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EVALUATION OF LEVEL CONTROL PERFORMANCE

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Abstract: Various kinds of control performance index have been studied widely during the last decade. In this paper a set of performance indices appropriate to level control is chosen and a fast, simple, online method is presented for monitoring the control loops. A graphical, user-friendly and interpretable display for operators to display performance indices has been developed that can be easily implemented for any control system. This tool can be used for statistical monitoring of control loops, predictive maintenance, fault diagnosis, and for tuning the controllers. Tests were performed with a pilot flotation cell, and the results are presented and discussed in this paper. *Copyright © 2002 IFAC*

Keywords: level control, performance indices, monitoring, oscillation, fault detection

1. INTRODUCTION

Most of the control loops used in industry are not operating efficiently. The consequences of this are increased raw material and energy consumption and a deterioration in product quality. Additional costs are also incurred as a result of malfunctions and the short life span of instruments because they are used defectively in control loops.

Most of the performance indices are based on statistical and stochastic information about a process, and they are mainly used in tuning controllers. The disadvantage of these methods is that they require knowledge of the process model, thus requiring a lot of tests that can be very expensive. Indices are also hard to interpret. A good index is dimensionless. Various dimensionless indices have been proposed in the literature. Control performance has been studied by Harris (1992), and Swanda and Seborg (1999). Åström *et al.* (1992), and Swanda and Seborg (1999) introduced the dimensionless rise time and IAE index.

It is recommended to monitor control performance on various levels of a control hierarchy. At the lowest level are the individual control loops, which are usually controlled by PID. The software tool for this kind of process monitoring has been developed f.ex. by Metso Co. (2000). On the second level are the higher control methods and production management.

The aim of the study presented in this paper is to evaluate different kinds of performance index and to use them to monitor the level control of a flotation cell. This paper focuses on the lower level loops but the results obtained for these indices can also be applied to control loops on higher levels. The goal is to create a practical tool that helps operators and engineers to monitor the controller's operation.

2. PROCESS DESCRIPTION

A flotation plant consists of a number of flotation cells in series. Pulp level control is a complex task because the operating limits are tight and the operating conditions vary. Each flotation cell has a mechanism for air injection; launders for collecting the froth located at the top of the cell, and structures to funnel the froth towards the launders. This causes severe non linearity's in the system's behaviour, as discussed by Jämsä-Jounela *et al.* (2001).

In theory the mathematical model for a single cell can be expressed with a differential equation derived from the total mass balance (Stephanopoulos, 1994). Assuming the density to be constant, then

$$A\frac{dh}{dt} = F_i - F_o = F_i - \frac{h}{R} \qquad (1)$$

where
A = cross-sectional area of the cell (m^2),

h = pulp level (m),
F_i = volumetric inflow (m^3/s)
F_o = volumetric outflow (m^3/s) and
R = resistance to flow due to a control valve.

If level differences are small, the F_o can be assumed to be linear with h.

The transfer function of the system is

$$G(s) = \frac{h(s)}{F_i(s)} = \frac{R}{RAs+1}, \qquad (2)$$

where

R = K_P = steady-state gain of the process, and
RA = τ = time constant of the process.

An estimate of the time constant can be calculated

$$t = \frac{Ah_s}{F_{i,s}}, \qquad (3)$$

where
h_s = the steady-state pulp level and
$F_{i,s}$ = the steady-state volumetric inflow.

3. CONTROL PERFORMANCE INDICES

Control performance has a great effect on variations in end product quality and thus on the productivity of the plant. During the last decades much effort has been placed on developing suitable indices for evaluating control performance. The monitoring can be based on calculating the deviations from the set point values, e.g. the integrals of the error, or on rigorous process models, which make the assessment more accurate but, at the same time, more complex.

The methods can be divided into two categories: stochastic and deterministic methods. The most commonly studied stochastic indices are those based on minimum variance calculation - the variance of the process measurement is compared to the smallest theoretically achievable variance, as discussed by Harris (1989). Deterministic indicators are more informative in the case of a sudden load disturbance or a set point change. They include the Idle Index developed by Hägglund (1999), which detects the sluggish control loops frequently found in industry.

It is also essential to detect oscillations present in the system, which can be caused by valve friction, bad controller tuning or an oscillating load disturbance. These oscillations can be identified by means of autocorrelation functions or spectral analyses (Thornhill and Hägglund, 1997). Horch (1999) demonstrated a method for detecting stiction in control valves based on cross-correlation. Hägglund (1995) presented an oscillation detection procedure that involved the calculation of IAE (Integral of Absolute Error).

