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Evaluation of control performance: methods, monitoring tool and applications in a flotation plant

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Abstract

Evaluating control performance has attracted considerable interest in recent years. A set of performance indices appropriate to monitoring and assessment in flotation cells is presented and discussed in this paper. A graphical, user-friendly and interpretable program for displaying performance indices for operators has been developed. Finally, the testing results from flotation cells in a zinc plant are presented and discussed.

Keywords

Process control, Mineral processing, Flotation machines

1. Introduction

A flotation plant consists of a number of flotation cells in series. 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. Pulp level control in the flotation cells is a complex task because the operating limits are tight and the operating conditions vary (Jämsä-Jounela et al., 2002).

Most of the control loops used in flotation cells 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.

During the last decades considerable effort has been placed on developing suitable indices for evaluating control performance. The evaluation methods can be divided into two categories: stochastic and deterministic methods. The most widely studied stochastic indices are those based on using of minimum variance controller (MVC) calculation as a benchmark. The variance of the process output is compared to the smallest, theoretically achievable variance, as initially discussed by Harris (1989). One advantage of these methods is that they require only output data from a controlled process and a priori knowledge of the dead time of the process or its estimation. Horch and Isaksson (1999) proposed a modified performance index that is more robust with non-stationary systems. Eriksson and Isaksson (1994) pointed out that a controller with a good MVC index does not necessarily have a good performance with respect to set-point changes. Overviews of the research carried out on minimum variance control during the past decade have been presented by Qin (1998).

Deterministic indicators are more informative in the case of a sudden load disturbance or a set-point change. Various dimensionless indices for set-point changes have been proposed in the literature, e.g. by Åström et al. (1992). Hägglund (1999) dealt with the rejection of step disturbances and described it by means of the idle index. Swanda and Seborg (1999) introduced the dimensionless rise time and the integral of absolute error (IAE) index. Two performance indices, the absolute performance index (API) and the robustness index (RI) were introduced by Shinskey (1990).

It is also essential to detect oscillations in the system, 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 between process input and output. Hägglund (1995) presented an oscillation detection procedure that involved the calculation of IAE.

A wide variety of commercial performance monitoring tools are nowadays available on the software market. Among the most common ones are the Process doctorTM from Matrikon, ABB's Loop Optimizer Suite, Honeywell's Loop ScoutTM. All of them use MVC-based indices, an autocorrelation function and various deterministic indices. Honeywell also uses data from similar production plants and is an offline product. The others have on-line versions as well.

The aim of this study is first to present methods that can be used in evaluation of performance of control loops in flotation and then to present a control performance monitoring program that uses these methods. Finally, the testing results from the flotation plant are presented and discussed.

2. Control performance indices

Monitoring can be based on calculating the deviations from the set-point values, e.g. the integrals of the error, or on process models, which make the assessment more accurate but, at the same time, more complex. Monitoring based on minimum variance control only requires output data from routine operation.

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

2.1. Performance indices for steady-state operation

Some indices are suitable for evaluating control performance in the case of a non-varying 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 the case of oscillation, it can be difficult to detect the difference from process trend displays. A value of the largest acceptable error between the values of the set-point and the process measurement, denoted here as e_{lim}, can be defined for each process. Thus an index for a permanent error can be calculated recursively as follows:

$$\mathrm{PE}_i = \gamma * \mathrm{PE}_{i-1} + (1-\gamma) * p_i \qquad (1)$$

where γ is the "forgetting factor", PE_{i-1} the previous value of the index, and

$$p_i = \begin{cases} -1, & e_i < e_{lim} \\ 0, & -e_{lim} < e_i < e_{lim} \\ 1, & e_i > e_{lim} \end{cases} \tag{2}$$

