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Axelsson, Jan Peter; Hagander, Per; Mandenius, Carl Fredrik; Mattiasson, Bo

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Author(s)
J.P. Axelsson, P. Hagander
Department of Automatic Control
C.F. Mandenius, B. Mattiasson
Department of Pure & Applied Biochemistry

Supervisor

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The paper was presented at the 1st IFAC workshop on Modelling and Control of Biotechnical Processes, Helsinki, Finland, August 17-19, 1982.

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COMPUTER CONTROL OF SUCROSE CONCENTRATION IN A FERMENTOR WTIH CONTINUOUS FLOW

J.P. Axelsson, P. Hagander

Department of Automatic Control, Lund Institute of Technology, Lund, Sweden.

C.F. Mandenius, B. Mattiasson

Department of Pure and Applied Biochemistry, Lund Institute of Technology, Lund, Sweden.

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1. INTRODUCTION

The control of fermentation processes is becoming more and more important, as continuous processes are developed. A high production rate and good utilization of the raw material requires, that substrate and product concentrations are maintained at specific levels despite disturbances in for instance substrate composition or cell condition.

This paper describes how a sucrose sensor is used to regulate the substrate concentration in the fermentation of sucrose to ethanol. A mathematical process model is determined and used in a Kalman filter to compensate for a measurement delaytime. The regulator is implemented on a digital computer. The modeling work facilitates the regulator design as compared with tuning of a PID-regulator at the process (Mandenius, Danielsson, Mattiasson, 1981).

2. THE PROCESS

In this section the process is described, modelled and analysed. Focus is on the limitations of control of the process.

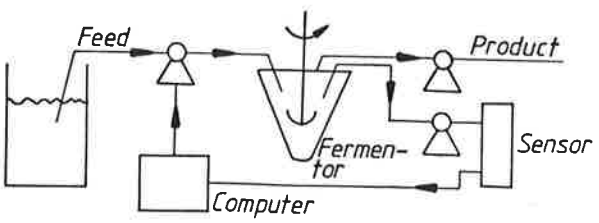

Fig. 2.1. The experimental set up.

A Description

The set up used, Fig. 2.1. consists of a continuous stirred tank reactor with alginate beads containing yeast cells and pumps for inflow and outflow. The inflow pump is computer controlled using the signal from an enzyme thermistor sensor measuring the sucrose concentration.

Yeast. The yeast species used in the process is baker's yeast, Saccharomyces cerevisiae, obtained from Jästbolaget, Sweden.

Kinetics. The hydrolysis of sucrose (S) is believed to be an extracellular process. The transport time is short to the sites of glycolysis in the cells. The rate of hydrolysis is often faster than the glycolysis, so there is a considerable amount of glucose (G) and fructose (F) in the reactor.

The enzyme capacity in the reactor is in excess, so that first order kinetics can be applied.

$$S + H_2O \underset{k_-}{\overset{k_+}{\rightleftharpoons}} G + F$$

The dynamics is given by

$$\frac{d[S]}{dt} = k_-[G]\cdot[F] - k_+[S]\cdot[H_2O] \qquad (2.1)$$

Excess of sucrose and water simplifies the dynamics to

$$\frac{d[S]}{dt} = -k_S[S] \quad \text{where } k_S = k_+\cdot[H_2O] \qquad (2.2)$$

In stationarity the rate of ethanol formation is proportional to the sucrose concentration over a wide range. At high ethanol concentrations, inhibition of cell activity occurs. However, the extracellular hydrolysis of sucrose is less influenced. This results in an increase in concentration of the intermediates fructose and glucose. Under such conditions a measurement of sucrose concentration is of less value for process control.

Immobilization of the yeast. The yeast cells are immobilized by entrapment in Ca-alginate beads with a diameter of 2-3 mm. The productivity of the immobilized yeast cells is preserved but growth is inhibited. Diffusion through the bead is fast (Cheetham, Blunt, Bucke, 1979).

The reactor. The reactor, Fig.2.2., with one inlet and two outlets has a volume of 5 ℓ. The inlet is 10 cm below the liquid surface, and the main outlet is at the surface. A minor outlet supplying the sensor is in the middle of the tank at 7 cm depth. In the reactor there is a stirrer rotating at about 180 rev/min. Mixing could be considered as immediate.

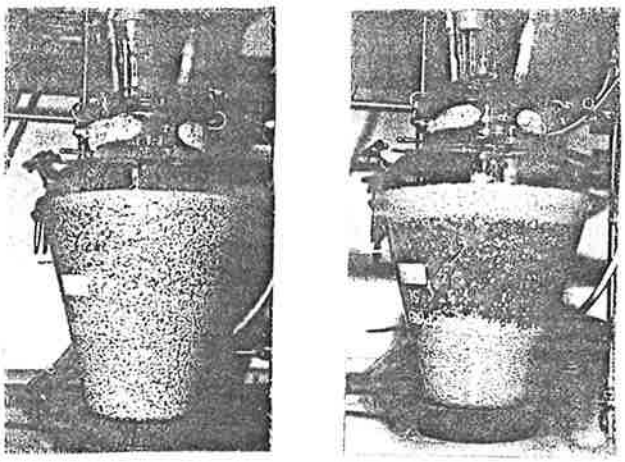

Fig. 2.2. The reactor filled with alginate beads with and without stirring.