3.1 CONTROL PERFORMANCE INDICES FOR LEVEL CONTROL OF FLOTATION CELL

The performance of level control loops can be examined in different ways. Important aspects for controlling the flotation process can be listed as follows:

- The accuracy of the controller describes the controllers ability to follow the set point value. Usually error integrals or variances are used to measure this quality.

- The speed of the controller demonstrates the amount of time the controller takes to change the process value when a set point changes. Rise time indices are used for this purpose.

- Disturbance tolerance characterizes the ability of the controller to cope with disturbances that can be measured.

- Noise sensitivity describes the stability of the controller reactions to sudden spikes or noise in process value measurements.

- Robustness of the controller describes the ability of the controller to act with wide range of process parameters.

- The valve capacity evaluates the validity of the actuator sizing.

The performance of the pulp level control loops are usually considered in two different states: a state with a set point change, and a normal operating state close to the steady-state conditions. Separate indices can be chosen to describe the control performance in these two different cases.

3.1.1 Performance indices for steady-state conditions

Three indices were chosen to evaluate control performance in the case of an unvarying set point. The permanent error (PE) between the set point and the measured process value is worth monitoring because it degrades the control loop performance and, in case of oscillation, the difference can be difficult to detect from process trend displays. For each process can be defined a value of the largest acceptable error between the values of the set point and the process measurement, denoted here as e_{lim}. Thus an index for a permanent error can be calculated recursively as follows:

$$PE_i = g \cdot PE_{i-1} + (1-g) \cdot p_i , \qquad (4)$$

where
γ = “the forgetting factor”,
PE_{i-1} = the previous value of the index, and

$$p_i = \begin{cases} -1, e_i < -e_{\lim} \\ 0, -e_{\lim} < e_i < e_{\lim} \\ 1, e_i > e_{\lim} \end{cases} \qquad (5)$$

The forgetting factor γ can be calculated as follows,

$$g = 1 - \frac{1}{5t} , \qquad (6)$$

where τ is an estimate of the time constant of the process. A good estimate of the time constant is the retention time of the flotation cell, which can be evaluated according to Eq. 3. When the process measurement equals the set point value, the index converges to zero. Values of the index near ±1 indicate that a permanent error is present, and the sign of the index shows whether the process value is above or below the desired set point value.

Oscillations around the set point can be detected by using the method developed by Hägglund (1995), which is based on monitoring the IAE values calculated between consecutive set point crossings of process value.

$$IAE_i = \int_0^{t_i} \left| y_{pv}(t) - y_{sp}(t) \right| dt \qquad (7)$$

where t_i is the times of successive set point (y_{sp}) crossing of y_{pv}.

If the value of the IAE_i exceeds the predefined value $IAE_{\lim}$, it can be concluded that a load disturbance has occurred. Because the process data are rather discrete, the $IAE_{\lim}$ was assumed to equal the area of a triangle with a height of $e_{\lim}/2$. Thus the $IAE_{\lim}$ can be calculated as

$$IAE_{\lim} = \frac{e_{\lim} t_{dis}}{4} , \qquad (8)$$

where t_{dis} is the duration of a single load disturbance which can be calculated if the frequency of the process is known. In an on-line application, the index can be calculated recursively by using the forgetting factor as in Eq. 4.

$$OSC_i = g \cdot OSC_{i-1} + (1-g) \cdot DIST_i , \qquad (9)$$

where

$$DIST = \begin{cases} 1, IAE_i \geq IAE_{\lim} \\ 0, IAE_i < IAE_{\lim} \end{cases} \qquad (10)$$

Stochastic variations around the set point value were selected for detection, e.g. by monitoring the integral of the squared error (ISE),

$$ISE_i = g \cdot ISE_{i-1} + (1-g) \cdot \left[y_{pv}(t) - y_{sp}(t) \right]^2 , \qquad (11)$$

which highlights the largest deviations. These variations may be too short-term to be detected by oscillation detection procedures, but they can be detected effectively with the ISE. The calculation can be carried out on-line by using a recursive algorithm.

3.1.2 Performance indices for set point change occurrences

Six indices were chosen to evaluate the control performance in a set point change. Monitoring was performed during a time period, the length of which was a multiple of the time constant estimated by an operator. A response to a step change in a set point value and the key figures determined from the process measurements in order to calculate the indices are illustrated in Fig 1.