The forgetting factor γ can be calculated as follows:

$$\gamma = 1 - \frac{1}{5\tau} \tag{3}$$

where τ is an estimate of the time constant of the process. When the process measurement equals the set-point value, the index converges to zero. Index values 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 the process value

$$\mathrm{IAE}_i = \int_0^{t_i} \left| y_{\mathrm{pv}}(t) - y_{\mathrm{sp}}(t) \right| \mathrm{d}t \tag{4}$$

where t_i are 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 discrete, the IAE_{lim} can be assumed to be equal to the area of a triangle with a height of $e_{lim}/2$. Thus the IAE_{lim} can be calculated as

$$\mathrm{IAE}_{lim} = \frac{e_{lim} t_{dis}}{4} \tag{5}$$

where t_{dis} is the duration of a single load disturbance that 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. (1)

$$\mathrm{OSC}_i = \gamma * \mathrm{OSC}_{i-1} + (1-\gamma) * \mathrm{DIST}_i \tag{6}$$

where

$$\mathrm{DIST} = \begin{cases} 1, \mathrm{IAE}_i \geq \mathrm{IAE}_{lim} \\ 0, \mathrm{IAE}_i < \mathrm{IAE}_{lim} \end{cases} \tag{7}$$

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

$$\mathrm{ISE}_i = \gamma * \mathrm{ISE}_{i-1} + (1-\gamma) * [y_{\mathrm{pv}}(t) - y_{\mathrm{sp}}(t)]^2 \tag{8}$$

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.

An index denoted as ISU can be used as a measure of how much the control action changes. It is similar to the index ISE.

$$\mathrm{ISU}_i = \gamma * \mathrm{ISU}_{i-1} + (1-\gamma) * [u(k) - u(k-1)]^2 \tag{8a}$$

The index will be large if the valve needs to move a considerable amount in order to maintain the set-point, and zero when no control action is necessary.

The most popular index is the dimensionless index based on minimum variance control. It describes, how close the actual output variance is compared to the minimum achievable variance, obtained if a minimum variance controller is employed.

The reason why we have minimal variance lies in the delay of the plant. The delay *d* prevents the controller from influencing the output immediately. During the first *d* steps the noise passes to the output and the minimum variance is therefore calculated from the first *d* elements of the noise-to-output impulse response of an estimated model.

$$\sigma_{mv}^2 = \sigma_a^2 (f_0^2 + f_1^2 + \cdots + f_{d-1}^2) \tag{9}$$

where f_i are the coefficients of the noise-to-output impulse response, σ_{mv}^2 is the minimum variance, σ_a^2 is the variance of the white noise disturbance.

The output variance σ_y^2 can be calculated from

$$\sigma_y^2 = \sigma_a^2 (f_0^2 + f_1^2 + \cdots + f_{d-1}^2 + \cdots) \tag{10}$$

The performance index based on minimum variance control is

$$\eta = \frac{\sigma_{mv}^2}{\sigma_y^2} \tag{11}$$

In order to calculate this index, the impulse response from the noise-to-output transfer function must be estimated, e.g. using an ARMA model, which can be estimated recursively for online operation. An ARMAX model can also be employed in order to find the time delay *d*.

2.2. Performance indices for set-point change occurrences

The following indices can be chosen for evaluating the control performance in a set-point change. Monitoring can be performed during a specific time period, the length of which is a multiple of the time constant. A response to a step change in a set-point value, and the key values that have to be determined from the process measurements in order to calculate the indices, are illustrated in Fig. 1.

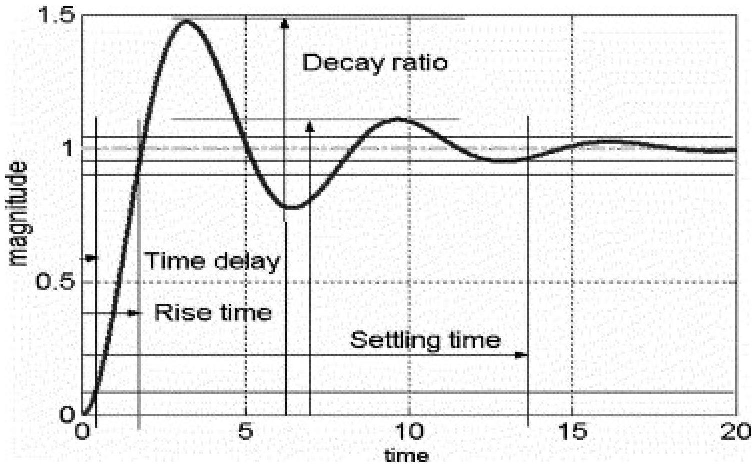

Fig. 1. Response to a step change in a set-point

Oscillations around the set-point can be observed using the method developed by Hägglund (1995).