The enzyme thermistor. The sucrose concentration is continuously measured with an enzyme thermistor. This is a flow microcalorimeter containing an immobilized enzyme, in this case invertase, β-D-fructofuranosidase [EC.3.2.1.26], which catalyses the exothermic hydrolysis of sucrose. The produced heat is measured with thermistors placed in close proximity to the enzyme. The thermistors are connected to a Wheatstone bridge. The calorimeter is carefully thermostated, and temperature changes less than 0.0001°C are possible to detect. The temperature difference normally measured in these experiments is of the magnitude of 0.01°C. The technique is developed by Danielsson, Mattiasson, Mosbach (1981).

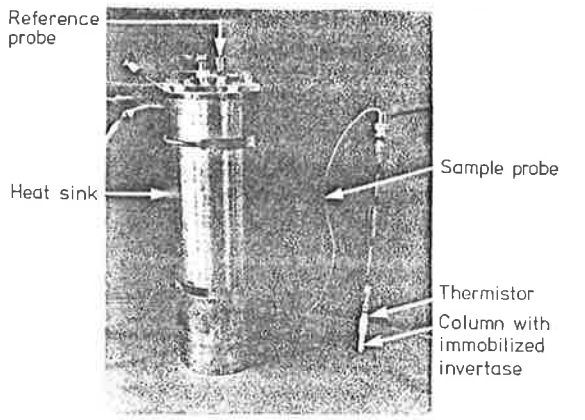

Fig. 2.3. The enzyme thermistor without its insulation.

The sensor is linear in the working range, which is expanded through dilution 1:10 to the range 0.01-0.035 mol/ℓ. The dynamics is characterized by a short rise time, about 1 min and a longer fall time, about 4 min. There is also a timedelay of 6.5 min due to the transport in the dilution system and in a heat exchanger in the calorimeter. There is a small drift, less than 0.00025 mol/ℓ·h. Normally the signal to noise ratio is about 30. A degasser is used to eliminate disturbances from bubbles in the flow. Another source of disturbances is the dilution system.

The sensor is well described by

$$\dot{y}(t) = \frac{1}{\tau}(S(t-\tau_d)-y(t)) \qquad \tau = \begin{cases} \tau_{up} \\ \tau_{down} \end{cases} \qquad (2.4)$$

Pumps. Three pumps are used. They are all peristaltic. A multichannel precision pump is used for the sensor and the dilution arrangement. The outlet pump is used in a special arrangement to give a constant tank level. The inflow is controlled with a voltage controlled pump. The capacity is in the range 0-150 mℓ/min. There is a severe nonlinearity in the pump characteristics around 10-40 mℓ/min, Fig 2.4. After calibration the pump is stable for weeks. Pump and tubing dynamics can be neglected.

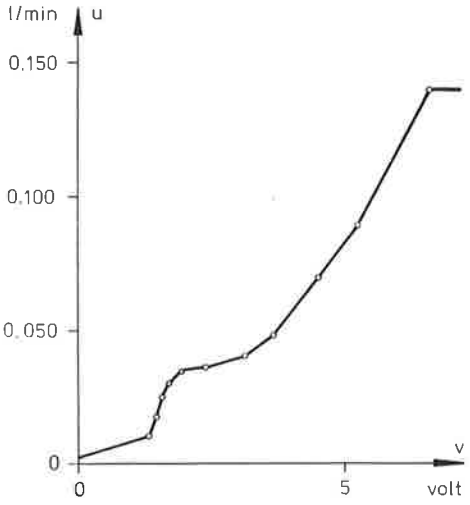

Fig. 2.4. Characteristics of the pump.

A Model of the System

Assumptions. Time constants less than 4 min are neglected. This implies the following simplifications:

- o Breakdown of sucrose is of first order.
- o Instantanous mixing.
- o Homogeneous reactor.
- o Perfect level control.
- o The flow could be changed immediately.
- o The sensor is characterized by its deadtime only.

The sensor falltime is the largest time-constant neglected.

Model. The assumptions made lead to the following model:

$$\frac{dS}{dt} = -k_S \cdot S - \frac{1}{V} \cdot S \cdot u + \frac{1}{V} \cdot S_{in} \cdot u \qquad (2.5)$$

$$y(t) = S(t-\tau_d) \qquad (2.6)$$

where

S = Reactor sucrose concentration (mol/ℓ).

u = Flow through the reactor (ℓ/min).

k_S = Reaction rate coefficient (0.020 1/min).

V = Working volume of the reactor (5.0 ℓ).