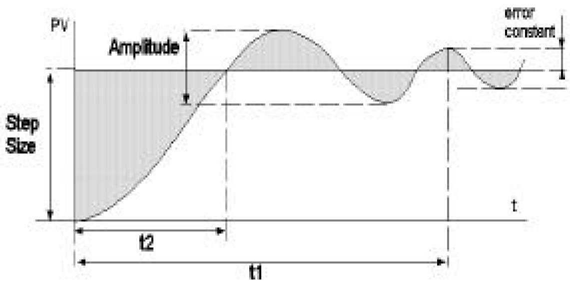

Fig. 1. Response to a step change in a set point. t1 is settling time and t2 is rise time.

Oscillations around the set point were observed by using the method developed by Hägglund (1995). Contrary to the on-line calculation discussed above the exponential weighting was not applied.

After a step change in a set point, there may be some oscillations before the process value settles to the steady state. By measuring the largest amplitude of the oscillation an index can be calculated to describe the size of the overshoot related to the step size:

$$AMP = \frac{y_{pv,\max} - y_{pv,\min}}{\Delta y_{sp}} , \qquad (12)$$

where $y_{pv,max/min}$ are the maximum and minimum values of the process measurement after a rise time and Δy_{sp} is the size of the set point change.

Long-term differences from the set point due to continuous oscillations or sluggish controller tuning were chosen to be monitored by calculating the integral of the time-weighted absolute error (ITAE)

$$ITAE = t\int_0^t \left| y_{pv}(t) - y_{sp}(t) \right| dt, \qquad (13)$$

which emphasizes long-term deviations. To obtain an independent and dimensionless index the value of the ITAE was related to the step size and to the sum of the arithmetic sequence, which follows from multiplication with time.

In order to characterize the rise time and settling time, Åström *et al.* (1992), Swanda and Seborg (1999) introduced procedures for calculating the normalized indices. In these studies an estimate of an apparent time delay was used to nondimensionalize the indices for a rise time and settling time. The dimensionless indices can also be calculated by relating the rise time and settling time to an approximation of a time constant τ. In the case of a flotation cell, the time constant can be assumed to correlate with the retention time of the cell accordingly to Eq. 3. The dimensionless indices for a rise time and settling time can therefore be expressed as follows:

$$SPD = \frac{t_{rise}}{t} \qquad (14)$$

and

$$TIME = \frac{t_{settling}}{t}. \qquad (15)$$

Undesirable performance of a control loop may also result from an inadequate actuator sizing, and not only from poor controller tuning. Therefore an index was developed to monitor the valve capacity. The value of the index describes the time t_{vc} that a valve opening is greater than 90 % or smaller than 10 % related to the time needed to carry out the set point change. The time constant defined by an operator is a rather valid approximation of the time spent to the set point change. The valve capacity index can therefore be calculated as

$$VC = \frac{\int_0^t t_{vc}\, dt}{t}, \qquad (16)$$

where

$$t_{vc} = \begin{cases} 0, x \in [0{,}1 \ldots 0{,}9] \\ 1, x < 0{,}1 \vee x > 0{,}9 \end{cases} \qquad (17)$$

and x is the valve opening. Values close to zero indicate a correct actuator sizing and while values close to one are a sign of a deficient valve sizing.

The Oscillation index for a set point change is calculated in the same way as the steady state oscillation index in eqs. 7,8 and 10 except that the final index is simply

$$OSC = \sum_i DIST_i\,. \qquad (18)$$

The OSC index is therefore the number of set point crossings where IAE_i has been larger than IAE_{lim}.

4. IMPLEMENTATION OF THE CONTROL PERFORMANCE INDICES

Different kinds of index were tested using a flotation cell process model that was programmed in Matlab Simulink toolbox. Indices that functioned well in the simulations and which were applicable to on-line monitoring were chosen to further development.

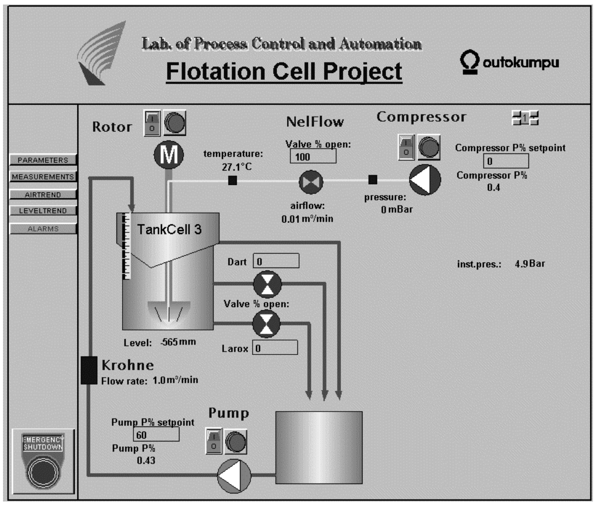

Fig. 2. Cimplicity user interface for a pilot flotation process.