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

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

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

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

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

which emphasizes long-term deviations. In order to obtain an independent and dimensionless index the value of the ITAE can be related to the step size and to the sum of the arithmetic sequence, which follows from multiplication by time.

In order to characterize the rise time and settling time, Åström et al. (1992) and Swanda and Seborg (1999) introduced procedures for calculating the normalized indices. In these studies an estimate of an apparent time delay was used to non-dimensionalize 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. The dimensionless indices for a rise time and settling time can therefore be expressed as follows:

$$\mathrm{SPD} = \frac{t_{\mathrm{rise}}}{\tau} \tag{14}$$

and

$$\mathrm{TIME} = \frac{t_{\mathrm{settling}}}{\tau} \tag{15}$$

The oscillation index for a set-point change can be calculated in the same way as the steady state oscillation index in Eqs. (4), (5) and (7) except that the final index is simply

$$\mathrm{OSC} = \sum_i \mathrm{DIST}_i \tag{16}$$

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

2.3. Performance index for disturbance rejection

Disturbance rejection can be detected by the idle index. The index is defined by

$$I_i = \frac{t_{pos} - t_{neg}}{t_{pos} + t_{neg}} \tag{17}$$

where the following procedures are updated every sampling instant:

$$t_{pos} = \begin{cases} t_{pos} + h & \text{if } \Delta u \Delta y > 0 \\ t_{pos} & \text{if } \Delta u \Delta y \leq 0 \end{cases}$$
$$t_{neg} = \begin{cases} t_{neg} + h & \text{if } \Delta u \Delta y < 0 \\ t_{neg} & \text{if } \Delta u \Delta y \geq 0 \end{cases} \qquad (18)$$

and h is the sampling period.

The index is bounded to the interval [−1, 1]. A positive value of I_i close to 1 means that the control is sluggish and negative value of I_i close to −1 is obtained in a well-tuned control loop.

2.4. Performance index for valve monitoring

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% with respect to the time needed to carry out the set-point change. The saturation index can therefore be calculated as

$$\mathrm{SI} = \frac{\int_0^\tau t_{\mathrm{vc}} \mathrm{d}t}{\tau} \qquad (19)$$

where

$$t_{vc} = \begin{cases} 0, x \in [0.1, \ldots, 0.9] \\ 1, x < 0.1 \ \vee x > 0.9 \end{cases} \qquad (20)$$

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

3. Control performance indices for level control of the flotation cells

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 controller's 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.

In the monitor program represented in this paper four indices for monitoring the performance of the level control loops in flotation are implemented as described above: a minimum variance index (MV), an ISE-index, a saturation index (SI), and an oscillation index (OCI).

4. Description of the monitoring tool

The program was developed using Visual Studio 6.0 software and programmed with Visual Basic. The program works as an OPC client and collects data from an OPC server at a specified rate, while calculating the performance indices at certain intervals for each configured loop using Matlab R12. The connection to the OPC server is established through TCP/IP network by giving an IP address of the OPC-server computer. After establishing the connection, the control loops are configured by giving a set-point, a measured value and an output of the control loop. The program then opens Matlab R12, creates matrices, and calculates the performance indices.

5. Testing results

The monitoring program was tested for the three level control loops of the flotation cells in Outokumpu Zinc plant in Kokkola. The loops discussed in more detail in this paper represent normal, oscillating and saturated controller behavior.

5.1. Normal controller behavior

The level control of the first of the six flotation cells in series in the zinc purification process has been taken as an example of a control loop describing normal behavior. Based on the output data shown in Fig. 2, it can be concluded that the process was relatively stable in the time interval 5000–11,000 s, when oscillation was less than 10% and the oscillation time approximately 500–1000 s. The index values show that the program predicts the normal behavior of the control loop in the interval 5000–11,000 s. Because of slight oscillative nature of the process, the values of the minimum variance index are close to zero, but however not zero. The values for the ISE and the oscillation indices are relatively normal, varying from 0.5 to 1 for ISE index and from 0.1 to 0.5 for oscillation index. Values for the saturation index are zero which indicates that the controller was sized correctly.