S_{in} = Feed sucrose concentration (0.50 mol/ℓ).

y(t) = Measurement signal.

τ_d = Delay time for the sensor (6.5 min).

Typical process parameters are given above in parentheses and will be referred to as (*). Working point is denoted (S°,u°) and typical values are:

S° = 0.15 mol/ℓ

u° = 0.043 ℓ/min

Verification. The process model is verified in different ways and seems to describe the reactor well. Parameters are estimated with simple experiments. The main problem is the reactor volume (V), chosen to be the total reactor volume, i.e. the volume of liquid and beads. The accuracy of the dynamics of the model (2.5)-(2.6) is tested by having the system oscillate using high gain proportional control, Fig. 2.5. Another check of the model is given when using a regulator based on Smith prediction. Here the prediction error gives a good idea of the validity of the model, Fig. 2.6.

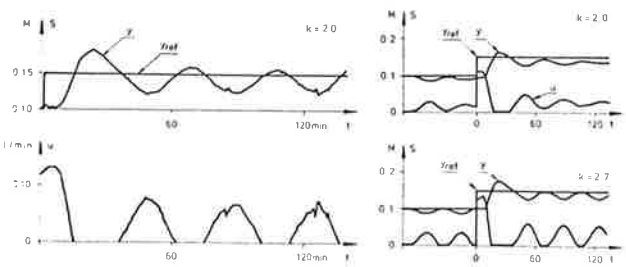

Fig. 2.5. Oscillations with high gain proportional feedback, measurements and simulations. Parameters (*) but k_s=0.014 1/min and τ_d=7.2 min.

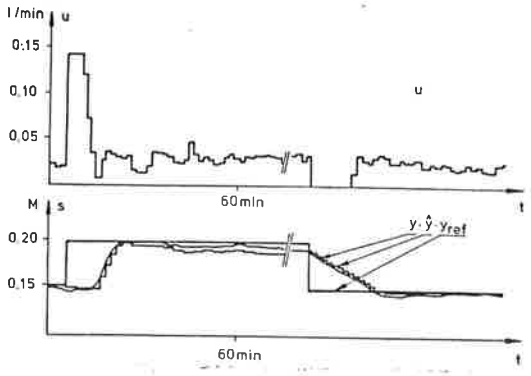

Fig. 2.6. Measured signal and model prediction using an Otto-Smith regulator.

Analysis of the Model

Different aspects of the process model (2.5)-(2.6) are now discussed.

Stationarity. The stationary point is

$$S^\circ = \frac{1}{1 + \frac{k_S V}{u^\circ}} \cdot S_{in} \qquad (2.7)$$

Linearisation around (S°,u°).

$$\dot{x} = ax + bu \qquad (2.8)$$

where

$x := S - S^\circ, \quad u := u - u^\circ$

$$a = -k_S - \frac{u^\circ}{V}, \quad b = \frac{S_{in} - S^\circ}{V}$$

Typical timeconstants with parameters (*).

space time in the reactor $\tau_r = \frac{V}{u^\circ} = 120$ min

time constant with yeast in the reactor $\tau = \frac{1}{k_S + \frac{u^\circ}{V}} = 35$ min

control authority

u=0 ℓ/min => $\dot{x}_{min} = -0.003$ mol/ℓ·min

u=0.15 ℓ/min => $\dot{x}_{max} = 0.007$ mol/ℓ·min

Bilinearisation around (S°,u°).

$$\dot{x} = ax + bu + cxu \qquad (2.9)$$

where

x,u,a,b are the same as in the linear case

$c = -(1/V)$

Aspects of the bilinearity. The bilinear structure of equation (2.5) is preserved under change of the stationary point. The cross term can be viewed as a pole moving with the input u:

u = 0 ℓ/min => a = -0.020 1/min

u = 0.15 ℓ/min => a = -0.050 1/min

It could also be considered as a gain, that varies with the sucrose concentration. It is easy to compensate for this by gain scheduling, cf section 4.

The effect of the timedelay. The limitations that the timedelay impose, can be studied in the frequency domain. Laplace transformation of the linearised system (2.9) gives

$$Y(s) = G_o(s) \cdot U(s)$$

where

$$G_o(s) = \frac{b}{s-a} e^{-s\tau_d} \qquad (2.10)$$

The timedelay results in a loss of phase margin, which will be a major difficulty in the control of the process.

The closed loop with feedback gain K is stable for

$$Re(G_o(i\omega_o)) = -\frac{b}{a}\cos(\omega_o \tau_d) > -\frac{1}{K}$$

and the critical gain is

$$K_{crit} = \frac{a}{b}\frac{1}{\cos(\omega_o \tau_d)} \qquad (2.11)$$

where ω_o is the smallest solution to

$$\tan(\omega_o \tau_d) = \omega_o \tau_d \cdot \frac{1}{a\tau_d} \qquad (2.12)$$

The parameters (*) give $K_{crit} = 3.7$.