The indices were implemented using the GE Fanuc Cimplicity and its script language. Cimplicity is a user interface program (Fig. 2.) for process control and is widely used in flotation processes. Outokumpu pilot Tankcell and the above-mentioned programs were used to test the indices in practice. The cell instruments were connected to the Foundation Fieldbus and controlled by Smar DFI 302. The controller unit was connected with Ethernet to the OPC (OLE (Object Linking and Embedding) for Process Control) server, and Cimplicity used the OPC port to control the flotation cell.

The index scripts were programmed in a separate window that monitors one control loop at a time. The operator gives the names of the measurement, the set point value and the PID-output that forms the control loop in a dialog box. Other information to be given to the program are the time constant of the process, the sampling frequency, and a numerical value that can be considered as a significant error. The time constant can be calculated by the steady state-method. The value of the significant error can be estimated by monitoring the process and using the variance of a stable situation. The value of significant error should preferably be too large than too small, because the effect of noise in the measurements and external interference cannot be removed by a more effective controller. This means that their influence must be eliminated if the goal is to monitor the performance of the controller.

Each index has it own bar in the user interface. The setpoint change index bars are green when their height is smaller than 1 and the rest of the bar is coloured red if it is longer. The steady-state indices are on the left in the screen and they have only one colour.

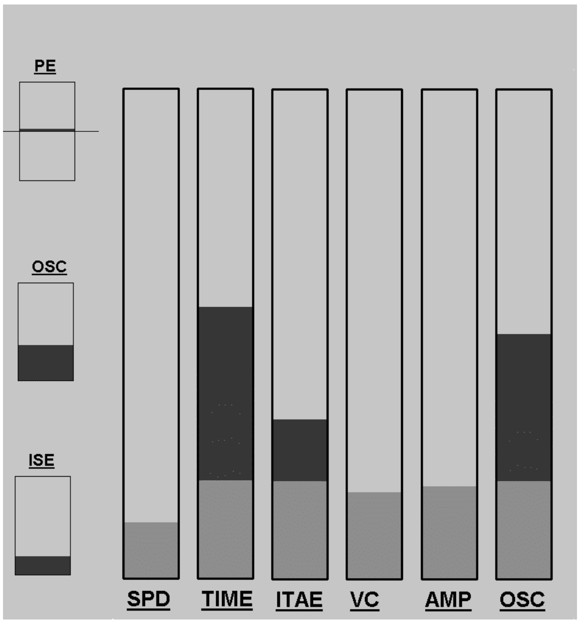

Fig. 3. Cimplicity user interface for control performance indices.

5. RESULTS

5.1 Set point change indices

The indices were tested with the laboratory's pilot Tankcell. First the PI controller of the outflow valve was tuned by the Ziegler-Nichols method. Then the indices were normalized so that each index gives values below one with an acceptable control performance. A response to a set point change with a properly tuned controller is presented in Fig. 4. In the trend displays the set point is marked in black, the process value in gray and a valve output in light gray.

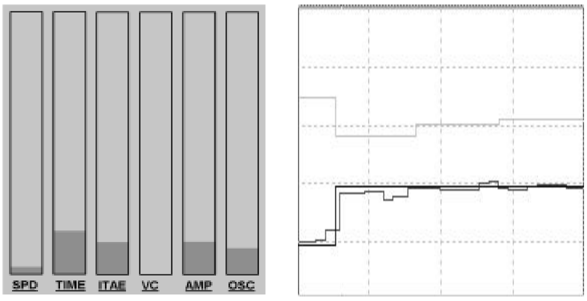

Fig. 4. Set point change indices and process trends with PI-parameters P = -3 and I = 15.

SPD = 0.1	TIME = 0.9	ITAE = 0.8
VC = 0	AMP = 0.7	OSC = 0.5

The PI parameters were changed in both directions in order to test the performance indices with all combinations of these parameters. The results achieved were in line with the hypothesis. With low proportional gains (P) and high integration times (I) the controller was sluggish, which resulted in high time indices and low amplitude and oscillation, as presented in Fig. 5.

