{C}_s(k-1) + (1-\alpha) C_{Cal}(k) \quad (18)$$

$$\frac{dC_o}{du} = C_d - \hat{C}_s(k) \quad (19)$$

where $C_{Cal}(k)$ is the calculated stock consistency at time k, $\hat{C}_s(k)$ is the filtered stock consistency and α is the tuning constant of the filter. The filter ignores the calculated consistency for α=0 and sets the consistency for unfiltered at α=1.

The time delay of the process model is calculated by Eq. 20. It is assumed in the calculation that a plug flow takes place in the pipe and the time delay varies proportional to the flow rate F_o. A tuning constant ε can be used for getting a better estimate for the dead time and it can be determined by identifying several dead times between the minimum and the maximum flow rate from the real process.

$$L = \varepsilon \frac{Volume}{F_o} \quad (20)$$

The dynamics of the dilution process is mainly governed by the dynamics of the dilution flow controller, that is, by the tuning of the control loop. In addition, the mixing process and the filter of the consistency sensor influence the overall process dynamics. If the process dynamics does not vary on an immense scale, the time constants of the process model can be kept fixed. Otherwise, the same type of scheduling can be performed for the process dynamics as illustrated for the time delay.

4. SIMULATION RESULTS

The proposed model predictive consistency control was investigated with the simulator developed in Matlab/Simulink™ environment. The process models used in the simulator are mainly based on the process models presented by Rao *et al.* (1993). Before presenting the simulation results, the needs for the ratio control strategy are demonstrated by process data and the simulations.

Figure 3 illustrates the behavior of the conventional consistency control where the consistency is controlled by the dilution valve. The Figure is from the broke system of the real paper machine. The objective is to show how severe the stability problem really is. The consistency control operates sufficiently until the broke flow to the blend chest is decreased from 1800 l/min to 1300 l/min. Then the stability of the consistency control loop is lost and the consistency begins to oscillate. The process gain of the consistency control loop increases nearly 30% when broke flow decreases 500 l/min.

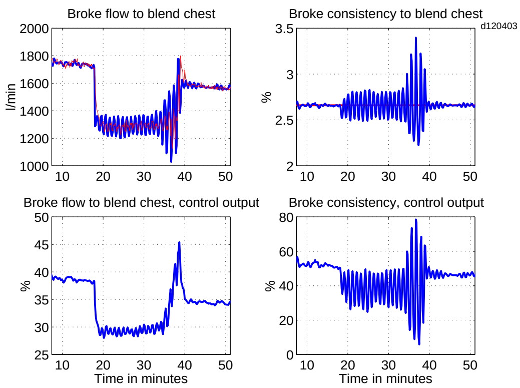

Fig. 3. An example of the consistency control performance from the real paper machine.

The same incident can be observed from the simulated response shown in upper Fig. 4 where the consistency controller is tuned at the high flow rate where the process gain is low and the process dead time short. The control performance in the tuning point is sufficient (0-300 seconds), but when the flow rate decreases, the process gain is increased and the consistency begins to oscillate as shown in Fig. 3. Lower Fig. 4 illustrates the simulation results for the consistency controller, which is tuned at low flow rate where the process gain is high and the time delay long. The control performance is very sluggish at the high flow rate and it takes a long time to reach the new setpoint (0-300 seconds). The control performance becomes acceptable when the flow rate is decreased to the level where the controller was initially tuned. In the same figures is also illustrated the control performance for the ratio consistency control. Figure demonstrates that the control performance is unaltered despite flow rate changes. In addition, the Figure shows that the disturbance caused by the change in flow rate in is considerably smaller compared with the traditional consistency control strategy due to the feedforward compensation from the flow rate.

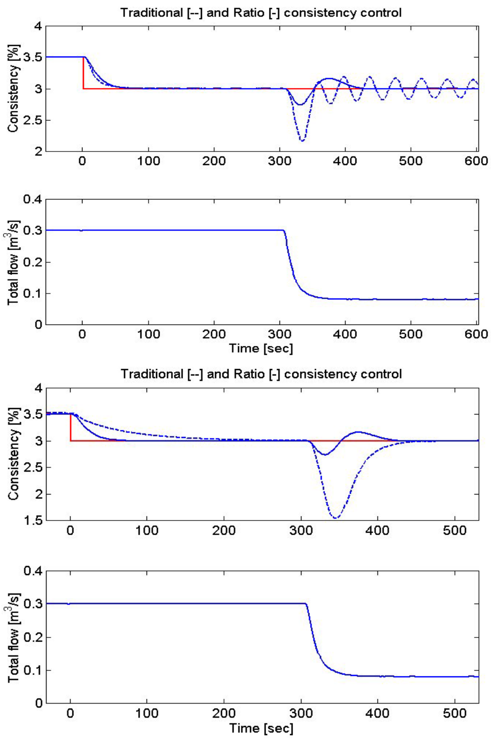

Fig. 4. Simulated responses of the conventional consistency control and the ratio control.

Fig. 5 illustrates the simulation results for the ratio control strategy in which the proposed model predictive control is used. Also, the Figure gives the simulation results for the PI-controller. The arrangement used for the model predictive control is the following. First, the process model used in the control is updated in two-second sampling intervals and the sampling interval can be considered as a tuning parameter. The process gain was calculated by equations 17-19 and the filter constant α was set at 0.5. The time delay of the process model was estimated by equation 20. Time constants of the process model were maintained fixed despite the changes in the flow rate. Based on the calculated process gain and the time delay, polynomials A and B used in the control were updated every two seconds while the polynomial C was kept fixed. The sampling time for the controller was set to 1 second and N_1 was changed in respect of the estimated time delay. The prediction horizon was kept at 30 samples and N_2 was changed on the basis of the prediction horizon and N_1. Control horizon N_u was set at 4.