For practical stability K is reduced considerably. For K=1.5, there is a phase margin of 68°. This corresponds to an overshoot of about 10%.

The effect of the severe nonlinearity of the pump. The characteristics of the pump, Fig. 2.4., gives for the operating point at u=0.035 ℓ/min, that small changes in u would result in an 18 fold change in gain. This gives drastic stability limitations on the feedback gain. Bounds on K are obtained from Off-Axis circle criterion e.g. Narendra and Taylor (1973). Practical stability requires further restrictions on K as determined by simulation.

$u^* = 0.043 \Rightarrow K_{crit} \approx 2.2 \quad K_{pract} \approx 0.7$

$u^* = 0.035 \Rightarrow K_{crit} \approx 1.2 \quad K_{pract} \approx 0.4$

After heavy disturbances or when a change of operating point is wanted, the pump will saturate, which should be considered in the design of regulators.

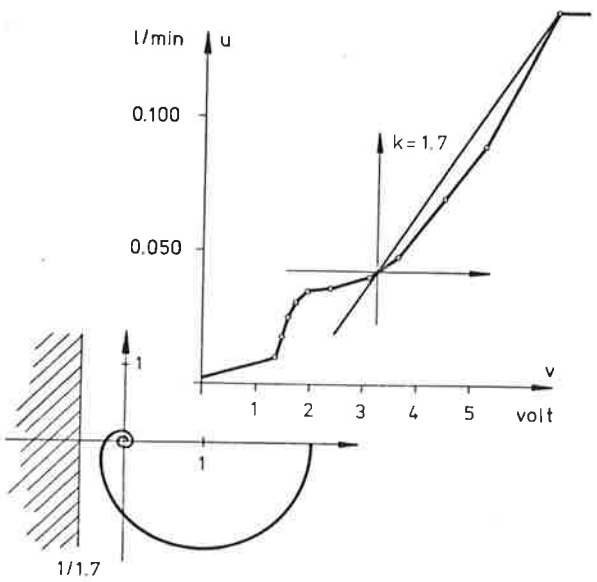

Fig. 2.7. Bounds on K obtained from the Off-Axis Circle criterion.

Sampling of the System

It is natural to study the sampled model of the process as a computer will be used to implement the regulator. Here the process model (2.5)-(2.6) is sampled with period h for both measurement and control signals. The control variable uses a zero order hold.

The sampled model is easier in two respects: the timedelay can be represented in state space form and the system becomes linear between sampling instants.

Let one state represent S in the reactor and let d states represent the delayed measurement signals.

$$x(t+h) = \Phi(u) \cdot x(t) + \Gamma(u) \cdot u(t) \qquad (2.13)$$
$$y(t) = \Theta(u) \cdot x(t)$$

where

$$\Phi = \begin{bmatrix} \varphi(u) & 0 & \cdots & 0 \\ 1 & 0 & & 0 \\ & 1 & \ddots & \\ & & \ddots & 0 \\ & & & 1 \quad 0 \end{bmatrix} \qquad \Gamma = \begin{bmatrix} \gamma(u) \\ 0 \\ \vdots \\ 0 \end{bmatrix}$$

$$\Theta = [0 \ldots \theta_{d-1}(u) \ \theta_d(u)]$$

$$\varphi(u) = e^{-(k_s + \frac{u}{V})\cdot h} \qquad \gamma(u) = \frac{S_{in}}{V} \int_0^h e^{-(k_s + \frac{u}{V})\cdot s} ds$$

$$\theta_{d-1}(u) = \frac{1}{\gamma(u)} \int_0^\tau e^{a(u)s} b(u) ds \qquad \theta_d(u) = \frac{1}{\gamma(u)} \int_\tau^h e^{a(u)s} b(u) ds$$

where $\tau_d = h \cdot (d-1) + \tau$

Choice of sample interval. It is convenient to have $\tau d = d \cdot h$. However, sometimes the sampling must be synchronized to other signals and the choice of h is restricted. Therefore the general case is treated.

Sampling introduces a zero on the negative real axes, inside or outside the unit circle. If h is chosen so that $\tau d = d \cdot h$, then the zero vanishes and there is a pure timedelay.

For the choice of h, rules of thumb give 4-17 samples per solution time of the closed loop system, i.e. 15-20 min. In our case the performance of the system is determined by how well the prediction of the measurement signal is done. A less accurate model needs a shorter sampling interval. A shorter sampling interval is also to prefer, if abrupt disturbances are likely.

Here a sample interval between 1 and 3 minutes is used.

Prefiltering. To avoid aliasing, an analogue lowpass filter is used at the input to the computer. The bandwidth must be chosen to maximum half of the sample frequency. Lowpass filters with bandwidth less than 0.1 Hz are difficult to implement in analogue technique, so the filter is divided in one analogue and one digital part. The analogue filter is of first order and has a bandwitdth of 0.1 Hz. The digital filter is a second order Butterworth filter with the sample interval 5 s. The bandwidth of the digital filter can easily be chosen according to the sample interval of the regulator.