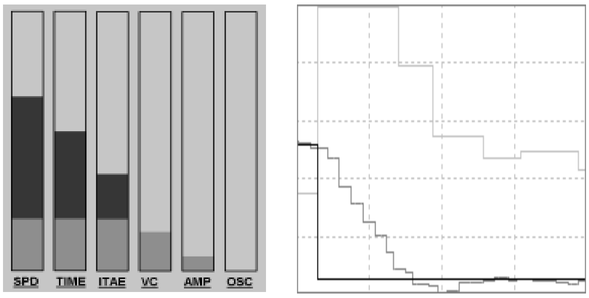

Fig. 5. Set point change indices and process trends with PI-parameters P = -2 and I = 40.

SPD = 3.3	TIME = 2.5	ITAE = 1.9
VC = 0.8	AMP = 0.3	OSC = 0

With low I and high P values the controller oscillated, but the speed index was low as can be seen in Fig. 6. With low P and low I values the amplitude and the ITAE and settling time were high.

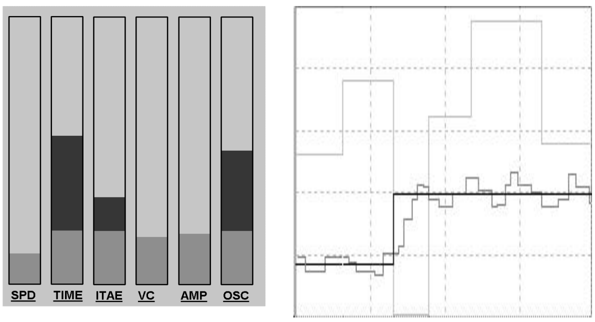

Fig. 6. Set point change indices and process trends with PI-parameters P = -10 and I = 7.

SPD = 0.5	TIME = 2.6	ITAE = 1.6
VC = 0.8	AMP = 0.9	OSC = 2.3

5.2 Steady state indices

The steady state indices were then tested and the controller was tuned to cause oscillation in process measurements. The indices and a trend display are presented in Fig. 7. The oscillation and ISE indices reached high values. The PE index was close to zero because the process value was oscillating around the set point and there was no permanent error present.

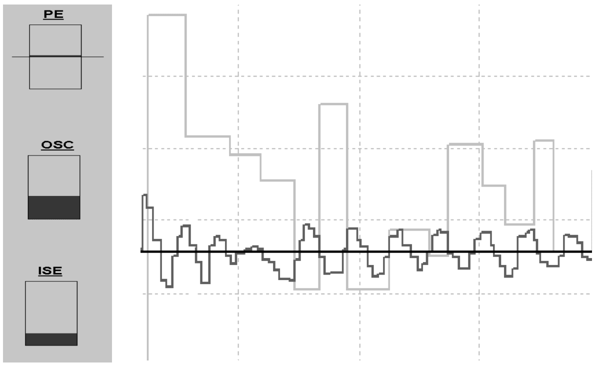

Fig. 7. Steady state indices when process is oscillating.

PE = 0.05	OSC = 0.3	ISE = 0.15

When the process value was unable to reach the set point the PE and ISE indices attained high values, as can be seen in Fig. 8.

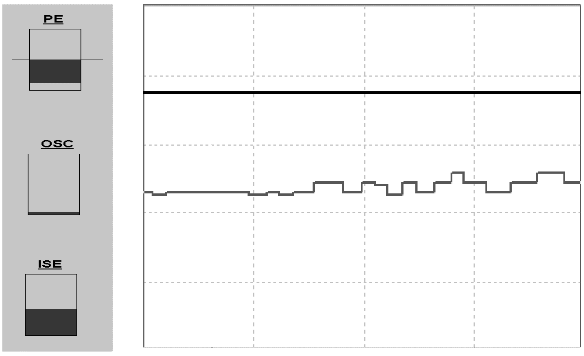

Fig. 8. Steady-state indices when there is a permanent error present.

PE = -0.8	OSC = 0.05	ISE = 0.4

6. CONCLUSIONS

Simulations and tests performed with the pilot flotation cell proved that the indices were sufficient to provide the necessary information about the control performance. In contrast to complex control evaluation methods the monitoring system described in this paper is easy to use and interpret. One of the benefits is that the measurement signals dot not need to be filtered.

This monitoring tool could be used in plants to monitor the key controllers where the indices are recorded in the SQL (Structured Query Language)-database. This tool could also be used as a predictive maintenance tool and in diagnostics.

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