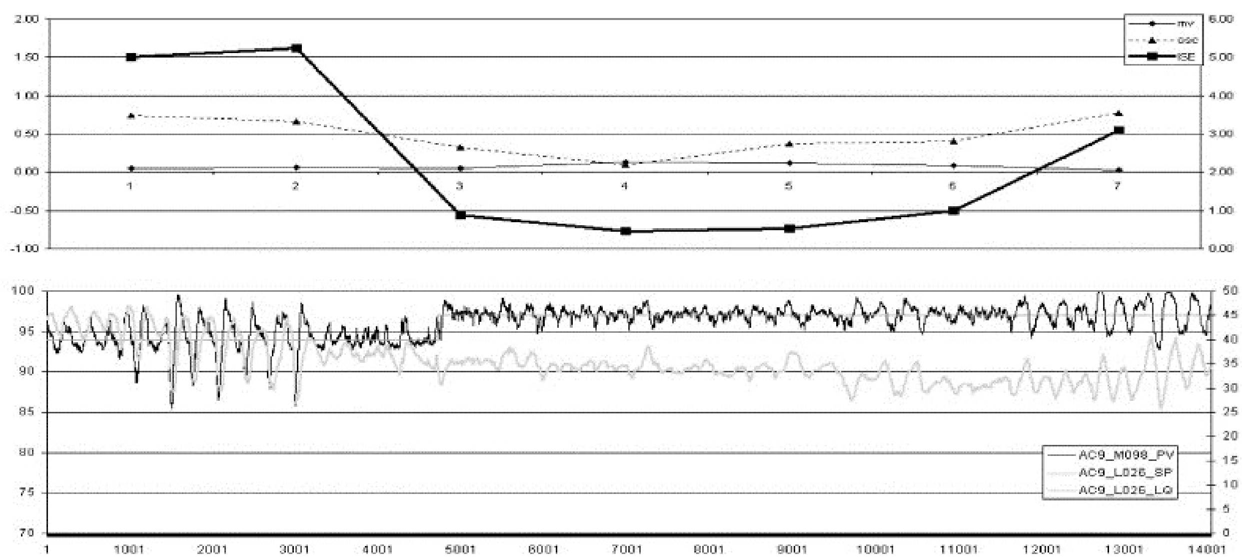

Fig. 2. Performance indices for the loop describing normal controller behavior

5.2.Oscillating controller behavior

The level control of the second of the six flotation cells in series has been taken as an example of a control loop describing oscillating behavior. Based on the output data shown in Fig. 3, it can be concluded that the process was relatively unstable, oscillation being almost 15% and the oscillation time approximately 500–1000 s. The index values show that the program predicts the oscillation of the control loop. Now, the values of the minimum variance index are zero, and the values of the ISE and the oscillation indices are much higher than in normal case above. The values of ISE-index vary from 2 to 3 and for oscillation index from 0.6 to 0.7. Saturation index was zero, thus the controller was sized correctly.