3. AIM WITH CONTROL

In this section is discussed the possibility with control to make the system less sensitive to parameter variations at steady state. The tank reactor could also be thought of as part of a larger system. Then a rapid response to setpoint changes from outer loops has to be considered. Possibilities and limitations are discussed.

Sensitivity to Changes in Process Parameters

The different parameters have the following characteristics:

k_S - The rate constant varies slowly due to several factors. A decrease of the constant by a factor 3-4 within a few days is normal, when nutrients are lacking.

V - The working volume of the reactor is fixed.

u - The flow through the reactor changes slowly due to drift in the pump, especially near the edge of the characteristics, Fig. 2.4.

S_{in} - Abrupt changes are likely in the sucrose concentration of the incoming feed, when a new batch of substrate enters.

Sensitivity to parameter changes is calculated for the stationary point, with and without proportional regulation.

$$S_p = \frac{d \log G(0)}{d \log p}$$

where $G(s) = \frac{Kb}{s+a+Kb}$

K = the gain in the feedback.

TABLE 3.1. Sensitivity to parameter changes around parameters (*)

p		k_S	u	S_{in}
S_p	K=0	-0.70	0.70	1.40
	K=1	-0.20	0.20	0.41
	K=4	-0.065	0.065	0.13

Change of Setpoint

Often a rapid change of operating point is wanted. Ideally, for this type of system, maximal flowrate (or no flow) is used until the desired concentration is obtained and then the flowrate is set to a corresponding stationary value. Such a strategy is called bang-bang control.

From the sampled model (2.13), the shortest time for change of operating point can be calculated. Consider a change $\pm\Delta S$ from S°.

$$\tau_{up} = \frac{1}{a} \ln \left[\frac{S^\circ + \Delta S - (b/a)\cdot u}{S^\circ - (b/a)\cdot u} \right] \qquad (3.1)$$

$$\tau_{down} = \frac{1}{k_S} \ln \left[\frac{S^\circ - \Delta S}{S^\circ} \right] \qquad (3.2)$$

TABLE 3.2. Shortest time for change of operating point around parameters (*), ΔS=0.05 mol/l

S°	τ_{up}	τ_{down}
0.10 mol/l	5.7 min	35 min
0.15	8.1	20
0.20	14	14

These times should be compared with the timedelay of the measurement signal. A fairly good process model should be used, if fast performance is wanted. Certain changes of operating point take a long time due to limited control authority. In these cases the accuracy of the model is not so critical.

Fast changes of the environment is not always wanted, beacause the microorganisms may come in a chock state. Here no such phenomena were observed. Another aspect is how the surrounding system reacts to large changes in flowrate. This has to be considered in an overall design.

4. DESIGN OF DIFFERENT REGULATORS

In this section different regulators are designed, analysed and tested. Tests are made using simulation and experiments on the real process.

Conventional PID Regulators

Analogue PID regulator. The possibilities with a straight forward analogue PID regulator is clearly limited due to nonlinearities of the pump and timedelay in the measurement signal, as seen in section 2. This was clear in earlier work on the process, Mandenius and co-workers (1981). However, if only slow disturbances e.g in the rate constant are considered and if there is no need for rapid response to setpoint changes, then the gain can be kept low and a reasonable performance is obtained with an analogue PI-regulator.

Computerised PID regulator. With a computer it is easy to implement a calibration curve for the pump and thus eliminate the gain variation. It is also easy to implement integrator anti reset windup, which is important, since the pump will be used over its whole range and is likely to saturate. The feedback gain can be increased and is now limitted by the timedelay only, and the performance of the simple PI-regulator is improved.

Use of D-part. The use of the derivative part of a PID-regulator is often of limited value and it needs careful tuning. Here the derivative time should be in the order of minutes, which motivates computer implementation. The timedelay limits the value of the D-part, but a certain improvement is obtained.

Results, see Fig. 4.1. and Fig. 4.2.

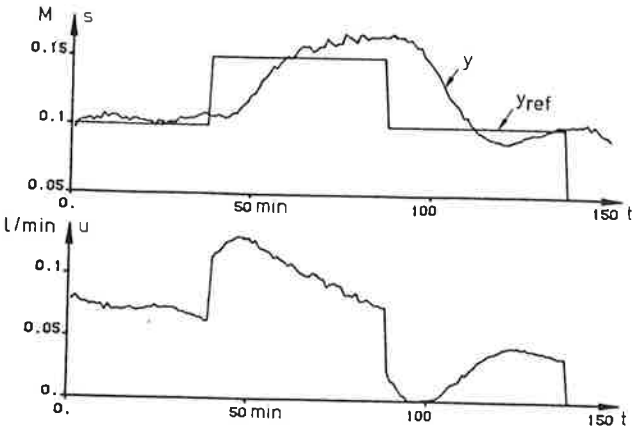

Fig. 4.1. PI-regulator on the real process. K=1.0 and Ti=15. Process parameters (*) but k_s=0.050 1/min.