In order to make a comparison possible between the model predictive control and the PI-control, the simulators for both control strategies were identical in all simulations, i.e. pipes, control valves, pumps, dilution flow controllers, sensors, etc. The only difference is between the consistency control algorithms. Fig. 5 presents the simulation results for both control configurations in 3 different simulation runs that consistency control is typically encountered in. The upper figure shows the control performance when the setpoint is changed from 4 to 3.5%. In the middle figure the flow rate is increased from 0.25 m^3/s to 0.3 m^3/s and the lower figure represents the case where the stock consistency in changed from 5% to 6.5% in a step while the flow rate is maintained at 0.25 m^3/s.

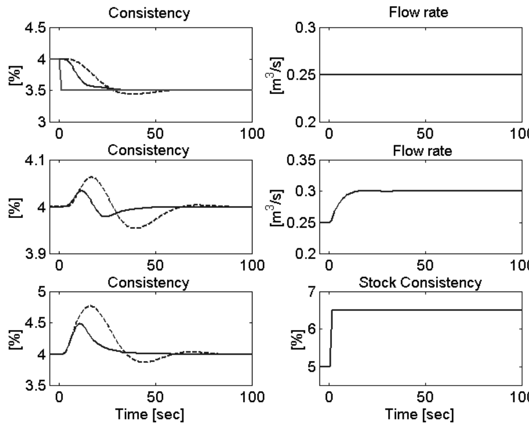

Fig. 5. Simulated responses of the ratio control strategy with the PI-controller [--] and with the adaptive model predictive controller [-].

The performance of the consistency control is considerably better with MPC compared with the PI-controller in all simulation cases. In the setpoint change there is no overshoot and the setpoint is also reached faster. However, it is worthy of note that setpoint changes hardly occur in the real processes and therefore one should not put too much emphasis on the setpoint behavior. In disturbance rejection, MPC outperforms the PI-controller. The disturbance caused by the flow rate change is eliminated considerably faster and the response is also better dampened. The disturbance caused by the change in feed consistency is also compensated faster with the MPC due to adaptive process models. When the stock consistency changes from 5% to 6.5%, the process gain decreases 30% and the decrease is not taken into account by any means in the PI-controller. In this simulation, the time of recovery from the disturbance was for the MPC less than half of the PI-controller. Fig. 6 presents the same simulation runs for the MPC as shown in Fig. 5, but large amplitude process noise is included in these simulations. The objective is to show that the process model updating and the model predictive control algorithm is robust even though there exist process noise and large variations in the stock consistency.

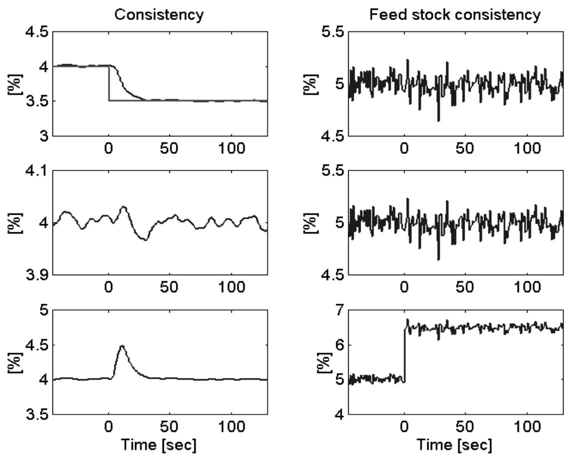

Fig. 6. Simulated responses of the model predictive control.

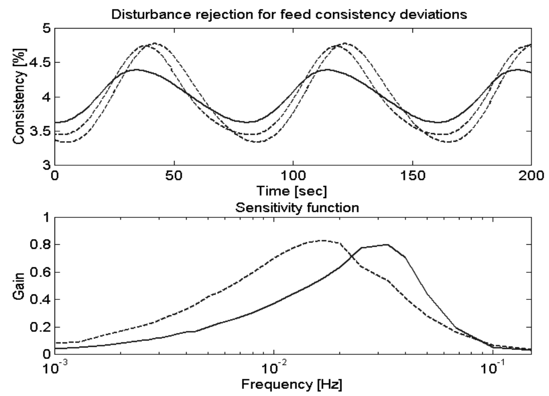

Fig. 7. Upper: Feed consistency changes periodically. Lower: Disturbance dampening of the process as function of frequency.

Fig. 7 shows the simulation results where the consistency of the stock to be diluted changes periodically between cycle times of 1 to 1500 seconds. The objective of this simulation is to determine disturbance attenuation properties of the consistency controls as a function of frequency. The upper figure shows the results for the one specific frequency. The Figure demonstrates that the MPC holds substantially better disturbance rejection properties compared with the PI-controller. In the lower figure the performances of the consistency controllers are shown as a function of the frequency at one operating point. The amplitude of the sensitivity function may vary between the operating conditions, but the results are very indicative. The maximum sensitivity for the MPC is close to 0.8 at 0.033 Hz and for the PI-controller at 0.02 Hz. The disturbance rejection in all frequencies with the MPC is substantially better than with the PI-controller.

5. CONCLUSIONS

In this paper an adaptive model predictive controller was applied to the control of consistency. The process gain was calculated by the mass balance equations and the process dynamics was updated on the basis of the flow rate. The proposed consistency control strategy was studied by simulation and the results were compared with the present control strategy in three different simulation runs. In the first simulation, the results of the setpoint change were shown and in the last two simulation runs disturbance attenuation properties of the consistency control were investigated. According to the simulations performed, the new control strategy possesses substantially better disturbance attenuation properties compared with conventional methods. This is made possible with an adaptive process model and a more sophisticated control algorithm. At the time of writing, the new control algorithm will be applied to the real plant for testing.

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