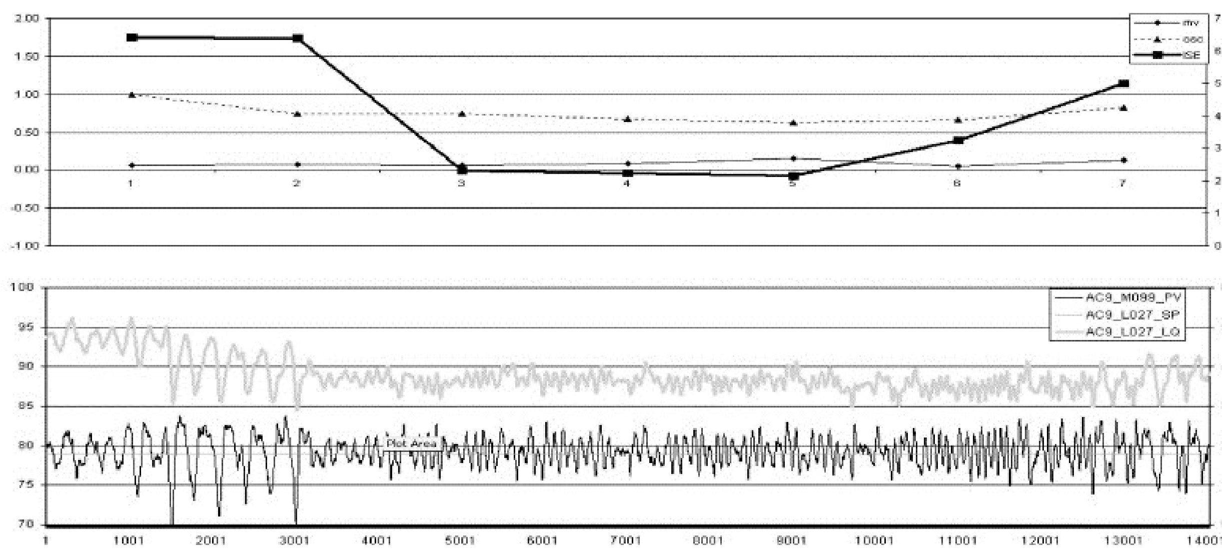

Fig. 3. Performance indices for the loop describing oscillating controller behavior

5.3.Saturated controller behavior

The level control of the last of the six flotation cells in series has been taken as an example of a control loop describing saturated behavior. According to the output data in Fig. 4, the process is very unstable, oscillation being almost 25% and the oscillation time approximately 1000 s. It can be concluded from the calculated index values that the control loop is saturated, because the saturation index differed greatly from zero. As expected, the values for the minimum variance index are zero, and the oscillation index is very high, varying from 0.5 to 1. The values for ISE index vary from 5 to 60. Based on these values it can be concluded that the process was far from the set-point value.

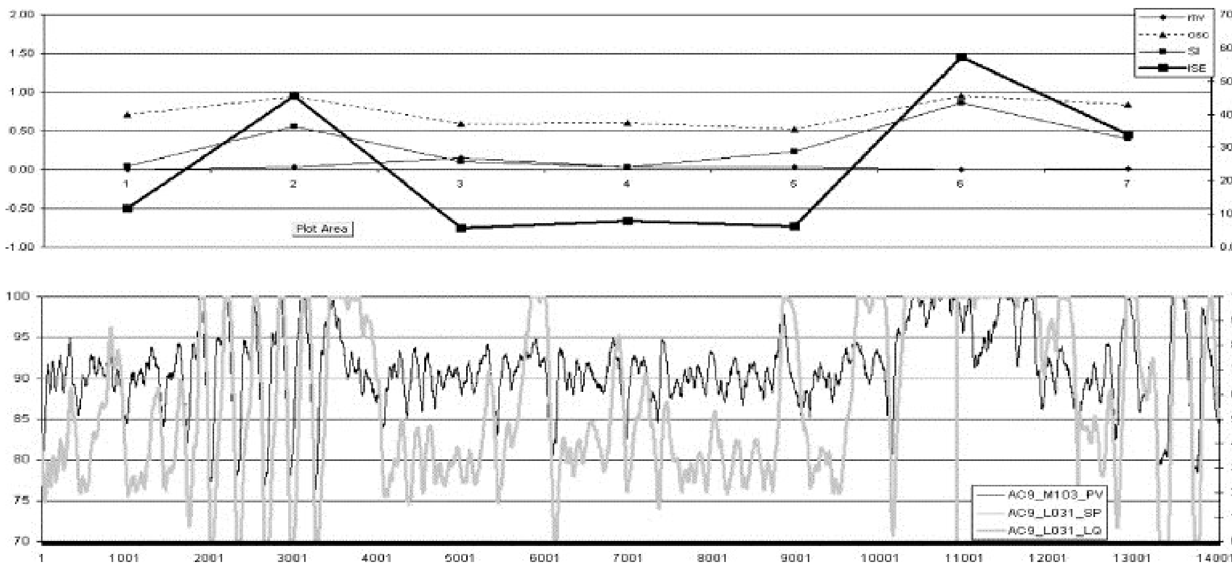

Fig. 4. Performance indices for the loop describing saturated controller behavior

6. Conclusions

A monitoring tool for calculating performance indices of the control loops has been presented and discussed in this paper. The simulations and tests demonstrated that the indices were sufficient to provide the necessary information about the control performance. In plant testing, the output from the flotation cells indicated which control loops were well tuned and which were not. The calculated performance indices supported these assumptions.

This monitoring tool could be used in flotation plants to monitor the key controllers for improving process control and product quality.

In future research, more attention will be paid to prioritize the importance of different control loops and to translate the control performance indices to economical measures.

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