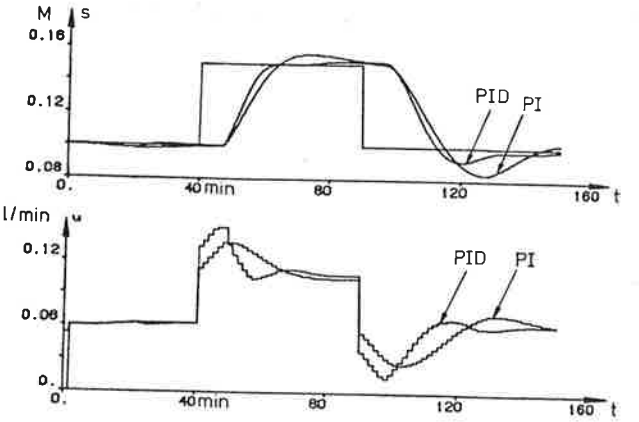

Fig. 4.2. Two different PID-regulators tested in simulation on the process in Fig. 4.1. Regulator parameters K=1.0 and Ti=15 as above and K=1.4, Ti=20 and Td=5. The performance is slightly improved by the D-part.

State feedback from a Kalman filter

Reconstruction of states of the process from measurement signals is usually done with a Kalman filter. This method can, in the discrete time case, also be applied to reconstruct actual states from timedelayed measurement signals.

Here the process model (2.13) is used for the Kalman filter. Performance is designed for a linear process. This process is slightly nonlinear. To be able to handle large setpoint changes, the nonlinearity should be accounted for in the filter. This is done by letting Φ, Γ and Θ vary with u, according to (2.13). The coefficients K and L can be kept constant or may also vary with u, to keep the closed loop performance unaltered as the operating point changes.

The regulator structure. Linear state feedback is done from a Kalman filter estimate.

The Kalman filter is

$$\hat{x}(t+h)=\Phi\hat{x}(t)+\Gamma u(t)+K[y(t)-\Theta\cdot\hat{x}(t)] \qquad (4.1)$$

The control signal is

$$u(t) = -L\cdot\hat{x}(t) + \frac{1}{G}\cdot y_{ref} \qquad (4.2)$$

The closed loop system on state space form is

$$\begin{bmatrix} x(t+h) \\ \tilde{x}(t+h) \end{bmatrix} = \begin{bmatrix} \Phi-\Gamma L & \Gamma L \\ 0 & \Phi-K\Theta \end{bmatrix} \begin{bmatrix} x(t) \\ \tilde{x}(t) \end{bmatrix} + \begin{bmatrix} \Gamma \\ 0 \end{bmatrix} \cdot \frac{1}{G}\cdot y_{ref} \qquad (4.3)$$

where $\tilde{x}(t) = x(t)-\hat{x}(t)$

G = Gain in the closed loop

Choice of poles. The poles for the process and for the Kalman filter are chosen independently. For the Kalman filter a trade off is done between sensitivity to noise and fast convergence. It is a good idea to have the Kalman filter to converge a little faster than the process. The poleconfiguration chosen, for both the Kalman filter and the process, is two control poles and the rest of the poles are left at the origin. The control poles are placed corresponding to ζ = 0.707 in continuous time terms.

Experiments show that it is possible to speed up the closed loop system a factor of four.

Calculation of the coefficients for the Kalman filter and for the linear state feedback is done in appendix 1. Note that the regulator is easy to tune.

Integrator. If the Kalman filter is well tuned, then the gain in the closed loop can be considerably increased and the stationary error reduced. The demand for an integrator to compensate for the stationary error is then reduced. All the same, it's a good idea to include an integrator.

The difference between measured y(t) and yref is integrated. The timedelay in the measurement signal may cause some trouble. Therefore the integrator time should be long compared with the timedelay. If the model in the Kalman filter were absolutely accurate, the difference between estimated y(t) and yref, should be integrated. It is likely to be a stationary error in the Kalman filter and therefore it is better to use the measured y(t).

The choice of integrator time is analysed in appendix 2.

Results on the real process. The regulator is tested in two ways. First a severe disturbance is given to the reactor and secondly change of operating point is tested. The disturbance is introduced by quickly pouring water into the reactor. This is not a very likely disturbance in an industrial process but it gives a good idea of the behavior of the system. Especially the Kalman filter can be studied. See Fig. 4.3. Changes of operating point are shown in Fig. 4.4.

Parameters of the regulator is specified by:

$$h = 3.25 \text{ min} \qquad d = 2$$

$$P_{\Phi-\Gamma L}(z) = z^3 - 0.6z^2 + 0.25z$$

$$P_{\Phi-K\Theta}(z) = z^3 - 0.4z^2 + 0.25z$$

No integrator was used and K and L were fix.

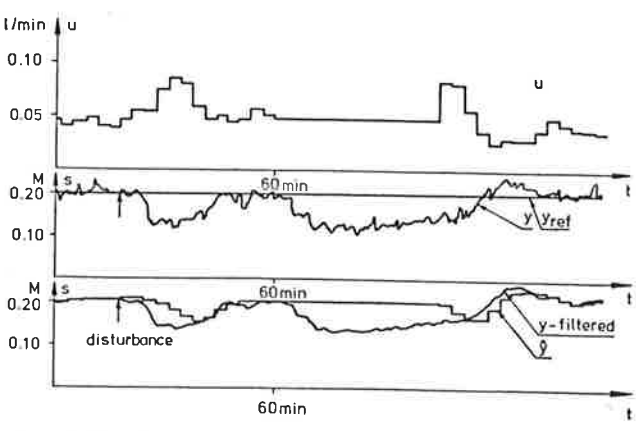

Fig. 4.3. Response to a heavy disturbance with regulator on and regulator off. Convergence of the Kalman filter is studied in the bottom diagram. Process parameters (*) but k_s=0.015 1/min.

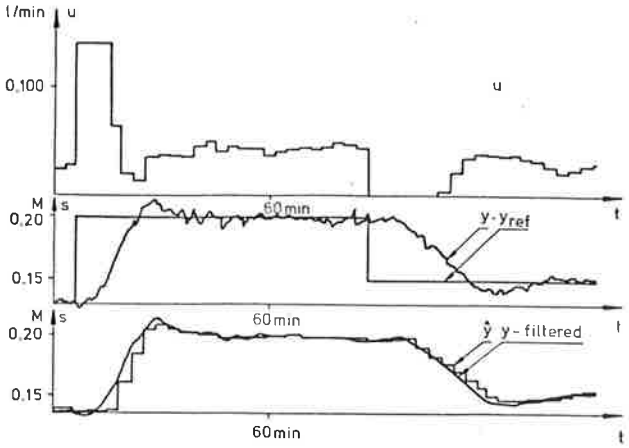

Fig. 4.4. Change of operating point. The same process parameters as in Fig. 4.3. Note that most of the control is performed before any change in measurement signal is recorded.

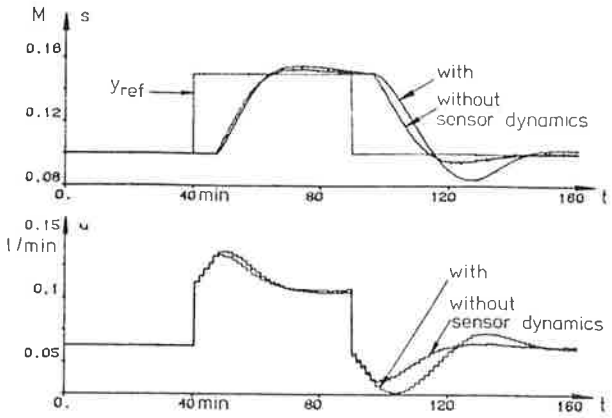

Fig. 4.5. Simulation of the process in Fig. 4.1. using the same PI-regulator. Here is seen what would have happened if the sensor had no dynamics. Notice the difference when going to the lower set-point. The regulator based on the Kalman filter is not that sensitive to sensor dynamics, although it could be a good idea to include it.

Compensation for Bilinearity

Another way to handle the bilinearity is evident from (2.5). Gainscaling of u(t) from v(t) gives a linear system that could be analysed in standard ways (Gutman, 1981). Consider the scaling

$$u(t) := \frac{V}{S_{in} - S(t)}\, v(t) \qquad (4.11)$$

gives

$$\dot{S}(t) = -k_S \cdot S(t) + v(t)$$

This design is convenient but has a weak point. The performance of the regulator might be sensitive to errors in gainscaling. Note that even for low cell activity S(t) is considerably lower than S_{in}.

Compensation for Sensor Dynamics

The Kalman filter can be expanded to include a state for sensor dynamics. This will be important if the rate constant of the processs is increased. The system matrices will then be

$$\Phi = \left[\begin{array}{cc|c} \varphi_{11} & 0 & \\ \varphi_{21} & \varphi_{22} & 0 \\ \hline & I & 0 \end{array}\right] \qquad \Gamma = \left[\begin{array}{c} \gamma_1 \\ \gamma_2 \\ \hline 0 \end{array}\right] \qquad (4.12)$$

where sensor dynamics can be approximated by

$$\varphi_{21} \approx \frac{h}{\tau}, \quad \varphi_{22} \approx 1 - \frac{h}{\tau}, \quad \gamma_2 \approx 0$$

to allow easy changes of τ

To account for the difference in rise and falltime of the sensor, the corresponding elements should be changed in Φ. It is important that the these changes are done at the right moment. This is easily done when a change of operating point is commanded and the slope is known. With this aid overshoots when going to a lower operating point can be eliminated.

5. IMPLEMENTATION

The regulators are implemented on a computer PDP11/03. Programs are written in PASCAL extended with a realtime kernel. The kernel incoperates facilities for mutual exclusion and synchronizing, which is necessary when concurrent activities are handled. See Elmqvist and Mattson (1982).

The computer is used not only for regulation but also for logging and testsignal generation. Aids for tuning the regulator is also included. Here is a list of facilities included:

- Digital filtering of measurement signal. A second order Butterworth filter is used. See prefiltering section 2.
- Calibration of input and output signal. Linear interpolation is used between arbitrarily spaced calibration points.
- Identification of the time constant for the process. Least square method is used.
- Computation of various regulator parameters from given process parameters and specification of the closed loop system.
- Regulation.
- Logging of data.
- Testsignal generation.
- Extensive operator comunication.

The program is divided in five concurrent activities:

- Filtering and calibration of measurement signal. Sampletime 5 s.
- Process control, sampletime 1 to 3 min.
- Logging of data. Arbitrary sampletime. Often synchronized with the regulator.
- Testsignal generator. Arbitrary sampletime.
- Operator communication. Waiting on interrrupt from operator console.

At last but not least important, all parameters are read from a number of different files at start up of the program.

6. CONCLUSIONS

This case study shows that computer control gives substantial advantages compared to analogue PID-control for a typical continuous fermentation process. A simple mathematical model of the process dynamics is of great help, when compensating for a timedelay in the measurement signal. The computer requirements are small if the computer is used only for regulation. A more sophisticated program system is necessary for efficient operator communication during start up, tuning and experiment documentation.

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APPENDIX 1

Calculations of the coefficients in the Kalman filter and in the linear state feedback are done. Note that the calculations are recursive and can easily be implemented for online tuning.

Kalman filter. The coefficints are given by

$$[I_\theta - \varphi \cdot D\ I_\theta I_n]\ K = C \qquad \text{(A1.1)}$$

where

$K = [1\ k_n \ \dots\ k_1]^T$ filter coefficients

$C = [1\ c_1 \ \dots\ c_n]^T$ coefficients of observer characteristic polynomial

$\varphi = e^{ah}$ time constant of the process

$$D = \begin{bmatrix} 0 & & & 0 \\ 1 & \ddots & & \\ & \ddots & \ddots & \\ 0 & & 1 & 0 \end{bmatrix}$$ backwards shiftmatrix

$$I_\theta = \begin{bmatrix} 1 & 0 & & 0 \\ & \theta_d & \theta_{d-1} & \\ & & \ddots & \theta_1 \\ 0 & & & \theta_0 \end{bmatrix}$$ representation of the timedelay

I_n = a diagonal matrix of dimension n+1 with elements 1 except for the bottom position which is 0.

Note that if $\tau_d = h \cdot d$ then $I_\theta = I$ and (A1.1) is easily solved:

$$k_n = c_1 + \varphi\ ,\ k_i = c_{n+1-i} + \varphi \cdot k_{i+1}\ ,\ i = n-1,\dots,1 \qquad \text{(A1.2)}$$

Linear state feedback. The coefficients are given by

$$l_1 = \frac{a_1 + \varphi}{\gamma}\ ,\ l_i = \frac{a_i}{\gamma}\ ,\ i = 2,\dots,n \qquad \text{(A1.3)}$$

where

$L = [l_1 \ \dots\ l_n]$ statefeedback coefficients

$A = [1\ a_1 \ \dots\ a_n]$ coefficients of desired process characteristic polynomial

APPENDIX 2

The choice of integrator time is analysed.

Introduce a new state x_{n+1} for the integrator.

$$x_{n+1}(t+h) = x_{n+1}(t) + [y_{ref} - y(t)] \qquad \text{(A2.1)}$$

State feedback from $x_1 \dots x_{n+1}$ gives the system matrix Φ'

$$\Phi' = \left[\begin{array}{c|c} \Phi - \Gamma L & \begin{matrix} -\gamma l_{n+1} \\ 0 \end{matrix} \\ \hline \begin{matrix} 0 & -1 \end{matrix} & 1 \end{array}\right] \qquad \text{(A2.2)}$$

Characteristic polynomial $P_{\Phi'}$ from $P_{\Phi - \Gamma L}$

$$P_{\Phi'}(\lambda) = (\lambda - 1) \cdot P_{\Phi - \Gamma L}(\lambda) + \gamma l_{n+1} \qquad \text{(A2.3)}$$

The state feedback coefficients are now given by

$$\begin{aligned} l_1 &= \frac{1}{\gamma} \cdot [a_1 + \varphi + 1]\ ,\ l_2 = \frac{1}{\gamma} \cdot [a_2 + \gamma l_1 - \varphi] \\ l_i &= \frac{1}{\gamma} \cdot [a_i + \gamma l_{j-1}]\ ,\ i = 3,\dots,n+1 \end{aligned} \qquad \text{(A2.4)}$$

Note that l_{n+1} can be interpreted as $-\frac{h}{T_i}$.